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Prepared for
NATIONAL AERONAUTICS AND SPACE ADMINISTRATION
Lewis Research Center
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U.S. DEPARTMENT OF ENERGY
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1.0 INTRODUCTION

1.1 Background

The National Aeronautics and Space Administration, Lewis Research Center (NASA-LeRC), as part of the national program for commercial development of magnetohydrodynamic (MHD) electric power generation technology, is conducting parallel studies of alternative MHD power systems. The development of these advanced energy conversion systems is consistent with the objective of the National Energy Program, i.e., to increase the generation of electricity from coal or coal-derived fuels in an energy efficient, economically competitive manner which conserves natural resources and minimizes adverse effects on the environment. In this study, Gilbert Associates, Inc., has parametrically assessed the potential performances, capital costs and costs of electricity of coal-fired closed cycle MHD (CCMHD) power plants.

Closed cycle MHD was one of the advanced energy conversion technologies considered in the Phase I Energy Conversion Alternatives Study (ECAS).[1,2,3*] Results from the ECAS study showed that CCMHD systems exhibited overall plant efficiencies which were among the highest of the advanced energy conversion systems considered; however, cost of electricity (COE) estimates were higher than other advanced technologies considered in ECAS Phase II. The high potential efficiency predicted for CCMHD power systems in ECAS, recent developments in combustion technology conducted by General Electric[4], and the need for consistent cost information required to evaluate the CCMHD development program has prompted NASA-LeRC to sponsor this study.

* Superscript numbers indicate references which are listed in Section 9.0

This study was conducted using the same ground rules as the AVCO and General Electric open cycle "Parametric Study of Potential Early Commercial Power Plants" (PSPEC) and the open cycle and closed cycle disk generator parametric studies conducted by Westinghouse. Results of these studies will provide a comparative assessment of the merits of alternative MHD systems.

A closed cycle MHD generator depends on the concept of non-equilibrium ionization, where the electron temperature is elevated above the gas temperature. This two-temperature plasma model results in a system that has the advantage of operating at relatively low gas temperatures compared to open cycle MHD (OCMHD) systems, which decreases the severity of the material problems while maintaining a high plasma electrical conductivity, since electrical conductivity is primarily a function of the electron temperature.

A closed cycle MHD plant operates at temperature levels which are comparable to conventional fossil power plants. Both CCMHD and conventional power plant combustors operate at a flame temperature of around 3500 F. OCMHD combustors require a flame temperature in excess of 4500 F. The plasma flowing through the MHD generator is primarily argon, which is less corrosive than the combustion gases in OCMHD and is completely slag free since combustion is external to the closed argon system. The absence of slag in the working fluid should simplify the channel design and lead to longer channel lifetimes (because of the absence of sulfur) and a less complex design for the heat recovery components.

The combustor in a CCMHD system can be fired directly with coal or by using clean fuel from a gasifier. If a direct coal-fired combustor is used, slag will be carried into the regenerative argon heat exchanger. Both open and closed cycle MHD systems require alkali seed materials in the MHD channel and downstream heat recovery components; therefore, the associated materials problems are not abated.

The environmental concerns for a CCMHD plant can be compared to both a conventional coal-fired power plant and to a direct coal-fired OCMHD plant. In comparison with a conventional plant, operating temperatures for both systems are essentially equivalent; therefore, NOx emission levels will be similar. SOx and particulate emission levels for a direct coal-fired CCMHD plant will be similar to those of a conventional plant but lower for a gasifier CCMHD plant.

In comparison with an OCMHD power plant, NOx effluents will be lower with CCMHD because of its lower operating temperature. SOx levels can be expected to be higher for a direct coal-fired CCMHD plant because seed is not mixed with the combustion products. Particulate emissions from both the CCMHD and the OCMHD systems will be approximately the same; however, the CCMHD exhaust will contain no seed.

A major problem in closed cycle MHD is maintaining the necessary level of non-equilibrium ionization in a plasma which can be highly turbulent and unstable. The concept of non-equilibrium conductivity depends on minimizing the number of electron collisions. For this reason, a noble gas such as argon, which has a relatively low collision cross section, is used as the working fluid. A potential difficulty anticipated in CCMHD systems is reduction in the degree of non-equilibrium ionization in the MHD generator due to contamination of the inert gas plasma by small quantities of residual combustion gases which may leak into the system in the regenerative heat exchanger.

Argon must be heated to a stagnation temperature of about 3100 F. To reach this temperature, a ceramic matrix regenerative heat exchanger is required. The maximum operating temperature of the ceramic (brick) is limited to about 3350 F. The ceramic cores of heat exchanger arrays are alternately heated by combustion products and cooled by argon. After the core bricks are heated, the combustion gases are purged and the passages evacuated before the argon enters the heat exchanger. Regardless of the evacuation pressure, a small quantity of combustion gases is carried over and mixed with the argon.

The contamination of the argon with molecular combustion species degrades the level of non-equilibrium ionization because these molecules have large collision cross sections, which increase the number of inelastic electron collisions which lowers the electron energy.

Recent experimental studies conducted by General Electric[5] have measured argon purity levels in a coal-fired regenerative heat exchanger test facility. Measured results from these tests indicate that the impurity levels for the major molecular species (N_2, CO, CO_2, and H_2O) are on the order of 100 ppm. These levels are lower than the theoretical values at which unacceptable generator performance degradation should occur.

1.2 Scope

This document reports the results of Phase I of the Parametric Analysis of Closed Cycle MHD Power Plants conducted by Gilbert Associates, Inc., under NASA Contract DEN 3-136. The parametric cases were selected to demonstrate, in a preliminary manner, the performance, cost and natural resource requirements plus the environmental impact of commercial scale coal-fired closed cycle MHD power plants.

Phase II, if funded, will consist of a detailed conceptual design study of specific closed cycle MHD power plant configurations identified in Phase I.

In this Phase I study, the technical feasibility, capital cost and cost of electricity for power plants using direct combustion of coal or coal derived fuel were parametrically evaluated. Three Reference Plants, differing primarily in selection of the heat source for the argon heat exchanger, were developed. Reference Plant 1 incorporates a direct coal-fired combustor having high slag rejection. Reference Plants 2 and 3 are systems which employ on-site integrated gasifiers to provide a clean fuel for combustion. Reference Plant 2 has an advanced technology pressurized gasifier and Reference Plant 3 uses a state-of-the-art atmospheric gasifier.

A total of 30 parametric cases were considered in this study, with performance and cost data generated for each plant. A complete description of each of the parametric cases is given in Section 3.0. These cases were, in general, based on Montana Rosebud coal using various hot gas and cold gas clean-up systems; a one stage and a two stage, atmospheric, direct fired coal combustor was analyzed, and a total of 6 different gasifier systems including both low Btu and medium Btu designs were included. Plant sizes were nominally 1000 MWe. The MHD plasma was argon seeded with 0.1% cesium.

1.3 Objective

The objective of this study was to develop preliminary information on the performance and cost for commercial scale coal-fired closed cycle MHD power plants and to assess the relative merits of various plant configurations. These plants were selected to reflect the best potential performance and cost of electricity for CCMHD plants.

1.4 Project Team

This study was conducted by a project team with Gilbert Associates, Inc., as the prime contractor and Program Manager; FluiDyne Engineering Corporation and TRW, Inc., were subcontractors.

FluiDyne Engineering Corporation provided performance, cost, material and development information for the high temperature ceramic argon heaters suggested for all the parametric cases studied, including those fired by slag-laden combustion products and clean fuels from combustion of gasified coal.

TRW, Inc., which is currently designing a two-stage pressurized coal-fired combustor for the DOE OCMHD program, provided combustor performance, cost, material, and development data for the Reference Plant 1 direct coal fired parametric cases.[a]

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[a]The subcontractor report to GAI contains proprietary information and is, therefore, not included in this report.

1.5 Ground Rules and Specifications

This section lists the ground rules and specified parameters that have been imposed or assumed in conducting this study.

For each of the plant configurations studied, the MHD nozzle, channel and diffuser were treated as an energy conversion device having a specified enthalpy extraction ratio and a specified isentropic generator efficiency. The assumption of treating the nozzle, channel and diffuser as an energy conversion device having a given "black box" performance, was specified in the Statement of Work of the initial contract. This assumption has been consistently applied throughout the main body of this report. A subsequent modification to the contract was received which required detailed channel calculations to substantiate the assumed performance parameters. Although these detailed channel results were not incorporated in the main body of this report, they are included in Appendix A.

Plant performance was based on an average day condition of 59 F, at an ambient pressure of 14.7 psia, and with a relative humidity of 60 percent. The ground rules for this study were selected such that a direct comparison of power plant construction and capital costs determined in this study could be made with results obtained in the open cycle PSPEC studies. All cost numbers are reported in mid-1978 (1978-1/2) dollars.

Montana Rosebud coal, which has a high moisture and low sulfur content, was the primary fuel. Illinois No. 6 coal was an alternate fuel having the general characteristic of low moisture and high sulfur content. The specified properties for Montana Rosebud and Illinois No. 6 coals are shown in Table 1-1. These coal properties are consistent with the properties specified for the Engineering Test Facility (ETF) conceptual design studies and the PSPEC studies sponsored by DOE/MHD.

The base fuel cost was assumed to be $1.05 per million Btu (MBtu). The sensitivity of the cost of electricity (COE) to inflationary increases in fuel cost from $1.05 to $1.50 per MBtu was investigated. In addition, the following range of fuel cost escalation was specified to allow for cost uncertainties:

Lower Limit: Fuel costs increase with general inflation of 6.5% per year. Base fuel cost remains constant in mid-1978 dollars.

Upper Limit: Fuel costs increase with general inflation of 6.5% per year plus a real cost increase of 3% per year.

A baseload plant with a 30 year life and an availability that permits a 65 percent capacity factor were specified. During plant construction, the capital cost was increased by applying an escalation factor of 6.5 percent per year on unused funds and an interest rate of 10 percent per year. The escalation and interest cost factors applied to plant construction are shown in Table 1-2. The specified cash flow during the construction period is given in Figure 1-1.

A labor rate of $14.20/hour, representative of a combined civil, mechanical, and electrical rate, was used for all construction site labor. This rate was based on a weighted average for a Middletown, USA construction

TABLE 1-1

COAL AND ASH ANALYSES

Ash Analysis, %	ILLINOIS # 6	MONTANA ROSEBUD
SiO_2	41.4 ± 5.4	37.6
Al_2O_3	19.3 ± 6.8	17.3
Fe_2O_3	22.3 ± 6.8	5.1
T_1O_2	0.9	0.7
P_2O_5	0.12	0.4
CaO	5.4 ± 3.3	11.0
MgO	1.7 ± 1.3	4.0
Na_2O	0.6 ± .2	3.1
K_2O	2.1 ± .4	0.5
SO_3	7.5 ± .6	17.5
Initial Deformation Temp. F	1960 ± 70	2190 ± 230
Softening Temp. F	2030 ± 70	2230 ± 240
Fluid Temp. F	2260 ± 200	2280 ± 240
Proximate Analysis, Coal, as rec'd, %		
Moisture	8.9	22.7
Volatile Matter	38.0	29.4
Fixed Carbon	41.7	39.2
Ash	11.4	8.7
Ultimate Analysis, %		
Hydrogen	5.4	6.0
Carbon	62.4	52.1
Nitrogen	1.2	.79
Oxygen	16.3	31.5
Sulfur	3.3	0.85
Heating Value, Wet, Btu/lb	11265	8920
Heating Value, Dry, Btu/lb	12370	11560
Coal Rank	HVCB	Subbit B

TABLE 1-2 - ESCALATION AND INTEREST COST FACTORS

[Escalation + Interest = Total. Annual rates: escalation, 6.5 percent; interest, 10 percent.]

Time from start of design to powerplant completion, yr	Escalation	Interest on obligated funds	Total
	Cost factor		
0	1.000	1.000	1.000
.5	1.018	1.022	1.040
1.0	1.037	1.044	1.081
1.5	1.056	1.069	1.125
2.0	1.076	1.094	1.170
2.5	1.096	1.122	1.218
3.0	1.116	1.151	1.267
3.5	1.137	1.182	1.319
4.0	1.158	1.214	1.372
4.5	1.179	1.249	1.428
5.0	1.202	1.285	1.487
5.5	1.224	1.324	1.548
6.0	1.247	1.365	1.612
6.5	1.270	1.409	1.679
7.0	1.294	1.454	1.748
7.5	1.319	1.503	1.522
8.0	1.344	1.554	1.698
8.5	1.369	1.609	1.978
9.0	1.395	1.666	2.061
9.5	1.422	1.726	2.148
10.0	1.449	1.790	2.239

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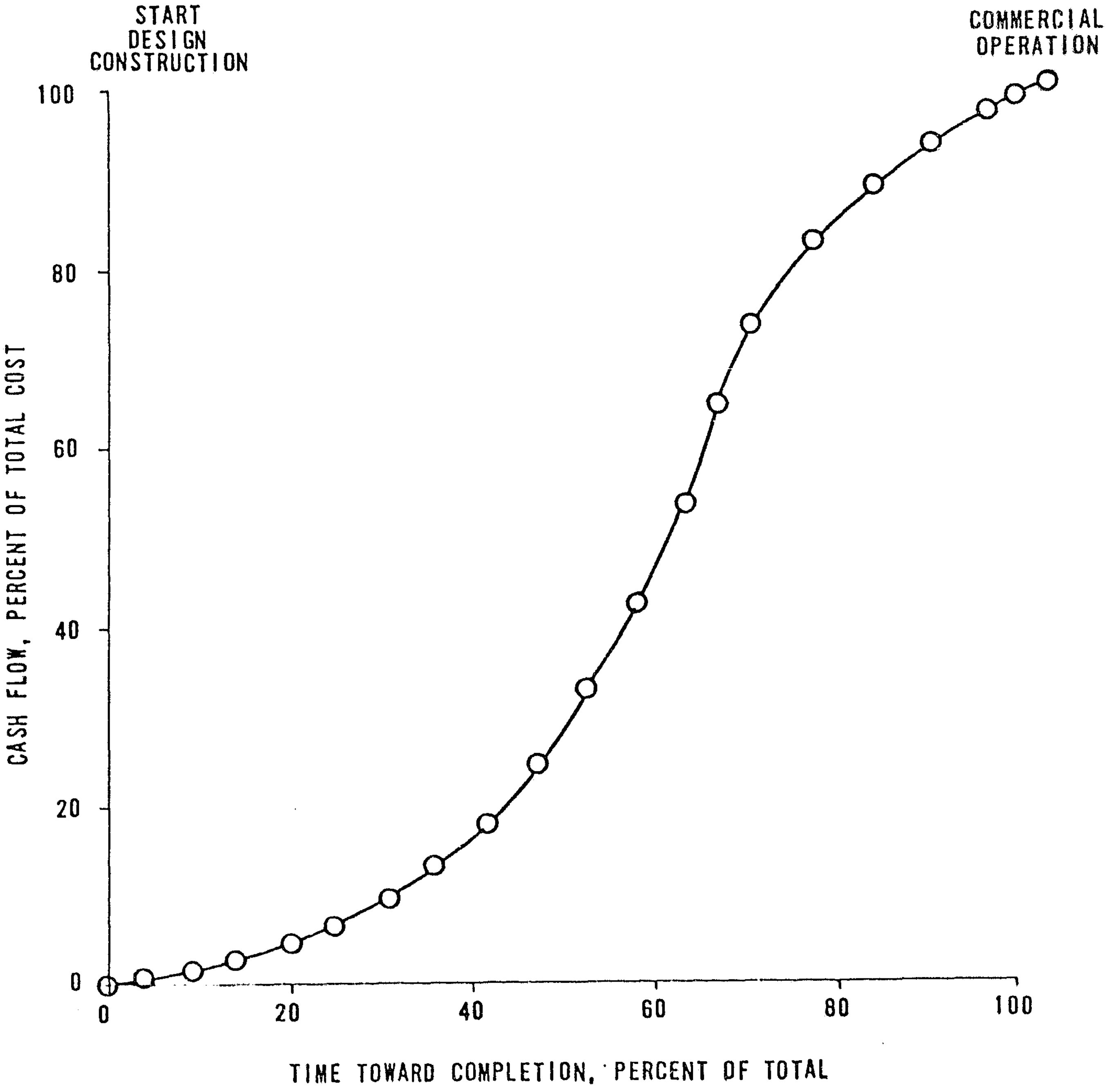

FIGURE 1-1
CASH FLOW DURING CONSTRUCTION

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site updated to mid-1978 dollars. An upper labor rate of $17.04/hour was used for sensitivity analyses.

A fixed charge rate of 18 percent per year was specified for estimating the capital cost contribution to the cost of electricity. This rate includes the cost of money, taxes, depreciation, insurance, and working capital.

The specified format for reporting cost numbers was based on the latest Department of Energy (DOE) Code of Accounts modified for closed cycle MHD plants.[6] All economic parameters specified for this study are shown in Table 1-3.

The environmental emission standards were based on the New Stationary Sources Performance Standards promulgated by the Environmental Protection Agency (EPA) in the June 11, 1979, Federal Register. The environmental standards applicable to this study for Montana Rosebud and Illinois No. 6 coals that are either direct fired (Reference Plant 1) or gasified (Reference Plants 2 and 3) are summarized in Table 1-4.

Table 1-3

SPECIFIED ECONOMIC PARAMETERS

Base Year	mid-1978
Labor Rate, $/hr	14.20
Interest, per annum	10%
Escalation, per annum	6.5%
Fixed Charge Rate, per annum	18%
Capacity Factor	0.65
Plant Life, yrs	30
Fuel Cost, $/MBtu	1.05
Levelizing Factor (w/o real escalation)	2.004
Ranges for Sensitivity Analysis	
Fuel Cost	1.05 to 1.50
Real Fuel Escalation Rate	1, 2, & 3%
Site Labor Rate	14.20 to 17.04

TABLE 1-4

SUMMARY OF ENVIRONMENTAL CONSTRAINTS

Pollutant	Type of Firing	Montana Rosebud Coal	Illinois No. 6 Coal
SO_2	All Cases	0.57 lb/MBTU (70% Removal)	0.60 lb/MBTU (90% Removal)
NOx	Direct Fired	0.50 lb/MBTU	0.60 lb/MBTU
	Gasification	0.50 lb/MBTU	0.50 lb/MBTU
Particulate	All Cases	0.03 lb/MBTU	0.03 lb/MBTU

2.0 SUMMARY

Gilbert Asociate, Inc., has conducted a parametric study of CCMHD plants which provides information that can be used to assess the relative merits of alternative MHD systems. The ECAS study indicated that CCMHD potentially had overall efficiencies which were among the highest of the advanced energy conversion systems considered; however, cost of electricity estimates were higher than other advanced technologies. The high efficiencies predicted in ECAS, recent developments in combustor technology and the need for updated consistent cost and performance information has motivated this study.

The closed cycle plants considered in this study consisted of an MHD topping cycle utilizing a non-equilibrium argon seeded with cesium plasma, a steam bottoming cycle and a combustion system which is external to both the argon and steam systems. The argon topping system and the steam bottoming system remained nearly unchanged for the majority of the configurations studied. The primary difference in the parametric cases was the method of firing the combustor for the high temperature regenerative argon heat exchanger.

Three Reference Plants were selected:

- Reference Plant 1 - Direct coal fired combustor with approximately 85% slag rejection.
- Reference Plant 2 - Advanced pressurized gasifier system
- Reference Plant 3 - State-of-the-art atmospheric gasifier system.

A base case for each reference plant was selected and a total of 30 parametric cases were defined in order to show the effect on overall plant performance of variations in operating parameters. The selected cases included variations in the following parameters:

Reference Plant 1

- Coal type
- Cleanup system
- Combustor stages
- Combustor pressure
- Steam condenser pressure
- Type of bottoming cycle

Reference Plant 2

- Type of gasifier system
- Coal type
- Cleanup system
- Gasifier pressure
- Plant size
- Oxidant
- Channel enthalpy extraction ratio, channel pressure

Reference Plant 3

- Type of gasifier system
- Coal type
- Cleanup system
- Plant size

The argon topping cycle, the steam bottoming cycle and the combustion system for the base cases of each reference plant were completely integrated and the performance predictions were optimized. For each parametric case, however, each plant variable noted earlier was changed independently of all other parameters; as a result, the overall plant performance for the parametric cases are not necessarily optimized.

For each case the combustion gas temperature was constrained by flue gas recirculation to 3350 F in order to limit the regenerative heat exchanger ceramic brick temperature to 3300 F. The stoichiometric ratio was 1.05. For plant configuration considered in the parametric study, the MHD nozzle, channel and diffuser were treated as an energy conversion device having a specified enthalpy extraction ratio and a specified isentropic generator efficiency. For the base cases and all but two parametric cases, the enthalpy extraction ratio was 36% and the channel efficiency was 78%. The steam bottoming cycle was supercritical having throttle conditions of 3500 psi/1000 F/1000 F.

Table 2-1 summarizes the performance and cost data for the three base cases. The efficiency of the direct coal fired system is higher than the efficiency of either of the gasifier plants. Gasifier system inefficiencies decrease the overall plant efficiency. The direct coal fired system, however, has the highest capital cost, primarily because of the cost of the regenerative argon heat exchanger. Slag carryover from the combustor into the heater necessitated a hot bottom design to minimize slag solidification in the core passages of the heater.

Table 2-1

Reference Plant Summary

Reference Plant	Combustion System	Power Output MWe	Plant Efficiency %	Heat Exchanger Cost $\$ \times 10^6$	Capital Cost $\$ \times 10^6$	Levelized COE mills/kW-hr
1	Direct Coal Fired	1000.0	43.2	244.6	967.2	53.90
2	Pressurized Gasifier	1012.6	39.4	54.3	958.7	54.85
3	Atmospheric Gasifier	994.5	36.1	106.6	873.6	54.05

These heaters require larger core brick hole sizes than those using clean fuel. The result is that coal fired heaters are larger and more expensive than those fired by gasifier fuel. The levelized cost of electricity was 53.9 mills/kw-hr. Slagging regenerative heat exchangers will require a significant development effort before they can be used.

With an advanced pressurized gasification system, the overall plant efficiency decreases to 39.4%. However, the capital cost also decreases to \$958.7 x 10^6 which tended to keep the levelized cost at about the same level as that for the direct coal fired case (about 1 mill/ kW-hr difference). With a clean pressurized combustion gas, the cold bottom regenerative heat exchanger system is more compact and less costly. The heat exchanger cost decreased to \$54.3 x 10^6 for the pressurized case compared to \$244.6 x 10^6 for the direct coal fired case. Although the heat exchanger is less expensive, the pressurized gasifier system was more expansive than a direct fired combustor and, in addition, the pressurized system utilizes an expansion turbine and associated equipment which is not present in atmospheric systems.

An atmospheric gasification system has an efficiency of only 36.1% which was the lowest of the plants studied. The cold bottom regenerative heat exchanger designed for clean atmospheric pressure gas has a cost of \$106.6 x 10^6 which is almost twice that for the pressurized case but is still much less than that for a direct fired case. The capital cost of the atmospheric system was lower than any of the other plants considered because of the relatively low regenerative heater cost, less expensive gasifier system (compared to the pressurized case) and absence of turbomachinery to expand the combustion gases to atmospheric pressure. The levelized cost of electricity was not significantly different for an atmospheric gasifier than for the other two case. Although efficiency decreased the cost of electricity did not change significantly because the capital cost also decreased.

Sulfur emission levels for the base case of Reference Plant 1, direct coal fired combustor were reduced to the NSPS limit of 0.57 lb/10^6 Btu of coal input using a spray drier (dry scrubber). The advanced gasifier system of Reference Plant 2 included a Morgantown iron oxide hot gas cleanup system. With this system, sulfur levels were reduced to 0.2 lb/10^6 Btu of coal input, well below the NSPS limit. The atmospheric gasifier plant (Reference Plant 3) included a cold gas Stretford cleanup system to reduce the sulfur level to a very low level of 6.2×10^{-4} lb/10^6 Btu of coal input. The absolute level of sulfur emission using a Morgantown iron oxide or a Stretford system, were based on published information and have to be considered preliminary pending verification or development.

Particulate emissions for all reference plants were controlled using a baghouse filter. The direct coal fired combustor system of Reference Plant 1 requires a baghouse with a particulate removal efficiency of 98.5% and the two gasifier systems require an efficiency of 97.4%. Both efficiencies are readily achievable with current technology.

Nitrous oxide emission levels are substantially more difficult to analyze because of the lack of experimental data. Combustion temperatures in closed cycle MHD systems are essentially equivalent to those expected in conventional fossil fired plants, therefore, NOx problems should not be more severe with closed cycle MHD than with conventional plants. Flame temperatures in closed cycle MHD are about 3500 F (compared to 4600 F for open cycle MHD). At this relatively low temperature there is some evidence that thermal NOx will not be formed, that only fuel-bound nitrogen will form NOx and that expected levels are not exceptionally high. Conventional NOx control techniques can be adapted for closed cycle systems which include: firing level and angle control, flue gas recirculation and stoichiometric selection. In addition, there is also the possibility that NOx decomposition may occur by catalytic reaction on the alumina refractory surface of the argon regenerative heat exchanger.

Parametric variations in plant operating parameters about each base case were considered and the results discussed in Section 3.0 of t' report. These studies show that performance of the direct coal fired of Reference Plant 1 could be improved by utilizing a single stage combustor with less slag rejection (however, this would compound the regenerative heater design problem), using a pressurized combustor or reducing the steam condenser pressure. The most significant increase in plant efficiency resulted from the use of a pressurized combustor. For a combustor pressure of 6 atm the overall plant efficiency was 44.9% (compared to 43.2% for the atmospheric combustor of base case 1.0). This system included a turboexpander in the combustion gas stream to lower the pressure to 1 atm before exiting the plant. Inclusion of a pressurized system, however, increases the plant cost and could increase the development cost of the system. Other parametric cases considered which lowered plant efficiency included the use of Illinois No. 6, a wet scrubber rather than a spray drier, and the use of a subcritical 2400 psi/1000 F/1000 F steam bottoming cycle.

Pressurized gasifiers were considered. For this specific application, the IGT and Westinghouse gasifier systems had the highest overall performance. The Texaco gasifier system did result in as high a plant efficiency as the IGT or Westinghouse gasifier systems; however, this was a parametric variation and as such, the system was not completely integrated. The use of a Stretford cold gas cleanup system in place of the Morgantown iron oxide system used with the IGT gasifier or the in-bed (hot gas) cleanup system of the Westinghouse gasifier resulted in a decrease in plant efficiency of 2.9 percentage points for the IGT system and 8.8 percentage points for the Westinghouse system. The use of oxygen to produce medium Btu gas in the gasifiers is not advantageous because of the energy penalty resulting from the air separation unit.

It has been suggested in other studies, that closed cycle MHD is more attractive for smaller plant sizes. This contention could not be varified in this study because the MHD generator was treated as an energy conversion device having a specified constant performance (independent of plant size). The specified enthalpy extraction ratio was 36% except for two case in which this ratio was arbitrarily increased to 38% and 40%. As expected, the overall plant performance was improved with these higher performance MHD generators.

Atmospheric gasifiers were considered in Reference Plant 3. The base case used a Combustion Engineering (CE) system; a Winkler and a Wellman gasifier were included as parametric cases. The CE system had slightly higher performance than the Winkler for a nominal plant size of 1000 MWe. For the Wellman gasifier which has a low capacity, the plant size was decreased to 100 MWe. The Wellman gasifier, adapted to a small plant, resulted in an overall plant efficiency of 43.6%. A spray drier gas cleanup system was also considered as a parametric case was the use of Illinois No. 6 coal.

Because of the uncertainties raised by the assumption of treating the MHD generator as an energy conversion devices having a specified performance, NASA-LeRC requested Gilbert Associates, Inc., to perform a series of non-equilibrium closed cycle MHD generator calculations as an add-on task to the original system engineering parametric study. The intent of this task was to evaluate whether the assumed generator performance could be achieved with the specified flow conditions. Results of this MHD generator study indicate that the specified channel performance is somewhat optimistic. The calculated enthalpy extraction ratio, with the stated flow conditions, were about 3 percentage points less than that specified by NASA-LeRC for a 1000 MWe plant. Further this study shows that if the plant size is reduced to 100 MWe, the enthalpy extraction ratio will decrease to about 31.6% (33% for a 1000 MWe plant). Details of this study are given in Appendix A.

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The viability of coal-fired closed cycle MHD depends primarily on the development of the regenerative argon heat exchanger and the non-equilibrium MHD channel. An atmospheric coal combustor or an atmospheric gasifier with cold gas cleanup are essentially state-of-the-art. Other major system components, such as compressors, heat exchangers, coal handling equipment, etc., are commercially available and require only a demonstration that they can be integrated into a total plant system.

From this study the following conclusions can be drawn:

- o Coal fired closed cycle MHD plants can be built which have efficiencies in the range of 40 to 45%. This efficiency level is slightly lower than oxygen enriched open cycle plants of the same size; however, direct-fired open cycle MHD plants are expected to have efficiencies of at least 50%. Therefore, closed cycle plant efficiencies compare favorably with oxygen enriched open cycle plants but are inferior to direct-fired open cycle plants.

- o The levelized cost of electricity (COE) in mid-1978 dollars is projected to be around 55 mills/kW-hr for the closed cycle system. For an oxygen enriched open cycle system the COE is about 42 mills/kW-hr. The direct-fired open cycle COE will be significantly less. Although the efficiency of closed cycle plants are comparable with oxygen enriched open cycle plants, the cost of electricity is significantly higher which confirms the ECAS conclusions.

- o The argon regenerative heater represents the key component which effects both the cost and performance of the plant. For a direct coal fired combustor with slag carryover, the development and technical problems are essentially identical to those of a direct-fired open cycle regenerative air heater. Regenerative argon heater development for gasifier systems will be less complex than for direct coal-fired systems and will essentially be analogous to the development of separately-fired open cycle air heaters. Regenerative heater development costs are expected to be high. Technical problems include, not only the basic heater development, but also a system which will minimize the amount of combustion gas (contamination) carried over into the argon during the cyclic operation.

- o Non-equilibrum MHD channel operation will have to be demonstrated. Steady operation of an unstable, turbulent plasma operation requires large scale verification, and long channel life-times will have to be demonstrated. The small scale closed cycle MHD channel tests planned at the Institute of Technology, Eindhoven, Netherlands should provide applicable design information.

- o Direct coal fired closed cycle MHD plants have the highest efficiency, but introduce regenerative argon heat exchanger problems and have a high capital cost.

- Advanced pressurized gasifier closed cycle MHD plants have acceptable performance with less expensive regenerative argon heat exchangers; however, the pressurized gasifier development problem has not been completely solved.

- Atmospheric gasifier closed cycle MHD plants project a near state-of-the-art configuration with minimum capital cost; however, the plant efficiency is very low.

Results of this study should be considered pre-conceptual. Phase II of this investigation should be continued if more accurate cost and peformance values are required. In Phase II, a more detailed conceptual design of a selected plant would be developed.

3.0 POWER PLANT DESIGN AND PERFORMANCE

In order to investigate the parameteric variations in performance, three reference plant configurations, differing primarily in the selection of the heat source for the argon heat exchanger, were defined:

Reference Plant 1. Power plant with a two stage atmospheric combustor directly fired with dried pulverized coal.

Reference Plant 2. Power plant with an advanced pressurized gasifier integrated with the MHD cycle having a hot gas cleanup system.

Reference Plant 3. Power plant with state-of-the-art atmospheric gasifiers integrated with the MHD cycle having a cold gas cleanup system.

A closed cycle MHD plant consists, essentially, of three basic systems: (1) the argon closed cycle topping system, (2) the steam bottoming cycle, and (3) the combustion system which is external to the primary argon and steam cycles. The argon and steam systems were essentially identical for all three reference plants. The combustion system and the integration of this system with the overall plant configuration represented the significant differences in the three reference plants. Variations in coal type, cleanup system, type of gasifier, pressurization of the combustor, plant size, channel performance and steam bottoming cycle design resulted in a total of 30 parametric cases.

The three reference plants were of a nominal 1000 MWe size with Montana Rosebud coal. The specified channel efficiency was 78% with an enthalpy extraction of 36%. The topping cycle working fluid was argon having a pressure of 10 atm and a temperature of 3100 F at the entrance to the MHD nozzle. The steam bottoming cycle was supercritical with throttle conditions of 3500 psi/1000 F/1000 F.

The assumed performance parameters for all parametric cases studied are given in Table 3-1.

3.1 Reference Plant 1 - Direct Coal-Fired Combustor

The assumed design conditions for the Base Case 1.0 and all parametric variations for Reference Plant 1 are defined in Table 3-2. The schematic diagram and the heat and mass balance for the base case are shown in Figure 3-1.

3.1.1 Topping Cycle

The argon-cesium plasma temperature of 3100 F and pressure of 10 atm (147 psia) entering the MHD generator were given as a ground rule for this study. The plasma mass flow rate of 2772 kg/sec (6113 lb/sec) entering the generator was based on the desired output power of 1000 MWe from the MHD inverter. With a specified channel efficiency of 78% and an enthalpy extraction ratio of 36%, the pressure ratio across the MHD generator was determined to be 4.8. The MHD generator is cooled using demineralized water in a cooling loop which is completely separate from the steam bottoming cycle. In this study, the low grade heat from the channel cooling water was transferred to the atmosphere.

Table 3-1

ASSUMED PERFORMANCE PARAMETERS

Stoichiometric Ratio	1.05
Inverter Efficiency	98.5%
Argon Compressor Efficiency	89%
Air Blower Efficiency	90%
Air Compressor Efficiency	90%
Boiler Feed Pump Efficiency	75%
Recirculation Fan Efficiency	90%
Argon Heat Exchanger Energy Loss	
Direct Coal Fired	2%
Gasifier	1%
High Pressure Turbine Efficiency	90%
Intermediate Pressure Turbine Efficiency	88%
Low Pressure Turbine Efficiency	85%

Table 3-2

REFERENCE PLANT 1 - DIRECT COAL FIRED

PARAMETRIC CASES

Case	1.0	1.1	1.2	1.3	1.4	1.5	1.6					
Variation	Base Case	Coal Type	Clean-Up	Combustor	Combustor Pressure	Steam Plant	Steam Plant					
Power Output, MWe	1000											
Combustion Loop												
Combustor	2 Stg			1 Stg	2 Stg							
Pressure, atm	1				6	1						
Coal Type	MR	Ill. #6		MR								
Cleanup	Spray Dryer		Other	Spray Dryer								
Argon Loop												
Channel Efficiency, %	78											
Enthalpy Extraction, %	36											
Pressure, atm	10											
Working Fluid	Argon											
Inlet Temp, °F	3100											
Steam Plant												
Throttle Conditions	*						***					
Heater Arrangement	**											
Back Pressure, in. Hg.	2.5					1.5	2.5					
Output	Balanced											

* 3500/1000/1000
** DA only or best possible arrangement
*** 2400/1000/1000

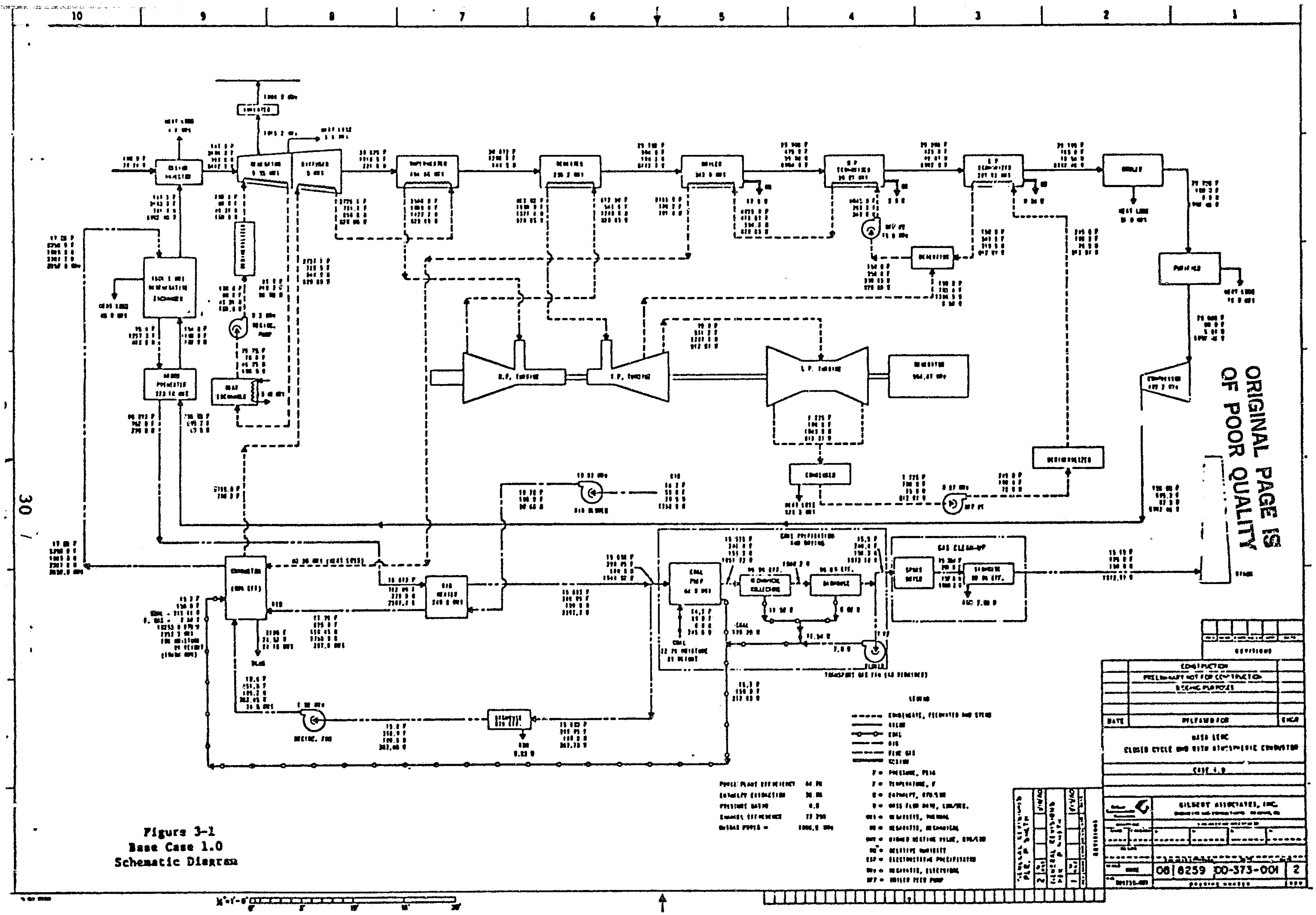

Figure 3-1
Base Case 1.0
Schematic Diagram

The approximately 1800 F plasma leaving the diffuser enters a series of heat exchangers where steam is produced, superheated and reheated. The cesium seed condenses and is collected in the boiler, the high pressure economizer and the low pressure economizer. The seed material is recirculated back to the cesium injection point.

In order to minimize the work of the compressor and to limit the compressor discharge temperature, the argon must be cooled to the lowest possible temperature at the inlet to the compressor. A cooler is placed in the flow stream to lower the argon temperature by extracting low grade heat which is then discharged to the environment. In this study, no attempt has been made to utilize low grade heat through cogeneration. Current compressor technology has an argon discharge temperature limitation of less than 600 F. However, it was felt that with moderate design changes and by extrapolating current technology to the time frame where CCMHD would be competative, a discharge temperature of about 700 F is reasonable. For all the parametric cases studied, the compressor inlet temperature has been constrained to 80 F in order to minimize the discharge temperature.

The pressure drop through the heat exchangers and across the MHD generator establishes the required pressure ratio of the argon compressor. For Reference Plant 1, the compressor pressure ratio was 5.44. The overall system pressure loss ratio was 0.118.

A single stage axial flow argon compressor was recommended. Three compressor configurations were considered: (1) single stage compressor, (2) two stage compressor with interstage cooling (interstage cooler heat transfer discharged to the environment), and (3) two stage compressor with interstage cooling using boiler feed water. Although the work of compression was less with a two stage compressor, the overall plant efficiency was also less than that predicted with a single stage compressor because of the energy loss by interstage cooling. With a two stage compressor, the discharge temperature could be limited to 600 F (current state-of-art). If the heat required for interstage cooling (low grade heat) is rejected to the environment, the required heat transfer in the regenerative heat exchanger must be increased in order to achieve the 3100 F MHD generator inlet temperature. The net effect is a decrease in overall plant efficiency of about one percentage point. An alternative configuration using a split low temperature economizer feed water flow for interstage cooling was also studied. With this arrangement, not only was the heat transfer requirement of the regenerative heater greater than that for a single stage compressor, but the heat rejected in the cooler was also increased because of the reduced feed water flow rate in the low temperature economizer (resulting in a higher argon temperature at the exit of the low temperature economizer). With a two stage split economizer flow interstage cooler configuration, the overall plant efficiency was 3.7 percentage points less than with a single stage compressor.

Upon leaving the compressor, the argon enters a metallic preheater where the temperature is increased to 1100 F. A ceramic hot bottom regenerative heat exchanger, fired by combustor products, then raises the argon temperature to slightly more than 3100 F at which time the cesium is injected.

The heat loss in the regenerative heat exchanger has been estimated to be between 1 and 5 percent of the heat exchanger duty. In this study, a 3% heat loss has been assumed for all direct coal-fired cases and a 1% heat loss for clean fuel. The system sensitivity analysis indicated that an overall plant efficiency decrease of 1.4 percentage points can be expected when the heat loss is increased from 1 to 5 percent. A complete description and the operating characteristics of the argon regenerative heat exchanger are given in Appendix B.

3.1.2 Steam Bottoming Cycle

The steam bottoming cycle is a supercritical 3500 psi/1000 F/1000 F unit with a condenser pressure of 2.5 in.Hg. The turbine-generator produces power to operate the argon compressor and to supplement the MHD generated power for the grid.

The demineralizer is operated at 245 psi. Steam from the intermediate pressure turbine is extracted for the deaerator which operates at 150 psi. Feed water is used to cool the combustor and the diffuser (the MHD channel has an independent cooling loop).

The boiler design in a CCMHD system differs from both a conventional design and from an OCMHD design primarily because of the mode of heat transfer. In a CCMHD boiler, the working fluid is an argon-cesium mixture free of particulates. Heat transfer occurs almost completely by convection with only a small contribution from gas radiation. In conventional and OCMHD systems, heat transfer is principally by particle radiation. A CCMHD boiler, therefore, requires much larger heat transfer surfaces.

3.1.3 Combustion System

A two stage, atmospheric combustor directly-fired with pulverized coal (70% through a 200 mesh screen) dried to 10% moisture was used in Reference Plant 1. The combustion gas exit temperature was constrained to 3350 F in order to limit the regenerative heat exchanger ceramic brick temperature to 3300 F. Combustion air was preheated in a metallic heat exchanger prior to entering the combustor. The stoichiometric ratio was 1.05. Flue gas recirculation to the combustor was used to limit the combustor flame temperature to 3350 F.

At a flame temperature of 3350 F, the environmental problems encountered in CCMHD systems are directly comparable with those of a conventional system. Thermal NOx does not pose a significant problem at these relatively low flame temperatures. The only NOx formation is from fuel bound nitrogen. With flue gas recirculation and staged combustion, NOx levels are predicted to be less than current environmental standards. SOx and particulate standards can be satisfied with the same equipment used in conventional plants, e.g., cyclones, baghouses and wet or dry scrubbers.

Hot combustion gases are used to heat the argon working fluid in the regenerative heater, the argon preheater and the combustion air heater prior to entering the coal preparation and drying subsystem. Montana Rosebud coal, dried from 22.7% moisture (as received) to 10% moisture (Reference Plant 1) using flue gas, is pulverized before entering the combustor. A cyclone separator and a baghouse remove the majority of the particulate before the flue gas enters the spray dryer (or dry scrubber). A final baghouse removes most of the remaining particulate before the gas is exhausted through the stack.

3.1.4 Reference Plant 1 Performance

The energy balance and performance summary for Base Case 1 and the six parametric cases described in Table 3-2 are given in Table 3-3 and 3-4, respectively.

In Table 3-4, the efficiency terms are defined as follows:

a. Thermodynamic Efficiency

The thermodynamic efficiency is defined as the gross power output of the combined cycle divided by the total heat input to the combined cycle.

$$\eta_{TH} = \frac{P_{MHD} + P_{STMG}}{Q_{AR} + Q_{STMC}}$$

where P_{MHD} is the MHD inverter power output, P_{STMG} is the net power output from the steam turbine-generator (gross power output minus the argon compressor power), Q_{AR} is the heat input to the argon in the regenerative heater and Q_{STMC} is the heat input to the steam cycle from the combustor.

b. Overall Plant Efficiency

The overall plant efficiency is defined as the ratio of the net power output to the grid to the total thermal input power of the coal.

TABLE 3-3
REFERENCE PLANT 1

ENERGY BALANCE

CASE	1.0	1.1	1.2	1.3	1.4	1.5	1.6
VARIATION	REFERENCE PLANT	ILL. #6 COAL	CLEAN-UP	1 STAGE COMB.	PRESSURIZED COMB.	1.5 IN. HG. COND. PRESS.	SUBCRITICAL STEAM PLANT
ENERGY INPUT							
ARGON COMPRESSOR, MW	492.2	492.2	492.2	492.2	492.2	492.2	492.2
AIR COMPRESSOR/BLOWER, MW	19.8	20.2	20.1	19.6	187.0	19.8	19.8
RECIRCULATION FAN, MW	1.9	4.6	2.4	1.8	0.8	1.9	1.9
BOILER FEED PUMPS, MW	16.6	16.6	16.6	16.6	16.6	16.3	12.8
MHD COOLING LOOP PUMP, MW	0.3	0.3	0.3	0.3	0.3	0.3	0.3
TOTAL POWER INPUT, MW	530.8	533.9	531.6	530.5	696.9	530.5	527.0
COAL (AS REC'D), MW	2316.7	2373.4	2354.7	2297.1	2410.1	2317.0	2316.5
COMBUSTION AIR, MW	38.0	38.8	38.6	37.7	39.5	38.0	38.0
TOTAL HEAT INPUT, MW	2354.7	2412.2	2393.3	2334.8	2449.6	2355.0	2354.5
TOTAL ENERGY INPUT, MW	2885.5	2946.1	2924.9	2865.3	3146.5	2885.5	2881.5
ENERGY OUTPUT							
MHD INVERTER, MW	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0
STEAM TURBINE, MW	554.4	554.4	554.4	554.2	555.1	568.4	540.8
GAS TURBINE, MW	-	-	-	-	248.7	-	-
TOTAL POWER OUTPUT, MW	1554.4	1554.4	1554.4	1554.2	1803.8	1568.4	1540.8
REGENERATIVE HEAT EXCHANGER, MW	48.0	48.0	48.0	48.0	48.0	48.0	48.0
ARGON COOLER, MW	52.0	52.0	52.0	52.0	52.0	52.0	52.0
ARGON PURIFIER, MW	16.0	16.0	16.0	16.0	16.0	16.0	16.0
CONDENSER, MW	828.4	828.4	828.4	828.1	829.5	813.9	838.2
SPRAY DRYER/CLEAN-UP, MW	13.1	59.2	29.7	16.1	13.5	13.1	13.1
SLAG, MW	24.2	25.5	24.2	0.0	24.8	23.8	23.8
STACK LOSS, MW	313.3	326.8	334.2	314.6	326.1	313.3	313.3
INVERTER, MW	15.2	15.2	15.2	15.2	15.2	15.2	15.2
CESIUM SYSTEM, MW	4.0	4.0	4.0	4.0	4.0	4.0	4.0
COMBUSTOR SYSTEM HEAT LOSS, MW	1.3	1.3	1.3	1.4	3.3	1.3	1.3
MHD GENERATOR LOSS, MW	5.5	5.5	5.5	5.5	5.5	5.5	5.5
MISCELLANEOUS, MW	10.1	9.8	12.0	10.2	4.8	11.0	10.3
TOTAL HEAT OUTPUT, MW	1331.1	1391.7	1370.5	1311.1	1342.7	1317.1	1340.7
TOTAL ENERGY OUTPUT, MW	2885.5	2946.1	2924.9	2865.3	3146.5	2885.5	2881.5

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TABLE 3-4
REFERENCE PLANT 1

PERFORMANCE SUMMARY

CASE	1.0	1.1	1.2	1.3	1.4	1.5	1.6
VARIATION	REFERENCE PLANT	ILL. #6 COAL	CLEAN-UP	1 STAGE COMB.	PRESSURIZED COMB.	1.5 IN. HG. COND. PRESS.	SUBCRITICAL STEAM PLANT
COAL INPUT (AS REC'D), MW	2316.7	2373.4	2354.7	2297.1	2410.1	2317.0	2316.5
TOTAL POWER OUTPUT, MW	1554.4	1554.4	1554.4	1554.4	1803.8	1568.4	1540.8
TOTAL POWER INPUT, MW	530.5	533.6	531.3	530.2	696.6	530.2	526.7
AUXILIARY POWER, MW	23.1	23.7	23.5	22.9	24.1	23.1	23.1
BUSBAR POWER, MW	1000.8	997.1	999.6	1001.1	1083.1	1015.1	991.0
RECIRCULATION, %	15.7	18.7	15.9	15.5	12.5	15.7	15.7
STACK GAS TEMPERATURE, F	195	195	245	195	195	195	195
STOICHEOMETRIC RATIO	1.05	1.05	1.05	1.05	1.05	1.05	1.05
COAL TYPE	MR	Ill. #6	MR	MR	MR	MR	MR
COAL MOISTURE, %	10	2	10	10	10	10	10
COAL HHV, BTU/lb	10,404	12,118	10,404	10,404	10,404	10,404	10,404
COMBUSTOR PRESSURE, psia	17.3	17.3	17.3	17.3	85.7	17.3	17.3
CLEAN-UP SYSTEM	Spray Dryer	Spray Dryer	Wet Scrubber	Spray Dryer	Spray Dryer	Spray Dryer	Spray Dryer
STEAM PLANT	Super Critical	Super Critical	Super Critical	Super Critical	Super Critical	Super Critical	Sub Critical
CONDENSER PRESSURE, In.Hg.	2.5	2.5	2.5	2.5	2.5	1.5	2.5
MHD THERMODYNAMIC EFF., %	52.7	52.7	52.7	52.7	52.7	52.7	52.7
OVERALL EFF., %	43.20	42.01	42.45	43.58	44.94	43.81	42.78
COMBUSTION EFF. %	82.6	80.6	81.2	83.3	79.4	82.6	82.6
STEAM PLANT EFF., %	39.8	39.8	39.8	39.8	39.8	39.8	39.8

$$\eta_o = \frac{P_{OUT} - P_{IN} - P_{AUX}}{HHV}$$

where P_{OUT} is the total power output from the plant, P_{IN} is total power input required by the combined cycle, P_{AUX} is the auxiliary power and HHV is the total thermal input of the coal.

c. Combustion Efficiency

The combustion efficiency is defined as the ratio of the energy which is transferred from the combustion gas to the argon and steam working fluids to the total thermal coal input.

$$\eta_c = \frac{Q_{AR} + Q_{STMC}}{HHV}$$

where Q_{AR} is the energy transferred from the combustion gas to the argon, Q_{STMC} is the heat transfer from the combustion gas to the steam bottoming cycle and HHV is the higher heating value of the coal.

d. Steam Plant Efficiency

The steam bottoming cycle plant efficiency is the ratio of the net power produced in the bottoming cycle to the total heat input to the steam cycle.

$$\eta_{STM} = \frac{P_{STMG} + P_C}{Q_{STM} + Q_{STMC}}$$

where P_{STMG} is the net power output from the steam turbine-generator, P_C is the argon compressor power, Q_{STM} is the total heat input to the bottoming cycle from the argon topping cycle and Q_{STMG} is the heat input to the steam bottoming cycle from the combustion gas.

The net busbar power for Base Case 1.0 was approximately 1000 MWe which is approximately the power output from the MHD topping cycle. In other words, this is a balanced plant with the total power produced in the bottoming cycle being used to drive the argon compressor, the various feedwater pumps and fans, and for powering all the plant auxiliaries (i.e., magnet power, hotel loads, miscellaneous pumps, etc.).

A flue gas recirculation rate of 15.7% was required to maintain a 3350 F combustor exit temperature with Montana Rosebud coal dried to 10% and burned at a stoichiometric ratio of 1.05. The stack gas temperature was 195 F.

The overall power plant efficiency (coal pile to busbar) for this plant was 43.2%, which compares favorably with an oxygen enriched open cycle plant of the same size. Parametric variations in plant operating parameters were independently varied and the results shown in Figure 3-2.

Case 1.1 - Illinois No. 6 Coal

The effect of coal type was considered in this case by substituting Illinois No. 6 coal for the Montana Rosebud which was used in the base case. Illinois No. 6 coal has an as-received moisture content of 8.9%

CASE	DESCRIPTION	PARAMETRIC CASE	REFERENCE CASE
1.0	REFERENCE PLANT	---	---
1.1	COAL TYPE	ILL. #6	MR
1.2	GAS CLEAN-UP	WET SCRUBBER	SPRAY DRYER
1.3	COMBUSTOR	1 STAGE	2 STAGE
1.4	COMBUSTOR PRESSURE	6 atm	1 atm
1.5	STEAM PLANT	1.5 in. Hg	2.5 in. Hg
1.6	STEAM PLANT	2400/1000/1000	3500/1000/1000

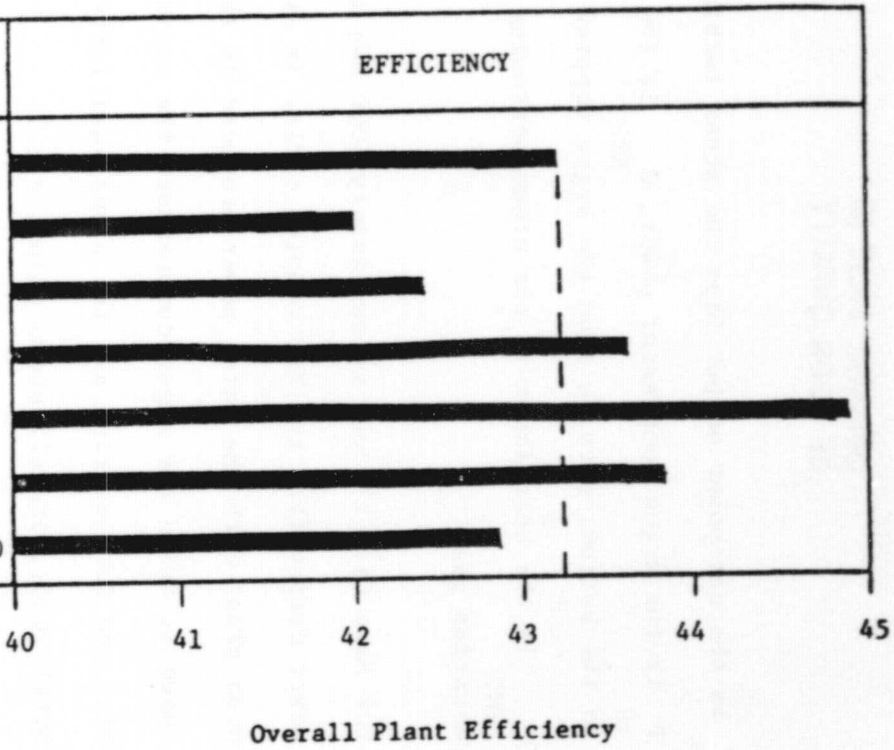

Figure 3-2

Reference Plant 1

Overall Plant Efficiency

(compared to 22.7% for Montana Rosebud) and a sulfur content of 3.3% (compared to 0.85% for Montana Rosebud). In Case 1.1, Illinois No. 6 coal was dried to 2% moisture. Combustion of Illinois No. 6 coal produces a higher flame temperature than Montana Rosebud which results in an increase in the amount of flue gas recirculation (18.7% compared to 15.7%) required to maintain the combustor exit temperature at 3350 F. Although the energy required for coal drying was less for Illinois No. 6 coal than for the base case, the energy requirement of the sulfur removal system was substantially greater for the high sulfur Illinois No. 6 coal compared with low sulfur Montana Rosebud coal. The net effect was that the overall plant efficiency was decreased by about 1.2 percentage points with Illinois No. 6 coal.

Case 1.2 - Wet Scrubbers

To show the influence of the type of cleanup system on overall plant performance, a wet scrubber was used to replace the spray dryer (or dry scrubber) considered in the base case. Both systems used Montana Rosebud coal. The overall power plant efficiency decreased by 0.75 percentage points when a wet scrubber was considered. The efficiency decrease was attributed to the higher stack gas temperature (245 F vs. 195 F) which resulted in a higher energy stack loss and an increase in the cleanup system energy requirements (29.7 MWe as opposed to 13.1 MWe in the base case).

Case 1.3 - Single Stage Combustor

A single stage combustor with 100% slag carryover results in a small increase in overall plant efficiency (0.38 percentage points) compared to the two stage slagging combustor used in the base case. Although there was a slight decrease in the heat loss from the combustor to the boiler feedwater, the primary reason for the increase in efficiency was due to energy conserved by not having a slag rejection loss. This configuration presents an interesting parametric case; however, the high slag carryover into the argon regenerative heat exchanger renders this case impractical.

Case 1.4 - Pressurized Combustor

A significant improvement in overall power plant efficiency (increase of 1.74 percentage points) was achieved with a two stage pressurized combustor. In this configuration, a gas turbine was placed downstream of the argon preheater to reduce the combustion gas pressure from about 6 atm to 1 atmosphere. The gas turbine produced an additional 248.7 MWe of power while the additional combustion air compressor required about 168 MWe of power. A pressurized combustion system is also attractive from a cost consideration. The combustor and the regenerative heat exchanger systems are significantly more compact when pressurized, which results in a less expensive plant.

Case 1.5 - Reduced Condenser Pressure

When the condenser pressure was reduced from 2.5 in.Hg. in the base case to 1.5 in.Hg., the overall power plant efficiency increased by 0.61 percentage points. This increase in efficiency was due to the increased power extracted from the stream turbines and the decrease in the energy rejected in the steam condenser.

Case 1.6 - Subcritical Bottoming Cycle

The base case utilized a supercritical 3500 psi/1000 F/1000 F steam bottoming cycle. A subcritical 2400 psi/1000 F/1000 F steam system was also considered. With a subcritical system the power required by the boiler feed pumps was less than for a supercritical system; however, the power generated in the steam turbines was also less. The energy rejected in the steam condenser was slightly greater with a subcritical system. The net effect of these differences is a decrease in the overall power plant efficiency of 0.42 percentage points with a subcritical system.

3.2 Reference Plant 2 - Pressurized Gasifiers

Reference Plant 2 is similar to Reference Plant 1 except for the method of firing the combustor. In Reference Plant 1, the combustor was directly fired with coal. In Reference Plant 2, an advanced technology pressurized gasifier with a hot gas cleanup system was used to produce clean fuel gas, which was then burned in a combustor for generation of hot gases to heat the argon in the regenerative heat exchanger. The argon topping cycle and the steam bottoming cycle are identical to those used in the base case of Reference Plant 1. The combustion system represents the only significant change.

The base case for Reference Plant 2 utilizes an air blown, 10 atmosphere pressurized, Institute of Gas Technology (IGT) fluidized bed gasifier to produce clean fuel gas. A Morgantown Iron Oxide hot gas cleanup system was used for SOx removal.

A description of the pertainent operating conditions for the base case and 14 parametric cases studies are summarized in Table 3-5. As indicated, both a Westinghouse fluidized bed gasifier with "in-bed" sulfur removal and a Texaco entrained bed gasifier with a cold gas cleanup system were included as parametric cases.

The schematic diagram containing the heat and mass balance for the base case of Reference Plant 2 is shown in Figure 3-3.

A description and the performance data for the gasifier and gas cleanup systems are included in Section 4 of the report. In this section, the overall plant performance for the base case and parametric cases is discussed.

3.2.1 <u>Reference Plant 2 Performance</u>

Reference Plant 2 is similar to Reference Plant 1 except that an IGT gasifier is used to produce clean fuel gas rather than a direct coal-fired combustor. The regenerative argon heater for Reference Plant 1 was a hot bottom design (argon temperature entering the heater was 1100 F) to facilitate removal of slag from the base of the heater. In Reference Plant 2, a cold bottom design was recommended, since slag is removed in the gasifier system and the heater should be slag free.

SOx is removed by either hot gas or cold gas cleanup which is part of the gasifier system. A cyclone and baghouse collector is used to remove particulate.

Table 3-5

REFERENCE PLANT 2 - PRESSURIZED GASIFIERS

PARAMETRIC CASES

Case	2.0	2.1	2.2	2.3	2.4	2.5	2.6	2.7	2.8	2.9	2.10	2.11	2.12	2.13	2.14
Variation	Base Case	Clean-Up	Coal	Gasifier Pressure	Plant Size	Plant Size	Oxidant	Channel EFF	Channel EFF	Channel Press	Gasifier Type	Clean-Up	Gasifier Type	Coal	Oxidant
Power Output, MWe	1000				500	250	1000								
Combustion Loop															
-- Gasifier	IGT										Westing-house		Texaco		
Oxidant	Air						Oxygen	Air							Oxygen
Gasifier Pressure, atm	10			15	10										
Coal Type	MR		Ill. #6	MR										Ill. #6	MR
Cleanup	Hot	Cold	Hot								In-bed	Cold			
Argon Loop															
Channel Efficiency, %	78							73	75	78					
Enthalpy Extraction, %	36							40	38	36					
Pressure, atm	10									12	10				
Working Fluid	Argon														
Inlet Temp, °F	3100														
Steam Plant															
Throttle Conditions	*														
Heater Arrangement	Best Possible														
Back Pressure, in. Hg.	2.5														
Output	Balanced														

* 3500/1000/1000

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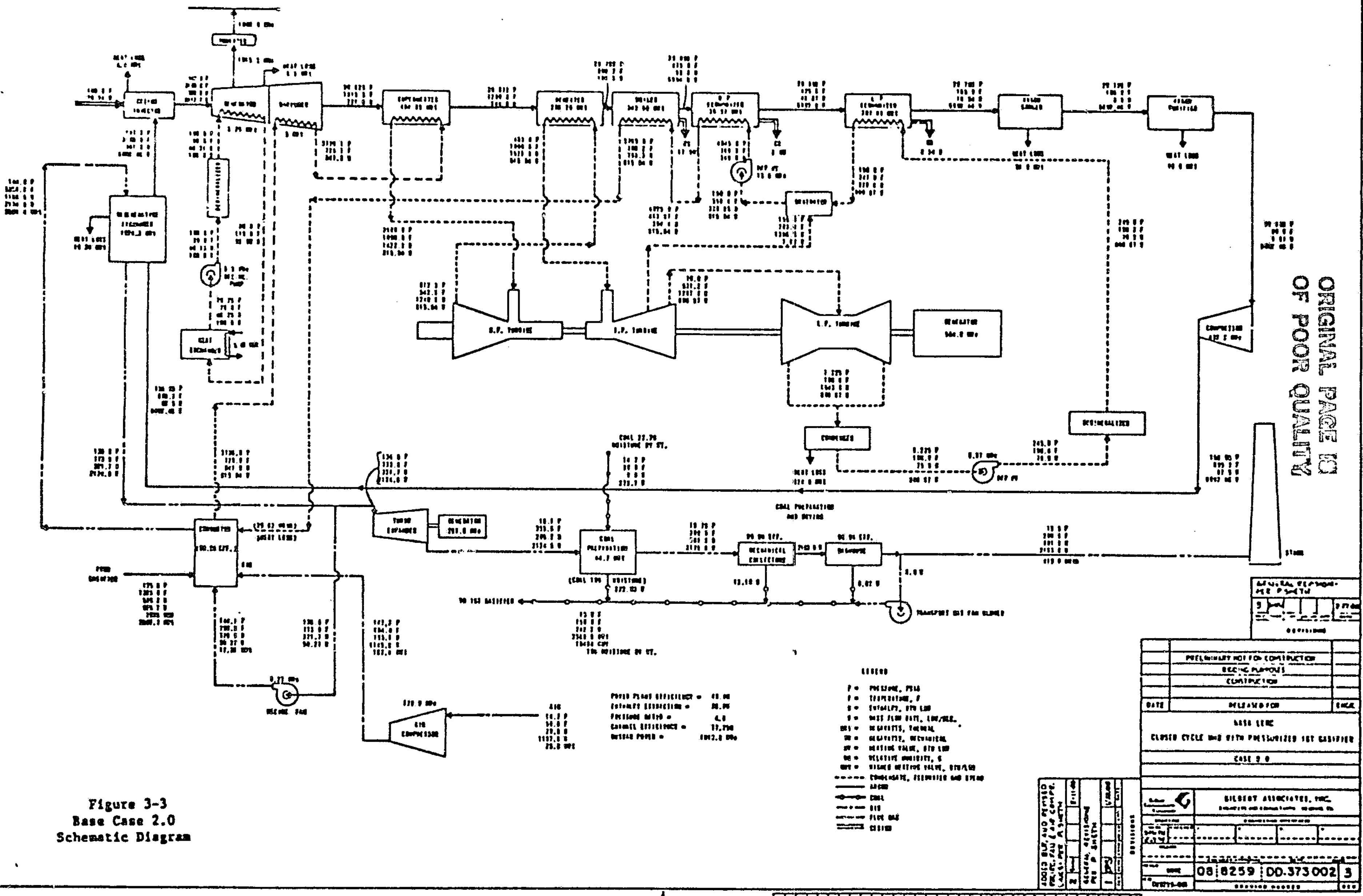

Figure 3-3
Base Case 2.0
Schematic Diagram

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The overall plant efficiency for Base Case 2.0 is 39.4% at a generated power level of 1012.6 MWe. The energy balances for Base Case 2.0 and for the 14 parametric cases studied are shown in Table 3-6 and a performance summary is given in Table 3-7. The overall plant efficiency for each parametric case is also shown in Figure 3-4. The gasifier system performance is summarized in Table 3-8.

Base Case 2.0 and parametric cases 2.1 through 2.9 were based on an IGT gasifier. Cases 2.10 and 2.11 incorporated a Westinghouse gasifier and cases 2.12 through 2.14 were based on a Texaco gasifier.

Case 2.1 - Cold Gas Cleanup System

A Stretford desulfurization system was used to replace the Morgantown iron oxide hot gas cleanup system used in the base case in order to show the effect on overall plant peformance which can be expected with alternate cleanup systems. With an IGT gasifier and a Stretford cleanup system, the overall plant efficiency was decreased from 39.4% for the base case to 36.5%. With the Morgantown iron oxide system, the temperature of the clean fuel gas entering the combustor was 1335 F; whereas, with the Stretford cold gas cleanup system, this temperature was only 105 F. This decrease in the sensible heat of the fuel requires an increase in the amount of coal input to the gasifier in order to maintain the same power output from the plant. The thermal input of coal to the gasifier for the base case (hot gas cleanup system) was 2550.8 MWt (10,038 ton/day) compared to 2710.3 MWt (10,666 ton/day) for the cold gas cleanup system. This increase in coal feed required an increase in the number of gasifier units (6 units were required for the base case and 7 units for the cold gas cleanup system).

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TABLE 3-6
REFERENCE PLANT 2
ENERGY BALANCE

CASE	2.0	2.1	2.2	2.3	2.4	2.5	2.6	2.7	2.8	2.9	2.10	2.11	2.12	2.13	2.14
VARIATION	REFERENCE PLANT	CLEAN-UP	ILL. #6 COAL	15 atm GAS.PRESS	500 MWe PLANT	250 MWe PLANT	O_2 BLOWN	EXTRACTION RATIO 40%	EXTRACTION RATIO 38%	CHANNEL PRESS 12 atm	WESTINGHOUSE GASIFIER	CLEAN-UP	TEXACO GASIFIER	ILL. #6 COAL	O_2 BLOWN
ENERGY INPUT															
ARGON COMPRESSOR, MW	492.2	492.2	492.2	492.2	246.0	123.0	492.2	442.9	466.2	492.2	492.2	492.2	492.2	492.2	492.2
AIR COMPRESSOR/BLOWER, MW	179.0	192.3	192.0	237.0	89.5	44.7	213.0	161.0	169.5	178.7	185.7	181.0	172.3	162.8	172.8
RECIRCULATION FAN, MW	0.3	0.4	0.2	0.2	0.1	0.1	1.3	0.2	0.2	0.3	0.2	0.5	0.1	0.1	1.9
MHD COOLING LOOP PUMP, MW	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3
BOILER FEED PUMP, MW	16.5	16.5	16.5	16.5	8.2	4.1	16.6	13.6	15.0	16.5	16.6	16.5	16.6	16.5	16.5
TOTAL POWER INPUT, MW	688.3	701.7	700.9	746.2	344.1	172.2	723.4	618.0	651.2	688.0	695.0	695.5	681.5	671.9	683.7
FUEL GAS HHV, MW	1980.4	2129.7	2072.1	2060.2	990.1	495.0	2370.6	1782.3	1876.1	1980.4	2088.4	2041.3	1988.9	1949.4	2012.9
FUEL GAS (SENSIBLE), MW	527.3	284.5	450.7	477.2	263.7	131.9	91.9	474.6	499.5	527.3	630.5	375.9	531.9	483.1	156.7
COMBUSTION AIR, MW	25.9	27.9	27.9	27.2	13.0	6.5	30.8	23.3	24.6	25.9	26.8	26.2	24.8	24.0	25.1
TOTAL HEAT INPUT, MW	2533.6	2442.1	2550.7	2564.6	1266.8	633.4	2493.3	2280.2	2400.2	2533.6	2745.7	[illegible]	2871.9*	2456.5	2194.7
TOTAL ENERGY INPUT, MW	3221.9	3143.8	3251.6	3310.8	1610.9	805.6	3216.7	2898.2	3051.4	3221.6	3440.7	3133.9	3513.3	3128.4	2878.4
ENERGY OUTPUT															
MHD INVERTER, MW	1000.0	1000.0	1000.0	1000.0	500.0	250.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0
STEAM TURBINE, MW	554.9	556.1	555.6	555.5	277.4	138.7	558.1	458.3	504.0	554.9	555.8	555.3	651.4	554.6	555.2
GAS TURBINE	261.0	264.6	284.2	326.6	130.5	65.2	237.3	234.9	247.3	261.0	335.8	275.0	320.5	286.2	149.8
TOTAL POWER OUTPUT, MW	1815.9	1820.7	1839.8	1882.1	907.9	453.9	1795.4	1693.2	1751.3	1815.9	1891.6	1830.3	1971.9	1840.8	1705.0
REGENERATIVE HEAT EXCAHNGER, MW	19.3	19.3	19.3	19.3	9.7	4.8	19.3	17.4	18.3	19.3	19.3	19.3	19.3	19.3	19.3
ARGON COOLER, MW	51.9	51.9	51.9	51.9	26.0	13.0	51.9	46.7	49.1	51.9	51.9	51.9	51.9	51.9	51.9
ARGON PURIFIER, MW	16.0	16.0	16.0	16.0	8.0	4.0	16.0	14.4	15.1	16.0	16.0	16.0	16.0	16.0	16.0
CONDENSER, MW	824.9	826.7	825.9	824.9	412.4	206.2	829.6	681.3	749.3	824.9	826.2	825.6	1018.8	824.5	825.5
COAL DRYER, MW	44.7	47.4	11.3	45.2	22.3	11.1	47.4	40.2	42.3	44.7	45.6	55.2	0.0	0.0	0.0
STACK LOSS, MW	415.0	326.2	452.0	434.2	207.5	102.1	420.8	373.4	393.1	415.0	551.8	300.5	405.2	340.8	221.8
INVERTER	15.2	15.2	15.2	15.2	7.6	3.8	15.2	15.2	15.2	15.2	15.2	15.2	15.2	15.2	15.2
CESIUM SYSTEM, MW	4.0	4.0	4.0	4.0	2.0	1.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0
COMBUSTION SYSTEM HEAT LOSS, MW	0.0	0.6	0.0	0.6	0.3	0.2	1.4	0.6	0.8	0.0	0.0	0.8	0.0	0.5	1.6
MHD GENERATOR LOSS, MW	5.5	5.5	5.5	5.5	2.75	1.4	5.5	5.5	5.5	5.5	5.5	5.5	5.5	5.5	5.5
MISCELLANEOUS, MW	9.5	10.3	10.7	11.9	4.45	4.1	10.2	10.3	7.4	9.2	13.6	9.6	5.5	9.9	12.6
TOTAL HEAT OUTPUT, MW	1406.0	1323.1	1411.8	1428.7	703.0	351.7	1421.3	1205.0	1300.1	1405.7	1549.1	1303.6	1541.4	1287.6	1173.4
TOTAL ENERGY OUTPUT, MW	3221.9	3143.8	3251.6	3310.8	1610.9	805.6	3216.7	2898.2	3051.4	3221.6	3440.7	3133.9	3513.3	3128.4	2878.4

* Includes 286.3 MW Waste Heat From Fuel Gas.

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TABLE 3-7
REFERENCE PLANT 2

PERFORMANCE SUMMARY

CASE	2.0	2.1	2.2	2.3	2.4	2.5	2.6	2.7	2.8	2.9	2.10	2.11	2.12	2.13	2.14
VARIATION	REFERENCE PLANT	CLEAN-UP	ILL. #6 COAL	15 atm GAS.PRESS	500 MWe PLANT	250 MWe PLANT	O_2 BLOWN	EXTRACTION RATIO 40%	EXTRACTION RATIO 38%	CHANNEL PRESS 12 atm	WESTINGHOUSE GASIFIER	CLEAN-UP	TEXACO GASIFIER	ILL. #6	O_2 BLOWN
PLANT SIZE, MW	1800.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0
COAL TYPE	MR	MR	ILL. #6	MR	MR	MR	MR	MR	MR	MR	MR	MR	MR	ILL. #6	MR
COAL MOISTURE, %	10	10	5	10	10	10	10	10	10	10	10	10	22.7	8.9	22.7
RECIRCULATION, %	2.3	3.7	1.9	1.6	2.3	2.3	11.5	2.3	2.3	2.3	0.88	4.2	0.03	0.7	23.0
STACK GAS TEMPERATURE, F	200	200	326.3	200	200	200	200	200	200	200	291	200	422.6	362	200
STOICHEOMETRIC RATIO	1.05	1.05	1.05	1.05	1.05	1.05	1.05	1.05	1.05	1.05	1.05	1.05	1.05	1.05	1.05
CLEAN-UP SYSTEM	HOT	COLD	HOT	HOT	HOT	HOT	HOT	HOT	HOT	HOT	HOT	COLD	COLD	COLD	COLD
FUEL INPUT TO COMBUSTOR, MW	2507.7	2433.6	2550.1	2544.8	1263.3	630.6	2397.7	2235.1	2394.5	2581.7	2718.9	2439.9	2520.8	2448.2	2192.9
COAL INPUT TO GASIFIER, MW	2569.9	2731.1	2851.7	2586.0	1284.7	641.3	2629.2	2313.5	2435.0	2625.3	2634.6	3179.0	3016.3	2904.8	2537.6
MHD SYSTEM:															
TOTAL POWER OUTPUT, MW	1815.9	1820.7	1839.8	1882.1	907.9	453.9	1795.4	1693.2	1751.3	1815.9	1891.6	1830.3	1971.9	1840.8	1705.0
TOTAL POWER INPUT, MW	688.3	701.7	700.9	746.2	344.1	172.2	723.4	618.0	651.2	688.0	695.0	690.5	681.5	671.9	684.1
AUXILIARY POWER, MW	25.7	27.3	28.4	26.0	13.0	6.6	27.3	23.1	24.4	25.7	26.3	31.7	30.1	30.9	27.6
GASIFIER SYSTEM:															
AIR COMPRESSOR, MW	89.2	94.7	98.0	113.8	44.5	22.2	16.0	80.3	84.5	67.7	113.8	112.8	163.0	153.1	25.2
OXYGEN PRODUCTION, MW	0.0	0.0	0.0	0.0	0.0	0.0	42.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	67.1
AUX. FED PUMP, MW	0.13	0.13	0.13	0.13	0.13	0.13	0.13	0.13	0.13	0.13	0.16	0.16	0.16	0.16	0.16
BUSBAR POWER, MW	1012.57	996.87	1012.37	995.9	506.2	252.7	985.9	971.6	991.0	1034.3	1056.34	995.1	1097.17	984.7	900.87
MHD THERMODYNAMIC EFF., %	52.7	52.7	52.7	52.7	52.7	52.7	52.7	58.6	55.6	52.6	52.7	52.6	52.7	52.7	52.7
OVERALL PLANT EFF., %	39.4	36.5	35.5	38.5	39.4	39.4	37.5	42.0	40.7	39.4	40.1	31.3	36.4	33.9	35.5
COMBUSTION EFF., %	74.9	70.5	67.7	74.1	74.9	74.9	70.4	74.9	74.9	74.9	73.0	60.6	63.8	66.2	71.0
STEAM PLANT EFF., %	39.8	39.8	39.8	39.8	39.8	39.8	39.8	39.8	39.8	39.8	39.8	39.8	39.8	39.8	39.8

CASE	DESCRIPTION	PARAMETRIC CASE	REFERENCE CASE
2.0	REFERENCE PLANT	---	---
2.1	GAS CLEAN-UP	COLD	HOT
2.2	COAL TYPE	Ill. #6	MR
2.3	GASIFIER PRESSURE	15 atm	10 atm
2.4	PLANT SIZE	500 MWe	1000 MWe
2.5	PLANT SIZE	250 MWe	1000 MWe
2.6	OXIDANT	OXYGEN	AIR
2.7	ENTHALPY EXTRACTION RATIO	40%	36%
2.8	ENTHALPY EXTRACTION RATIO	38%	36%
2.9	CHANNEL PRESSURE	12 atm	10 atm
2.10	GASIFIER	Westinghouse	IGT
2.11	GAS CLEAN-UP	COLD	HOT
2.12	GASIFIER	TEXACO	IGT
2.13	COAL TYPE	Ill. #6	MR
2.14	OXIDANT	OXYGEN	AIR

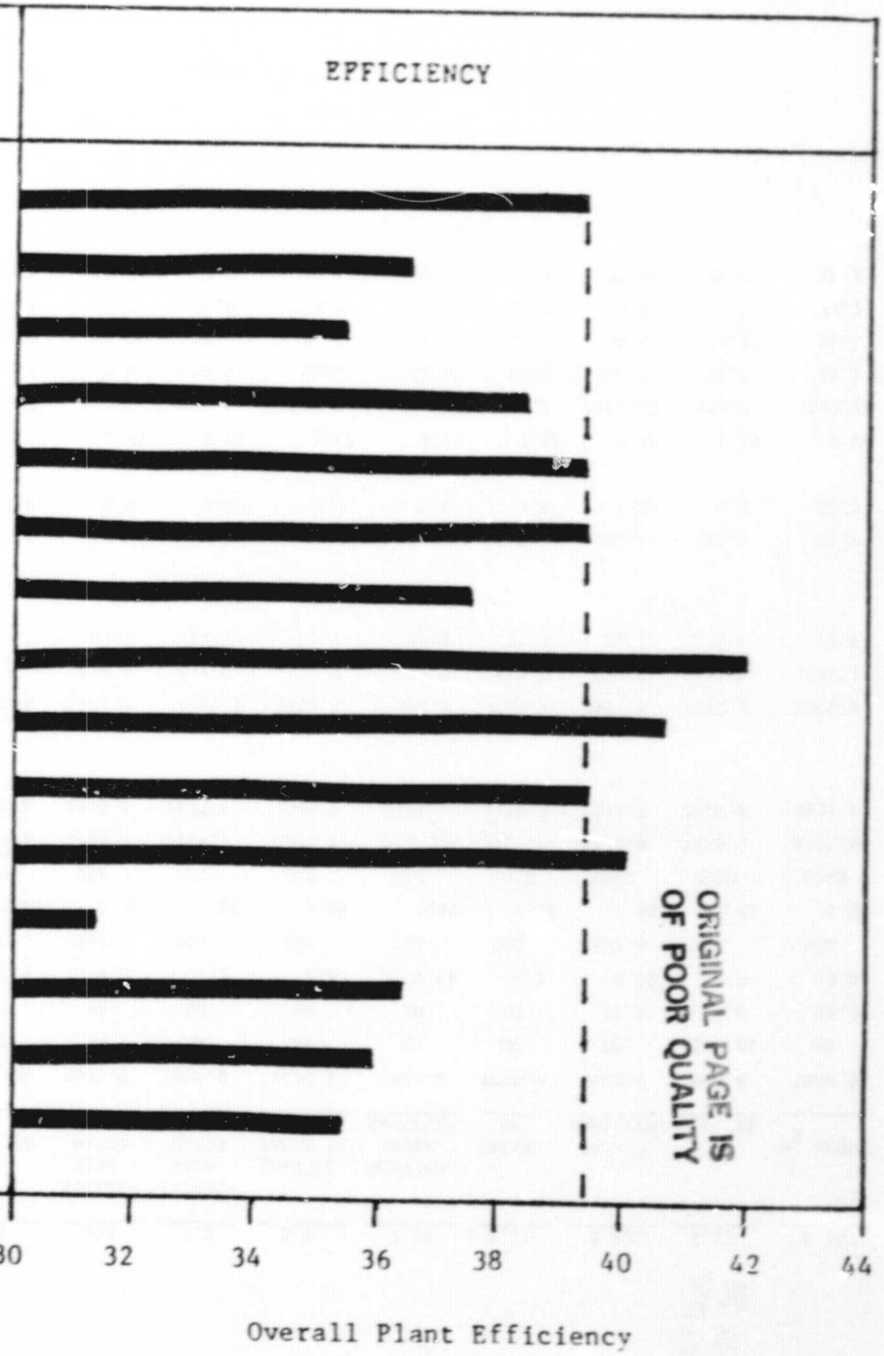

Figure 3-4

Reference Plant 2
Overall Plant Efficiency

Table 3-8

REFERENCE PLANT 2

GASIFIER SYSTEM SUMMARY

CASE	2.0	2.1	2.2	2.3	2.4	2.5	2.6	2.7	2.8	2.9	2.10	2.11	2.12	2.13	2.14
VARIATION	REFERENCE PLANT	CLEAN-UP	ILL. #6 COAL	15 atm GAS.PRESS	500 MW_e PLANT	250 MW_e PLANT	O_2 BLOWN	EXTRACTION RATIO 40%	EXTRACTION RATIO 38%	CHANNEL PRESS 12 atm	WESTINGHOUSE GASIFIER	CLEAN-UP	TEXACO GASIFIER	ILL. # 6 COAL	O_2 BLOWN
PLANT SIZE, MW_e	1000	1000	1000	1000	500	250	1000	1000	1000	1000	1000	1000	1000	1000	1000
GASIFIER	IGT	IGT	IGT	IGT	IGT	IGT	IGT	IGT	IGT	IGT	W	W	TEXACO	TEXACO	TEXACO
GASIFIER TYPE	FLUIDIZED	FLUIDIZED	FLUIDIZED	FLUIDIZED	FLUIDIZED	FLUIDIZED	FLUIDIZED	FLUIDIZED	FLUIDIZED	FLUIDIZED	FLUIDIZED	FLUIDIZED	ENTRAINED	ENTRAINED	ENTRAINED
OXIDANT	AIR	AIR	AIR	AIR	AIR	AIR	OXYGEN	AIR	AIR	AIR	AIR	AIR	AIR	AIR	OXYGEN
GASIFIER PRESSURE, atm	10	10	10	15	10	10	10	10	10	10	10	10	10	10	10
COAL TYPE	MR	MR	ILL. # 6	MR	MR	MR	MR	MR	MR	MR	MR	MR	MR	ILL. #6	MR
COAL MOISTURE, %	10	10	5	10	10	10	10	10	10	10	10	10	22.7	8.9	22.7
CLEAN-UP SYSTEM	FeO	FeO	FeO	FeO	FeO	FeO	FeO	FeO	FeO	FeO	In-Bed	STRETFORD	STRETFORD	STRETFORD	STRETFORD
COAL SIZE	0-1/4in	0-1/4in	0-1/4in	0-1/4in	0-1/4in	0-1/4in	0-1/4in	0-1/4in	0-1/4in	0-1/4in	0-1/4in	0-1/4in	70-200mesh	70-200mesh	70-200mesh
lb AIR OR OXYGEN/lb MAF COAL	3.16	3.16	3.41	3.16	3.16	3.16	0.60	3.16	3.16	3.16	3.16	3.16	5.20	5.20	0.95
lb STEAM/lb MAF COAL	0.66	0.66	0.62	0.66	0.66	0.66	0.61	0.66	0.66	0.66	0.66	0.66	0.50	0.50	0.60
PRESSURE OF RAW GAS, psia	147	147	147	221	147	147	147	147	147	147	-	147	147	147	147
TEMPERATURE OF RAW GAS, F	1765	1765	1500	1560	1765	1765	1855	1765	1765	1765	-	1600	1800	200	2100
PRESSURE OF CLEAN GAS, psia	130	130	130	200	130	130	130	130	130	130	147	130	130	130	130
TEMPERATURE OF CLEAN GAS, F	1335	105	1200	1200	1335	1335	1160	1335	1335	1335	1600	105	105	105	105
HIGHER HEATING VALUE, BTU/SCF	151	151	154	156	151	151	302	151	151	151	131	131	98	103	262.6
COAL INPUT TO GASIFIER, MW_t	2550.8	2710.3	2822.1	2579.4	1275.4	637.7	2712.3	2295.8	2416.6	2550.8	3226.9	3150.6	2961.3	2887.1	2690.2
COAL FLOW, ton/day	10,038	10,666	9,836	10,150	5,019	2,510	10,673	9,034	9,510	10,038	12,699	12,398	13,592	10,493	12,3[illegible]8
NUMBER OF UNITS	6	7	6	7	3	2	7	6	6	6	11	11	10	8	7
COMBUSTION EFFICIENCY, %	75.6	71.2	68.4	74.8	75.6	75.6	71.1	75.6	75.6	75.6	59.8	61.2	65.1	66.8	71.7
POWER PLANT EFFICIENCY, %	41.4	38.5	37.5	40.5	41.4	41.4	39.5	44.0	42.7	41.4	34.3	33.3	35.4	35.9	37.5

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Case 2.2 - Illinois No. 6 Coal

Illinois No. 6 coal having an as-received moisture content of 8.9% was dried to 5% moisture and fired in an IGT gasifier having a Morgantown iron oxide hot gas cleanup system. The overall plant efficiency decreased by 3.9 percentage points when Illinois No. 6 coal was used instead of Montana Rosebud. A major contributor to the lower efficiency was the increase in the stack gas loss with Illinois No. 6 coal. The coal drying energy requirement was reduced considerably for Illinois No. 6 coal (11.3 MWt for Illinois No. 6 coal compared to 44.7 MWt for Montana Rosebud). Because less energy is extracted from the flue gas in the dryer, the stack gas temperature and, therefore, the stack gas energy loss increased.

Case 2.3 - Gasifier Pressure of 15 atm

With a gasifier pressure of 15 atm, the size of the gasifiers was decreased and the number of gasifier units was reduced from 6 (at 10 atm) to 5. However, the temperature of the clean fuel gas was reduced from 1765 F (at 10 atm) to 1560 F. This decreases the sensible heat of the fuel entering the combustor. The net effect of reduced fuel gas temperature and other minor differences resulted in a decrease in overall plant performance. The net power plant efficiency decreased from 39.4% for the base case to 38.5% if the gasifier were operated at 15 atm.

Case 2.4 - 500 MWe Plant Size

This parametric case was not representative of what would be expected in a smaller plant because the channel efficiency and enthalpy extraction ratio were not changed from the 1000 MWe base case. Treating the channel as an energy converison device with the same specified performance (specified by NASA-LeRC) regardless of plant size resulted in a linear scale down in power output from 1000 MWe to 500 MWe at the same overall plant efficiency of 39.4%.

Case 2.5 - 250 MWe Plant Size

This case did not show the parametric variation in plant size for the same reasons stated in Case 2.4. In smaller size plants, the channel efficiency and enthalpy extraction ratio are expected to be reduced. In this case, these operating parameters were the same for the 250 MWe and the 1000 MWe plant. The overall plant efficiency was therefore the same regardless of plant size.

Case 2.6 - IGT Oxygen Blown Gasifier

The IGT gasifier was considered to be oxygen blown in this parametric case in order to produce a medium Btu fuel gas having a higher heating value of 302 Btu/SCF compared to the low Btu gas used in the air blown base case of 151 Btu/SCF. A Lotepro air separation plant was assumed which required 212 kW/hr per ton of equivalent pure oxygen. The temperature of the raw gas leaving the gasifier increased from 1765 F to 1855 F when oxygen was used as the oxident. The temperature of the combustion gases leaving the combustor and entering the regenerative heater was still limited to 3350 F because of the temperature limitation of the refractory brick material. Because of the absence of nitrogen in the combustion gas and the higher heating value of the clean fuel gas, the flow rate of the combustion gases was less with an oxygen blown gasifier than with an air blown system. This reduced flow rate resulted in a decrease in the power produced in the expansion gas turbine from 261.0 MWe for the base case to 237.3 MWe for the oxygen blown case. The net busbar power output of the plant decreased from 1012.7 MWe to 985.9 MWe. The overall plant efficiency decreased from 39.4% for the air blown base case to 37.5% for the oxygen blown case. The advantages of an oxygen blown system are the reduction in the total number of gasifier units (from 6 to 5), smaller size gasifier units, and a reduction in the size of piping and components in the combustion system.

Case 2.7 - Enthalpy Extraction Ratio of 40% and Channel Efficiency of 73%

The enthalpy extraction ratio was arbitrarily increased from 36% to 40% and the channel efficiency was decreased from 78% to 73% for this case. This case shows that the overall power plant efficiency, for these assumed channel conditions, could be increased from 39.4% for the base case to 42%. No attempt was made in this part of the study to assess the realism of these channel performance numbers (channel treated as an energy conversion device); but rather, the performance numbers were used as given values.

Case 2.8 - Enthalpy Extraction Ratio of 38% and Channel Efficiency of 75%

This case was similar to Case 2.7 except that the enthalpy extraction ratio was 38% (compared to 36% for the base case) and the channel efficiency was 75% (compared to 78% for the base case). Provided these specified conditions could be achieved in the channel, the overall power plant efficiency would be 40.7% compared to 39.4% for the base case.

Case 2.9 - Channel Pressure of 12 atm

This parametric case did not effectively show the influence of argon pressure entering the MHD channel because the enthalpy extraction ratio and channel efficiency were also specified. In reality, the argon pressure would influence the electrical conductivity of the plasma which would effect the enthalpy extraction ratio and channel efficiency. With the channel performance specified, the conductivity does not enter into the analysis. Therefore, the overall plant efficiency was the same for this case and the base case.

Case 2.10 - Westinghouse Gasifier

In Case 2.10, a Westinghouse fluidized bed gasifier having in-bed desulfurization was substituted for the IGT gasifier used in the base case. A schematic diagram showing pertinent state points is given as Figure 3-5. The higher heating value of the clean fuel gas was 131 Btu/SCF (compared to 151 Btu/SCF for the base case) using Montana Rosebud coal dried to 10% moisture. This lower heating value required a larger coal input than the reference case and the addition of a combustion air preheater to raise the temperature of the compressor discharge air from 604 F to 1000 F at the inlet to the combustor. The combustion gas flow rate was larger for the Westinghouse gasifier than for the IGT gasifier. This increased flow rate resulted in an increase in the power extracted from the expansion turbine (237.3 MWe for Case 2.10 compared to 261.0 MWe for the base case). The busbar power output from the plant with a Westinghouse gasifier was 1056.3 MWe and 1012.6 MWe with an IGT gasifier system. The overall net power plant efficiency was 40.1% with a Westinghouse gasifier compared to 39.4% for the IGT system used for the reference case.

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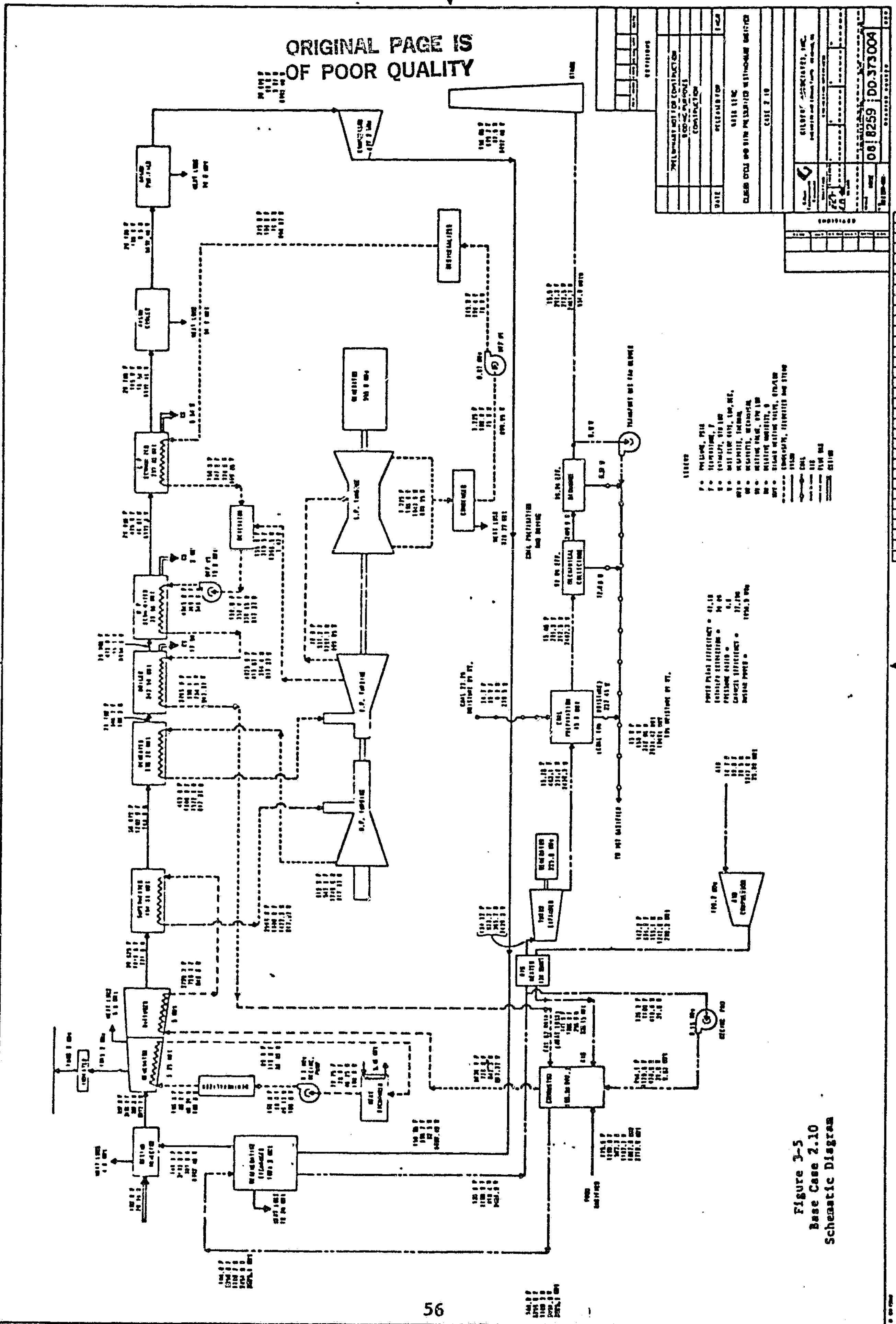

Figure 3-5
Base Case 2.10
Schematic Diagram

Case 2.11 - Cleanup System

If a Stretford cold gas cleanup system could be assumed to be used with the Westinghouse gasifier for sulfur control rather than the dolomite sorbent, the overall plant efficiency would decrease to 31.3% compared to 40.1% using the dolomite inbed desulfurization scheme. Principally, this is caused by the loss of sensible heat of the clean fuel gas entering the combustor.

Case 2.12 - Texaco Gasifier

A Texaco entrained bed gasifier was considered as a parametric variation in place of the IGT gasifier used in the base case. A Stretford cold gas cleanup system was used for sulfur control. Pulverized coal (70% through 200 mesh) was required with this entrained bed gasifier. The IGT and Westinghouse gasifiers are of the fluidized bed type which require crushed coal (0-1/4 inch). The higher heating value of the clean fuel gas was 98 Btu/SCF (151 Btu/SCF for the base case). Figure 3-6 shows the schematic diagram and pertinent state points for this case. Coal drying is not required. The coal can be fired directly as it leaves the pulverizer. The busbar power for this plant was 1097.2 MWe with an overall plant efficiency of 36.4% (compared to 1012.6 MWe and 39.4% for the base case).

Case 2.13 - Illinois No. 6 Coal

For this case, the type of coal was changed from Montana Rosebud to Illinois No. 6. Pulverized Illinois No. 6 coal was fired as-received with 8.9% moisture. The overall plant efficiency was 33.9% and the busbar power output was 984.7 MWe.

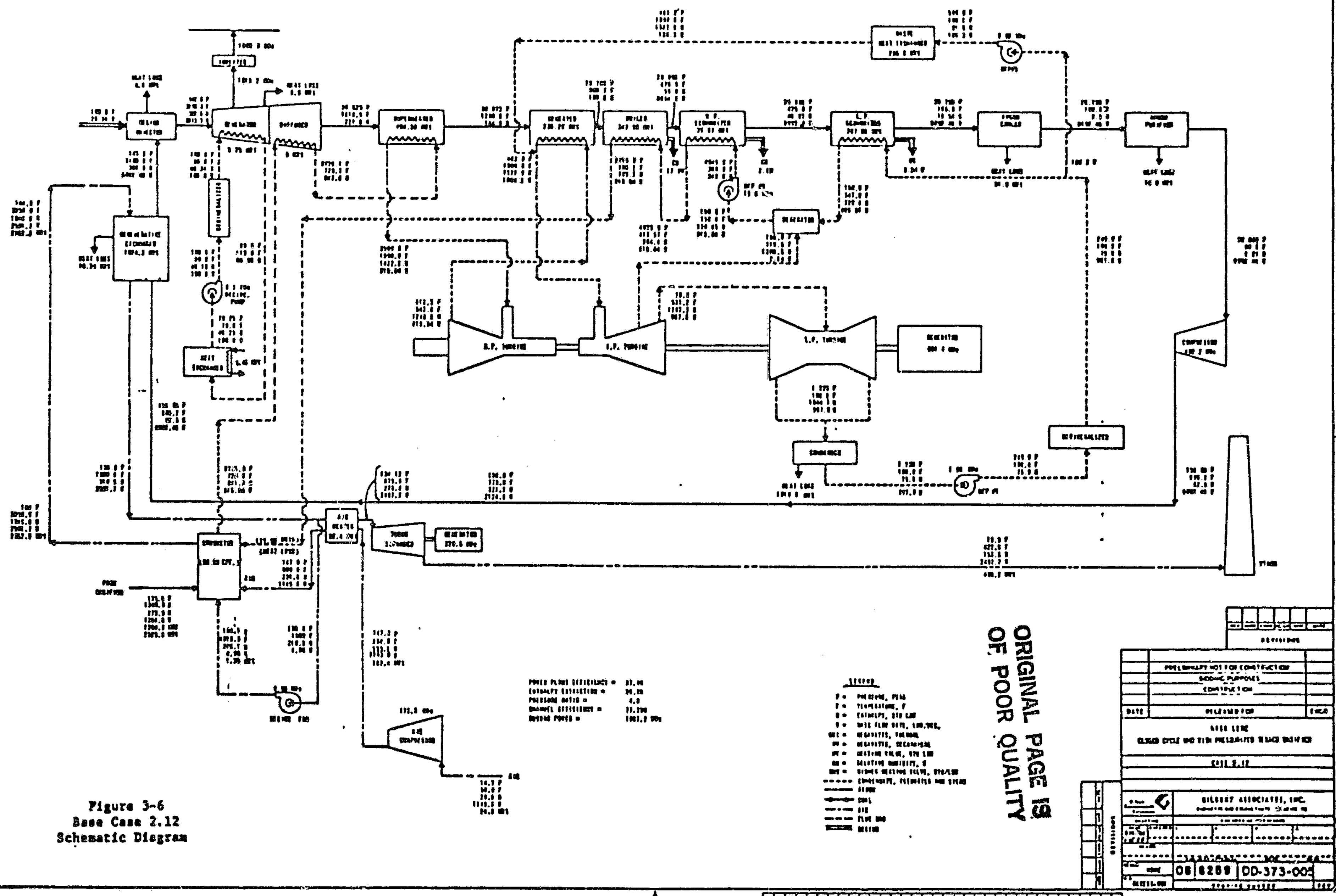

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Figure 3-6
Base Case 2.12
Schematic Diagram

Case 2.14 - Oxygen Blown Texaco Gasifier

The Texaco gasifier was assumed to be oxygen blown in this parametric case. The higher heating value of the clean fuel gas using oxygen was 262.6 Btu/SCF compared to 98 Btu/SCF for the air blown Texaco gasifier. The power output from the expansion turbine was 149.8 MWe for the oxygen blown case compared to 320.5 MWe for the air blown Texaco gasifier system. This was due to the reduced mass flow of combustion gases. The busbar power was 900 MWe for Case 2.14 compared to 1097 MWe for the air blown Texaco gasifier case. The overall plant efficiency was 35.5% for the oxygen blown system.

3.3 Reference Plant 3 - Atmospheric Gasifiers

Reference Plant 3 is similar to Reference Plant 2 except that a state-of-the-art atmospheric gasifier with a cold gas cleanup system was used to produce clean fuel gas. The argon topping cycle and the steam bottoming cycles are essentially identical to those used in Reference Plants 1 and 2. The combustion system represents the only significant change.

The base case for Reference Plant 3 utilizes an air blown, atmospheric Combustion Engineering (CE) entrained bed gasifier. A Stretford cold gas cleanup system was used for sulfur removal.

The pertinent operating conditions for the base case and 7 parametric cases are summarized in Table 3-9. As indicated, a Winkler fluidized bed gasifier and a Wellman fixed bed gasifier were considered as parametric cases.

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Table 3-9

REFERENCE PLANT 3 - ATMOSPHERIC GASIFIER

PARAMETRIC CASES

Case	3.0	3.1	3.2	3.3	3.4	3.5	3.6	3.7				
Variation	Base Case	Clean-Up	Coal	Plant Size	Gasifier Type	Coal	Clean-Up	Gasifier Type				
Power Output, MWe	1000			500	1000			100				
Combustion Loop												
Gasifier	CE				Winkler			Wellman				
Oxidant	Air											
Gasifier Pressure, atm	1											
Coal Type	MR		Ill. #6	MR		Ill. #6	MR					
Cleanup	Cold	Hot	Cold				Hot	Cold				
Argon Loop												
Channel Efficiency, %	78											
Enthalpy Extraction, %	36											
Pressure, atm	10											
Working Fluid	Argon											
Inlet Temp, °F	3100											
Steam Plant												
Throttle Conditions	*											
Heater Arrangement	**											
Back Pressure, in. Hg.	2.5											
Output	Balanced											

* 3500/1000/1000

** DA only or best possible arrangement

A schematic diagram containing the heat and mass balance for the base case of Reference Plant 3 is shown in Figure 3-7.

A description and performance data for the gasifiers and gas cleanup systems are included in Section 4. In this section, overall plant performance for the base case and parametric cases is discussed.

3.3.1 Reference Plant 3 Performance

The regenerative argon heat exchanger for this case utilizes a cold bottom where the argon leaving the compressor enters directly into the heat exchanger.

A Stretford cold gas cleanup system was recommended for sulfur control. Development of this system is further along than any of the hot gas cleanup systems currently being considered. A cyclone and baghouse collector was used for particulate control.

Montana Rosebud coal pulverized to 70% through 200 mesh and dried to 5% was used for Base Case 3.0. The overall plant efficiency was 36.1% at a net generated power level of 994.5 MWe. The energy balance for Base Case 3.0 and the 7 parametric cases studied are shown in Table 3-10 and a performance summary is given in Table 3-11. The overall plant efficiency for each parametric case is also shown in Figure 3-8. The gasifier system performance is summarized in Table 3-12.

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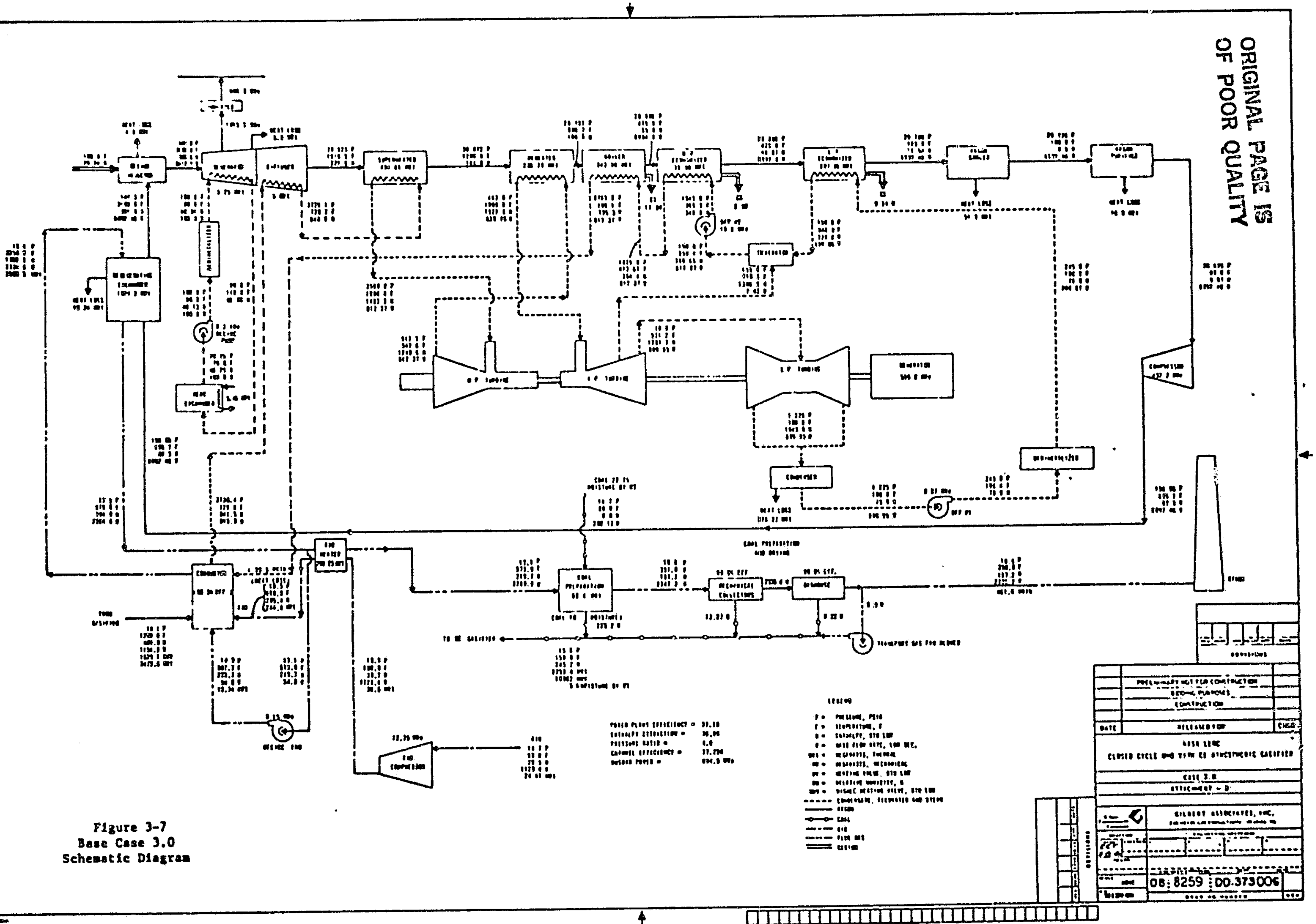

Figure 3-7
Base Case 3.0
Schematic Diagram

TABLE 3-10

REFERENCE PLANT 3

PERFORMANCE SUMMARY

CASE	3.0	3.1	3.2	3.3	3.4	3.5	3.6	3.7
VARIATION	REFERENCE PLANT	CLEAN-UP	ILL. Ø6 COAL	500 MWe PLANT	WINKLER GASIFIER	ILL. Ø6 COAL	CLEAN-UP	WELLMAN GASIFIER
PLANT SIZE, MW	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0
COAL TYPE	MR	MR	ILL. Ø6	MR	MR	ILL. Ø6	MR	MR
COAL MOISTURE, %	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0
RECIRCULATION, %	2.3	2.3	2.7	2.3	2.4	3.7	2.4	20.7
STACK GAS TEMPERATURE, F	350.0	350.0	530.0	350.0	350.0	513.0	350.0	395.0
STOICHEOMETRIC RATIO	1.05	1.05	1.05	1.05	1.05	1.05	1.05	1.05
CLEAN-UP SYSTEM	COLD	HOT	COLD	COLD	COLD	COLD	COLD	COLD
FUEL INPUT TO COMBUSTOR, MW	2472.6	2472.6	2252.4	1117.1	2244.0	2240.0	2240.4	225.9
COAL INPUT TO GASIFIER, MW	2753.4	2753.4	2621.0	1377.8	2790.0	2789.0	2789.0	230.0
MHD SYSTEM:								
TOTAL POWER OUTPUT, MW	1555.8	1555.8	1555.8	777.9	1555.8	1555.8	1555.8	155.6
TOTAL POWER INPUT, MW	522.7	522.7	522.7	261.3	522.8	523.0	522.8	52.2
AUXILIARY POWER, MW	27.5	27.5	26.2	13.7	27.9	27.8	27.9	2.3
GASIFIER SYSTEM:								
AIR BLOWER, MW	11.1	11.1	10.9	5.5	9.0	9.2	9.0	0.9
BUSBAR POWER, MW	994.5	994.5	996.0	497.4	996.1	995.8	995.8	100.2
MHD THERMODYNAMIC EFF., %	52.7	52.7	52.7	52.7	52.7	52.7	52.7	52.7
OVERALL PLANT EFF., %	36.1	36.1	38.0	36.1	35.7	35.7	35.7	43.6
COMBUSTION EFF., %	69.8	69.8	73.9	69.8	69.0	69.0	69.0	84.3
STEAM PLANT EFF., %	39.8	39.8	39.8	39.8	39.8	39.8	39.8	39.8

TABLE 3-11
REFERENCE PLANT 3

ENERGY BALANCE

CASE	3.0	3.1	3.2	3.3	3.4	3.5	3.6	3.7
VARIATION	REFERENCE PLANT	CLEAN-UP	ILL. #6 COAL	500 MWe PLANT	WINKLER GASIFIER	ILL. #6 COAL	CLEAN-UP	WELLMAN GASIFIER
ENERGY INPUT								
ARGON COMPRESSOR, MW	492.2	492.2	492.2	246.1	492.2	492.2	492.2	49.2
AIR COMPRESSOR/BLOWER, MW	12.7	12.7	12.7	6.3	13.1	13.1	13.1	1.3
RECIRCULATION FAN, MW	0.2	0.2	0.2	0.1	0.2	0.4	0.2	-
BOILER FEED PUMP, MW	17.3	17.3	17.3	8.6	17.3	17.3	17.3	1.7
MHD COOLING LOOP PUMP, MW	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3
TOTAL POWER INPUT, MW	522.7	522.7	522.7	261.3	522.8	523.0	522.8	52.2
FUEL GAS HHV, MW	1978.3	1978.3	1978.3	989.1	2006.7	2013.6	2006.7	203.1
FUEL GAS (SENSIBLE), MW	494.3	494.3	496.3	247.0	478.4	466.6	478.4	40.1
COMBUSTION AIR, MW	24.4	24.4	24.4	12.2	25.0	25.1	24.4	2.0
TOTAL HEAT INPUT, MW	2497.0	2497.0	2499.0	1248.3	2510.1	2505.3	2509.5	245.2
TOTAL ENERGY INPUT, MW	3019.7	3019.7	3021.7	1509.6	3032.9	3028.3	3032.3	297.4
ENERGY OUTPUT								
MHD INVERTER, MW	1000.0	1000.0	1000.0	500.0	1000.0	1000.0	1000.0	100.0
STEAM TURBINE, MW	555.8	555.8	555.8	277.9	555.8	555.8	555.8	55.6
GAS TURBINE, MW	0.0	0.0	0.0	0.0	0.0	0.0	0.0	[illegible]
TOTAL POWER OUTPUT, MW	1555.8	1555.8	1555.8	777.9	1555.8	1555.8	1555.8	155.6
REGENERATIVE HEAT EXCHANGER, MW	19.3	19.3	19.3	9.7	19.3	19.3	19.3	1.9
ARGON COOLER, MW	51.9	51.9	51.9	26.0	51.9	51.9	51.9	5.2
ARGON PURIFIER, MW	16.0	16.0	16.0	8.0	16.0	16.0	16.0	1.6
CONDENSER, MW	809.9	809.9	809.9	404.9	809.9	809.9	809.9	80.9
COAL DRYER, MW	66.4	66.4	16.6	33.2	59.3	17.7	59.3	5.4
STACK LOSS, MW	461.0	461.0	520.9	230.5	482.1	524.9	482.0	49.5
INVERTER, MW	15.2	15.2	15.2	7.6	15.2	15.2	15.2	1.5
CESIUM SYSTEM, MW	4.0	4.0	4.0	2.0	4.0	4.0	4.0	0.4
COMBUSTOR SYSTEM HEAT LOSS, MW	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MHD GENERATOR LOSS, MW	5.5	5.5	5.5	2.75	5.5	5.5	5.5	0.55
MISCELLANEOUS, MW	14.7	14.7	6.6	7.05	13.9	8.1	13.4	3.85
TOTAL HEAT OUTPUT, MW	1463.9	1463.9	1465.9	731.7	1477.1	1472.5	1476.5	141.8
TOTAL ENERGY OUTPUT, MW	3019.7	3019.7	3021.7	1509.6	3032.9	3028.3	3032.3	297.4

CASE	DESCRIPTION	PARAMETRIC CASE	REFERENCE CASE
3.0	REFERENCE PLANT	---	---
3.1	GAS CLEAN-UP	SPRAY DRYER	STRETFORD
3.2	COAL TYPE	ILL. #6	MR
3.3	PLANT SIZE	500 MWe	1000 MWe
3.4	GASIFIER	WINKLER	CE
3.5	COAL TYPE	ILL. #6	MR
3.6	GAS CLEAN-UP	SPRAY DRYER	STRETFORD
3.7	GASIFIER	WELLMAN	CE

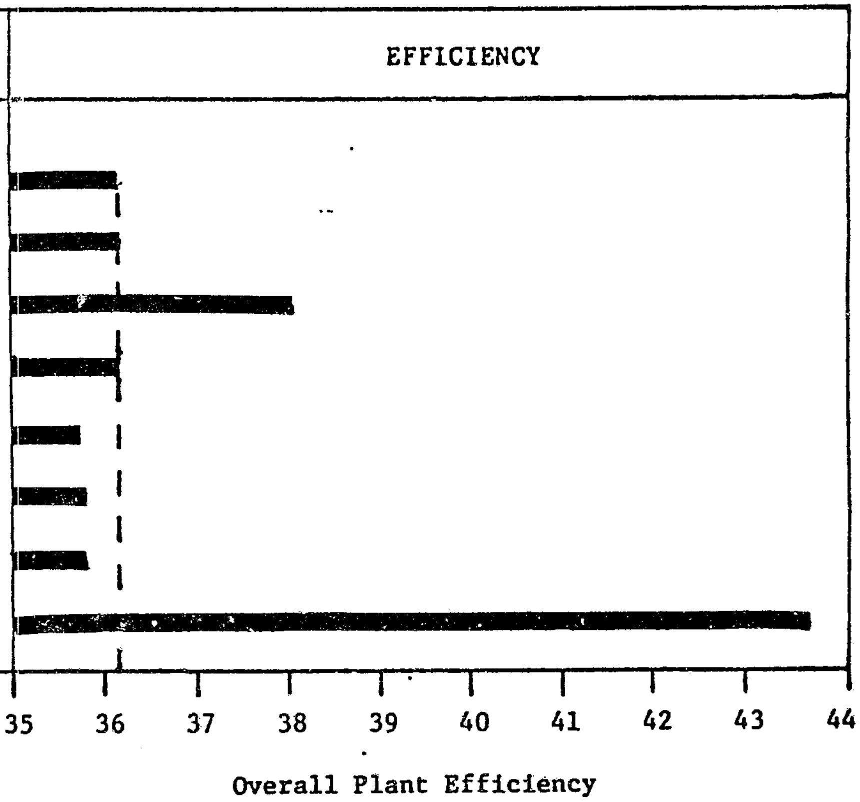

FIGURE 3-8

Reference Plant 3
Overall Plant Efficiency

Table 3-12

REFERENCE PLANT 3

GASIFIEP SYSTEM SUMMARY

CASE	3.0	3.1	3.2	.3	3.4	3.5	3.6	3.7				
VARIATION	REFERENCE PLANT	CLEAN-UP	ILL. # 6 COAL	500 MW_e PLANT	WINKLER GASIFIER	ILL. # 6 COAL	CLEAN-UP	WELLMAN GASIFIER				
PLANT SIZE, MW_e	1000	1000	1000	500	1000	1000	1000	100				
GASIFIER	CE	CE	CE	CE	WINKLER	WINKLER	WINKLER	WELLMAN				
GASIFIER TYPE	ENTRAINED	ENTRAINED	ENTRAINED	ENTRAINED	FLUIDIZED	FLUIDIZED	FLUIDIZED	FIXED				
OXIDANT	AIR	AIR	AIR	AIR	AIR	AIR	AIR	AIR				
GASIFIER PRESSURE, atm	1	1	1	1	1	1	1	1				
COAL TYPE	MR	MR	ILL # 6	MR	MR	ILL. # 6	MR	MR				
COAL MOISTURE, %	5	5	5	5	5	5	5	5				
CLEAN-UP SYSTEM	STRETFORD	SPRAY DRYER	STRETFORD	STRETFORD	STRETFORD	STRETFORD	SPRAY DRYER	STRETFORD				
COAL SIZE	70-200 mesh	70-200 mesh	70-200 mesh	70-200 mesh	0-3/8 in	0-3/8 in	0-3/8 in	1 1/4-2in				
lb AIR OR OXYGEN/LB MAF COAL	4.31	4.31	4.88	4.31	3.50	3.78	3.50	3.31				
lb STEAM/lb MAF COAL	0	0	0	0	0.67	0.67	0.67	0.66				
PRESSURE OF RAW GAS, psia	14.7	14.7	14.7	14.7	14.7	14.7	14.7	14.7				
TEMPERATURE OF RAW GAS, F	300	300	300	300	350	350	350	1000				
PRESSURE OF CLEAN GAS, F	14.7	14.7	14.7	14.7	14.7	14.7	14.7	14.7				
TEMPERATURE OF CLEAN GAS, F	105	105	105	105	105	105	105	105				
HIGHER HEATING VALUE, BTU/SCF	118	118	116	118	125	128	125	142				
COAL INPUT TO GASIFIER, MW_t	3084.4	3084.4	2912.8	1542.2	3125.6	3123.2	3125.6	250.5				
COAL FLOW, ton/day	11,167	11,167	9,841	5,583	11,673	10,552	11,673	907				
NUMBER OF UNITS	4	4	4	2	11	10	11	11				
COMBUSTION EFFICIENCY, %	62.5	62.5	66.2	62.5	61.7	61.7	61.7	77.0				
POWER P' ANT EFFICIENCY, %	33.3	33.3	35.2	33.3	32.9	32.9	32.9	41.2				

Case 3.1 - Hot Gas Cleanup System

A spray dryer system was considered with the CE gasifier system in this parametric case. However, it was felt that this arrangement would not be satisfactory since the raw fuel gas leaving the gasifier should be cleaned and the sulfur removed before the gas entered the combustor and particularly the ceramic regenerative argon heater. For this reason, the gas was still cleaned with a Stretford cleanup system, resulting in the same performance as the base case.

Case 3.2 - Illinois No. 6 Coal

The overall plant efficiency for the CE gasifier using Illinois No. 6 coal was 38% (base case was 36.1% with Montana Rosebud coal). The higher heating value of the clean fuel gas was 116 Btu/SCF with Illinois No. 6 coal which compares to 118 Btu/SCF with Montana Rosebud coal. The busbar power with Illinois No. 6 and Montana Rosebud coal were essentially the same; however, because of the higher heating value for Illinois No. 6 coal, the amount of coal input to the plant was decreased.

Case 3.3 - 500 MWe Plant Size

As in Cases 2.4 and 2.5, this parametric case does not represent the influence of plant size on performance because of the assumed channel enthalpy extraction ratio and channel efficiency. These performance parameters are related to plant size; however, in this study these quantities were specified by NASA-LeRC to be the same as the 1000 MWe base case. Therefore, plant performance was linearly scaled and the overall plant efficiency was identical regardless of the size of plant.

Case 3.4 - Winkler Gasifier

A Winkler fluidized bed gasifier with a Stretford cold gas cleanup system was used in Case 3.4. Montana Rosebud coal, dried to 5% moisture and crushed to 0-3/8 inch size, was considered. The higher heating value of the clean fuel gas was 125 Btu/SCF with the Winkler gasifier compared to 118 Btu/SCF with the CE gasifier. A total of 11 Winkler gasifier units are required for this 1000 MWe plant; whereas, only 4 CE gasifier units are required. The overall power plant efficiency was 35.7% (compared to 36.1% for the CE gasifier plant).

Case 3.5 - Illinois No. 6 Coal

With the Winkler gasifier, the use of Illinois No. 6 coal instead of Montana Rosebud coal has little effect on overall plant performance. The higher heating value of the clean fuel gas was 128 Btu/SCF with Illinois No. 6 coal compared to 125 Btu/SCF using Montana Rosebud. Ten gasifier units were required instead of 11 with Montana Rosebud coal.

Case 3.6 - Spray Dryer

For the reasons stated in Case 3.1, a Stretford cleanup system was recommended to clean the raw fuel gas before it entered the combustor and the argon regenerative heater. With this configuration the performance was the same as Case 3.4.

Case 3.7 - Wellman Gasifier

The Wellman fixed bed gasifier is a relatively small capacity unit which is advantagous for small size plants. For Case 3.7, the plant size was reduced to 100 MWe instead of the typical 1000 MWe plants considered in the majority of the parametric cases. Montana Rosebud coal dried to 5% moisture and crushed to 1 1/4 to 2 inch size was considered. The higher heating value of the clean fuel gas was 142 Btu/SCF. A total of 11 gasifier units were required for this nominal 100 MWe plant. The overall plant efficiency was 43.6%; however, the specified enthalpy extraction ratio was 36% and the channel efficiency was 78%. These channel performance parameters were not adjusted for plant size. Treating the channel as an energy conversion device having the same enthalpy extraction ratio and channel efficiency as a 1000 MWe plant is misleading. The channel performance of a small channel is not expected to be as high as for large channels.

4.0 MAJOR COMPONENTS/SUBSYSTEMS

To attain the system performance in Section 3.0, it was necessary to consider several types of components/subsystems. The following is a summary of those components/subsystems considered and their applicability for CCMHD use.

4.1 Firing Systems

A different firing system was selected for each reference case. In Reference Case 1.0 a direct fired coal combustor was used. Reference Case 2.0 used a high pressure gasifier while a low pressure gasifier was used for Reference Case 3.0. A description of each system follows:

4.1.1 Direct-Fired Combustor - Reference Case 1.0

The direct-fired coal combustor used for Case 1.0 is similar to those proposed for OCMHD power plants. The combustors can be designed to operate either pressurized or at atmospheric pressure and consist of one or two stages. The combustor selected for Case 1.0 has two stages and operates at one atmosphere.

4.1.2 Pressurized Gasifier - Reference Case 2.0

The reference plant 2.0 cycles differ from those in Case 1.0 mainly in that gasifier systems are used instead of a direct coal fired combustor. Three pressurized gasifiers were investigated for Reference Plant 2 - IGT, Westinghouse and Texaco. Discussions of these gasifier systems are given in the following section. Tables 4-1 through 4-20 give a detailed

Table 4-1
Summary of Results
CASE 2.0

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GASIFICATION	
Operation Conditions	
Gasifier Type	IGT
Pressure, psia	147
Bed Temp., ^{o}F	1900-2000
Exit Temp., ^{o}F	1765
Reactants	
Feed Coal*	Montana Rosebud (10% Mois.)
FC + VM	7,987
Ash	1,013
Moisture	1,000
Air or Oxygen, lb/lb maf coal	3.16 (air)
Steam, lb/lb maf coal	0.66
CLEAN-UP	
Process	Morgantown Iron Oxide
Sulfur removed, %	90
Pressure, psia	135
Inlet/Outlet Temp., ^{o}F	1335
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	1.48: 0.67: 1.0
Coal carbon conv. to CH_4, Atom %	8.86
Cold gas efficiency, % coal HHV	76
Cold clean gas (Dry)	
SCF/Ton maf coal	131,300
Mol. Wt.	25.36
HHV, Btu/scf	151
SOLID RESIDUE	
Carbon, Wt.% in ash	20
OTHERS	
Steam decomposition**, %	42

* Based on 7987 lb.maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

TABLE 4-2

GAS COMPOSITION

CASE 2.0

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Basis: 7987 lbs. M.R. Coal (MAF)
10,000 lbs. M.R. Coal (10% Moisture)

Component	Hot Raw Gas At Gasifier Exit			Clean Fuel Gas To Burner		
	Lbs.	Moles	Mole%	Lbs.	Moles	Mole %
CO	8,492	303.2	18.80	8,382	299.2	18.84
CO_2	5,988	136.1	8.44	5,910	134.3	8.46
H_2	413	204.5	12.68	408	201.8	12.71
H_2O	4,068	225.7	14.00	4,015	222.8	14.03
CH_4	720	44.9	2.78	711	44.3	2.79
H_2S	105	3.1	0.19	11	0.3	0.02
NH_3	2.4	0.1	0.01	2.4	0.1	0.01
N_2+A_2	19,456	694.9	43.10	19,203	685.3	43.14
Total	39,244	1,612.5	100.00	38,642	1,588.1	100.00
M.W. (avg.)		24.35			24.33	
Temp., °F		1765			1335	
Press., psia		147			130	
HHV, Btu/SCF		130			130	
Gas Eff., % Coal HHV		103			97	

TABLE 4-3
SUMMARY OF RESULTS
CASE 2.1

GASIFICATION	
Operation Conditions	
Gasifier Type	IGT
Pressure, psia	147
Bed Temp., ^oF	1900-2000
Exit Temp., ^oF	1765
Reactants	
Feed Coal*	Montana Rosebud (10% Moisture)
FC + VM	7,987
Ash	1,013
Particle Size	0 in x 1/4 in
Air or Oxygen, lb/lb maf coal	3.16 (air)
Steam, lb/lb maf coal	0.66
CLEAN-UP	
Process	Stretford
Sulfur removed, %	>99
Pressure, psia	130
Inlet/Outlet Temp., ^oF	105/105
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	1.48:0.67:1.00
Coal carbon conv. to CH_4, atom %	8.86
Cold gas efficiency, % coal HHV	76
Cold clean gas (Dry)	
SCF/Ton maf coal	131,300
Mol. Wt.	25.36
HHV, Btu/scf	151
SOLID RESIDUE	
Carbon, Wt.% in ash	20
OTHERS	
Steam decomposition**, %	42
Carbon conversion, %	95.7

* Based on 7,987 maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

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TABLE 4-4

GAS COMPOSITION
CASE 2.1

Basis: 7,987 lbs. M.R. Coal (MAF)
10,000 lbs. M.R. Coal (10% Moisture)

Component	Hot Raw Gas At Gasifier Exit			Clean Fuel Gas To Burner		
	Lbs.	Moles	Mole%	Lbs.	Moles	Mole%
CO	8,492	303.2	18.80	8,492	303.2	21.73
CO_2	5,988	136.1	8.44	5,988	136.1	9.76
H_2	413	204.5	12.68	413	204.5	14.66
H_2O	4,068	225.7	14.00	214	11.86	0.85
CH_4	720	44.9	2.78	720	44.9	3.22
H_2S	105	3.1	0.19	0	0	10 ppm
NH_3	2.4	0.1	0.01	2.4	0.1	0.01
N_2+A_2	19,456	694.9	43.10	19,456	694.9	49.77
Total	39,244	1,612.5	100.00	35,285	1,395.56	100.00
M.W. (avg.)		24.35			25.29	
Temp., °F		1765			105	
Press., psia		147			130	
HHV, Btu/SCF		130			150	
Gas Eff., % Coal HHV		103			78	

TABLE 4-5
SUMMARY OF RESULTS
CASE 2.2

GASIFICATION	
Operation Conditions	
Gasifier Type	IGT
Pressure, psia	147
Bed Temp., ^oF	1900-2000
Exit Temp., ^oF	1500
Reactants, lbs.	
Feed Coal*	Ill. No. 6 (5% Moisture)
FC + VM	8,311
Ash	1,189
Coal Size	0 in x 1/4 in.
Air or Oxygen, lb/lb maf coal	3.41 (Air)
Steam, lb/lb maf coal	0.62
CLEAN-UP	
Process	Morgantown Iron Oxide
Sulfur removed, %	90
Pressure, psia	135
Inlet/Outlet Temp., ^oF	1200/1200
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	1.47:0.76:1.0
Coal carbon conv. to CH_4, atom %	13.19
Cold gas efficiency, % coal HHV	75
Cold clean gas (Dry)	
SCF/Ton maf coal	154,300
Mol. Wt.	25.59
HHV, Btu/scf	154
SOLID RESIDUE	
Carbon, Wt.% in ash	20
OTHERS	
Steam decomposition**, %	46
Steam Export, 10^6 Btu	
From Pretreater	0.25
From Heat Recov. Unit	3.42

* Based on 8311 lbs maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

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TABLE 4-6
GAS COMPOSITION
CASE 2.2

Basis: 10,000 lbs. Ill. No.6 Coal (5% Moisture)
8,311 lbs. Ill. No.6 Coal (MAF)

Component	Hot Raw Gas At Gasifier Exit			Clean Fuel Gas To Burner		
	Lbs.	Moles	Mole%	Lbs.	Moles	Mole%
CO	8,228	293.8	17.37	7,872	277.4	17.47
CO_2	6,684	151.9	8.98	6,312	143.4	9.03
H_2	405	200.5	11.85	383	189.3	11.92
H_2O	3,273	181.6	10.73	3,090	171.4	10.79
CH_4	1,146	71.5	4.23	1,083	67.5	4.25
H_2S	366	10.7	0.63	33	1.0	0.06
NH_3	2.4	0.14	0.01	2.3	0.1	0.01
N_2	21,903	781.7	46.20	20,682	738.1	46.47
Total	42,007	1,691.84	100.00	39,457	1,588.2	100.00
M.W. (avg.)		24.83			24.78	
Temp., °F		1500			1200	
Press., psia		147			130	
HHV, Btu/SCF		137			138	
Gas Eff., % Coal HHV		97			94	

TABLE 4-7
SUMMARY OF RESULTS

CASE 2.3

GASIFICATION	
Operation Conditions	
Gasifier Type	IGT
Pressure, psia	221
Bed Temp., °F	1900/2000
Exit Temp., °F	1560
Reactants, lbs.	
Feed Coal*	Montana Rosebud (10% Moisture)
FC + VM	7,987
Ash	1,013
Coal Size	0 in. x 1/4 in.
Air or Oxygen, lb/lb maf coal	3.16 (air)
Steam, lb/lb maf coal	0.66
CLEAN-UP	
Process	Morgantown Iron Oxide
Sulfur removed, %	90
Pressure, psia	206
Inlet/Outlet Temp., °F	1200/1200
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	1.44:0.67:1.0
Coal carbon conv. to CH_4, atom %	11.0
Cold gas efficiency, % coal HHV	78.5
Cold clean gas (Dry)	
SCF/Ton maf coal	153,600
Mol. Wt.	25.28
HHV, Btu/scf	156
SOLID RESIDUE	
Carbon, Wt.% in ash	20
OTHERS	
Steam decomposition**, %	40

* Based on 7987 lbs maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

TABLE : 4-8

GAS COMPOSITION

CASE 2.3

Basis: 7,987 lbs. M.R. Coal (MAF)
10,000 lbs. M.R. Coal (10% Moisture)

Component	Hot Raw Gas At Gasifier Exit			Clean Fuel Gas To Burner		
	Lbs.	Moles	Mole%	Lbs.	Moles	Mole%
CO	8,237	294.1	18.17	8,130	290.3	18.21
CO_2	5,988	136.1	8.41	5,910	134.3	8.43
H_2	413	204.5	12.64	408	201.8	12.66
H_2O	4,172	231.5	14.31	4,118	228.5	14.34
CH_4	864	53.9	3.33	853	53.2	3.33
H_2S	105	3.1	0.19	11	0.3	0.02
NH_3	2.4	0.1	0.01	2.4	0.1	0.01
N_2+A_2	19,456	694.9	42.94	19,203	685.3	43.00
Total	39,237	1,618.2	100.00	38,635	1,593.8	100.00
M.W. (avg.)		24.26			24.24	
Temp., ^{o}F		1560			1200	
Press., psia		221			200	
HHV, Btu/SCF		133			133	
Gas Eff., % Coal HHV		103			95	

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TABLE 4-9
SUMMARY OF RESULTS
CASE 2.6

GASIFICATION	
Operation Conditions	
Gasifier Type	IGT
Pressure, psia	147
Bed Temp., ^oF	1900-2000
Exit Temp., ^oF	1855
Reactants, lbs.	
Feed Coal*	Montana Rosebud (10% Moisture)
FC + VM	7,987
Ash	1,013
Coal Size	0 in. x 1/4 in.
Air or Oxygen, lb/lb maf coal	0.60 (Oxygen)
Steam, lb/lb maf coal	0.61
CLEAN-UP	
Process	Morgantown Iron Oxide
Sulfur removed, %	90
Pressure, psia	135
Inlet/Outlet Temp., ^oF	1160
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	1.15:0.49:1.00
Coal carbon conv. to CH_4, atom %	8.2
Cold gas efficiency, % coal HHV	83.5
Cold clean gas (Dry)	
SCF/Ton maf coal	85,600
Mol. Wt.	20.90
HHV, Btu/scf	302
SOLID RESIDUE	
Carbon, Wt.% in ash	20
OTHERS	
Steam decomposition**, %	70

* Based on 7987 lbs maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

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TABLE 4-10
GAS COMPOSITION
CASE 2.6

Basis:	10,000 lbs.	M.R.	Coal	(10% Moisture)
	7,987 lbs.	M.R.	Coal	(MAF)

Component	Hot Raw Gas At Gasifier Exit			Clean Fuel Gas To Burner		
	Lbs.	Moles	Mole%	Lbs.	Moles	Mole%
CO	8,720	311.3	34.52	8,621	307.8	34.57
CO_2	5,768	131.1	14.53	5,703	129.6	14.56
H_2	545	269.8	29.91	538	266.3	29.92
H_2O	2,493	138.3	15.33	2,465	138.3	15.54
CH_4	668	41.6	4.61	660	41.1	4.62
H_2S	105	3.1	0.34	11	0.3	0.03
NH_3	2.4	0.14	0.02	2.4	0.14	0.02
N_2+A_2	188	6.7	0.74	186	6.6	0.74
Total	18,489	902.0	100.00	18,186	890.14	100.00
M.W. (avg.)		20.50			20.46	
Temp., ^oF		1855			1160	
Press., psia		147			130	
HHV, Btu/SCF		255			255	
Gas Eff., % Coal HHV		101			94	

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TABLE 4-11
SUMMARY OF RESULTS
CASE 2.10

GASIFICATION	
Operation Conditions	
Gasifier Type	Westinghouse two-stage
Pressure, psia	147
Bed Temp., °F	2000-2100
Exit Temp., °F	1600-1700
Reactants	
Feed Coal*, Lbs.	Montana Rosebud (10% moisture)
FC + VM	7,987
Ash	1,013
Particle Size	1/8 - 1/4 in. x 0 in.
Air or Oxygen, lb/lb maf coal	3.16 (air)
Steam, lb/lb maf coal	0.66
CLEAN-UP	
Process	Dolomite In-Bed
Sulfur removed, %	85
Pressure, psia	147
Inlet/Outlet Temp., °F	1600-1800/1600-1800
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	2.11:1.05:1.0
Coal carbon conv. to CH_4, atom %	6
Cold gas efficiency, % coal HHV	70
Cold clean gas (Dry)	
SCF/Ton maf coal	150,700
HHV, Btu/scf	131.1
SOLID RESIDUE	
Carbon, Wt.% in ash	20
OTHERS	
Steam decomposition**, %	15.5
Heat Available for Export, 10^6 Btu	—
HHV of Char, 10^6 Btu	

* Based on 7987 lbs maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

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TABLE 4-12

GAS COMPOSITION

CASE 2.10

Basis: 10,000 lbs. M.R. Coal (10% moisture)
7,987 lbs. M.R. Coal (MAF)

Component	Hot Raw Gas At Gasifier Exit			Clean Fuel Gas To Burner		
	Lbs.	Moles	Mole%	Lbs.	Moles	Mole%
CO				9,679	345.6	19.42
CO_2				4,837	109.9	6.18
H_2				331	163.9	9.21
H_2O				5,488	304.6	17.12
CH_4		Not Applicable		458	28.6	1.61
H_2S *				16	0.5	0.03
NH_3				-	-	-
N_2				23,147	826.1	46.43
Total				43,956	1,779.2	100.00
M.W. (avg.)					24.71	
Temp., ^oF					1600-1800	
Press., psia					147	
HHV, Btu/SCF					108.56	
Gas Eff., % Coal HHV					99	

* 85% of sulfur removal assumed

TABLE 4-13
SUMMARY OF RESULTS
CASE 2.11

GASIFICATION

Operation Conditions	
Gasifier Type	Westinghouse two-stage
Pressure, psia	147
Bed Temp., ^oF	2000-2100
Exit Temp., ^oF	1600-1700
Reactants	
Feed Coal*, lbs.	Montana Rosebud (10% moisture)
FC + VM	7.987
Ash	1,013
Particle Size	1/8 - 1/4 in. x 0 in.
Air or Oxygen, lb/lb maf coal	3.16 (air)
Steam, lb/lb maf coal	0.66

CLEAN-UP

Process	Stretford
Sulfur removed, %	>99
Pressure, psia	147
Inlet/Outlet Temp., ^oF	105/105

PRODUCT GAS

$CO:CO_2:H_2$ Mole ratio	2.11:1.05:1.0
Coal carbon conv. to CH_4, atom %	6
Cold gas efficiency, % coal HHV	70
Cold clean gas (Dry)	
SCF/Ton maf coal	169,100
HHV, Btu/scf	131.1

SOLID RESIDUE

Carbon, Wt.% in ash	2.0

OTHERS

Steam decomposition**, %	15.5
Heat Available for Export, 10^6 Btu	14.359
HHV of Char, 10^6 Btu	—

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* Based on 7987 lbs maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

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TABLE 4-14

GAS COMPOSITION

CASE 2.11

Basis: 10,000 lbs. M.R. Coal (10% moisture)
7,987 lbs. M.R. Coal (MAF)

Component	Hot Raw Gas At Gasifier Exit Lbs.	Moles	Mole%	Clean Fuel Gas To Burner Lbs.	Moles	Mole%
CO	9,679	345.6	19.40	9,679	345.6	23.25
CO_2	4,837	109.9	6.17	4,837	109.9	7.39
H_2	331	163.9	9.20	331	163.9	11.02
H_2O	5,488	304.6	17.10	228	12.6	0.85
CH_4	458	28.6	1.61	458	28.6	1.93
H_2S	105	3.1	0.17	0	0	0
NH_3	-	-	-	-	-	-
N_2	23,147	826.1	46.35	23,147	826.1	55.56
Total	44,045	1,781.8	100.00	38,680	1,486.7	100.00
M.W. (avg.)		24.72			26.02	
Temp., ^{o}F		1600-1800			105	
Press., psia		147			130	
HHV, Btu/SCF		108.47			129.95	
Gas Eff., % Coal HHV		99			71	

TABLE 4-15

SUMMARY OF RESULTS

CASE 2.12

GASIFICATION	
Operation Conditions	
Gasifier Type	Texaco
Pressure, psia	147
Bed Temp., ^oF	~3000
Exit Temp., ^oF	1800
Reactants, Lbs.	
Feed Coal*	Montana Rosebud (22.7% moisture)
FC + VM	7,987
Ash	1,013
Particle Size	70% - 200 mexh
Air or Oxygen, lb/lb maf coal	5.20 (air)
Water, lb/lb maf coal	0.50
CLEAN-UP	
Process	Stretford
Sulfur removed, %	>99
Pressure, psia	130
Inlet/Outlet Temp., ^oF	105/105
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	2.11:067:1.0
Coal carbon conv. to CH_4, atom %	
Cold gas efficiency, % coal HHV	65
Cold clean gas (Dry)	
SCF/Ton maf coal	210,400
HHV, Btu/scf	98
SOLID RESIDUE	
Carbon, Wt.% in ash	2.0
OTHERS	
Steam decomposition**, %	not applicable
Steam Export, 10^6 Btu	27.96

* Based on 7,987 lbs. maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

TABLE 4-16
GAS COMPOSITION
CASE 2.12

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Basis: 11,643 lbs. M.R. Coal (22.7 % moisture)
7,987 lbs. M.R. Coal (MAF)

Component	Hot Raw Gas At Gasifier Exit			Clean Fuel Gas To Burner		
	Lbs.	Moles	Mole%	Lbs.	Moles	Mole%
CO	9,560	341.3	15.40	9,560	341.3	18.62
CO_2	7,076	134.6	6.07	7,076	134.6	7.34
H_2	413	204.6	9.23	413	204.6	11.17
H_2O	7,143	396.4	17.88	281	15.6	0.85
CH_4	29	1.8	0.08	29	1.8	0.10
H_2S	105	3.1	0.13	0	0	0
NH_3	-	-	-	-	-	-
N_2	31,805	1,135.1	51.20	31,805	1,135.1	61.92
Total	56,131	2,216.9	100.00	49,164	1,833.0	100.00
M.W. (avg.)		24.80			26.19	
Temp., ^{o}F		1800			105	
Press., psia		147			130	
HHV, Btu/SCF		80.28			97.08	
Gas Eff., % Coal HHV		103.4			66.1	

TABLE 4-17

SUMMARY OF RESULTS

CASE 2.13

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GASIFICATION	
Operation Conditions	
Gasifier Type	Texaco
Pressure, psia	147
Bed Temp., ^{o}F	3000
Exit Temp., ^{o}F	2000
Reactants	
Feed Coal*, lbs.	Illinois No. 6 (8.9% moisture)
FC + VM	8311
Ash	1189
Particle Size	70%-200 mesh
Air or Oxygen, lb/lb maf coal	5.20
Water, lb/lb maf coal	0.72
CLEAN-UP	
Process	Stretford
Sulfur removed, %	>99
Pressure, psia	130
Inlet/Outlet Temp., ^{o}F	105/105
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	1.67:0.66:1.0
Coal carbon conv. to CH_4, atom %	1
Cold gas efficiency, % coal HHV	65
Cold clean gas (Dry)	
SCF/Ton maf coal	214,100
HHV, Btu/scf	102.8
SOLID RESIDUE	
Carbon, Wt.% in ash	2.0
OTHERS	
Steam decomposition**, %	not applicable
Heat Available for Export, 10^6 Btu	32.911

* Based on 8311 lbs maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

***Hot raw gas cooled down to 300^{o}F is assumed.

TABLE : 4-18
GAS COMPOSITION
CASE 2.13

Basis: 10,428 lbs. Ill. No. 6 Coal (8.9% moisture)
8,311 lbs. Ill. No. 6 Coal (MAF)

Component	Hot Raw Gas At Gasifier Exit			Clean Fuel Gas To Burner		
	Lbs.	Moles	Mole%	Lbs.	Moles	Mole%
CO	10,801	385.6	16.43	10,801	385.6	19.58
CO_2	6,695	152.1	6.48	6,695	152.1	7.72
H_2	466	230.7	9.83	466	230.7	11.71
H_2O	6,921	384.1	16.36	302	16.7	0.85
CH_4	33	2.1	0.09	33	2.1	0.10
H_2S	366	10.7	0.46	0	0	0
NH_3	-	-	-	-	-	-
N_2	33,128	1,182.3	50.35	33,128	1,182.3	60.04
Total	58,410	2,347.6	100.00	51,425	1,969.5	100.00
M.W. (avg.)		24.88			26.11	
Temp. °F		200			105	
Press., psia		147			130	
HHV, Btu/SCF		85.5			101.92	
Gas Eff., % Coal HHV		103			66	

TABLE 4-19

SUMMARY OF RESULTS

CASE 2.14

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GASIFICATION	
Operation Conditions	
Gasifier Type	Texaco
Pressure, psia	147
Bed Temp., °F	~3000
Exit Temp., °F	2100
Reactants	
Feed Coal*, lbs.	Montana Rosebud (22.7% moisture)
FC + VM	7,987
Ash	1,013
Particle Size	70% - 200 mesh
Air or Oxygen, lb/lb maf coal	0.95 (100% oxygen)
, lb/lb maf coal	0.60
CLEAN-UP	
Process	Stretford
Sulfur removed, %	>99
Pressure, psia	130
Inlet/Outlet Temp., °F	105/105
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	1.47:0.54:1.0
Coal carbon conv. to CH_4, atom %	0.13
Cold gas efficiency, % coal HHV	73
Cold clean gas (Dry)	
SCF/Ton maf coal	104,600
HHV, Btu/scf	262.6
SOLID RESIDUE	
Carbon, Wt.% in ash	2
OTHERS	
Steam decomposition**, %	not applicable
Heat Available for Export, 10^6 Btu	17.524

* Based on 7,987 lbs maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

TABLE 4-20

GAS COMPOSITION

CASE 2.14

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Basis: 11,643 lbs. M.R. Coal (22.7% moisture)
7,987 lbs. M.R. Coal (MAF)

Component	Hot Raw Gas At Gasifier Exit Lbs.	Moles	Mole%	Clean Fuel Gas To Burner Lbs.	Moles	Mole%
CO	10,327	368.7	33.47	10,327	368.7	47.89
CO_2	5,922	134.6	12.21	5,922	134.6	17.49
H_2	506	250.5	22.73	506	250.5	32.54
H_2O	6,047	335.6	30.45	118	6.5	0.85
CH_4	11	0.7	0.06	11	0.7	0.09
H_2S	105	3.1	0.28	0	0	0
NH_3	-	-	-	-	-	-
N_2	247	8.8	0.80	247	8.8	1.14
Total	23,165	1,102.0	100.00	17,131	769.8	100.00
M.W. (avg.)		21.02			22.26	
Temp., ^{o}F		2100			105	
Press., psia		147			130	
HHV, Btu/SCF		181.84			260.39	
Gas Eff., % Coal HHV		98			73.5	

summary of the gasification systems considered and the predicted raw gas and clean fuel gas compositions. Tables for Cases 2.4, 2.5, 2.7, 2.8 and 2.9 have been omitted because they are the same as those for Case 2. Methodology used to perform these analyses is decribed below.

In order to perform parametric analyses of closed cycle MHD power plants, it was necessary to predict various gasifier performances for each type of coal and for each type of oxidant (air or oxygen). The following assumptions were used to determine the product gas compositions and yields:

1. Coal requirements were based on 7987 lbs of Moisture and Ash Free (MAF) Montana Rosebud coal. The coal and ash analyses are given in Table 1-1. Extents of coal drying were varied with gasifier type and feed type (dry or slurry feed). For dry feed, drying of coal feed to 5 wt. % was assumed for fixed-bed gasifiers, 5-10 wt. % for fluidized-bed gasifiers, and 2-5 wt. % for entrained-bed gasifiers. Drying is not required for slurry feed, i.e., in the Texaco gasification cases.

2. Oxidant to coal ratio, steam (or water) to coal ratio, CO:CO_2:H_2 ratio, amount of CH_4 formed, gas heating values, and the coal gas efficiency were based on the published data judged to be representative for each gasifier. Heat losses from the gasifiers were based on 1.0% of coal HHV for most gasifiers and 0.5% for the C.E. gasifier.

3. All sulfur in the coal was assumed to be converted to H_2S. Carbon content in the ash was 2-35 wt.% based on published data projected for commercial operation.

4. Material and energy balances were carried out to determine the amount of gases produced for a given amount of coal. The gas compositions were adjusted to fit all the above assumptions. The extent of sulfur removal was assumed to be 90% for Morgantown iron oxide process, 85% for in-situ hot cleanup process, greater than 99% for the Stretford desulfurization process, and 80% for a dry FGD process.

For each case, the higher heating values (HHV, Btu/SCF) are given for hot raw gas, clean fuel gas, and dry cold clean gas. The gas efficiency is defined as:

$$\text{Gas Efficiency (\%)} = \frac{\text{(Sensible Heat + Chemical Heat) x 100}}{\text{Coal HHV}}$$

Gas efficiency as defined above can exceed 100% since coal HHV represents only a part of the total heat input, i.e., the oxidant sensible heat and the steam latent heat are not included.

The cold gas efficiency is defined below:

$$\text{Cold Gas Efficiency (\%)} = \frac{\text{Chemical Heat in the Gas x 100}}{\text{Coal HHV}}$$

C-2

Cold gas efficiency is always less than 100%, since the sensible heat and gasifier heat loss are not included.

Table 4-21 summarizes the overall gasifier performance, including the physical dimensions and characteristics, for each gasifier system considered in Reference Plant 2.

4.1.2.1 IGT Gasifier

Developer: The Institute of Gas Technology (IGT) Chicago, Illinois

Description: The U-Gas fluid-bed gasifier is of vertical cylindrical construction with an internal cyclone for returning the elutriated fines to the bed. A sloped grid at the bottom, containing one or more inverted cones, serves as the air and steam distributer and the agglomerated ash outlet. A schematic of the IGT U-Gas system is shown in Figure 4-1 and a process schematic for MHD power generation is in Figure 4-2.

Crushed coal (1/4 in x 0 in) is pressurized in a lock-hopper operated at 50-350 psi and 800 F, reacted with air then fed to a fluidized-bed gasifier operating at 50-350 psi and 1900 F. Air and steam are introduced at the base of the gasifiers. The coal is rapidly gasified without slagging, and the high temperature inhibits formation of tar oil, or phenols.

Gases from the gasifier pass through heat recovery and sulfur removal systems. Gas produced is about 155 Btu/SCF; substituting oxygen for air produces a medium Btu fuel gas (~300 Btu/SCF). The ash, which contains 5-20 wt. % carbon, is selectively removed from the fluidized bed bottom by agglomeration. The agglomerates fall into a water-filled

TABLE 4-21

Reference Plant 2.0

GASIFIER OVERALL PERFORMANCES

Cases	2.0	2.1	2.2	2.3	2.4	2.5	2.6	2.7	2.8	2.9	2.10	2.11	2.12	2.13	2.14
Power Output, MWe	1,000	1,000	1,000	1,000	500	250	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000
Feed Coal															
Type	M.R.	M.R.	Ill.	M.R.	M.R.	M.R.	M.R.	M.R.	M.R.	M.R.	M.R.	M.R.	M.R.	Ill.	M.R.
Mois., Wt. %	10	10	5	10	10	10	10	10	10	10	10	10	22.7	8.9	22.7
TPD	10,038	10,666	9,836	10,150	5,019	2,510	10,673	9,034	9,510	10,038	12,699	12,398	13,592	10,493	12,348
Gasifier															
Type	IGT										Westinghouse		TEXACO		
Oxidant	Air	Air	Air	Air	Air	Air	Oxygen	Air	Air	Air	Air	Air	Air	Air	Oxygen
Pressure, ATM	10	10	10	15	10	10	10	10	10	10	10	10	10	10	10
Size, Ft	22ID x 30H										13.5ID x 95H		90D x 15H		
Thruput,(1) TPD/Unit	1,650	1,650	1,650	1,910	1,650	1,650	2,200	1,650	1,650	1,650	1,200	1,200	1,430	1,430	1,900
Constru. Mat'l.	Carbon Steel														
Refractory	Yes														
Units Req'd(2)	6	7	6	7	3	2	7	6	6	6	11	11	10	8	7

(1) Gasifier throughputs have been adjusted for the oxidant type and pressure effect.

(2) Numbers indicate operating units; spare gasifiers are not included.

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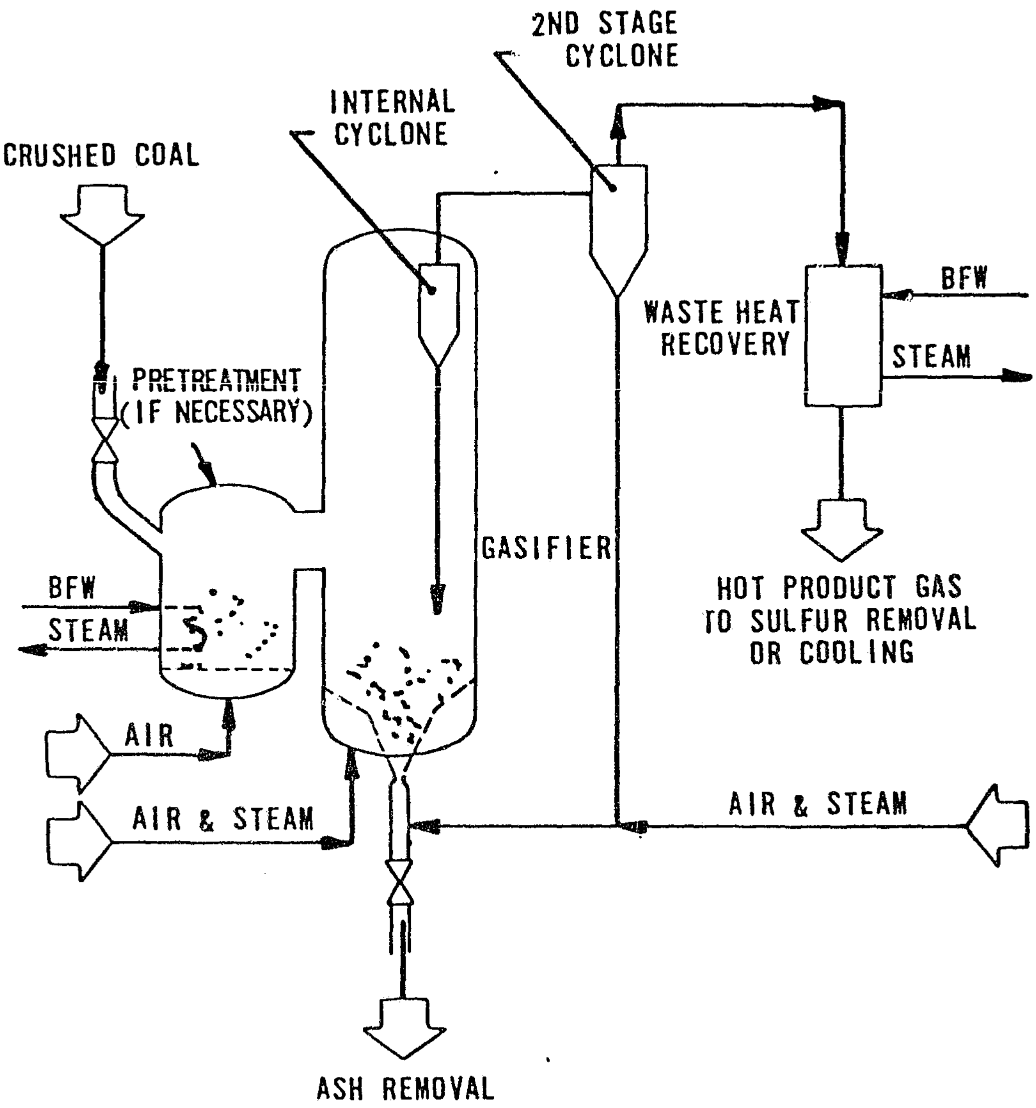

Figure 4-1

IGT COAL GASIFICATION SYSTEM

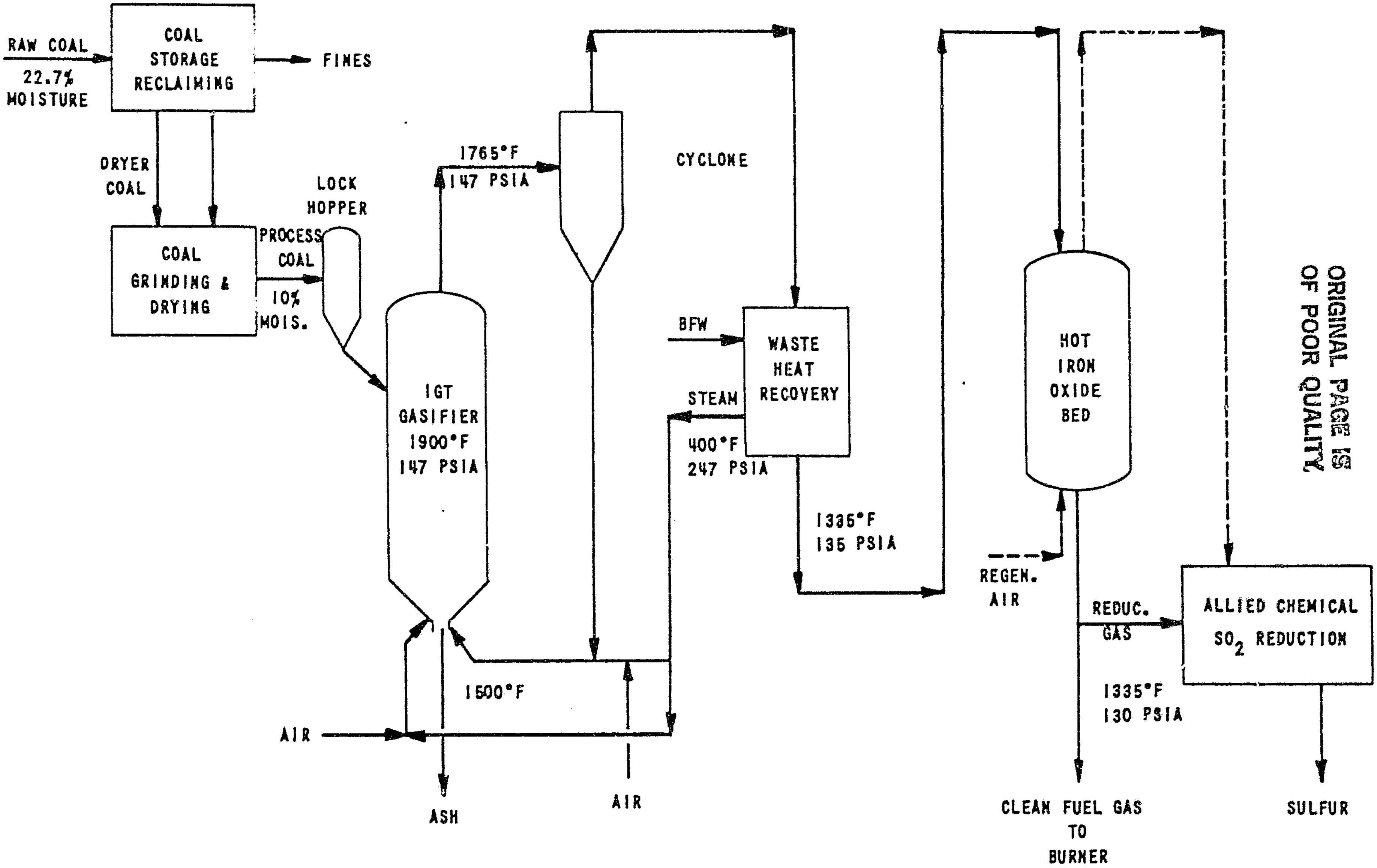

Figure 4-2

PROCESS SCHEMATIC FOR IGT GASIFICATION SYSTEM CASE 2.0

hopper and are withdrawn as a slurry. Unburned coal moving out of the reactor is separated from the gas stream by cyclones and returned to the bed.

Status: The research and development of this process is co-sponsored by the American Gas Association and DOE. The process has been tested in an air-blown 485 lb/hr unit showing suitability for both combined cycle power generation and a "grass roots" source of industrial and power generation energy. A design study was performed for a 10-35 TPH pilot plant sufficient to fuel a 100 MWe power utility.

4.1.2.2 Westinghouse Gasifier

Developer: Westinghouse Electric Corp., Research and Development Center, Pittsburgh, Pa.

Description: The Westinghouse multistage fluid-bed gasification process consists of two vertical cylindrical vessels - a recirculating bed devolatilizer/desulfurizer and an agglomerating fluidized bed combustor/gasifier. The schematic is shown in Figure 4-3.

Crushed (1/8 - 1/4 in x 0 in), dried coal is fed into a central draft tube of the devolatilizer/desulfurizer unit. Coal and internally recycled solids flowing at a velocity greater than 15 fps are carried upward in the draft tube by hot gases from a combustor. Recycled solids flow downward in a fluidized bed surrounding the draft tube at rates up to 100 times the coal feed rate. They dilute the coal feed to prevent agglomeration as it devolatizes. Heat required for the coal-system gasification reactions is provided by hot gases produced in the combustor, which is operated at 1900-2100 F and 130-200 psig.

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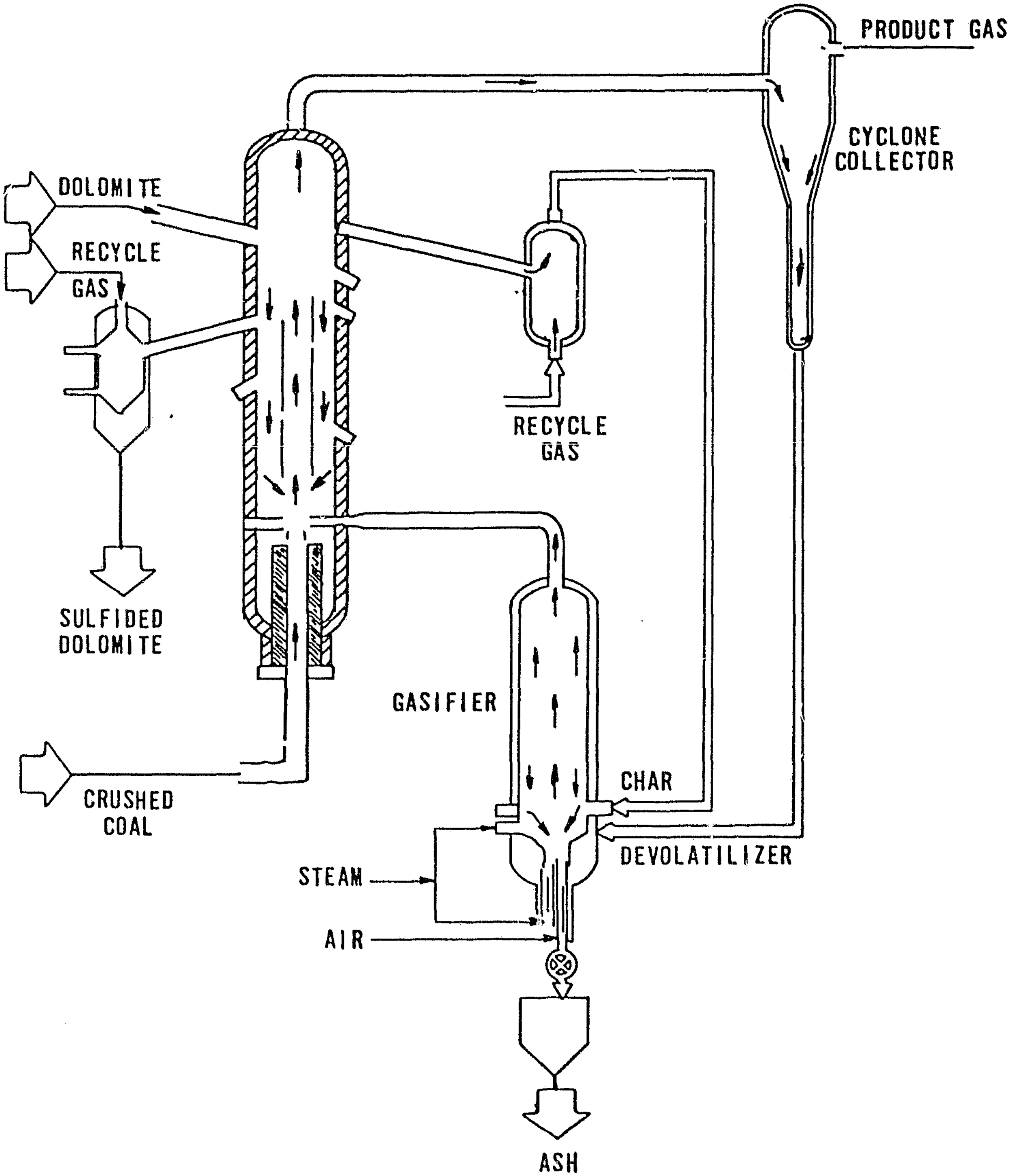

Figure 4-3

WESTINGHOUSE GASIFICATION PROCESS

A lime sorbent is added to the devolatilizer/desulfurizer reactor, operated at 1600-1800 F and 130-200 psig, to remove sulfur which is present as hydrogen sulfide in the gas. Spent sorbent is withdrawn from the reactor after stripping out the char. Spent sorbent is regenerated or discarded. Char is withdrawn from the top section of the devolatilizer/desulfurizer and fed to the combustor. There, char is gasified with air and steam at 1900-2100 F in the combustor/gasifier. Ash agglomerates at this temperature and is removed.

Raw product gas (about 135 Btu/SCF) from the devolatilizer/desulfurizer unit passes through a cyclone to remove fines and then through a heat recovery unit. Fines are recycled to the combustor. Oils and tars are not produced in this process. No excess steam is generated in the two stage gasifier sections.

Status: Development of this process began in 1972 and a 15 TPD Process Development Unit (PDU) (for the combustor/gasifier section) has been operating since 1975. The initial concept for the Westinghouse coal gasification system was a two-reactor system (as described above) which includes absorption of hydrogen sulfide with dolomite. From the development work on the gasifier, the concept of a single-stage system has evolved in addition to the original two-reactor system. The two-stage concept is currently not under consideration for initial commercialization; however, work in this area is still being explored at the Westinghouse Research Laboratories.

4.1.2.3 Texaco Gasifier

Developer: Texaco, Inc., Montebello, California

Description: The Texaco entrained-bed gasifier is a vertical, cylindrical pressure vessel having a carbon steel shell. In this downward, concurrent, entrained-bed reactor, pulverized coal (70% through 200 mesh) is continuously fed to the reactor at temperatures of 2000 to 3000 F. Since the coal particles move through the reactor in a dilute phase, they are essentially not in contact with each other and hence both caking and non-caking coals can be used. Residence time is as low as a few seconds, which results in high throughput.

Due to the high temperatures, by-product tars, phenols, and heavy hydrocarbons are not formed. H_2S in the gas is removed in downstream desulfurization units.

A concentrated blend of coal and water feeds the Texaco gasifier. Coal slurry is fed through a special burner where it is mixed with oxygen and an additional temperature moderator (such as steam), if required. After leaving the refractory-lined reactor vessel, slag is quenched in a water tank and the gas stream enters a heat recovery section.

Status: The Texaco gasifier is commercially proven with hydrocarbon feedstocks. Texaco's Montebello, Calif., research laboratory is operating two pilot gasifiers, each capable of converting 15 to 20 TPD of coal. Tests on various coals have been conducted at pressures up to 1200 psi. Also, a 150-tons/day plant recently began operating in Obenhausen, Germany.

Preliminary engineering work on a Texaco coal gasification demonstration plant in Southern California is nearing completion; this plant will be capable of generating about 90 MW of electric power.

4.1.3 Atmospheric Gasifier - Reference Plant 3.0

Case 3.0 is primarily a Case 1.0 design with an atmospheric gasifier replacing the coal-fired combustor. A Combustion Engineering (CE) gasifier was used for the base case 3.0. A Winkler and a Wellman atmospheric gasifier were also considered in Cases 3.4 and 3.7, respectively. A description of each gasifier is included in Sections 4.1.3.1 through 4.1.3.3.

Tables 4-22 through 4-31 are a Summary of Results showing design criteria and an exit gas composition for each Reference Case 3.0 low pressure gasifier system. Tables for Cases 3.1, 3.3 and 3.6 have been omitted because they are identical with Case 3.0.

Table 4-32 gives a summary of the overall gasifier performance, including the physical dimensions and characteristics, for each gasifier system used in Reference Plant 3.

4.1.3.1 Combustion Engineering Gasifier

Developer: Combustion Engineering Inc., Windsor, Connecticut

Description: The Combustion Engineering entrained-bed gasifier is of vertical, cylindrical construction and is designed for atmospheric pressure operation. A combustion section, consisting of tangentially oriented combustor nozzles, is at the bottom of the structure. Directly

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TABLE 4-22

SUMMARY OF RESULTS

CASE 3.0

GASIFICATION	
Operation Conditions	
Gasifier Type	Combus. Eng.
Pressure, psia	14.7
Bed Temp., °F	3000-3400
Exit Temp., °F	300
Reactants	
Feed Coal*	Montana Rosebud (5% Mois.)
FC + VM	7,987
Ash	1,013
Particle Size	70% minus 200 mesh
Air or Oxygen, lb/lb maf coal	4.31 (air)
Steam, lb/lb maf coal	0
CLEAN-UP	
Process	Stretford
Sulfur removed, %	>99
Pressure, psia	14.7
Inlet/Outlet Temp., °F	105/105
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	2.00:0.52:1.00
Coal carbon conv. to CH_4, atom %	0
Cold gas efficiency, % coal HHV	70
Cold clean gas (Dry)	
SCF/Ton maf coal	156,100
Mol. Wt.	25.97
HHV, Btu/scf	118
SOLID RESIDUE	
Carbon, Wt.% in ash	2.0
OTHERS	
Steam decomposition**, %	133
Steam generated, % coal HHV	28.87
Carbon conversion, %	99.7

* Based on 7987 lb. maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

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TABLE 4-23

GAS COMPOSITION

CASE 3.0

Basis: 7987 lbs. M.R. Coal (MAF)
9474 lbs. M.R. Coal (5% Mois.)

Component	Hot Raw Gas At Gasifier Exit			Clean Fuel Gas * To Burner		
	Lbs.	Moles	Mole%	Lbs.	Moles	Mole%
CO	11,204	400.0	24.18	11,204	400.0	22.49
CO_2	4,365	103.9	6.28	4,365	103.9	5.84
H_2	404	200.0	12.09	404	200.0	11.24
H_2O	113	6.3	0.38	2,404	133.4	7.50
CH_4	-	-	-	-	-	-
H_2S	105	3.1	0.19	0	0	-
NH_3	-	-	-	-	-	-
N_2+A_2	26,375	941.3	56.86	26,375	941.3	52.93
Total	42,566	1,654.6	100.00	44,752	1,778.6	100.00
M.W. (avg.)		25.85			25.28	
Temp., °F		300			105	
Press., psia		14.7			14.7	
HHV, Btu/SCF		117			109	
Gas Eff., % Coal HHV		72			70	

* 10 ppm H_2S in volume remained in the fuel gas

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TABLE 4-24

SUMMARY OF RESULTS

CASE 3.2

GASIFICATION	
Operation Conditions	
Gasifier Type	Combustion Engineering
Pressure, psia	14.7
Bed Temp., ^oF	3000-3400
Exit Temp., ^oF	300
Reactants	
Feed Coal* , lbs.	Ill. No. 6 (2% moisture)
FC + VM	8,311
Ash	1,189
Particle Size	70% - 200 mesh
Air or Oxygen, lb/lb maf coal	4.88
Steam, lb/lb maf coal	none
CLEAN-UP	
Process	Stretford
Sulfur removed, %	>99
Pressure, psia	14.7
Inlet/Outlet Temp., ^oF	105/105
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	1.98:0.39:1.0
Coal carbon conv. to CH_4, atom %	0
Cold gas efficiency, % coal HHV	70.44
Cold clean gas (Dry)	
SCF/Ton maf coal	175,000
HHV, Btu/scf	116.37
SOLID RESIDUE	
Carbon, Wt.% in ash	2
OTHERS	
Steam decomposition**, %	not applicable
Heat Available for Export, 10^6 Btu	20.959
HHV of Char, 10^6 Btu	—

* Based on 8311 lbs maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

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TABLE 4-25
GAS COMPOSITION
CASE 3.2

Basis: 9,694 lbs. Ill. No. 6 Coal (2% moisture)
8,311 lbs. Ill. No. 6 Coal (MAF)

Component	Hot Raw Gas At Gasifier Exit			Clean Fuel Gas To Burner		
	Lbs.	Moles	Mole%	Lbs.	Moles	Mole%
CO	12,610	450.2	23.46	12,610	450.2	22.19
CO_2	3,943	89.6	4.67	3,943	89.6	4.42
H_2	459	227.4	11.85	459	227.4	11.21
H_2O	563	31.2	1.63	307	152.2	7.50
CH_4	-	-	-	-	-	-
H_2S	366	10.7	0.56	0	0	0
NH_3	-	-	-	-	-	-
N_2	31,098	1,109.9	57.83	31,098	1,109.9	34.68
Total	49,039	1,919.0	100.00	48,417	2,029.3	100.00
M.W. (avg.)		25.55			25.06	
Temp., ^oF		300			105	
Press., psia		14.7			14.7	
HHV, Btu/SCF		113.82			107.66	
Gas Eff., % Coal HHV		74			74	

TABLE 4-26
SUMMARY OF RESULTS
CASE 3.4

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GASIFICATION	
Operation Conditions	
Gasifier Type	Winkler
Pressure, psia	14.7
Bed Temp., ^{o}F	1500-1800
Exit Temp., ^{o}F	350-400
Reactants	
Feed Coal*, lbs.	Montana Rosebud (5% moisture)
FC + VM	7,987
Ash	1,013
Particle Size	0 in. x 3/8 in.
Air or Oxygen, lb/lb maf coal	3.50 (air)
Steam, lb/lb maf coal	0.67
CLEAN-UP	
Process	Stretford
Sulfur removed, %	>99
Pressure, psia	147
Inlet/Outlet Temp., ^{o}F	105/105
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	1.59:0.64:1.0
Coal carbon conv. to CH_4, atom %	2.34
Cold gas efficiency, % coal HHV	65
Cold clean gas (Dry)	
SCF/Ton maf coal	159,000
HHV, Btu/scf	125.03
SOLID RESIDUE	
Carbon, Wt.% in ash	37
OTHERS	
Steam decomposition**, %	26.22
Heat Available for Export, 10^6 Btu	10.761
HHV of Char, 10^6 Btu	8.385

* Based on 7987 lbs maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

TABLE 4-27
GAS COMPOSITION
CASE 3.4

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Basis: 9,474 lbs. M.R. Coal (5 % moisture)
7,987 lbs. M.R. Coal (MAF)

Component	Hot Raw Gas At Gasifier Exit			Clean Fuel Gas To Burner		
	Lbs.	Moles	Mole%	Lbs.	Moles	Mole%
CO	8,866	316.5	18.89	8,866	316.5	20.53
CO_2	5,621	127.7	7.62	5,621	127.7	8.28
H_2	403	199.5	11.91	403	199.5	12.94
H_2O	4,434	246.1	14.69	2,084	115.6	7.50
CH_4	189	11.8	0.70	189	11.8	0.76
H_2S	105	3.1	0.19	0	0	0
NH_3	-	-	-	-	-	-
N_2	21,593	770.6	46.00	21,593	770.6	49.99
Total	42,819	1,675.3	100.00	38,756	1,541.7	100.00
M.W. (avg.)		24.60			25.14	
Temp., ^{o}F		350-400			105	
Press., psia		14.7			14.7	
HHV, Btu/SCF		106.43			115.65	
Gas Eff., % Coal HHV		74			68	

TABLE 4-28

SUMMARY OF RESULTS

CASE 3.5

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GASIFICATION	
Operation Conditions	
Gasifier Type	Winkler
Pressure, psia	14.7
Bed Temp., °F	1500-1800
Exit Temp., °F	350-400
Reactants	
Feed Coal*, lbs.	Illinois No. 6 (5% moisture)
FC + VM	7,987
Ash	1,013
Particle Size	0 in. x 3/8 in.
Air or Oxygen, lb/lb maf coal	3.78 (air)
Steam, lb/lb maf coal	0.67
CLEAN-UP	
Process	Stretford
Sulfur removed, %	> 99
Pressure, psia	147
Inlet/Outlet Temp., °F	105/105
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	1.58:0.51:1.0
Coal carbon conv. to CH_4, atom %	2.46
Cold gas efficiency, % coal HHV	65.29
Cold clean gas (Dry)	
SCF/Ton maf coal	168,800
HHV, Btu/scf	127.87
SOLID RESIDUE	
Carbon, Wt.% in ash	35
OTHERS	
Steam decomposition**, %	25.86
Heat Available for Export, 10^6 Btu	12.151
HHV of Char, 10^6 Btu	8.977

* Based on 8311 lbs maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

TABLE 4-29
GAS COMPOSITION
CASE 3.5

Basis: 9,500 lbs. Ill. No. 6 Coal (5 % moisture)
8,311 lbs. Ill. No. 6 Coal (MAF)

	Hot Raw Gas At Gasifier Exit			Clean Fuel Gas To Burner		
Component	Lbs.	Moles	Mole%	Lbs.	Moles	Mole%
CO	10,061	359.2	19.41	10,061	359.2	20.98
CO_2	5,116	116.2	6.28	5,116	116.2	6.79
H_2	458	226.7	12.25	458	226.7	13.24
H_2O	4,628	256.8	13.88	259	128.4	7.50
CH_4	214	13.3	0.72	214	13.3	0.78
H_2S	366	10.7	0.56	0	0	0
NH_3	-	-	-	-	-	-
N_2	24,315	867.8	46.90	24,315	867.8	50.71
Total	45,158	1,850.7	100.00	40,423	1,711.6	100.00
M.W. (avg.)		24.40			24.82	
Temp., °F		350-400			105	
Press., psia		14.7			14.7	
HHV, Btu/SCF		109.41			118.27	
Gas Eff., % Coal HHV		73.5			66.0	

TABLE 4-30

SUMMARY OF RESULTS

CASE 3.7

GASIFICATION	
Operation Conditions	
Gasifier Type	Wellman
Pressure, psia	14.7
Bed Temp., ^{o}F	2000-2500^{o}F
Exit Temp., ^{o}F	1000
Reactants	
Feed Coal*, lbs.	9,500 (5% moisture)
FC + VM	8,311
Ash	1,189
Particle Size	1-1/4" x 2"
Air or Oxygen, lb/lb maf coal	3.31 (air)
Steam, lb/lb maf coal	0.66
CLEAN-UP	
Process	Stretford
Sulfur removed, %	>99
Pressure, psia	14.7
Inlet/Outlet Temp., ^{o}F	110/110
PRODUCT GAS	
$CO:CO_2:H_2$ Mole ratio	1.33:0.51:1.0
Coal carbon conv. to CH_4, atom %	7.0
Cold gas efficiency, % coal HHV	75.0
Cold clean gas (Dry)	
SCF/Ton maf coal	156,400
HHV, Btu/scf	141.5
SOLID RESIDUE	
Carbon, Wt.% in ash	12.0
OTHERS	
Steam decomposition**, %	30.26
Tar/Oil HHV, 10^6 Btu	0.621

* Based on 7987 lbs maf coal

** 100 - (H_2O in hot raw gas - H_2O in coal feed) (100.0)/(steam + H_2O in air)

TABLE 4-31
GAS COMPOSITION
CASE 3.7

Basis: 9,500 lbs. M.R. Coal (5% moisture)
7,987 lbs. M.R. Coal (MAF)

	Hot Raw Gas At Gasifier Exit			Clean Fuel Gas To Burner		
Component	Lbs.	Moles	Mole%	Lbs.	Moles	Mole%
CO	8,107	289.4	17.62	8,107	289.4	19.02
CO_2	5,797	131.7	8.02	5,797	131.7	8.66
H_2	440	217.8	13.26	440	217.8	14.32
H_2O	4,103	227.7	13.86	2,057	114.2	7.51
CH_4	558	34.8	2.12	558	34.8	2.29
C_2H_4	122	4.3	0.26	122	4.3	0.28
H_2S	105	3.1	0.19	0	0	0
NH_3	49	2.9	0.18	10	.9	0.06
N_2	20,406	728.3	44.32	20,406	728.3	47.86
Tar & Oil	426	2.8	0.17	0	0	0
Total	40,113	1,642.8	100.00	37,520	1,521.4	100.0
M.W. (avg.)		24.41			24.65	
Temp., ^oF		1000			105	
Press., psia		14.7			14.7	
HHV (dry, clean gas)		125.28			135.1	
Gas Eff., % Coal HHV		91.0			77.4	

Table 4-32

Reference Plant 3

GASIFIER SYSTEM SUMMARY

Cases	3.0	3.1	3.2	3.3	3.4	3.5	3.6	3.7
Power Output, MWe	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000
Feed Coal								
Type	M.R.	M.R.	Ill.	M.R.	M.R.	Ill.	M.R.	M.R.
Mois., Wt. %	5	5	2	5	5	5	5	5
TPD	11,167	11,167	9,841	5,583	11,673	10,552	11,673	907
Gasifier								
Type	COMB. ENG.				WINKLER			WELLMAN
Oxidant	Air	Air	Air	Air	Air	Air	Air	Air
Pressure, ATM	1	1	1	1	1	1	1	1
Size, Ft	30W x 40L x 170H				18ID x 75H			9-12D x 12H
Thruput,[1] TPD/unit	2,760	2,760	2,760	2,760	1,100	1,100	1,100	84
Constr. Mat'l.	CARBON STEEL							
Refractory	YES							
Units Req'd.[2]	4	4	4	2	11	10	11	11

(1) Gasifier throughputs have been adjusted for the oxident type and pressure effect.

(2) Numbers indicate operating units; spare units are not included.

above the combustion section is the reduction section where steam (if used) and additional feed coal is fed into the gasifier. Hot gases from the combustion section and the steam feed (if required) entrain and gasify the feed coal as it passes vertically through the unit.

A process schematic for a C.E. gasification system used for MHD power generation is shown in Figure 4.4.

Pulverized coal (70% through 200 mesh) and recycled char are fed through the combustor nozzles and oxidized at 3000-3400 F with a near stoichiometric quantity of air. The resulting hot gases rise into the reduction section while molten slag formed in the combustor is removed from the bottom and quenched. Steam and pulverized coal are injected tangentially through the reduction nozzles into the hot gases rising from the combustion section. Feed coal is devolatilized and the volatiles are cracked in the lower, high temeprature portion of the reducing section of the gasifier. As the gases rise through the remainder of the gasifier, they are cooled to 1600-1700 F by the endothermic gasification reactions.

Gases exiting the top of the reduction section are directed downward into a waste heat recovery unit (heat exchanger section), where their temperature is reduced by tubular heat transfer surfaces. These surfaces recover heat in the form of saturated steam, superheated steam, and sensible heat in a transfer medium.

The product gases leave the heat exchanger sections at 300 F and enter a spray dryer, a cyclone, and a scrubber which remove the particulate matter. Char and ash thus collected are recycled to the combustor coal

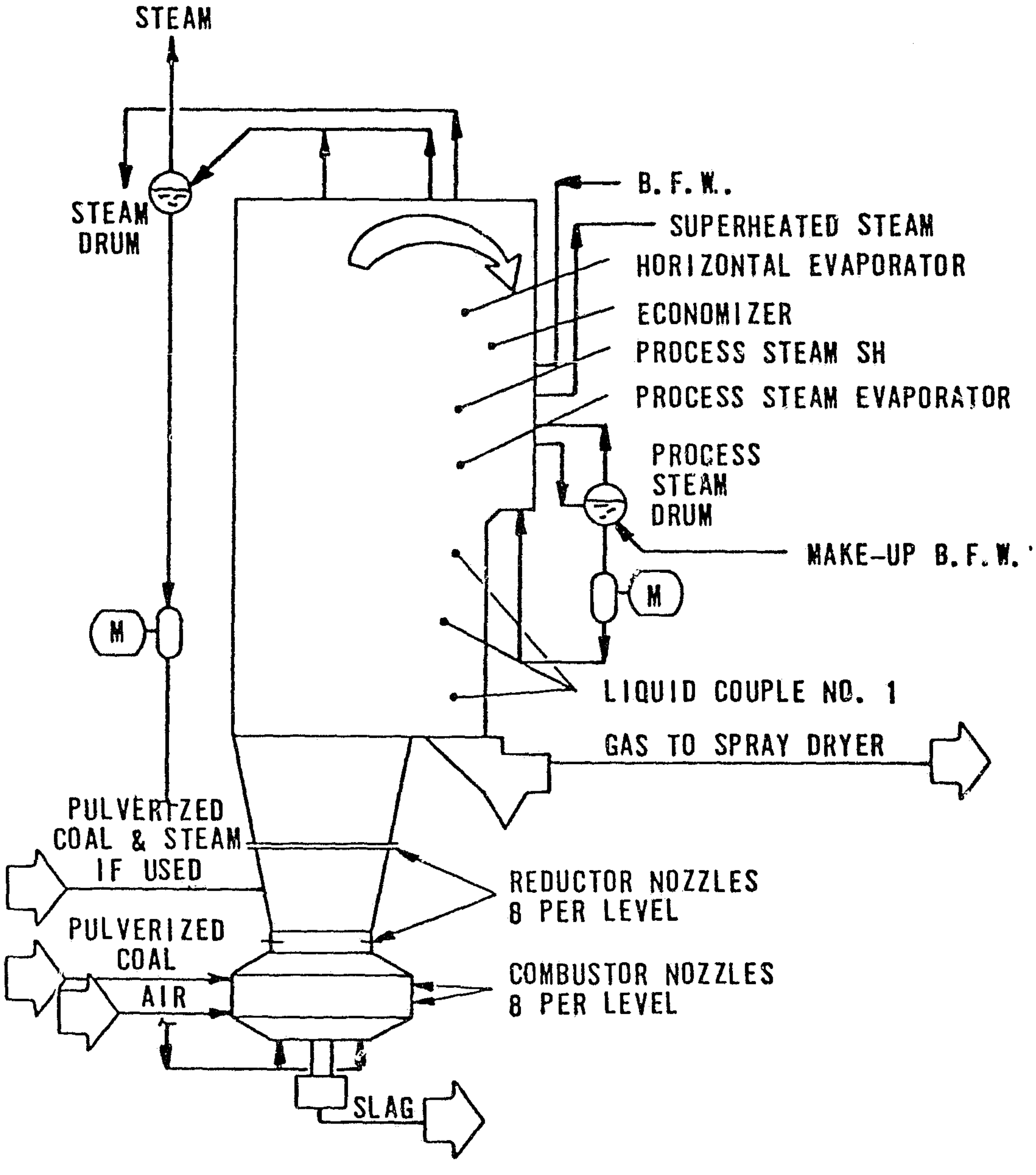

Figure 4-4

COMBUSTION ENGINEERING GASIFIER

pulverizer. Product gases are then sent to a sulfur removal unit, resulting in a low Btu (about 127 Btu/SCF) fuel gas. By substituting oxygen for air in the gasification, a medium Btu gas (about 285-316 Btu/SCF) can be produced.

Status: The Combustion Engineering gasifier is not yet commercially available. If the planned 120 TPD PDU operations are successful, the gasifier may become available for integrated coal gasification/electric power generation.

For the system used, the following were assumed:

1. Steam is not used in the C.E. process and the moisture in the coal feed must be reduced to 5% in order to achieve a high bed temperature (3000-3400 F).

2. 30.04×10^6 Btu of heat (equivalent to 28.8% of coal HHV) is available for steam generation.

3. The gas composition predicted is speculative, since air-blown slurry feed has not been tested.

4.1.3.2 Winkler Gasifier

Developer: Davy Power gas, Inc., Lakeland, Florida

Description: The Winkler fluidized-bed gasifier has a vertical cyclindrical construction with a steel shell lined on the inside with refractory. A schematic of the gasifier is shown in Figure 4-5.

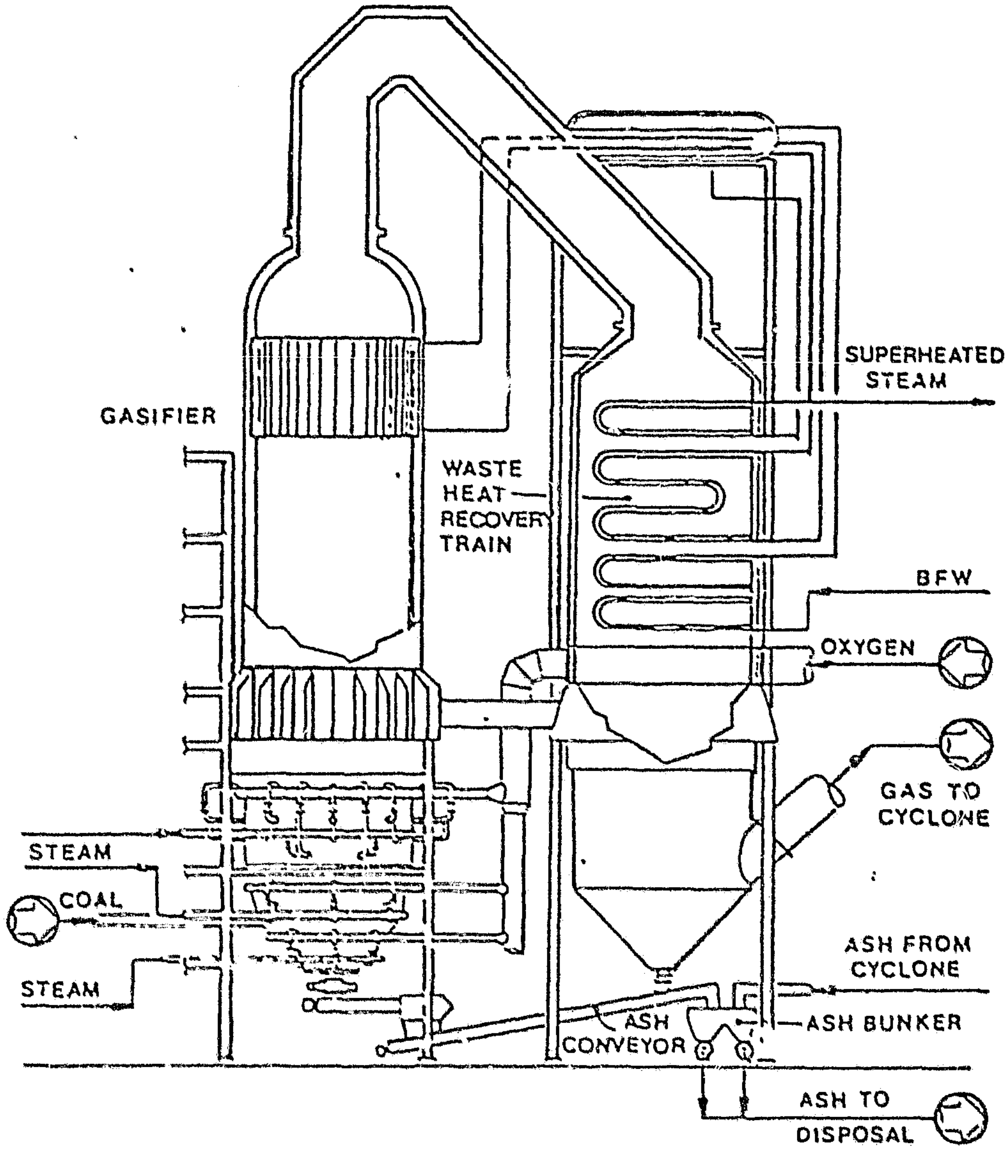

Figure 4-5 Winkler Gasifier

Crushed coal (0 in x 3/8 in) is dried and fed to the atmospheric bed gasifier through a variable-speed screw feeder. Coal reacts with air (or oxygen) and steam to produce a raw gas rich in carbon monoxide and hydrogen. Because of the high temperature (1500-1800 F), all tars and heavy hydrocarbons are reacted.

About 70% of the ash is carried over by the gas and 30% is removed from the bottom of the gasifier by the ash screw. Unreacted carbon carried over by gas is converted by secondary steam and oxygen in the space above the fluidized bed. As a result, maximum temperature occurs above the fluidized bed. To prevent ash particles from melting and forming deposits in the exit duct, the gas is cooled in a radiant boiler section before it leaves the gasifier.

Raw gas leaving the gasifier is passed through a heat recovery section. As a result, the gas temperature is reduced to about 350-400 F. Fly ash is removed by cyclones, wet scrubbers and an electr[illegible]atic precipitator. H_2S in the gas is removed by a desulfurization unit. T[illegible]duct gas has a heating value of about 125 Btu/SCF when air is used as an oxidant, and about 280 Btu/SCF when oxygen is used.

Status: The Winkler gasifier is commercially available. This process was developed over fifty years ago. The process has been used commercially at 16 plants in a number of countries, using a total of 36 generators. Most previous experience was with German brown coals and their coke. Davy Powergas Inc. is currently developing a high pressure modification (up to 210 psig) of the Winkler process which should increase the thermal efficiency.

4.1.3.3 Wellman Gasifier

Developer: McDowell - Wellman Co.

Description: There are two types of Wellman fixed-bed gasifiers: the standard type and the agitated type. The rated capacity of an agitated gasifier is about 25% higher than that of the standard gasifier of the same size, and unlike the standard gasifiers, it can handle caking bituminous coals. The agitated gasifier, as shown schematically in Figure 4-6, is described below:

Crushed coal (3/16 in x 5/16 in) is fed from the top while an air (or oxygen) - steam mixture is introduced through a revolving grate at the bottom. The agitator gasifier, operated at 1 atm, has a slowly revolving horizontal arm which spirals vertically below the surface of the fuel bed. The agitator reduces channeling and maintains a uniform bed. Crude gas leaving the gasifier between 1000 and 1200 F contains tar, oil, phenols, and particulates. Ash is removed continuously through a slowly revolving eccentric grate at the bottom of the reactor.

After leaving the gasifier, the hot raw gas is passed through a gas cooling and purification section. Ash, carried over by gas, and tar/oil are removed by scrubbing. The gas containing H_2S is then sent to a desulfurization unit. The product gas has a heating value of 120-160 Btu/SCF when air used as oxident.

Status: This process is commercial and has been in use for over 35 years.

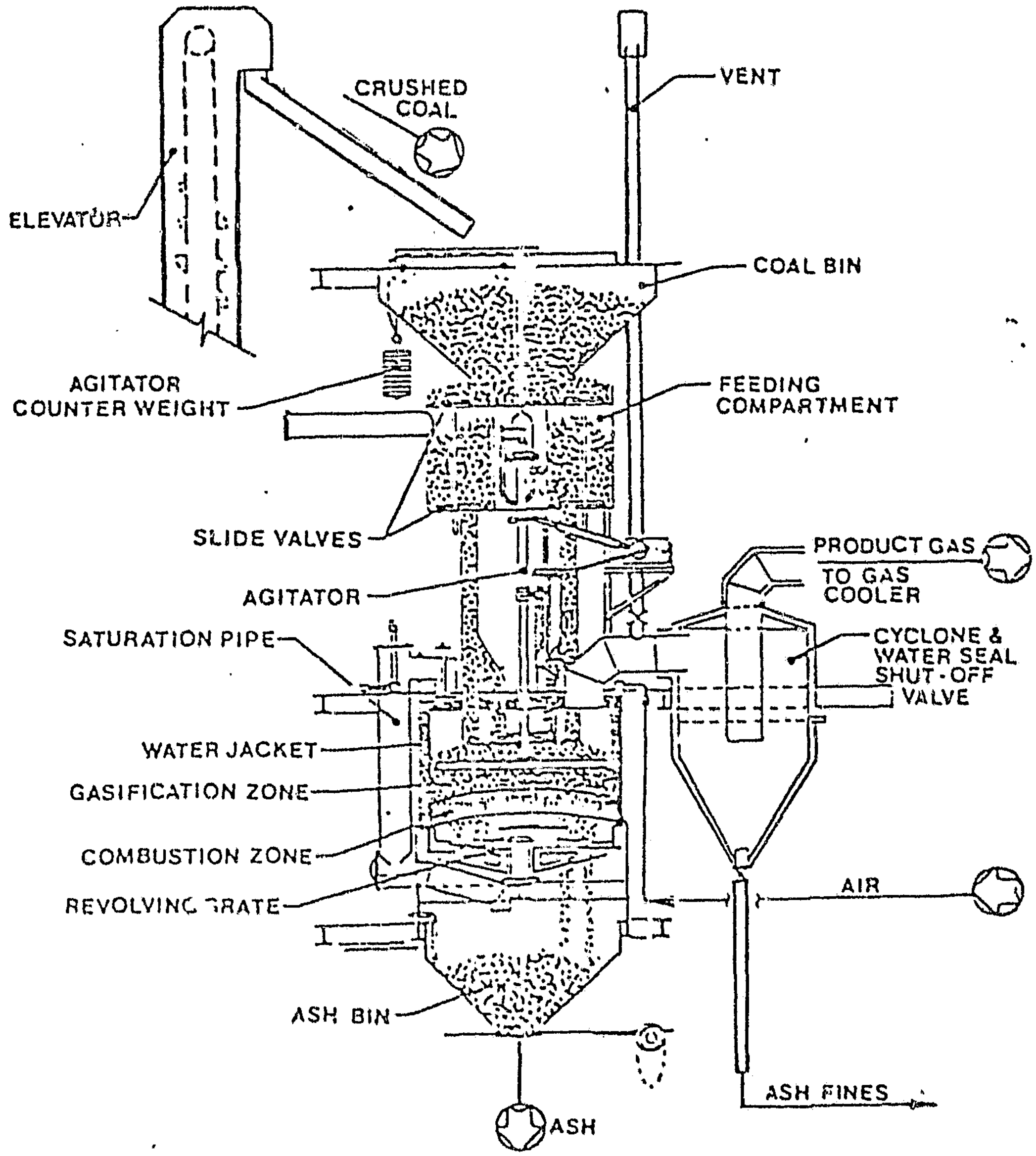

Figure 4-6 Wellman-Galusha Gasifier

4.2 Gas Cleanup Systems

Four approaches to gas cleanup were considered for the gasifier systems used, namely cold gas cleanup, hot gas cleanup, in-bed cleanup and flue gas desulphurization. Each of these processes is described below: Table 4-33 is a summary of the systems used in each case.

4.2.1 Cold Gas Cleanup

Four cold gas cleanup processes were considered for removing H_2S from the fuel gas and converting it to elemental sulfur. The candidate processes were the Stretford, Amine Absorption, Benfield, and Selexol processes. The Stretford process was selected for use in the CCMHD studies and is therefore the only cold gas cleanup process discussed.

The Stretford process has been proven on coal-derived gas and natural gas under low pressure operation (about 1 atm). It can be modified for high pressure operation. A successful test has been demonstrated in a coal liquefaction plant at Cresap, West Virginia to remove 1.6 tons of sulfur under 200 psig. Some of the CO_2 will be absorbed under pressurized conditions; the extent of the CO_2 absorption requires further investigation.

4.2.1.1 Stretford Desulfurization Process

The process flow sheet is shown in Figure 4-7. The product gas from the gasifier enters the hyrdogen sulfide (H_2S) absorber where nearly all the H_2S is removed. In the absorber unit, the gas contacts the Stretford solution, which is an aqueous solution of a vanadium salt, anthraquinone disulfonic acid (ADA). The H_2S in the producer gas is oxidized by the vanadium to elemental sulfur, while the vanadium is reduced from a V^{+5}

TABLE 4-33

DESULFURIZER OVERALL PERFORMANCES

Cases	2.0	2.1	2.2	2.3	2.4	2.5	2.6	2.7	2.8	2.9	2.10	2.11
Power Output, MWe	1,000	.1,000	1,000	1,000	500	250	1,000	1,000	1,000	1,000	1,000	1,000
Feed Coal												
Type	M.R.	M.R.	Ill.	M.R.	M.R.	M.R.	M.R.	M.R.	M.R.	M.R.	M.R.	M.R.
TPD	10,038	10,666	9,836	10,150	5,019	2,510	10,673	9,034	9,510	10,038	12,599	12,398
Sul. Cont., TPD	99.4	105.6	338.4	100.5	49.7	24.8	105.7	89.4	94.2	99.4	125.7	122.7
Desulfurizer												
Type	Hot Iron Oxide	Stret'd	← Hot Iron Oxide →								In-Bed	Stret'd
Size, Ft		-	← 14D x 24H →								-	-
Constru. Mat'l.	Special Alloys	C.S	← Special Alloys →								C.S.	C.S.
Thruput TPD Sulfur (1)	-	45	-	-	-	-	-	-	-	-	-	45
Total Sul. Remov., TPD	89.5	104.5	304.6	90.5	44.7	22.3	95.1	80.5	84.8	89.5	106.9	121.5
Units Req'd. (2,3)	20	2	46	22	10	6	20	18	20	20	-	3

(1) Various thruputs of Stretford desulfurization units are used.

(2) Numbers indicate oprating units; spare units are not included.

(3) The units required have been adjusted for 8-hr cycle time by considering the flow rates and the sulfur contents in the gas.

TABLE 4-33 (Cont'd.)

DESULFURIZER OVERALL PERFORMANCES

Cases	2.12	2.13	2.14	3.0	3.1	3.2	3.3	3.4	3.5	3.6	3.7
Power Output, MWe	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000
Feed Coal											
Type	M.R.	Ill.	M.R.	M.R.	M.R.	Ill.	M.R.	M.R.	Ill.	M.R.	M.R.
TPD	13,592	10,493	12,348	11,167	11,167	9,841	5,583	11,673	10,552	11,673	907
Sul. Cont., TPD	115.6	346.1	105.0	120.4	120.4	349.2	60.2	122.0	382.1	122.0	9.5
Desulfurizer											
Type	Stretford	Stretford	Stretford	Stretford	Dry FGD	Stretford	Stretford	Stretford	Stretford	Dry FGD	Stret'd
Size, Ft	-	-	-	-	-	-	-	-	-	-	-
Constru. Mat'l	C.S	C.S	C.S	C.S	C.S	C.S	C.S	C.S	C.S	C.S	C.S
Thruput TPD Sulfur(1)	45	90	45	45	-	90	30	45	90	-	10
Total Sul. Remov., TPD	114.5	342.6	104.6	119.2	108.4	345.7	59.6	120.8	378.3	109.8	9.4
Units Req'd.(2,3)	3	4	3	3	-	4	2	3	4	-	1

(1) Various thruputs of Stretford desulfurization units are used.

(2) Numbers indicate oprating units; spare units are not included.

(3) The units required have been adjusted for 8-hr cycle time by considering the flow rates and the sulfur contents in the gas.

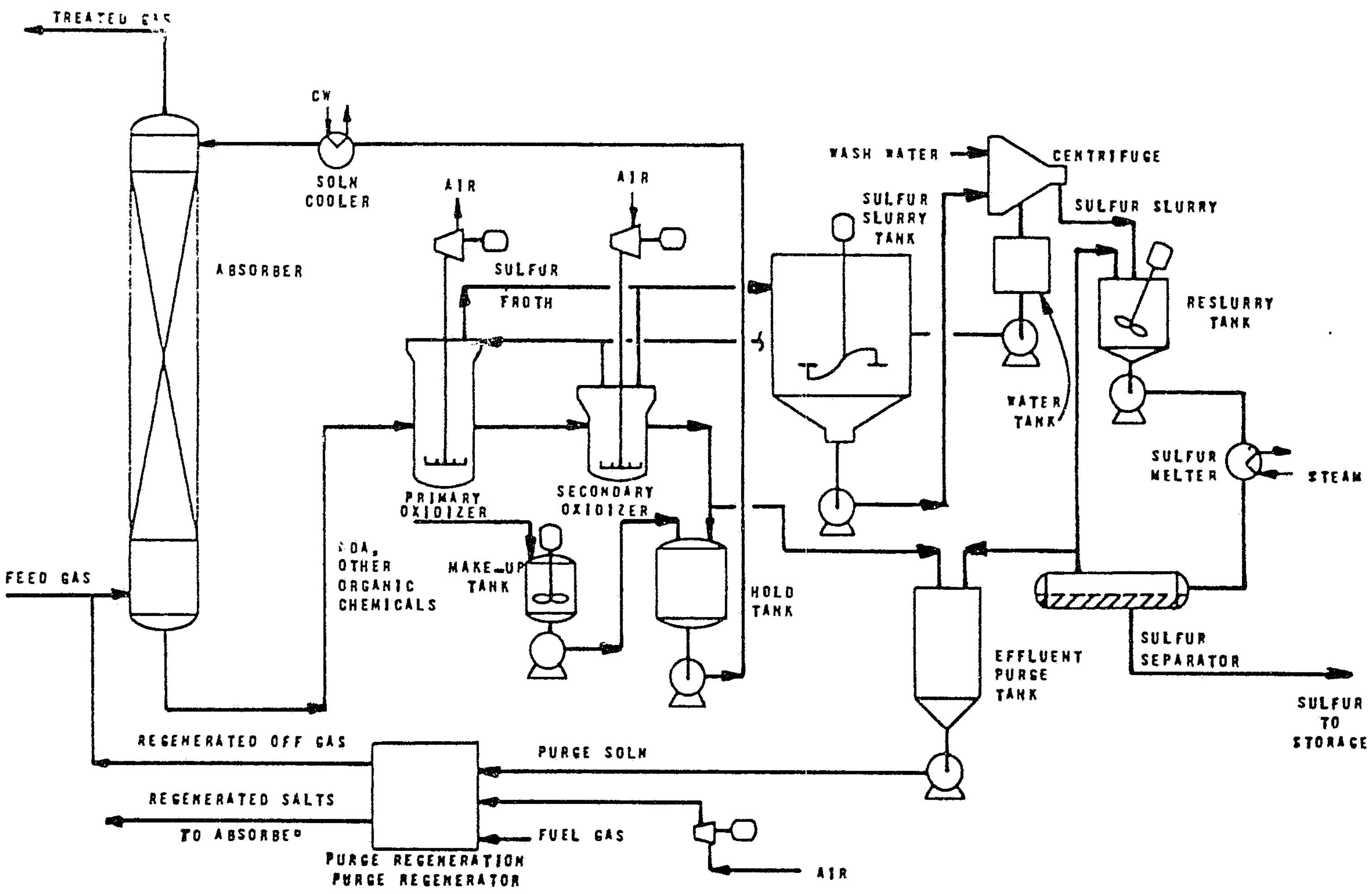

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Figure 4-7

STRETFORD DESULFURIZATION SYSTEM

to a V^{+4} state. The delay tank at the bottom of the absorber provides the necessary residence time for chemical reaction to proceed before the solution goes to the oxidizer tank.

In the oxidizer tanks, air provided by blowers oxidizes the vanadium from the V^{+4} back to the V^{+5} state (the ADA acts as a catalyst for this oxidation). The air also acts as a flotation agent, forming a sulfur/liquor/air froth that floats to the top of the oxidizer. The clear, regenerated liquor flows from the bottom and passes to a pump tank from which it is recycled to the absorber column. The sulfur froth, or slurry, overflows from the oxidizer to the slurry tank. The filter operates periodically and produces sulfur in the form of a washed, wet filter cake.

This process is capable of reducing the sulfur level in the producer gas to below four parts per million. This sulfur level is comparable to that in natural gas and should be more than adequate to meet the EPA standards.

Over 50 Stretford desulfurization plants are operating worldwide, with plant capacities from 0.1×10^6 to 200×10^6 SCFD (atmospheric operation). Sulfur removal rates range from 0.5 TPD to 90 TPD. In this study, sulfur removal rates of 10, 30, 45, and 90 TPD are used. Carbon steel construction is suitable for the Stretford process; however, inert linings such as epoxy resins are required for oxidizer and sulfur slurry tanks. Use of stainless steel linings for solution and sulfur slurry pumps is also recommended.

4.2.2 Hot Gas Cleanup

Several processes are under development to remove H_2S from fuel gases at high temperatures (1000-1500 F). This would avoid the thermal penalty of cooling the fuel gas before introduction to a low-temperature gas cleanup process. The following processes represent the major areas of current interest and development in hot gas cleanup: iron oxide sorption, solid dolomite sorption, molten salt systems, and zinc oxide absorption.

Iron oxide sorption, which was developed by Morgantown Energy Research Center (MERC), was selected for CCMHD hot gas cleanup. It was judged to be at a more advanced stage of development than other high temperature processes. A description of this process follows.

4.2.2.1 Morgantown Iron Oxide Process

The MERC fired-bed iron oxide process has not been commercialized and therefore proven flow sheets for the process are not yet available. A conceptual flow sheet proposed for the MERC process is shown in Figure 4-8. Feed gas is shown entering at the top of the H_2S removal reactor and exiting at the bottom; the regeneration gas flows in the opposite direction. Periodic reversal of the flows may be required to prevent particulate buildup at the top of the reactor.

A potential improvement in the MERC process would be opeation in a moving-bed, continuous mode rather than in the cyclic, fixed-bed operation thus far demonstrated. One major advantage for attempting moving or fluidized bed operation is the superior temperature control possible

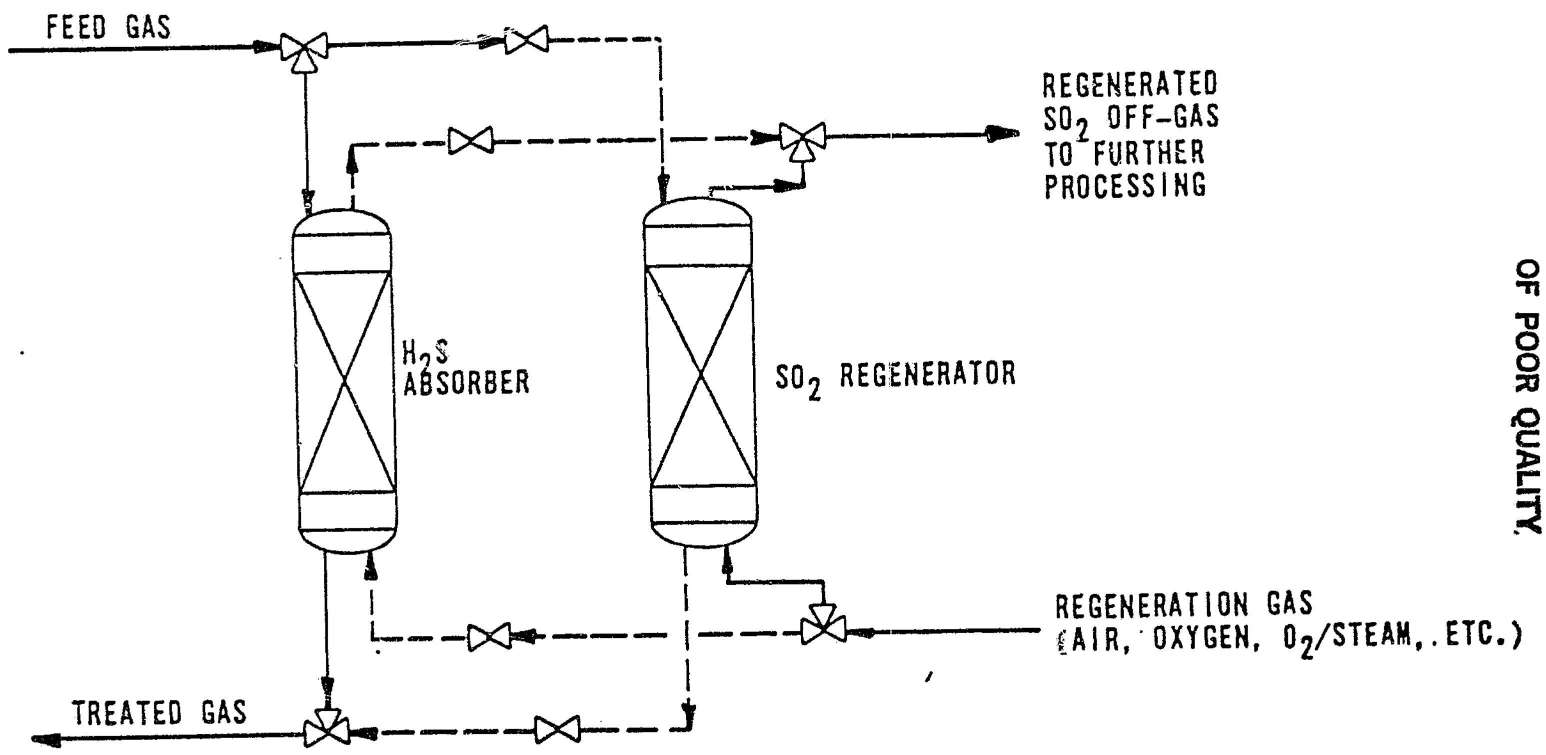

Figure 4-8

MORGANTOWN IRON OXIDE PROCESS

during regeneration. In addition, operation would be simplified by elimination of the manifolding and valves required for cyclic fixed-bed systems. The processing impediment to such operation is sorbent durability (e.g., minimal attrition and sorbent degradation required). This is especially difficult to achieve in a fluid-bed system; hence, extension of the MERC process to moving-bed operation is more compatible with sorbent strength.

The basic chemistry involved in the MERC process is sulfidation of iron oxide during absorption and reoxidation of the iron sulfide during regeneration. The proposed sequence of reactions representing this process begins with the fresh sorbent as Fe_2O_3. In the presence of hydrogen and temperatures above 650 F, the iron oxide is expected to be reduced to Fe_3O_4. Formation of further reduced species is also possible. The absorption reaction can then be represented as:

$$Fe_3O_4 + 3H_2S + H_2 \rightarrow 3FeS + 4H_2O$$

Regeneration of the sulfided iron oxide can be represented by:

$$3FeS + 5O_2 \rightarrow Fe_3O_4 + 3SO_2$$

A stoichiometric excess of oxygen then leads to further oxidation of the iron oxide to Fe_2O_3. Basing the overall reactions on Fe_2O_3, the absorption and regeneration steps can be represented by:

$$Fe_2O_3 + 2H_2S + H_2 \rightarrow 2FeS + 3H_2O,$$

$$2FeS + 7/2\ O_2 \rightarrow Fe_2O_3 + 2SO_2\ .$$

Other reactions may also occur in the process. For example, the water-gas shift reaction

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

is known to occur under absorber conditions.

The hot iron oxide reactor would be exposed to both corrosive-reducing and corrosive-oxidizing atmospheres at high temperature. Use of refractory lining, stainless steel weld overlays, and special alloys (such as 309SS, 310SS, 310SS aluminized, Incoloy 800, Incoloy 800 aluminized, and 18-8 austenitic stainless steels) will likely be required. Avoidance of aqueous phase corrosion by using proper startup and shutdown procedures is advisable in order to prevent additional materials problems. In addition, there is a critical need downstream of the hot gas cleanup system for efficient and low pressure drop filters to remove traces of alkali metals and other impurities.

A major problem with this process is the regeneration of off-gas, where air generation yields a dilute SO_2 stream. Because sulfur is presumed to be recovered in solid form, costly treating of the off-gas would be required. This is highly dependent on local market conditions.

The SO_2 off-gas from the hot iron oxide reactor requires further treating, and several options have been proposed. When a market is available, the off-gas is suitable feed for a sulfuric acid plant. Production of elemental sulfur is the other major option, and it requires a chemical reduction of the SO_2 in the off-gas. Studies by TVA and Stone & Webster proposed using the Allied Chemical Corporation's SO_2 reduction technology to reduce the off-gas SO_2 to elemental sulfur. This system requires a reducing gas for the reaction with SO_2; natural gas is used commercially. The TVA study assumed use of such natural gas, while the Stone & Webster study relied on Allied Chemical's assessment that use of a coal-derived fuel gas would be feasible.

For the CCMHD study, coal-derived fuel gas is also assumed to be the reducing gas. A portion of the sulfur dioxide in the feed gas is reacted with the CO/H_2 reductant in a fixed bed catalytic reactor, yielding a mixture of elemental sulfur, hydrogen sulfide and some unreacted SO_2, as well as CO_2 and water vapor:

$$2CO + SO_2 \rightarrow 2CO + S$$

$$2H_2 + SO_2 \rightarrow 2H_2O + S$$

$$3H_2 + SO_2 \rightarrow 2H_2O + H_2S$$

The H_2S and unreacted SO_2 from the reduction system are then reacted in a Claus plant to give additional elemental sulfur and water vapor:

$$2H_2S + SO_2 \rightarrow 2H_2O + 3S$$

4.2.3 In-Bed Cleanup

The Westinghouse two-stage gasification system employs dolomite in-bed desulfurization. In this fluidized bed system (Case 2.10), crushed dolomite (approx. 1.8 in) is dried and partially calcined in a heater. The partially calcined dolomite is fed into the top section of the desulfurizer where calcination proceeds to completion. The calcined dolomite reacts with sulfur compounds in the gas to form $(MgO)_x \cdot CaS$. The desulfurizer and the dolomite regenerator (if used) could be constructed of carbon steel with a refractory lining.

4.2.4 Flue Gas Desulfurization (FGD)

For those cases that use low sulfur coal for hot raw gas combustion, non-regenerable dry FGD appears to be applicable for meeting the new emission standard (70% sulfur removal). It is not recommended for high sulfur coal cases, since it appears to be economical only at low SO_2 concentrations in flue gas. This is because the dry methods (dry injection, spray dryer) use expensive sorbents - lime, sodium carbonate, or such naturally occurring carbonates as Trona (a hydrous sodium carbonate) and nahcolite (sodium bicarbonate). Wet scrubbing, the most widely used FGD method, employs cheaper limestone, so it has an operating cost edge at high SO_2 removal levels.

4.3 Topping Cycle

The CCMHD topping cycle is a closed argon loop in which the following functions occur:

1. Argon heating in regenerative heat exchangers.
2. Cesium seeding of the argon stream.
3. Energy extraction in the channel.
4. Argon cool-down and seed recovery in downstream heat exchangers.
5. Purification of argon and cesium.

Brief descriptions of the argon heat exchanger system, the argon purification system and the cesium system are given below. The argon regenerators are similar to those proposed for OCMHD indirect-fired systems. The argon purification system and the cesium system both represent preconceptual designs. The argon loop components decribed herein are typical for all CCMHD cases. Discussion of the non-equilibrium channel performance is presented in Appendix A.

4.3.1 Argon Heat Exchanger System

The argon heat exchanger system consists of an array of several high temperature, regenerative, ceramic matrix heat exchangers. Heat from the direct combustion of coal is transferred to the argon gas in each regenerative ceramic matrix. The ceramic core of the heat exchanger is alternately heated by coal or coal-derived gas combustion products and cooled by argon that has been pressurized to about 10 atmospheres. The heat exchangers are cycled through the following operating modes: heating by combustion gases, purge by combustion gases, argon blowdown, and argon recovery.

The ceramic passages in the core of the heat exchangers are purged and evacuated after the heating cycle to minimize combustion gas carryover and eliminate contamination of the argon with impurities. Contamination of the argon reduces the degree of non-equilibrium ionization in the MHD generator because of inelastic collisons between the molecular species and free electrons. Recent experimental studies have shown that measured levels of combustion gas species in the argon plasma can be controlled to less than 100 ppm, which is less than the predicted level where degradation of non-equilibrium ionization is significant.

Argon is heated to a stagnation temperature of 3100 F in the CCMHD regenerative heat exchangers. The operating surface temperature limit of the alumina ceramic brick is 3350 F. This material temperature limit necessitates moderating the combustor flame temperature with flue gas recirculation. Design data for the CCMHD regenerative heat exchangers is given in Appendix B.

4.3.2 Argon Purification System

In order to minimize contamination of the argon, the regenerative heat exchanger system is evacuated to a low pressure by a vacuum pump system. Inevitably, residual combustion gas products will be picked up by the argon. To maintain the impurities at a low and constant level, an argon purification system is proposed. Since the argon remaining in the heater after blowdown can not be economically vented to the atmosphere, it is purged into the argon purifiers and returned to the compressor inlet.

A detailed design of the purification system has not been performed in this study. Among the purification systems considered were a cryogenic system which would condense the combustion impurities, an adsorption system (activated charcoal and molecular sieve), and a getter system. None of these systems will remove all of the expected impurities, so a combination of these systems would probably be required. An adsorption system was conceptualized in this study for cost estimation purposes.

4.3.3 Cesium System

As the plasma leaves the MHD diffuser and is cooled in the heat recovery section, the 0.1% cesium seed is condensed and separated from the argon

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in the downstream components. These downstream heat exchangers are conceptualized to be of the shell and tube type, with the cesium condensing on the outside of the tubes and collected at the bottom of the heat exchanger as a liquid. After collection, the cesium enters a liquid metal pump and is pressurized to about 10 atmospheres; the cesium is then cleaned in a purifier and injected through a series of spray nozzles into the argon leaving the regenerative heat exchanger.

Although argon is inert with respect to cesium, reactions are possible between the cesium and residual combustion gas impurities that can be picked up by the argon in the regenerative heat exchanger. While the contamination level is expected to be small, it is necessary to reprocess the cesium to prevent a contaminating accumulation that can occur in a closed system. The cesium cleanup system is not well defined at this time, but conceptually it would consist of a combination of mechanical filtration and chemical reaction vessels.

4.4 Coal Handling And Drying

Coal handling and drying information received from MHD contractors was evaluated. This information, plus that from published technical papers, forms the basis for the following summary of key findings.

1. Moisture must be removed from the selected Rosebud Montana coal (from 25.5% on "as received" basis to 10% moisture by weight) to facilitate pulverizing and to improve combustion efficiency. Additional drying of the reference coal from 10% to 2% by weight moisture improves the efficiency of the MHD cycle by 0.3%.

2. Using projected drying techniques, the reference coal can be dried to 2% by weight with dryer gas inlet temperatures of about 800 F (700 K). There is more commercial experience with surface moisture removal than with bed moisture removal. The coal will be ground to 70% through 200 mesh. The hardgrove grindabilities of the reference coal are 52 with 25% moisture and 60 with 10% moisture.

3. Of the three major methods of coal drying, namely steam drying, oil drying and thermal drying, thermal drying is presently the moist economical process.

4. Use of low moisture coal improves flame temperature and allows the use of shorter channels for the same power extraction. Drying coal from 10% to 2% moisture by weight would increase the flame temperature from 25 F (14 K) to 75 F (42 K), depending upon the specific oxidant conditions.

5. Western coal is typically dried to 5% moisture and Eastern coal to about 2% moisture. Eastern coals can be dried more efficiencly than Western coal because they are more dense, have less moisture, and a higher percent of the total moisture exists as surface moisture. Approximately 2.5 to 5% of the coal thermal input (1 - 2 efficiency points) is used in drying coal.

6. To avoid fires at atmospheric operating pressure, the oxygen content by volume in the pulverizer inert gas should not be more than 6% with N_2 or 9% with CO_2.

7. About 2 to 3% by weight of the incoming coal is expected to be removed by the flue gas used for drying. The flue gas temperature leaving the coal dryer should be about 280 F (411K). To avoid the formation of sulfuric acid in the downstream components, to control the emission levels of SOx and NOx, and to maintain minimum thermal loss to the stack, the final gas temperature after mixing of the coal dryer gas should be approximately 220 F.

Drying of Western coal to 5% moisture by weight is practical, but drying to 2% by weight is questionable. Drying equipment will add complexity to the system and increase the auxiliary power required. The performance of the coal drying system will affect the performance of the complete system; however, if either 4% or 6% rather than 5% moisture is obtained, system performance will not be strongly effected.

Thermal drying of coal can be achieved by tapping the flue gas from downstream of the regenerative heat exchanger, downstream of the electrostatic precipitator or by mixing the two gas streams from downstream of the superheater and main air compressor gas turbine. The first method seems to be most appropriate for MHD plants because it utilizes the minimum auxiliary components and it requires the minimum amount of flue gas per pound of coal. This is the method selected by Gilbert Associates in all of the CCMHD cases evaluated.

According to vendors, drying equipment to dry coal to low moisture levels is available. Whether drying to 2% moisture will result in additional coal heat loss (2 to 3% of the heating value) is still unresolved. If this loss occurs, then drying coal to 2% moisture would not serve the purpose of obtaining higher flame temperature (with constant fuel rate).

The three major methods of coal drying are thermal drying, steam drying and oil drying. The selection of a dryer involves a careful evaluation of cost, types of processes (continuous or batch) and desired residence time.

4.4.1 Thermal Drying

The different types of thermal dryers are rotary, fluid bed, suspension and continuous tray type. Based on a drying efficiency of about 65%, the average energy required to remove one pound of moisture is about 1700 Btu. Drying gas is usually taken from the main process stream. Downstream of the pulverizer, pulverized coal is separated from the gas stream by a combination of separation equipment such as cyclones and bag houses. The dry gas is then mixed with the main gas stream. Though this process requires higher energy than the two processes described below, it is a continuous process and utilizes the low level heat present in the MHD system.

Thermal drying was selected by Gilbert for all CCMHD cases.

4.4.2 Steam Drying

The steam drying process involves heating the coal with saturated steam until the lumps are heated through to their centers. The pressure is then released and a vacuum is applied which cools the coal through evaporation of moisture. This process is repeated. Part of the natural bed moisture is forced from the coal as a liquid under the conditions imposed by heating with saturated steam. The thermal requirements for steam drying

(750 to 950 Btu/lb of moisture removed) are less than that for flue gas drying because this mechanical drying does not require latent heat of vaporization (970 Btu/lb) to remove the liquid water. A major advantage of the steam drying is that less particle degradation occurs than with thermal drying.

The amount of steam required to remove a unit quantity of water from coal decreases as the total amount of water increases. With 400 psig saturated steam, about 0.773 lb of steam per lb of water removed is required. The ratio of corresponding weights of raw coal to dry coal virtually equals the ratios of their gross heating values from volatile matter and fixed carbon. This ratio is called the "improvement ratio." Commercial experience with steam drying exists in Austria, Czechoslovakia and Hungary. Steam drying has not attained commercial status in the United States, probably because it is a batch process, which is more expensive and cumbersome than continuous processes.

4.4.3 Oil Drying

Oil drying involves heating low rank coal in an oil batch to temperatures between 428 F (493 K) to 752 F (673 K), which causes the coal to release its water and produces an oil soaked product. The energy consumption for oil dehydration will be the same as thermal drying because water must still be evaporated from the interior of the coal particle. The high price of oil and the requirement of a batch process are major disadvantages of this process.

5.0 DEVELOPMENT ISSUES

The scope of this study includes only the development issues specific to closed cycle MHD. The following list of requirements is pertinent to CCMHD technology.

1. Design, fabrication, performance and transient control of the combustor, channel and magnet.

2. Scale-up verification of the MHD power system.

3. Selection, development and utilization of diagnostic equipment.

4. Design and operating behavior of a high temperature regenerative preheat system utilizing coal combustion gases.

5. Design and installation of the channel cooling system.

6. Electrical isolation of the channel.

7. Compliance with New Source Performance Standards with regard to atmospheric emissions.

5.1 State-of-the-Art Technology

5.1.1 High Temperature Heater

At the present time, regenerative heat exchangers are used with clean combustion products in the steel industry to heat combustion air to 2450 F at 95 psia for blast furnace applications. The heat transfer matrix is primarily alumina. The burner is generally installed in one leg of a U tube configuration; the heated gas flows across the dome and down through the flues in the ceramic matrix.

In a direct-fired closed cycle MHD application, the combustion process results in a dirty gas; the valving designs required to sequence hot gas flow for these requirements have not yet been demonstrated. Additionally, high purity of the argon working fluid must be maintained through the generator. Argon contamination from hot side deposits or surface contact would penalize performance. Another problem involves the heat transfer mechanism. Absence of radiation heat transfer will reduce overall heat transfer to the argon, and this effect will be most pronounced at the high temperature, top portion of the exchanger. This condition results in large, expensive heaters that produce a negative economic impact on CCMHD.

5.1.2 MHD Generator

Open cycle MHD generators have been run at AVCO for many continous hours under realistic operating conditions. The Arnold Engineering Development Center (AEDC) is operating a channel which will verify the high enthalpy extraction, in the range of 20%, essential for commercial operation. The Soviets, using natural gas as fuel, have extracted power levels above 12 MWe from their U-25 channel installation.

Closed cycle channels use a clean monatomic working fluid (argon). Very high enthalpy extraction ratios and corresponding high power densities are expected. Durability of electrodes and insulation at high current densities and voltage gradients must be demonstrated to accomplish these feats. The planned CCMHD generator experiments currently planned at the University of Technology, Eindhoven, Netherlands should provide valuable information on the design of closed cycle channels.

5.1.3 Steam Generators

Firing steam boilers with gases containing the combination of combustion products, alkali salts, high temperatures and high heat fluxes required in open cycle MHD applications have not been demonstrated. Present advances in boiler design, and specialized designs such as the "black liquor" boilers used in paper processing, support the belief that adequate steam generating subsystems can be provided. This is supported by the short duration tests performed at the University of Tennessee Space Institute.

Closed cycle applications present the problem of having gaseous or liquid alkali metal on the steam generator gas side; temperatures are compatible with conventional plant operation. Argon, the hot gas working fluid, will provide only convective heat transfer. Heat transfer rates will have to be verified. Effects of minimal radiation heat transfer on the argon side could be mitigated, if necessary, by using a secondary combustion loop for boiling and auxiliary superheating or reheating. The secondary loop boiler would have oal combustion products, but no seed; however, designs will be analogous to conventional coal fired boilers.

5.1.4 Compressor

MHD applications, even in demonstration sizes of 500 MWt, utilize air flow rates and compression ratios in excess of those commercially available for axial compressors. For a given system power rating, use of argon will result in a several fold increase in flow rates. However, by use of multiple compressor units, which is normal practice, and reasonable extension of present designs, systems could be provided with minimal development risk. Use of argon requires blade modifications and adequate sealing, but neither of these requirements is expected to impose significant penalties. Present compressor design temperature limits of about 450 - 500 F could be extended to 600 F with minor modifications to off-the-shelf designs.

5.2 Development Needs

The development of closed cycle components and subsystems have not progressed to the level of open cycle systems. The emphasis on MHD development has definitely been placed on open cycle systems. For this reason, the development needs of the components comprising the closed cycle system are not well defined at this time. Fortunately, many of the components are similar or equivalent to those used in an open system. In this section, only those major components that are judged significantly different from those in an open cycle MHD system are included.

5.2.1 High Temperature Argon Heater

The following sections consider development needs for two types of heaters; those fired directly with coal and those fired with the product of a coal gasifier.

5.2.1.1 Coal Fired Heaters

Development work is needed in four areas to show technological feasibility and to develop the data base for design. These areas are (1) ceramic materials, (2) operability of the heater system, i.e. preventing clogging of the heater passages with slag, (3) bed support, and (4) valves.

Materials: The heater bed, hot gas inlet ducts and manifolds, upper vessel dome, vessel liner, and lower vessel plenum must be constructed of a material which will resist corrosion/erosion by molten slag at temperatures up to 3370 F, as well as being cost effective with acceptabe mechanical properties. The major problem is the very high temperatures. Operating experience and test data at these conditions is very limited.

Chrome-bearing, fused cast refractories usually have the best slag resistant properties at very high service temperatures. However, they are expensive and have poor thermal shock resistance. Preliminary tests of high alumina materials have shown some swelling due to slag pick-up, but continuing tests at Montana State University indicate that this material has considerble potential. Further tests at subscale and with various slags are needed. Also needed is a suitable method of anchoring the castable liner to the castable back-up layers and further tests at subscale.

Operability: Slag present in the hot gas stream will accumulate in the argon heaters and asociated ducting. A clean-out procedure will be required in which the slag is melted and flows out of the heaters. The frequency and duration of the clean-out cycle must be determined, and will be dependent on many factors including the specific ash characteristics of the coal.

Preliminary tests with a subscale heater (20 ft high bed) are being done at Montana State University. Additional testing at this scale would be needed, followed by tests at a larger scale in order to progressively move to the commercial plant size.

Bed Support: A bed support is required which will endure temperatures up to free flowing slag temperatures while not obstructing the slag flow. Development work is needed to determine the accumulation of slag on cooled surfaces during the clean-out cycle, materials of construction, and overall structural integrity. Tests at or near full-scale will be needed to verify satisfactory operation during both normal and clean-out thermal cycling. The use of fluxing additives to reduce the slag melting point and viscosity would also need to be tested.

Valves: Six valves are required for the operation of each heater in the regenerative heater system. These are: combustion products inlet and outlet, argon inlet and outlet, and two smaller valves to accommodate fluid changeover. The combustion gas inlet valve has the most severe service. This valve will require some development and the others will require verification testing.

5.2.1.2 Clean Fired Heaters

The development needs for clean-fired argon heaters are significantly reduced when compared to the direct coal-fired heaters, the most similar industrial equipment being blast furnace stoves. Significant differences are the higher temperature, higher thermal effectiveness, larger physical size, and the operating requirements associated with electric power generation.

Measurements of high temperature creep are needed to assure satisfactory life of the ceramics. Tests of blast furnace valves at high temperatures are needed. Tests of the proposed castable insulation are needed. The castable insulation could be replaced with bricks. This would increase the cost of Case 2.0 (1000 MWe) by about 15 million dollars.

5.2.2 Combustor

As with the high temperature argon heater, two types of combustors are required; those fired directly with coal and those fired with the product of a coal gasifier.

5.2.2.1 Coal Fired Combustors

Because it is to be run at a much lower temperature, development of the combustor for a CCMHD system is expected to be much easier than its OCMHD counterpart. This combustor, however, is not state-of-the-art. Coal injection schemes, diagnostic techniques, insulation, cooling schemes and techniques for maximizing flame stability and minimizing heat loss must be developed. Scaleup from the current 20 MWt development size to the required 500 MWt size will be required in at least 5:1 steps.

5.2.2.2 Gas Fired Combustors

Gas fired combustors are also not state-of-the-art but are not expected to be a significant development problem. The major problems will be minimizing heat loss and guaranteeing complete combustion.

5.2.3 Coal Gasifiers

Coal gasifiers of two types were considered in this study; low pressure and high pressure. The development problems related to each as discussed in the following sections.

5.2.3.1 Low Pressure Gasifiers

The present fixed bed low pressure gasfier has a limitation in its ability to handle caking coals. A mechanical stirrer or slagging fixed bed gasifier is to be tested with strongly caking coals to study feasibility of removing this limitation. The lock hopper dry coal feeder presently used commercially with the fixed bed gasifier is found to be expensive. A screw feeder which has a good potential for application at low pressure appears to be a good candidate as a dry coal feeder, if particle size limitations can be overcome.

For the entrained bed systems, use of dry solid feed (lock hopper) systems are under investigation; the use of screw feeders for this low pressure application should be given high priority due to its potential for improved economics.

Presently, there are no fluidized bed gasifiers commercially available that operate at pressures above atmospheric. This gasifier is also limited to non-caking coal applications. In addition to development of a low pressure dry coal feeder for use with the fluidized bed, it is also necessary to provide a means of pretreating caking coals at low pressure before its introduction to the gasifier to eliminate agglomerating characteristics of coal.

For all three gasifiers considered above, it will also be necessary to develop suitable burners for use with low/medium Btu gas.

5.2.3.2 High Pressure Gasifiers

Because of the high pressure, only the entrained bed and fluidized bed systems can be considered. Entrained bed systems require the development of lock hoppers, a slurry feed system, an extrusion feed system and a gas cleanup system. Fluidized bed gasifiers require development of the above plus a coal pretreatment system.

5.2.4 MHD Generator

A closed cycle MHD generator depends on the concept of non-equilibrium ionization, where the electron temperature is elevated above the gas temperature. The plasma flowing through the generator is argon seeded with cesium and is essentially slag free since combustion is external to the argon system. In one respect the absence of slag in the working fluid simplifies the channel design; but in another sense the problem of heat transfer is more critical since a slag layer acts as a thermal insulation. A major problem in the closed cycle generator is maintaining the necessary level of non-equilibrium ionization in a plasma, which can be highly turbulent and unstable. Arcing between adjacent electrodes through the interelectrode insulating material and the boundary layer and arch discharges across the channel remain an unresolved question.

The planned closed cycle generator tests at the University of Technology, Eindhoven, Netherlands should clearify the development requirements of the generator.

5.2.5 Steam Generator

The design of the steam generators does not impose a significant development problem. The temperatures are similar to those encountered in conventional fossil power plants. The differences are due to the corrosive nature of the cesium seed and the need to varify the surface area design for an argon side working substance where the heat transfer mode is almost entirely convective.

5.2.6 Argon Compressor

No significant development problems are anticipated for an argon compressor to satisfy the requirements for this specific application. Even through the required flow rates and compression ratios for this application exceed present commercially available compressor capabilities, the design of an argon system should be no more difficult than for air.

Scaling to larger sizes is more an economic problem than a technical problem. Development costs are expected to be relatively low and would include checkout of blading angles and proper selection of materials.

6.0 ESTIMATED COSTS

6.1 Capital Cost Estimates

Estimates of capital cost for the power plant cases studied were based on a combination of scaling procedures, vendor quotations and engineering estimates. Capital cost estimates are presented in the ETF (DOE/MHD) Code of Accounts format as modified by NASA-LeRC for closed cycle MHD power plants. All economic parameters used in estimating capital costs are consistent with those stipulated in Section 1.5.

The primary data source used to estimate costs of the various closed cycle MHD plants was the Cost Estimating Procedure (CEP) developed by GAI for DOE/MHD. The CEP consists of cost equations for the accounts/subaccounts as defined in the DOE/MHD Code of Accounts. Cost equations are in the form, $C = KM^x$, where:

C = cost, \$ x 10^3 (mid-1978)

K = derived constant

M = power plant rating, MWt

x = scaling exponent

Costs calculated from the CEP equations are total installed costs (TIC) and are based on mature technology. The cost elements contributing to the total installed cost displayed in the DOE/MHD Code of Accounts format (component cost, installation cost, etc.) are contained within the CEP equations. For each account or subaccount, however, there is a different relationship for individual cost elements as a percentage of the total installed cost. Component costs, for instance, could represent either 20% or 70% of the total installed cost. Consequently, there is no viable method available for extracting the cost elements from the total installed cost. All costs presented in the Code of Accounts summaries are therefore given as total costs only, except where specific cost breakdowns are available.

The cost basis for all CEP cost equations is power plant thermal input. Since the contents of closed cycle accounts are not always consistent with the open cycle accounts used in the CEP model, many CEP cost estimates required adjustment in order to be representative of closed cycle account costs. These adjustments were made through analytical procedures involving re-sizing or re-configuration of components as dictated by flow, pressure or other parametric requirements.

Cost estimates obtained from sources other than the CEP or other cost equations fall into four categories: engineering estimate, literature cost data, contractor cost data or vendor cost data. Engineering estimates refer to GAI cost estimates based on either conceptual designs or previous cost estimates for similar items. Literature cost data generally includes information found in trade journals or technical publications. Contractor cost data refers only to information available from MHD reports (ETF, PSPEC, etc.), while vendor cost data includes both specific quotations (e.g., FluiDyne heat exchangers) or costs scaled from previous quotations.
The cost basis column in the Code of Accounts cost summaries was used to indicate the source of the cost estimate for each account or subaccount

(see Table 6-1). Accounts having more than one cost source either contain multiple subsystems or components that required individual cost analysis.

As shown in the Code of Account cost summaries, addition of the total cost columns plus a 10% charge for Engineering Services results in the Total Estimated Cost, or Overnight Construction Cost (OCC). The Total Capital Cost (TCC) is obtained by applying the interest and escalation multiplier (cost factor) to the OCC. The cost factor is a function of the design and construction period for each power plant based on fixed interest and escalation rates of 10% and 6.5%, respectively. A period of 6 1/2 years from the start of design to the end of construction was estimated for the closed cycle MHD plants costed in this study, resulting in a cost factor of 1.679.

Capital costs for the closed cycle MHD cases studied are given in Tables 6-2 through 6-6. For reference purposes, lists of equipment for Cases 1 and 2 are given in Appendix C.

6.2 Cost of Electricity

Cost of electricity (COE) calculations were based on two methodologies: levelized COE (LEV) and escalated levelized (LEV') COE. In addition, COE's based on the ECAS method of calculation were compiled since the baseline COE values for capital, fuel and O&M used to compute LEV are used to calculate the "ECAS" COE. Levelized COE's are based on a levelizing factor of 2.004. The escalated levelized COE's represent an exercise in fuel cost sensitivity and were calculated for real fuel escalation rates of 1, 2 and 3 percent. In addition, COE's were calculated based

Table 6-1

Code of Accounts Cost Basis Code

Code Number	Cost Source
1	CEP* cost equation
2	CEP and/or CCDB # equations, adjusted by analysis
3	Engineering estimate
4	Literature cost data, scaled
5	Contractor cost data, scaled
6	Vendor quotation
7	Vendor cost data, scaled

* Cost Estimation Procedure

\# Component Cost Data Bank

DOE CODE OF ACCOUNTS

CLOSED CYCLE MHD PLANTS

Mid 1978 Dollars x 10^{-3}

CASE 1.0

Account No.	Account Description	Cost Basis Code	Quantity	Material and Installation Cost	Indirect Cost and Contingency	Total Cost
310.0	Land and Land Rights	1	N.A.			1100
311.0	Structures and Improvements	1	N.A.			55320
311.1	Improvements to Site	-	-			-
311.2	Main Building	-	-			-
311.3	Steam Turbine and MHD Compressor	-	-			-
311.4	Coal Bunker and Processing Area	-	-			-
311.5	Service Buildings	-	-			-
311.6	Other Buildings and Structures	-	-			-
312.0	Steam Boiler Equipment	-	-			123595
312.1	Account Moved to 313.1	-	-			-
312.2	Account Moved to 313.5	-	-			-
312.3	Account Not Used	-	-			-
312.4	Steam Generation Sections	2,3	1			102238
312.5	Account Moved to 313.6	-	-			-
312.6	Auxiliary Boiler Systems	1	1			2983
312.7	Other Boiler Plant Systems	1	1			18374
313.0	Combustion System Equipment	-	-			99173
313.1	Coal Handling and Processing	1	1			19610
313.2	Combustor Subsystem	2,6	1			28158
313.3	Coal Gasifier Subsystem	-	-			-
313.4	Air/Oxidizer System	2,7	1			4720
313.5	Slag and Ash Handling	4	1			31500
313.6	Effluent Control	4,5	1			
314.0	Turbocompressor, Turbogenerator	2,5	1			93280
314.1	Argon Compressors, Steam Turbines	-	-			-
314.2	Condenser and Auxiliaries	-	-			-
314.3	Circulating Water Systems	-	-			-
314.4	Steam Piping Systems	-	-			-
314.5	Other Turbine Plant Equipment	-	-			-

DOE CODE OF ACCOUNTS

CLOSED CYCLE MHD PLANTS

Mid 1978 Dollars x 10^{-3}

CASE 1.0

Account No.	Account Description	Cost Basis Code	Quantity	Material and Installation Cost	Indirect Cost and Contingency	Total Cost
315.0	Accessory Electric Equipment	1	1			37422
316.0	Miscellaneous Power Plant Equip.	1	1			6173
317.0	MHD Topping Cycle Equipment	-	-			456830
317.1	Account Moved to 313	-	-			-
317.2	MHD Generator Subsystem	2,3	1			16370
317.3	Magnet Subsystem	5	1			61824
317.4	Inverters and Electrode Controls	2	1			54568
317.5	Regenerative Heat Exchanger Subsys.	2,6	16	243818	73146	316964
317.6	Seed Subsystem	3	1			201
317.7	Other MHD Topping Cycle Support	2,3	1			6883
318.0	Account Not Used	-	-			-
319.0	Account Not Used	-	-			-
350.0	Transmission Plant	1	1			6412
	Subtotal - Direct Accounts					877305
	Engineering Services					87931
	Total Estimated Cost					961236

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DOE CODE OF ACCOUNTS

CLOSED CYCLE MHD PLANTS — Mid 1978 Dollars x 10^{-3}

CASE 2.0

Account No.	Account Description	Cost Basis Code	Quantity	Material and Installation Cost	Indirect Cost and Contingency	Total Cost
310.0	Land and Land Rights	1	N.A.			1203
311.0	Structures and Improvements	1	N.A.			62176
311.1	Improvements to Site	-	-			-
311.2	Main Building	-	-			-
311.3	Steam Turbine and MHD Compressor	-	-			-
311.4	Coal Bunker and Processing Area	-	-			-
311.5	Service Buildings	-	-			-
311.6	Other Buildings and Structures	-	-			-
312.0	Steam Boiler Equipment	-	-			132793
312.1	Account Moved to 313.1	-	-			-
312.2	Account Moved to 313.5	-	-			-
312.3	Account Not Used	-	-			-
312.4	Steam Generation Sections	2,3	1			110408
312.5	Account Moved to 313.6	-	-			-
312.6	Auxiliary Boiler Systems	1	1			3105
312.7	Other Boiler Plant Systems	1	1			19280
313.0	Combustion System Equipment	-	-			279320
313.1	Coal Handling and Processing	1	1			21880
313.2	Combustor Subsystem	2	1			14551
313.3	Coal Gasifier Subsystem	3	1			218303
313.4	Air/Oxidizer System	5	1			10916
313.5	Slag and Ash Handling	-	-			-
313.6	Effluent Control	5	1			7670
314.0	Turbocompressor, Turbogenerator	2,5,6	1			133430
314.1	Argon Compressors, Steam Turbines	-	-			-
314.2	Condenser and Auxiliaries	-	-			-
314.3	Circulating Water Systems	-	-			-
314.4	Steam Piping Systems	-	-			-
314.5	Other Turbine Plant Equipment	-	-			-

Table 6-3 (continued)

DOE CODE OF ACCOUNTS

CLOSED CYCLE MHD PLANTS

Mid 1978 Dollars x 10^{-3}

CASE 2.0

Account No.	Account Description	Cost Basis Code	Quantity	Material and Installation Cost	Indirect Cost and Contingency	Total Cost
315.0	Accessory Electric Equipment	1	1			37330
316.0	Miscellaneous Power Plant Equip.	1	1			6400
317.0	MHD Topping Cycle Equipment	-	-			210244
317.1	Account Moved to 313	-	-			-
317.2	MHD Generator Subsystem	2,3	1			16370
317.3	Magnet Subsystem	5	1			61824
317.4	Inverters and Electrode Controls	2	1			58646
317.5	Regenerative Heat Exchanger Subsys.	6	16	51000	15300	66300
317.6	Seed Subsystem	3	1			201
317.7	Other MHD Topping Cycle Support	2,3	1			6883
318.0	Account Not Used	-	-			-
319.0	Account Not Used	-	-			-
350.0	Transmission Plant	1	1			6609
	Subtotal - Direct Accounts					871565
	Engineering Services					87157
	Total Estimated Cost					958722

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Table 6-4

DOE CODE OF ACCOUNTS

CLOSED CYCLE MHD PLANTS

Mid 1978 Dollars x 10^{-3}

CASE 2.10

Account No.	Account Description	Cost Basis Code	Quantity	Material and Installation Cost	Indirect Cost and Contingency	Total Cost
310.0	Land and Land Rights	1	N.A.			1203
311.0	Structures and Improvements	1	N.A.			62116
311.1	Improvements to Site	-	-			-
311.2	Main Building	-	-			-
311.3	Steam Turbine and MHD Compressor	-	-			-
311.4	Coal Bunker and Processing Area	-	-			-
311.5	Service Buildings	-	-			-
311.6	Other Buildings and Structures	-	-			-
312.0	Steam Boiler Equipment	-	-			128702
312.1	Account Moved to 313.1	-	-			-
312.2	Account Moved to 313.5	-	-			-
312.3	Account Not Used	-	-			-
312.4	Steam Generation Sections	2,3	1			105231
312.5	Account Moved to 313.6	-	-			-
312.6	Auxiliary Boiler Systems	1	1			3233
312.7	Other Boiler Plant Systems	1	1			20238
313.0	Combustion System Equipment	-	-			328021
313.1	Coal Handling and Processing	1	1			27780
313.2	Combustor Subsystem	2	1			15361
313.3	Coal Gasifier Subsystem	3	1			260113
313.4	Air/Oxidizer System	5	1			17485
313.5	Slag and Ash Handling	-	-			-
313.6	Effluent Control	5	1			8182
314.0	Turbocompressor, Turbogenerator	2,5,6	1			133430
314.1	Argon Compressors, Steam Turbines	-	-			-
314.2	Condenser and Auxiliaries	-	-			-
314.3	Circulating Water Systems	-	-			-
314.4	Steam Piping Systems	-	-			-
314.5	Other Turbine Plant Equipment	-	-			-

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DOE CODE OF ACCOUNTS

CLOSED CYCLE MHD PLANTS

Mid 1978 Dollars x 10^{-3}

CASE 2.10

Account No.	Account Description	Cost Basis Code	Quantity	Material and Installation Cost	Indirect Cost and Contingency	Total Cost
315.0	Accessory Electric Equipment	1	1			41350
316.0	Miscellaneous Power Plant Equip.	1	1			6637
317.0	MHD Topping Cycle Equipment	-	-			210244
317.1	Account Moved to 313	-	-			-
317.2	MHD Generator Subsystem	2,3	1			16370
317.3	Magnet Subsystem	5	1			61824
317.4	Inverters and Electrode Controls	2	1			58646
317.5	Regenerative Heat Exchanger Subsys.	6	16	51000	15300	66300
317.6	Seed Subsystem	3	1			201
317.7	Other MHD Topping Cycle Support	2,3	1			6883
318.0	Account Not Used	-	-			-
319.0	Account Not Used	-	-			-
350.0	Transmission Plant	1	1			6738
	Subtotal - Direct Accounts					919601
	Engineering Services					91960
	Total Estimated Cost					1011561

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Table 6-5

DOE CODE OF ACCOUNTS

CLOSED CYCLE MHD PLANTS

Mid 1978 Dollars x 10^{-3}

CASE 2.12

Account No.	Account Description	Cost Basis Code	Quantity	Material and Installation Cost	Indirect Cost and Contingency	Total Cost
310.0	Land and Land Rights	1	N.A.			1203
311.0	Structures and Improvements	1	N.A.			62176
311.1	Improvements to Site	-	-			-
311.2	Main Building	-	-			-
311.3	Steam Turbine and MHD Compressor	-	-			-
311.4	Coal Bunker and Processing Area	-	-			-
311.5	Service Buildings	-	-			-
311.6	Other Buildings and Structures	-	-			-
312.0	Steam Boiler Equipment	-	-			132773
312.1	Account Moved to 313.1	-	-			-
312.2	Account Moved to 313.5	-	-			-
312.3	Account Not Used	-	-			-
312.4	Steam Generation Sections	2,3	1			110408
312.5	Account Moved to 313.6	-	-			-
312.6	Auxiliary Boiler Systems	1	1			3105
312.7	Other Boiler Plant Systems	1	1			19280
313.0	Combustion System Equipment	-	-			384077
313.1	Coal Handling and Processing	1	1			21880
313.2	Combustor Subsystem	2	1			1212
313.3	Coal Gasifier Subsystem	3	1			332605
313.4	Air/Oxidizer System	5	1			20632
313.5	Slag and Ash Handling	-	-			-
313.6	Effluent Control	5	1			7670
314.0	Turbocompressor, Turbogenerator	2,5,6	1			147209
314.1	Argon Compressors, Steam Turbines	-	-			-
314.2	Condenser and Auxiliaries	-	-			-
314.3	Circulating Water Systems	-	-			-
314.4	Steam Piping Systems	-	-			-
314.5	Other Turbine Plant Equipment	-	-			-

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DOE CODE OF ACCOUNTS

CLOSED CYCLE MHD PLANTS

Mid 1978 Dollars x 10^{-3}

CASE 2.12

Account No.	Account Description	Cost Basis Code	Quantity	Material and Installation Cost	Indirect Cost and Contingency	Total Cost
315.0	Accessory Electric Equipment	1	1			39330
316.0	Miscellaneous Power Plant Equip.	1	1			6400
317.0	MHD Topping Cycle Equipment	-	-			217213
317.1	Account Moved to 313	-	-			-
317.2	MHD Generator Subsystem	2,3	1			15370
317.3	Magnet Subsystem	5	1			61824
317.4	Inverters and Electrode Controls	2	1			58646
317.5	Regenerative Heat Exchanger Subsys.	6	16	51000	15300	66300
317.6	Seed Subsystem	3	1			201
317.7	Other MHD Topping Cycle Support	2,3	1			13852
318.0	Account Not Used	-	-			-
319.0	Account Not Used	-	-			-
350.0	Transmission Plant					6667
	Subtotal - Direct Accounts					997070
	Engineering Services					99707
	Total Estimated Cost					1076777

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DOE CODE OF ACCOUNTS

CLOSED CYCLE MHD PLANTS

Mid 1978 Dollars x 10^{-3}

CASE 3.0

Account No.	Account Description	Cost Basis Code	Quantity	Material and Installation Cost	Indirect Cost and Contingency	Total Cost
310.0	Land and Land Rights	1	N.A.			1164
311.0	Structures and Improvements	1	N.A.			57187
311.1	Improvements to Site	-	-			-
311.2	Main Building	-	-			-
311.3	Steam Turbine and MHD Compressor	-	-			-
311.4	Coal Bunker and Processing Area	-	-			-
311.5	Service Buildings	-	-			-
311.6	Other Buildings and Structures	-	-			-
312.0	Steam Boiler Equipment	-	-			129790
312.1	Account Moved to 313.1	-	-			-
312.2	Account Moved to 313.5	-	-			-
312.3	Account Not Used	-	-			-
312.4	Steam Generation Sections	2,3	1			107549
312.5	Account Moved to 313.5	-	-			-
312.6	Auxiliary Boiler Systems	1	1			3003
312.7	Other Boiler Plant Systems	1	1			19118
313.0	Combustion System Equipment	-	-			183952
313.1	Coal Handling and Processing	1	1			21708
313.2	Combustor Subsystem	2	1			14415
313.3	Coal Gasifier Subsystem	3	1			132211
313.4	Air/Oxidizer System	5	1			1913
313.5	Slag and Ash Handling	-	-			-
313.6	Effluent Control	5	1			7585
314.0	Turbocompressor, Turbogenerator	2,5	1			96034
314.1	Argon Compressors, Steam Turbines	-	-			-
314.2	Condenser and Auxiliaries	-	-			-
314.3	Circulating Water Systems	-	-			-
314.4	Steam Piping Systems	-	-			-
314.5	Other Turbine Plant Equipment	-	-			-

DOE CODE OF ACCOUNTS

CLOSED CYCLE MHD PLANTS

Mid 1978 Dollars x 10^{-3}

CASE 3.0

Account No.	Account Description	Cost Basis Code	Quantity	Material and Installation Cost	Indirect Cost and Contingency	Total Cost
315.0	Accessory Electric Equipment	1	1			38985
316.0	Miscellaneous Power Plant Equip.	1	1			6359
317.0	MHD Topping Cycle Equipment	-	-			274064
317.1	Account Moved to 313	-	-			-
317.2	MHD Generator Subsystem	2,3	1			17125
317.3	Magnet Subsystem	5	1			61824
317.4	Inverters and Electrode Controls	2	1			57909
317.5	Regenerative Heat Exchanger Subsys.	6	20	100094	30028	130122
317.6	Seed Subsystem	3	1			201
317.7	Other MHD Topping Cycle Support	2,3	1			6883
318.0	Account Not Used	-	-			-
319.0	Account Not Used	-	-			-
350.0	Transmission Plant					6623
	Subtotal - Direct Accounts					794163
	Engineering Services					79416
	Total Estimated Cost					873577

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on a range of fuel and labor costs as specified by NASA-LeRC. Tables 6-7 through 6-10 give the COE's for the cases studied based on four conditions of fuel and labor costs - low fuel, low labor (baseline); high fuel only; high labor only; and high fuel, high labor.

6.3 Capital Cost Comparisons

CCMHD capital costs have been compared with those of a selected PSPEC case. The AVCO-PSPEC Case II-1 was selected as the basis of comparison since it uses both a) a regenerative high temperature heat exchanger (HTAH) and b) advanced design gasifier systems, which are common to four out of five of the CCMHD cases studied. Detailed comparisons of all individual accounts or subaccounts have not been made due to a lack of commonality caused by the modification of the CCMHD Code of Accounts and the dearth of detailed cost information for the CCMHD components and subsystems. Only major accounts or components, having similar characteristics, have been compared. The cost comparison rationale is presented below; comparison results are given in Table 6-11.

6.3.1 Cost Comparison Rationale

Comparison of capital costs involves three sets of cost data: individual cost accounts presented as the total installed cost (TIC) in the Code of Accounts format, overnight construction costs and total capital costs. Since the total capital costs are based on overnight construction cost, cost escalations related to MHD plant design and construction time factors, comparisons of absolute total installed cost values are not valid unless all plants have identical design and construction times. Design and

Table 6-7

SUMMARY OF COE's

FUEL = $\$1.05/10^6$ BTU

LABOR = $14.20/HR

COE METHOD	COE COMPONENT	Case 1.0	Case 2.0	Case 2.10	Case 2.12	Case 3.0
ECAS	Capital	51.26	50.23	50.79	53.03	46.57
	Fuel	8.29	9.10	8.94	9.85	9.92
	O&M	1.62	1.62	1.65	1.68	1.61
	Total	61.17	60.95	61.38	64.56	58.12
LEV (F-2.004)	Capital	34.04	33.36	33.73	35.22	30.74
	Fuel	16.61	18.24	17.92	19.74	19.88
	O&M	3.25	3.25	3.31	3.37	3.23
	Total	53.90	54.85	54.96	58.33	54.05
LEV' (1% escalation)	Capital	34.04	33.36	33.73	35.22	30.94
	Fuel	18.98	20.84	20.47	22.56	22.72
	O&M	3.25	3.25	3.31	3.37	3.23
	Total	56.27	57.45	57.51	61.15	56.89
LEV' (2% escalation)	Capital	34.04	33.96	33.73	35.22	30.94
	Fuel	21.84	23.97	23.55	25.94	26.13
	O&M	3.25	3.25	3.31	3.37	3.23
	Total	59.13	60.58	60.57	64.53	60.30
LEV' (3% escalation)	Capital	34.04	33.36	33.73	35.22	30.94
	Fuel	25.28	27.75	21.26	30.03	30.25
	O&M	3.25	3.75	3.31	3.37	3.23
	Total	62.57	64.36	64.30	68.62	64.42

Table 6-8

SUMMARY OF COE's

FUEL = $\$1.05/10^6$ BTU

LABOR = $17.04/HR

COE METHOD	COE COMPONENT	Case 1.0	Case 2.0	Case 2.10	Case 2.12	Case 3.0
ECAS	Capital	51.26	50.23	50.79	53.03	46.57
	Fuel	8.29	9.10	8.94	9.85	9.92
	O&M	1.94	1.95	1.98	2.01	1.93
	Total	61.49	61.28	61.71	64.89	58.44
LEV	Capital	34.04	33.36	33.73	35.22	30.94
	Fuel	16.61	18.24	17.92	19.74	19.88
(F-2.004)	O&M	3.89	3.91	3.97	4.03	3.87
	Total	54.54	55.51	62.04	58.99	54.69
LEV'	Capital	34.04	33.36	33.73	35.22	30.94
	Fuel	18.98	20.84	20.47	22.56	22.72
(1% escalation)	O&M	3.89	3.91	3.97	4.03	3.87
	Total	56.91	58.11	58.17	61.81	57.53
LEV'	Capital	34.04	33.36	33.73	35.22	30.94
	Fuel	21.84	23.97	23.55	25.94	26.13
(2% escalation)	O&M	3.89	3.91	3.97	4.03	3.87
	Total	59.77	61.24	61.25	65.19	60.94
LEV'	Capital	34.04	33.36	33.73	35.22	30.94
	Fuel	25.28	27.75	27.26	30.03	30.25
(3% escalation)	O&M	3.89	3.91	3.97	4.03	3.87
	Total	63.21	65.02	64.96	69.28	65.06

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Table 6-9

SUMMARY OF COE's

FUEL = $ 1.50/10^6 BTU

LABOR = $ 17.04/HR

COE METHOD	COE COMPONENT	Case 1.0	Case 2.0	Case 2.10	Case 2.12	Case 3.0
ECAS	Capital	51.26	50.23	50.79	53.03	46.59
	Fuel	11.84	13.00	12.77	14.07	14.17
	O&M	1.95	1.95	1.98	2.01	1.93
	Total	65.04	65.18	65.54	69.11	62.69
LEV	Capital	34.04	33.36	33.73	35.22	30.94
	Fuel	23.73	[illegible]	25.59	28.20	28.40
(F-2.004)	O&M	3.89	[illegible]	3.97	4.03	3.87
	Total	61.66	63.32	63.29	67.45	63.21
LEV'	Capital	34.04	33.36	33.73	35.22	30.94
	Fuel	27.11	29.77	29.24	32.22	32.45
(1% escalation)	O&M	3.89	3.91	3.97	4.03	3.87
	Total	65.04	67.04	66.94	71.47	67.26
LEV'	Capital	34.04	33.36	33.73	35.22	30.94
	Fuel	31.17	34.24	33.64	27.06	37.32
(2% escalation)	O&M	3.87	3.91	3.77	4.03	3.87
	Total	69.12	71.51	71.34	76.31	72.13
LEV'	Capital	34.04	33.36	33.73	35.22	30.94
	Fuel	36.10	39.64	38.94	42.90	43.20
(3% escalation)	O&M	3.89	3.91	3.97	4.03	3.87
	Total	74.03	76.91	76.64	82.15	78.01

Table 6-10

SUMMARY OF COE's

FUEL = $1.50/10^6 BTU

LABOR = $14.20/HR

COE METHOD	COE COMPONENT	Case 1.0	Case 2.0	Case 2.10	Case 2.12	Case 3.0
ECAS	Capital	51.26	50.23	50.79	53.03	46.57
	Fuel	11.84	13.00	12.77	14.07	14.17
	O&M	1.62	1.62	1.65	1.68	1.61
	Total	64.72	64.85	65.21	68.78	62.37
LEV	Capital	34.04	33.36	33.73	35.22	30.74
	Fuel	23.73	26.05	25.59	28.20	28.40
(F-2.004)	O&M	3.25	3.25	3.31	3.37	3.23
	Total	61.02	62.66	62.63	66.79	62.57
LEV'	Capital	34.04	33.36	33.73	35.22	30.94
	Fuel	27.11	29.77	29.24	32.22	32.45
(1% escalation)	O&M	3.25	3.25	3.31	3.37	3.23
	Total	64.40	66.38	66.28	70.81	66.62
LEV'	Capital	34.04	33.36	33.73	35.22	30.74
	Fuel	31.19	34.24	33.64	37.06	37.32
(2% escalation)	O&M	3.25	3.25	3.31	3.37	3.23
	Total	68.48	70.85	70.68	75.65	71.49
LEV'	Capital	34.04	33.36	33.73	35.22	30.94
	Fuel	36.10	39.64	38.94	42.90	43.20
(3% escalation)	O&M	3.25	3.25	3.31	3.37	3.23
	Total	78.39	76.25	75.98	81.49	77.37

Table 6-11

Comparison of Capital Costs

	AVCO-PSPEC Case II-1 1034 MWe	Case 1.0 1001 MWe	Case 2.0 1012 MWe	Case 210 1056 MWe	Case 2.12 1097 MWe	Case 3.0 995 MWe
Overnight Const. Cost (OCC)	$ 725,949	$ 967,236	$ 958,722	$ 1,011,561	$1,096,777	$ 873,579
Δ From AVCO-PSPEC		+241,287	+232,773	+285,612	+370,828	+147,630
Major Accounts Cost						
Regen. HX		316,964	66,300	66,300	66,300	130,122
Gasifiers		---	218,303	260,113	332,685	132,271
Turbomachinery		93,280	133,430	133,430	147,207	96,034
Total Cost	$ 219,709	$ 410,244	$ 418,033	$ 459,843	$ 546,192	$ 358,427
		+190,535	+198,324	+240,134	+326,483	+138,718
ECAS - COE, mills/kWh*	34.68	61.17	60.95	61.38	64.56	58.12
Δ From AVCO-PSPEC		+26.49	+26.27	+26.70	+29.88	+23.44

* PSPEC COE based on $\$1.05/10^6$ Btu and an average labor cost of \$21/hr

GAI COE based on $\$1.05/10^6$ Btu and an average labor cost of \$14.20/hr

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construction times of 6.5 and 5.75 years were assumed for the CCMHD and PSPEC plants, respectively, so only overnight construction cost comparisons for the nominal 1000 MWe plants were made. Total capital costs, however, are (somewhat) normalized when cost of electricity (COE) is calculated; therefore, comparisons of ECAS-COE are presented in lieu of total capital cost comparisons for CCMHD and PSPEC power plants.

For individual accounts, the most significant cost differences on a relative basis are with the gasifiers and the regenerative heat exchangers. In both instances, the CCMHD costs are larger than the OCMHD costs. Other cost account differences worthy of note are:

1. Turbocompressor, Turbogenerator (Acct. 314) - the CCMHD account includes an expensive argon compressor not included in the OCMHD plant, plus a turboexpander in the 2.0 cases.

2. Other Topping Cycle Equipment (Acct. 317.7) - the CCMHD account contains only an argon cooler and purifier while the OCMHD account contains the more expensive gasifiers.

Since each of the CCMHD cases contains different components, comparisons with the selected PSPEC case could only be made on the basis of like cost items in major accounts. For CCMHD Case 1.0, only the regenerative heat exchanger falls into this category, while for the remaining cases the major accounts are represented by the gasifiers, regenerative heat exchangers and turbo-machinery. Cost comparisons are therefore presented as cost differences between a) overnight construction cost and b) major accounts total installed cost.

7.0 ENVIRONMENTAL INTRUSION

7.1 Introduction

One of the main objectives of the national program to develop MHD power generation for utility applications is to assure that MHD power plants have minimum adverse effects on the environment.

The environmental emissions from the closed cycle MHD system come from the combustor, which burns either coal or coal-derived gaseous fuels. The function of the combustor is to provide a flow of high temperature combustion gases to the argon heat exchanger. The major pollutants from the combustor are SOx, NOx and particulates. Only these emissions are considered in the present evaluation, although other environmental impacts could arise from support systems such as cesium seed handling and processing or ash and waste disposal, and would have to be addressed in an Environmental Impact Statement.

In this study, the three base cases were investigated. The combustor of the first case is direct coal-fired, while the combustors of the second and third cases use gaseous fuel supplied from pressurized and atmospheric gasifiers, respectively. While most of the combustion gases are exhausted into the stack, a small portion of the combustion gas is recycled into the argon heat exchanger. Stack gas emissions such as SOx, NOx and particulates are assessed and compared with the EPA New Source Performance Standards (NSPS).

7.2 SOx Emissions

7.2.1 Base Case 1.0 - Direct Coal Fired

The combustion products of the closed cycle MHD system do not have the inherent sulfur control mechanism found in open cycle MHD systems. The reason is that the hot combustion gases in the closed cycle are only used to provide heat to the seeded MHD working fluid (argon), whereas in the open cycle system the combustion gas, which contains the seed material, is also the MHD working fluid. The seed in the OCMHD plasma reacts chemically with the sulfur in the gases to reduce the SOx emissions to an acceptable level. The SOx emissions from the combustor of the closed cycle MHD system, therefore, require controlling, which is accomplished by providing sulfur removal equipment in the gas flow path.

The CCMHD combustor when fueled with Montana Rosebud coal, has 1.1 percent sulfur on a dry basis with a higher heating value of 11,560 Btu/lb. The potential SOx emission rate for this coal is 1.903 lb/10^6 Btu. The 1979 EPA NSPS limit (Figure 7-1) requires that 70 percent of the potential SOx emissions be removed; this is equivalent to an allowable SOx emission level of 0.57 lb/10^6 Btu.

The NSPS limit can be attained by removing SO from the gases with either a typical wet scrubber or a dry scrubber system that is just entering the utility market. The wet scrubber operates with a reactive alkali medium such as lime or limestone slurry and precipitates the sulfur out of the flue gas as insoluble calcium sulfite and calcium sulfate. The scrubber sludge is then dewatered and discarded.

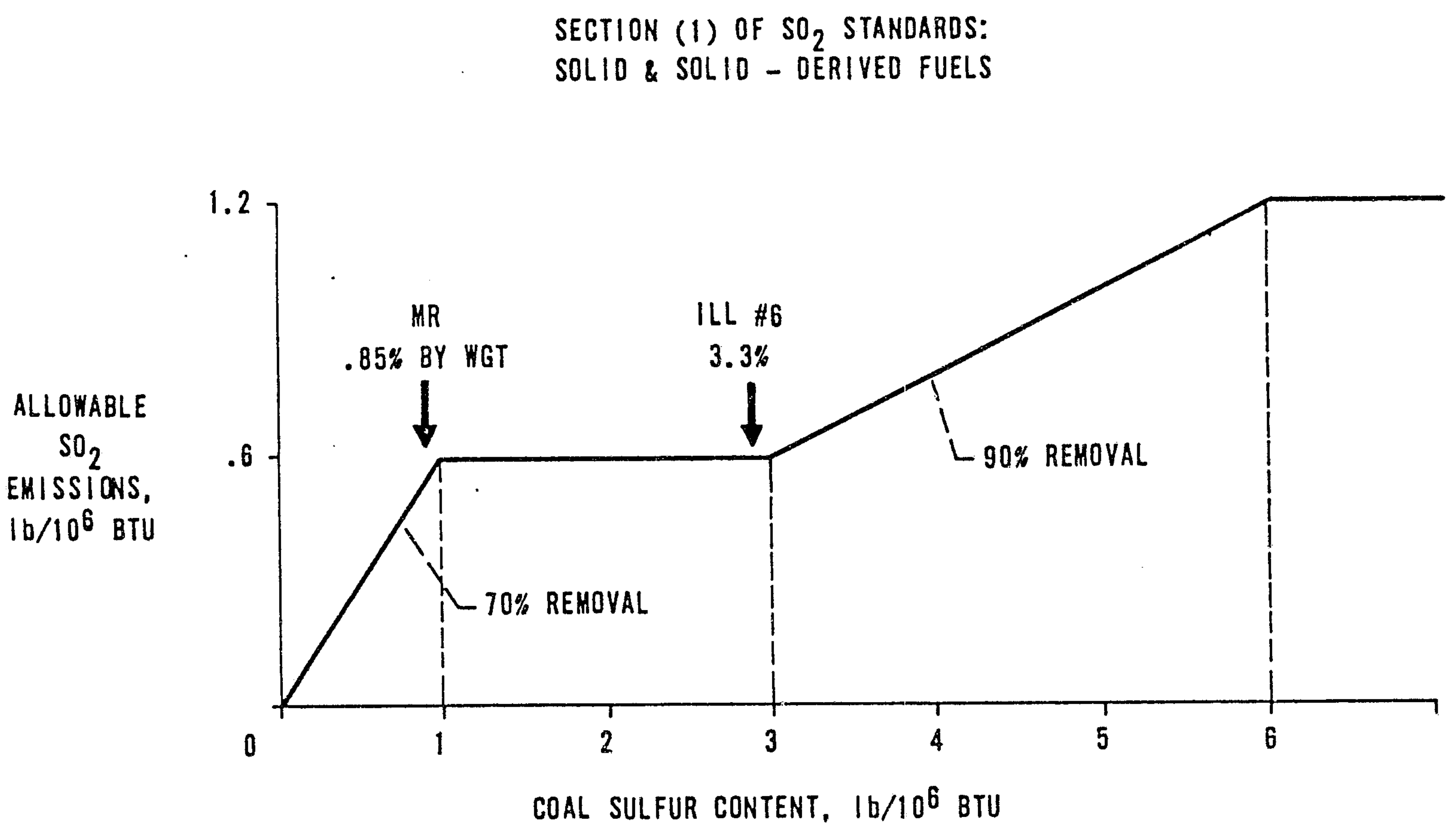

FIGURE 7-1

EPA NEW SOURCE PERFORMANCE STANDARD

In the dry scrubber, also called a spray dryer, the reactant (typically finely pulverized lime or atomized lime slurry) is sprayed into the flue gases as a "nearly dry" slurry. The moisture in the slurry evaporates and the solvent material reacts with the SO_2 in the flue gas to form dry calcium sulfate and sulfite powders. The dried spent chemicals, along with some flyash, are collected in powder form at the bottom of the spray dryer. The flue gases exiting from the spray dryer absorber are further cleaned in a baghouse filter or electrostatic precipitator before they are exhausted into the stack.

7.2.2 Base Case 2.0 - Pressurized Gasifier

The combustor in this case study is fueled with gaseous fuel supplied from a pressurized gasifier. The gasifier is an IGT design and operates at 10 atmospheres pressure. The hot gas produced by the gasifier is cleaned by passing the gas through a hot gas cleanup system. The hot gas cleanup system removes nearly 90 percent of the sulfur (which is in the form of H_2S) in the fuel gas. Cleaned gas is then delivered to the MHD combustor at 1335 F and 135 psia.

The product raw gas from the gasifier has an equivalent sulfur content of 0.25 percent. The potential SOx emission from the hot raw gases is 1.9 lb/10^6 Btu of coal heat input to the gasifier. After removing about 90 percent sulfur from the raw gas, the sulfur content in the clean gas is reduced to 0.027 percent; this is equivalent to a SOx emission of 0.2 lb/10^6 Btu of coal heat input, which is well below the 1979 EPA NSPS limit of 0.57 lb/10^6 Btu.

The recommended hot gas cleanup system is the iron oxide sorption method developed by the Morgantown Energy Research Center. This process consists of removing the H_2S from the raw hot gas by passing the gases through a regenerative sorption reactor containing iron oxide. The reaction mechanism is chemisorption, with H_2S diffusing into the sorbent particle and reacting with iron oxide to form iron sulfide. The sulfided absorbent is regenerated with air, steam or a mixture of O_2 and steam at temperatures of about 1000-1500 F, producing an offgas containing SO_2 and reusable iron oxide. The SO_2 gas from the sorption reactor is converted to recoverable sulfur in an Allied Chemical SO_2 reducer. Cases 2.0 and 2.10 both employ hot gas cleanup.

7.2.3 Case 2.12 - Cold Gas Cleanup System

For this case, the gaseous fuel from the pressurized gasifier is cleaned using a cold cleanup system. In a cold gas cleanup system, the cold gas is first passed through a venturi scrubber to remove essentially all the solid particulates, and then through a Stretford desulfurizer to remove the sulfur.

The venturi scrubber operates in the following manner: The gases are passed through a venturi tube to which low pressure water is added at the throat. Gas velocities at the throat are from 15,000 to 20,000 fpm, and the pressure drops are from 10 to 30 inches of water. The high turbulence in the venturi promotes intimate contact between the water droplets and the solid particulates in the gas. The wetted particles and droplets are then directed to a cyclone spray separator where the particulates are removed.

The particulate free gas is passed through a Stretford system for sulfur removal (detailed description of this process is in Section 4.2.1.1). This process involves passing the fuel gas containing H_2S through an absorber where nearly all the H_2S is removed. In the absorber unit, the gas comes in contact with the Stretford solution, which is an aqueous solution of a vanadium salt with anthraquinone disulfuric acid. The H_2S in the gas is oxidized by the vanadium to elemental sulfur which is then removed from the solution. This process is capable of reducing the sulfur level in the fuel gas to under 4 ppm, reducing SOx emissions well below the EPA-NSPS limit.

7.2.4 Base Case 3.0 - Atmospheric Gasifier

The combustor in this case study is fueled with gaseous fuel supplied from a CE atmospheric gasifier. The gasifier system utilizes 5 percent moisture Montana Rosebud coal. The fuel gas from the gasifier is cooled and then cleaned in a cold cleanup process which involves a venturi scrubber and a Stretford desulfurizer. The resultant clean fuel gas, containing about 10 ppm sulfur at 14.7 psia and 105 F, is then delivered to the CCMHD system combustor. The cold, clean gas is reheated to 1200 F before going to the combustor.

The hot raw gas from the gasifier has 0.23 percent sulfur content, which is equivalent to a potential SOx emission of 1.9 lb/10^6 Btu. After Stretford cleanup, the clean gas has an estimated 10 ppm sulfur content or a potential SOx emission level of 6.2×10^{-4} lb/10^6 Btu (equivalent to a total sulfur reduction of 99.97 percent). This cleaned fuel gas is thus far below the EPA-NSPS SOx limit of 0.57 lb/10^6 Btu.

7.3 NOx Emissions

Evaluation of the amounts of NOx produced by high temperature coal combustion is difficult. In general, the NOx production level is dependent upon various parameters such as the level of fuel-bound nitrogen, combustion temperature, stoichiometry of combustion, recirculation of flue gas and the intensity of mixing within the combustion chamber.

For the closed cycle MHD cases studied, the combustion flame temperature was in the range of 3400 - 3850 F, which is much lower than encountered in an open cycle MHD systems (~4500 F). Table 7-1 illustrates the type of fuel, stoichiometry and the estimated flame temperatures for the three CCMHD base cases studied. The flame temperatures are tempered with recirculated flue gas to a maximum temperature of 3350 F prior to entering the regenerative heat exchanger.

Laboratory experiments by Pershing and Wendt[7] have shown from coal combustion tests that the actual NOx emission level originates from fuel and thermal NOx. The types of coals investigated were Pittsburg, Western Kentucky, Colorado and the Montana coals. Tests were conducted under controlled conditions which allowed them to maintain a self-sustained pulverized coal flame and develop a methodology to separate the relative levels of thermal and fuel NOx contributions to the total NOx formation. Their conclusion was that the fuel NOx contributed at least 75 percent of the total NOx emissions and was not significantly altered by variations in primary air percentage, secondary air swirl and burner throat velocity.

Table 7-1

Estimated Flame Temperatures

Base Case	Combustor Fuel	Combustor Stoichiometry	Flame Temperature F
1.0	Direct coal firing, MR	1.05	3843
2.0	Gaseous fuel derived from a high pressure gasifier	1.05	3406
3.0	Gaseous fuel derived from atmospheric gasifier	1.05	3410

They also concluded that although a change in mixing significantly altered total emission levels, the dominant NOx producing mechanism was still through oxidation of the fuel-bound nitrogen.

Variation of the flame temperature had a significant effect on the total NOx and very little effect on the contribution from fuel NOx. From test results on Montana coal with flame temperature ranging from 3400 - 3800 F, the total NOx emission level would be in the order of 700 - 1000 ppm. If uncontrolled, this level would be in excess of the EPA-NSPS NOx limits shown in Table 7-2. Potential NOx emissions can be reduced by incorporating NOx controlling devices which are currently being applied in conventional fossil thermal power plants.

7.3.1 NOx Control Options

Several methods for controlling NOx emissions have been identified:

(a) Combustion modification appears to be a promising method for reducing NOx in the effluent stack gases. This method involves techniques such as initial fuel-rich combustion, downstream adjustment of the fuel-air mixture to complete combustion, and regulation of exhaust gas residence times in downstream components to enhance decomposition of NOx. Strom[8] reports that NOx emissions can be kept below applicable standards by burning coal at 85 percent (substoichiometric) oxidant conditions and controlling the radiant boiler residence times.

Table 7-2

1979 EPA NSPS LIMITS FOR NOx

Fuel	Type	$lbm/10^6 BTU$	PPM
Montana Rosebud	Solid	0.5	350
Montana Rosebud	Gaseous	0.5	350
Ill. #6	Solid	0.6	450
Ill. #6	Gaseous	0.5	350

(b) Modification of firing has been claimed by Combustion Engineering to produce low NOx levels. By using tangential firing, the flame temperature attained is lower and less NOx is produced than from in-front and opposed firing. This lower temperature is caused by better heat transfer resulting from a larger furnace volume.

(c) Recirculation of flue gas which lowers the flame temperature and thus contributes to a lower NOx production level.

(d) NOx decomposition work done in Japan by Mori and Taira[9] indicates that decomposition occurs after reacting with the alumina refractory surface of the regenerative heat exchanger.

More research is required for effective NOx control. NOx control techniques for conventional power plants are in various stages of development and should be applicable to MHD technology.

7.4 Particulate Emissions

7.4.1 Base Case 1.0 - Direct Coal Fired

Ash, in the form of slag, will condense out of the gas phase in the primary combustor. The primary combustor is designed to remove 70 to 90 percent of the slag. Based on flow rate analyses, the hot gases exiting from the air heater are expected to carry about 7742 lb/hr of particulates. About 15 percent of the gas flow rate exiting from the air heater is recycled into the argon regenerative heat exchanger (see detailed heat and mass balance diagram for Case 1.0). The remainder of the gas flow is first passed through the coal dryer and then exhausted to the stack. It is estimated[10] that in the coal drying process, the flue gas will pick up about 20 lbs of particulates (coal dust) for every ton of as received coal (coal is dried from 22.7 percent to 10 percent moisture). Thus, the total particulate loading is estimated to be 15,862 lb/hr in the flue gases exiting the coal dryer; this is equivalent to 2.0 lb/10^6 Btu of heat input. Before the flue gases are exhausted into the stack, the particulate level must be reduced to satisfy the EPA-NSPS limit of 0.03 lb/10^6 Btu. A bag house filter of about 98.5 percent particulate removal efficiency is capable of controlling the emissions to the federal regulatory limits, and commercially available equipment is adequate for this purpose.

7.4.2 Base Case 2.0 - Pressurized Gasifier

An IGT pressurized gasifier is used to provide hot, clean fuel gas to the combustor. It is estimated that 85 percent of the ash content in the coal will be removed at the bottom of the gasifier. In the IGT gasifier system, double cyclone separators with an expected particulate removal efficiency of 80 percent are utilized. The solids are separated from the cyclones and injected back into the gasifier.

The remaining particulates will be carried in the fuel gas. About 70 percent of the solid particulates will be entrained in the iron-oxide hot cleanup system. Thus, the clean fuel gas derived from Montana Rosebud coal is expected to have a particulate loading of about 0.0175 lbs/10^6 Btu of heat input.

The clean fuel gas is burned in the CCMHD system combustor. The combustion products passing through the argon heat exchanger and a turbo-expander are also utilized for coal drying. The flue gases will pick up additional solid particulates in the coal dryer, about 20 lbs of particulates for every ton of coal input. The total particulate loading in the flue gases is therefore expected to be about 1.14 lb/10^6 Btu. Before exhausting the stack gases, particulates have to be reduced to the EFA-NSPS limit of 0.03 lb/10^6 Btu. A bag house filter operating at an efficiency of 97.4 percent would be required for this purpose.

7.4.3 Base Case 3.0 - Atmospheric Gasifier

The hot raw gas from the CE atmospheric gasifier is cooled and cleaned by a venturi scrubber and the Stretford sulfur scrubber. All the particulates are essentially removed in the venturi scrubber. Thus, the fuel delivered to the combustor is free from particulates.

The combustion products, after passing through the heat exchangers, entrain the solid particulates while the gases go through the coal dryer. In the coal dryer, the as-received Montana Rosebud coal is dried from 22.7 percent to 5 percent moisture. The estimated solid particulates carried by the flue gases would be about 10,920 lb/hr, which is equivalent to 1.121 lb/10^6 Btu. Particulates in the stack gases have to be reduced to the 1979 EPA NSPS limit of 0.03 lb/10^6 Btu. A bag house filter with an operating efficiency of 97.3 percent would be required to meet the federal limits.

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8.0 CONCLUSIONS

This parametric evaluation of closed cycle MHD systems provides performance and cost information that can be utilized to compare design configurations for alternate MHD similar power plants. Because of the preconceptual nature of many of the designs for CCMHD subsystems and components, absolute values for performance and cost are particularly tenuous. On a relative basis, performance estimates are likely to be more accurate than cost estimates.

Comparison of performance and cost estimates of the CCMHD system designs showed that the atmospheric coal-fired combustor, Case 1.0, had the highest (43.2%) overall efficiency and the lowest levelized COE, which emphasizes efficiency in terms of fuel savings. Case 3.0, atmospheric gasifier, had the lowest efficiency (36.1%) but also the lowest ECAS method COE, reflecting the low capital cost of this power plant design. Since the MHD topping cycle and the steam bottoming cycles were similar in all the parametric cases studied, the variation in plant efficiency was primarily caused by the configuration of the combustion system.

From this study the following conclusions can be drawn:

- Coal fired closed cycle MHD plants can be built which have efficiencies in the range of 40 to 45%. This efficiency level is slightly lower than oxygen enriched open cycle plants of the same size; however, direct-fired open cycle MHD plants are expected to have efficiencies of at least 50%. Therefore, closed cycle plant efficiencies compare favorably with oxygen enriched open cycle plants but are inferior to direct-fired open cycle plants.

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- The levelized cost of electricity (COE) in mid-1978 dollars is projected to be around 55 mills/kW-hr for the closed cycle system. For an oxygen enriched open cycle system the COE is about 42 mills/kW-hr. The direct-fired open cycle COE will be significantly less. Although the efficiency of closed cycle plants are comparable with oxygen enriched open cycle plants, the cost of electricity is significantly higher which confirms the ECAS conclusions.

- The argon regenerative heater represents the key component which effects both the cost and performance of the plant. For a direct coal fired combustor with slag carryover, the development and technical problems are essentially identical to those of a direct-fired open cycle regenerative air heater. Regenerative argon heater development for gasifier systems will be less complex than for direct coal-fired systems and will essentially be analogous to the development of separately-fired open cycle air heaters. Regenerative heater development costs are expected to be high. Technical problems include, not only the basic heater development, but also a system which will minimize the amount of combustion gas (contamination) carried over into the argon during the cyclic operation.

- Direct coal fired closed cycle MHD plants have the highest efficiency, but introduce regenerative argon heat exchanger problems and have a high capital cost.

- Non-equilibrum MHD channel operation will have to be demonstrated. Steady operation of an unstable, turbulent plasma operation requires large scale verification, and long channel life-times will have to be demonstrated. The small scale closed cycle MHD channel tests planned at the Institute of Technology, Eindhoven, Netherlands should provide applicable design information.

- Advanced pressurized gasifier closed cycle MHD plants have acceptable performance with less expensive regenerative argon heat exchangers; however, the pressurized gasifier development problem has not been completely solved.

- Atmospheric gasifier closed cycle MHD plants project a near state-of-the-art configuration with minimum capital cost; however, the plant efficiency is very low.

Results of this study should be considered pre-conceptual. Phase II of this investigation should be continued if more accurate cost and peformance values are required. In Phase II, a more detailed conceptual design of a selected plant would be developed.

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26. "First Commercial Dry-Scrubbing System in U.S.," Chemical Engineering, Vol. 86, No. 17, August 13, 1979.

APPENDIX - A
NON-EQUILIBRIUM MHD CHANNEL STUDIES

A.1 INTRODUCTION

At the request of the NASA/LeRec, Gilbert Associates, Inc., performed non-equilibrium MHD generator study as an add-on task to the earlier investigations on the closed cycle MHD plant parametric analysis.

In the earlier plant studies, the detailed generator calculations were not performed, instead, the generator was considered as a "black box" energy conversion device. Assumptions for the power extraction (enthalpy extraction) and generator efficiency were made by NASA to establish the power output and flow conditions.

The intent of this study, therefore, was to perform non-equilibrium generator calculations and verify and confirm whether or not the generator could produce the preestablished electrical power output for the specified flow conditions. The results presented here confirm that the originally assumed values are reasonable and within 13 percent of quoted values.

The objectives of the task consist of the following:

1) Calculate generator performance (isentropic channel efficiency, power extraction, and channel overall efficiency by considering generator parameters and plant flow conditions).

2) Determine the approximate physical dimensions necessary for costing the MHD generator and the magnet.

3) Verify the "black box" generator performance assumptions that were made in the earlier plant studies.

4) Provide generator performance for plant cases where size and inlet stagnation temperature change.

MHD generator analyses was performed for the plant cases listed in Table A-1. Each of the five channel cases was analyzed by considering the best values for channel parameters selected from the suggested range listed in Table A-2 (selection of the channel parameters was based on the past experience, the literature and engineering judgement). Furthermore, an assessment was made of the effect of varying the Mach number and the electrical load parameter on the channel performance and geometry.

A.2 ASSUMPTIONS

The assumptions, that were used in the channel performance calculations, are given in Table A-3.

The plasma inlet and exit stagnation pressures were held at the values same as those used in the earlier plant studies. Except for the plant channel case 2.9, all of the remaining channel case studies were performed with an inlet stagnation temperature of 3100° F.

The composition of the impurities in the plasma represents the composition of the combustion products which are used in the regenerative heating process. The plasma is assumed to attain the required inlet stagnation temperature, by the concept of heating the Argon gas (noble gas) in a ceramic core regenerative heat exchanger. The ceramic matrix comprising the core of the heat exchanger is cyclically heated by combustion products from the combustor and cooled by the noble gas which is used in the primary channel loop. After the core bricks are heated, the combustion gases are purged and the passages evacuated before the noble gas enters the heat exchanger. Regardless of the evacuation process, a small quantity of combustion gases will be carried over and mixed with the noble gas.

C-3

TABLE A-1: System Variations

Plant Case	Parameter	
2.0	1000 MWe nominal output from MHD generator (base case)	
2.4	500 MWe	
2.5	250 MWe	Alternate Sizes
3.7	100 MWe	
2.9	3000° F, Inlet Stagnation Temperature	

TABLE A-2: NASA Suggested Channel Parameters

Parameter	Suggested Volume
Seed Fraction	0.1% vol
Magnetic Field	6 Tesla (max) - taper as required by Hall voltage limitations
Electrical Load Parameter	As Required
Turbulence factor	0.2 - 0.5
Mach No.	Per Design Approach
Wall Temp	200° F less than bulk gas temperature
Diffuser coeff.	0.6 - 0.7

TABLE A-3: Assumptions

Inlet Stagnation Pressure:	10 Atm
Inlet Stagnation Temperature:	3100° F*
Exit Stagnation Pressure (approx):	2 Atm
Diffuser Pressure Recovery Coeff.	0.6
Plasma Turbulence Factor:	0.2
Carrier Gas:	Argon
Seed Material:	Cesium, .1% (vol)
Impurities:	50 ppm

(Composition of Impurities:
N_2 = 75.41%, CO_2 = 14.82%, CO = 0.26%
H_2O = 9.51%)

Magnetic Field (max):	6 Tesla
Maximum Hall Field Limit:	4000 V/M
Wall Temperature:	200° F less than bulk gas temperature

* Channel of plant case 2-9 was analyzed with 3000° F temperature.

A.3 ANALYSIS PROCEDURE

The channel analysis was performed by using GAI non-equilibrium channel code which has the following features:

- o one dimensional
- o subsonic or supersonic
- o specified velocity
- o boundary layer effects considered by wall skin friction and heat transfer
- o plasma turbulence effect
- o inelastic collision effects

The channel geometry and performance characteristics were obtained by simultaneously solving the mass, energy, momentum and electron balance equations along with the non-equilibrium plasma properties. These equations are listed in Table A-4. Equations (1), (2) and (3) describe the fluid mechanical aspects in the channel flow subjected to the magnetic field interaction, whereas, equation (4) describes a two-temperature model[1] used to represent the non-equilibrium effects in the seeded noble gas. Before coupling equation (4) with the other channel equations, equation (4) was independently solved and the results were compared with that of Cool & Zukoski[2] for a potassium seeded argon plasma. Comparison of current density vs. electrical conductivity, based on the elastic collision model analysis, is in good agreement with Cool's prediction (also based on elastic collision) as shown in Figure A-1.

In the channel calculations, inelastic collision effects due to the presence of molecular impurities were included; these effects were modeled by use of δ_h-factors[3] (see equation (4)), whose values appear in Table A-5.

TABLE A-4: Governing Equations for Non-Equilibrium MHD Channel Flow

Momentum: $$\frac{dp}{dx} = - jxB - F - \rho V \frac{dv}{dx} \quad (1)$$

Energy: $$\frac{dh}{dx} = \frac{1}{\rho V} \{J \cdot E - Q - \rho V^2 \frac{dv}{dx}\} \quad (2)$$

Continuity: $$\frac{d}{dx} (\rho AV) = o \quad (3)$$

Electron Balance: $$\frac{Jy^2}{\sigma_{eff}} = 3N_e k M_e \ (T_e - T_g) \ \Sigma \frac{\gamma_{eh}}{m_h} \cdot \delta_h \quad (4)$$

Equation of State: $$h = h(P, T_g) \quad (5)$$

where

$N_e = f\ (T_e, N_s)$ Saha'a equation

$\gamma_{eh} = f\ (Q_n, N_h)$ collision frequency

$\sigma = eN_e\ \mu_{eh}$

$Jy = \sigma_{eff}\ (Ey - VB)$ segmented Faraday configuration

δ_h = models the inelastic collision effects

$\sigma_{eff} = f\ (\sigma, \beta, \xi)$

NON EQUILIBRIUM MODEL

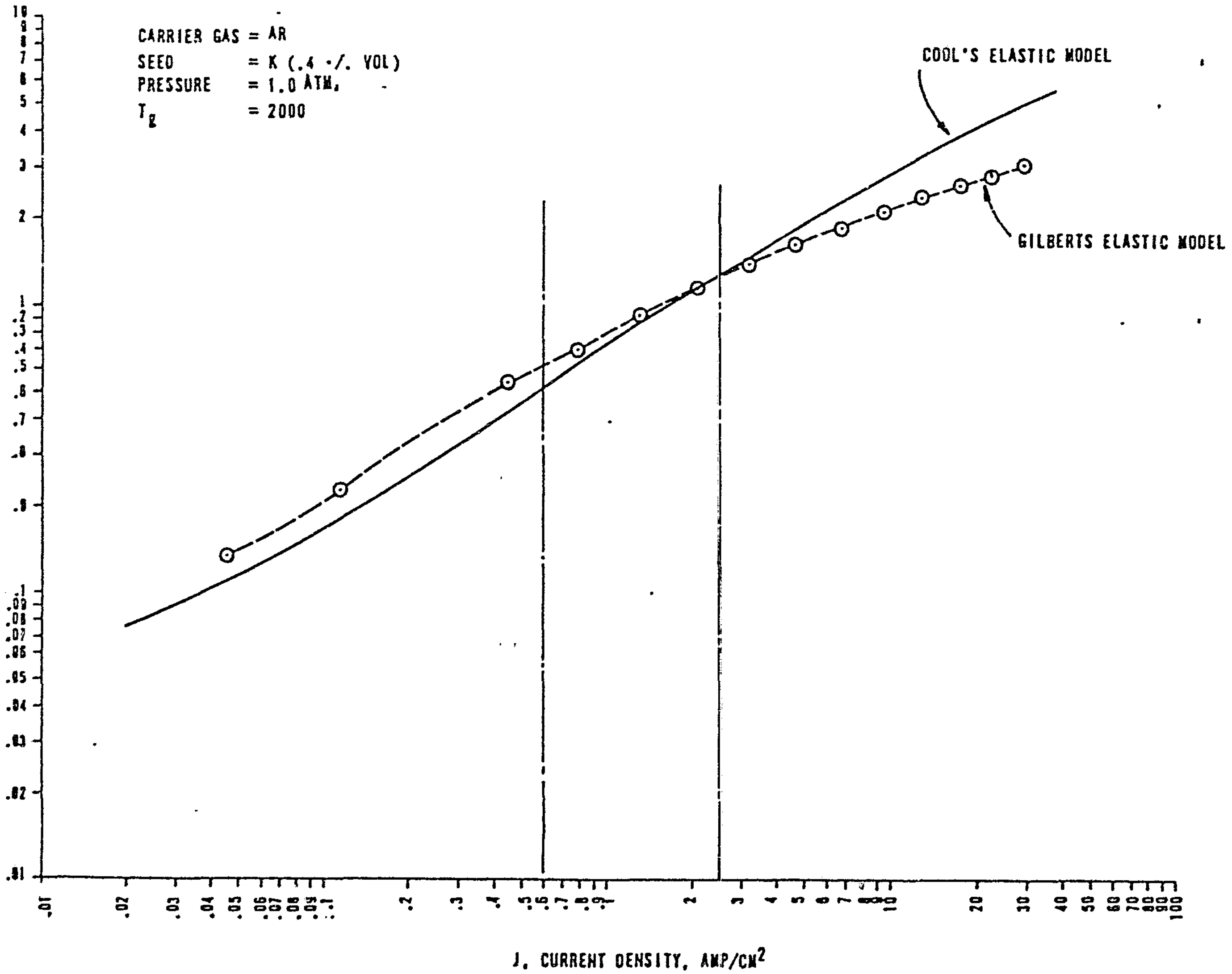

Figure A-1

TABLE A-5: Assumed Values of δ_h

Gas Species	δ_h
CO_2	1000
H_2O	1000
CO	100
N_2	7.8
Cs	1.0
Ar	1.0

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After validating the non-equilibrium plasma relationship (j vs σ), the entire set of equations (1) through (5) were simultaneously solved to give the required channel results. With the specified values for velocity, inlet stagnation pressure and temperature, and electrical loading factor, the calculations were repetitively performed by varying channel length until a specified diffuser exit pressure was obtained.

A.4 DISCUSSION OF RESULTS

In this section, channel results obtained from the analyses of the base case, and the alternate conditions discussed.

A.4.1 Base Case Studies, Plant Identification 2.0

Several channel studies were conducted in this plant size (1000 MW_e nominal). The variables that were used as the parameters are the velocity, channel L/D and the electrical loading factor. For ease of comparison with other cases, one channel run was designated the base case which has the parameters listed in Table A-6.

Base Case Channel Results

The results of the base channel case are summarized in Table A-7. This channel operating at supersonic velocity, with Mach number varying from 1.06 at inlet to 1.41 at exit, produces 868 MW_e of electrical output. To produce this power, the electrical loading factor was held at 0.8 with a predicted channel length of 11.8 meters with an L/D ratio of 6.7. This channel length is sufficient to drop the stagnation pressure from 10 atmospheres at nozzle inlet to 2.08 atmospheres at the diffuser exit (this pressure is within 4% of the specified value of 2 atmospheres). The power extraction, the isentropic channel efficiency and the overall efficiency for the channel are predicted to be 33%, 87.9% and 70.7% respectively (The definitions of these parameters are given in Figure A-2).

TABLE A-6: Base Case Parameters

Thermal Input	2465 MW_t*
Plasma Flow Rate	2778.5 Kg/sec
Velocity (const)	747 M/sec
Electrical Loading Factor	0.8
Supersonic Channel	

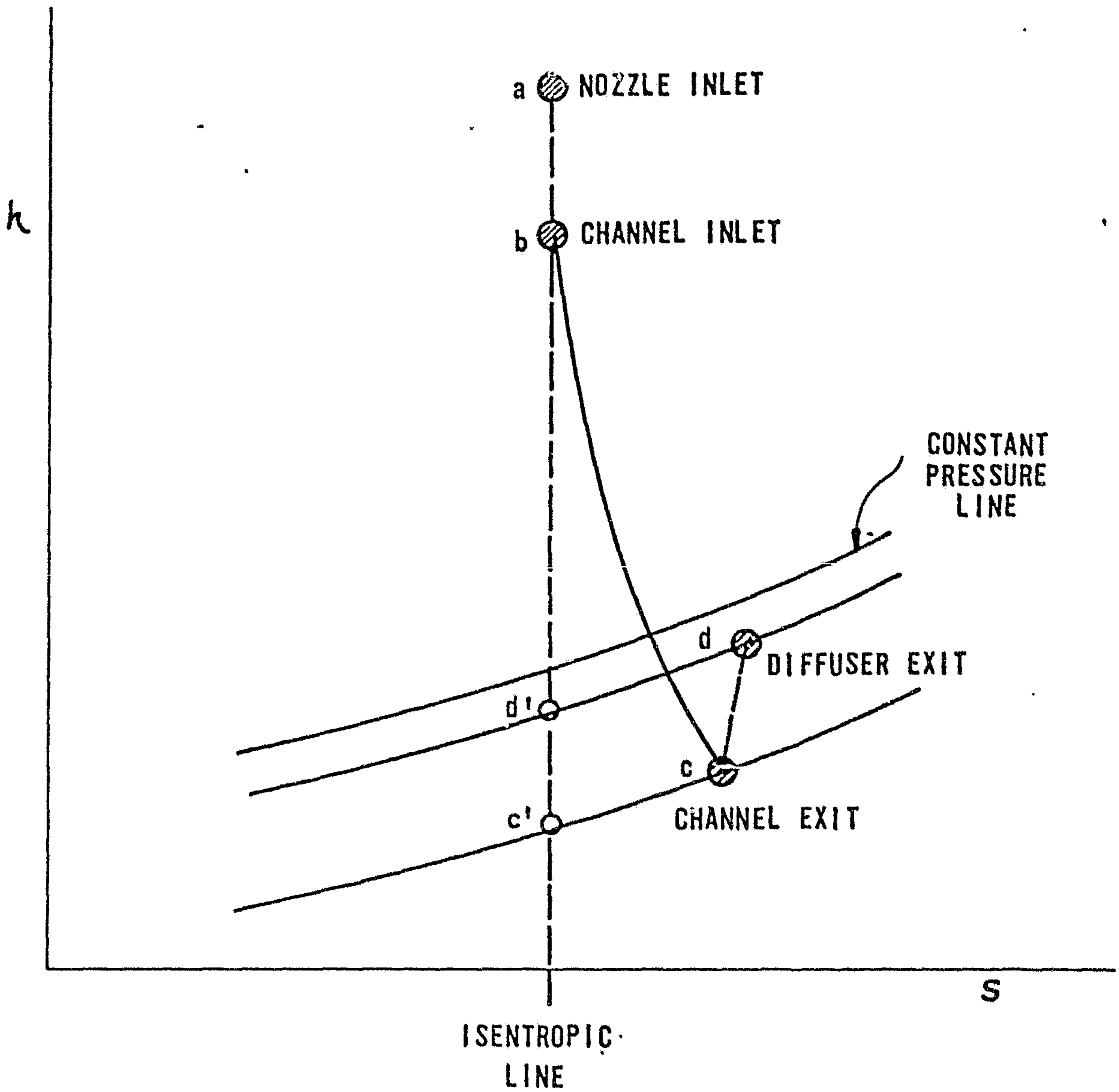
* This is the energy associated with the plasma flow at the entrance of nozzle (product of flow times stagnation enthalpy at inlet).

TABLE A-7: Base Channel Case

Thermal Input	265 MW_t
Inlet Stagnation	10 Atm, 3100° F
Exit Stagnation	2 Atm (approx.)
Velocity	747 M/sec
Electrical Loading Factor	0.8

L (M)	L/D	Channel Area In. (M^2)	Channel Area Ex. (M^2)	Mach No. In.	Mach No. Ex.	Power (MW_e)	Power Extraction %	Channel Eff. (%)	Overall Eff. (%)
11.8	6.7	2.43	7.88	1.057	1.411	868	33	87.9	70.7

EFFICIENCY & POWER EXTRACTION DEFINITIONS



$$\text{POWER EXTRACTION} = \frac{(\Delta T)_{ad}}{T_a}$$

$$\eta^{ISEN}_{CH.} = \frac{(\Delta h)_{bc}}{(\Delta h)_{bc'}}$$

$$\eta_{OVERALL} = \frac{(\Delta h)_{ad}}{(\Delta h)_{ad'}}$$

Figure A-2

The thermodynamic state points across the base case channel are given in Figure A-3. Other details of the results such as the static temperature, static pressure, physical cross sectional area, Mach number, Hall parameter, current density, electron temperature, and electrical conductivity variations along the channel are described in Figures A-4 through A-11. The electron temperature (and hence the electrical conductivity) was observed to go through a rapid increase near the end of the channel, as seen in Figure A-9. This can be attributed to the fact that the electrons experience fewer collisions at the reduced pressures, and therefore lose less energy. This higher energy state results in increased electron temperature.

Several parametric variations of the base case are made to investigate the effect of the electrical loading and gas velocity on channel performance. The first case examined the performance changes as the gas velocity was reduced to a subsonic value while keeping the channel geometry constant. The second case investigated the effect of changing the velocity while keeping the loading parameter constant, and the third case reversed these rolls with loading parameter varying and gas velocity held constant.

<u>Velocity Effect with Constant L/D</u>. In this case, the channel analysis was performed with the constant subsonic velocity of 635 meters/second, as compared to the supersonic velocity for the base case. The electrical loading parameter, however, was reduced to 0.755, from the value of 0.8 that was used in the base case analysis, in order to maintain essentially at the same L/D value as that of the base case channel.

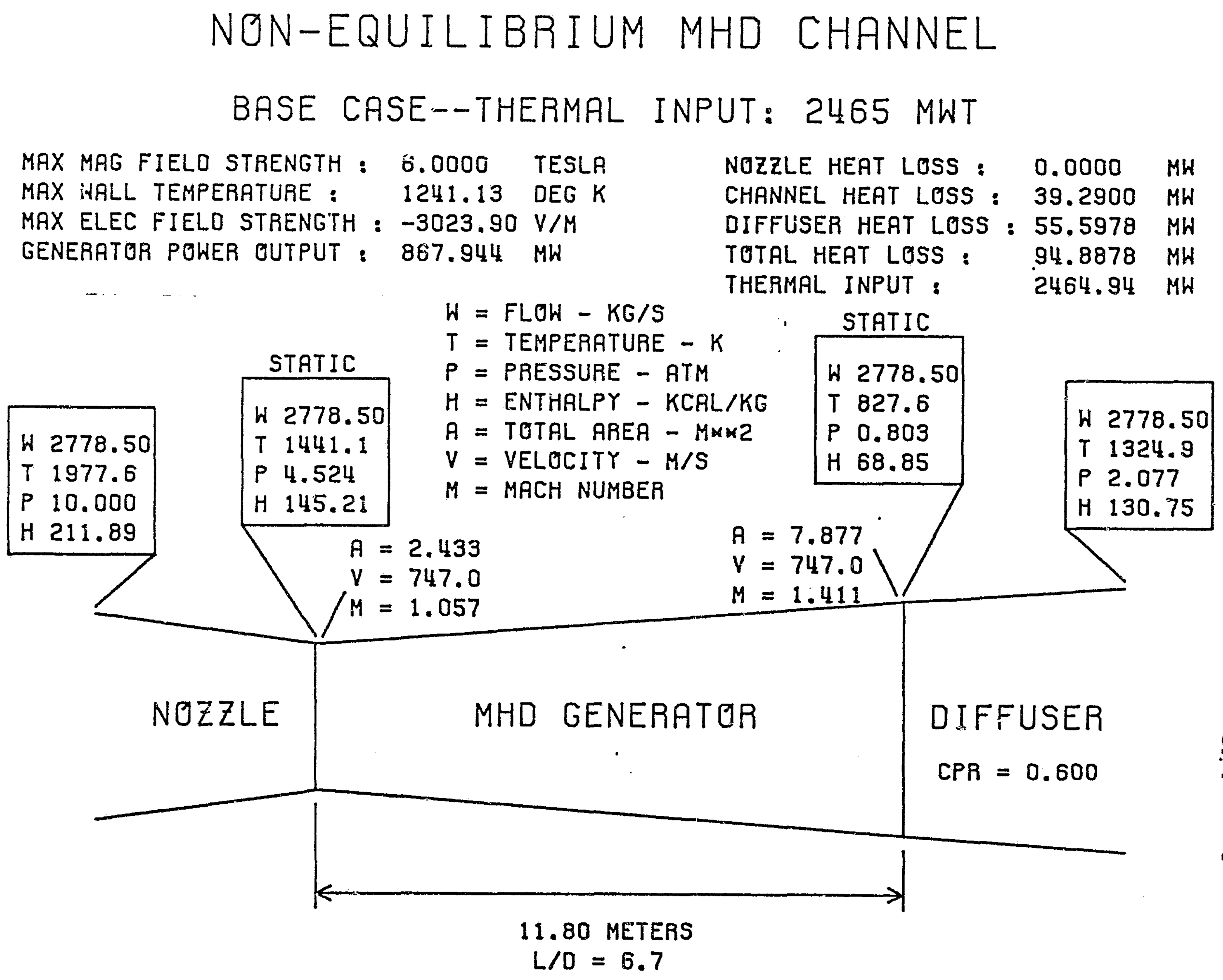

NØN-EQUILIBRIUM MHD CHANNEL
BASE CASE--THERMAL INPUT: 2465 MWT
MAX MAG FIELD STRENGTH : 6.0000 TESLA
MAX WALL TEMPERATURE : 1241.13 DEG K
MAX ELEC FIELD STRENGTH : -3023.90 V/M
GENERATØR PØWER ØUTPUT : 867.944 MW
NØZZLE HEAT LØSS : 0.0000 MW
CHANNEL HEAT LØSS : 39.2900 MW
DIFFUSER HEAT LØSS : 55.5978 MW
TØTAL HEAT LØSS : 94.8878 MW
THERMAL INPUT : 2464.94 MW
W = FLØW - KG/S
T = TEMPERATURE - K
P = PRESSURE - ATM
H = ENTHALPY - KCAL/KG
A = TØTAL AREA - M**2
V = VELØCITY - M/S
M = MACH NUMBER
W 2778.50
T 1977.6
P 10.000
H 211.89
STATIC
W 2778.50
T 1441.1
P 4.524
H 145.21
A = 2.433
V = 747.0
M = 1.057
STATIC
W 2778.50
T 827.6
P 0.803
H 68.85
A = 7.877
V = 747.0
M = 1.411
W 2778.50
T 1324.9
P 2.077
H 130.75
NØZZLE
MHD GENERATØR
DIFFUSER
CPR = 0.600
11.80 METERS
L/D = 6.7

Figure A-3

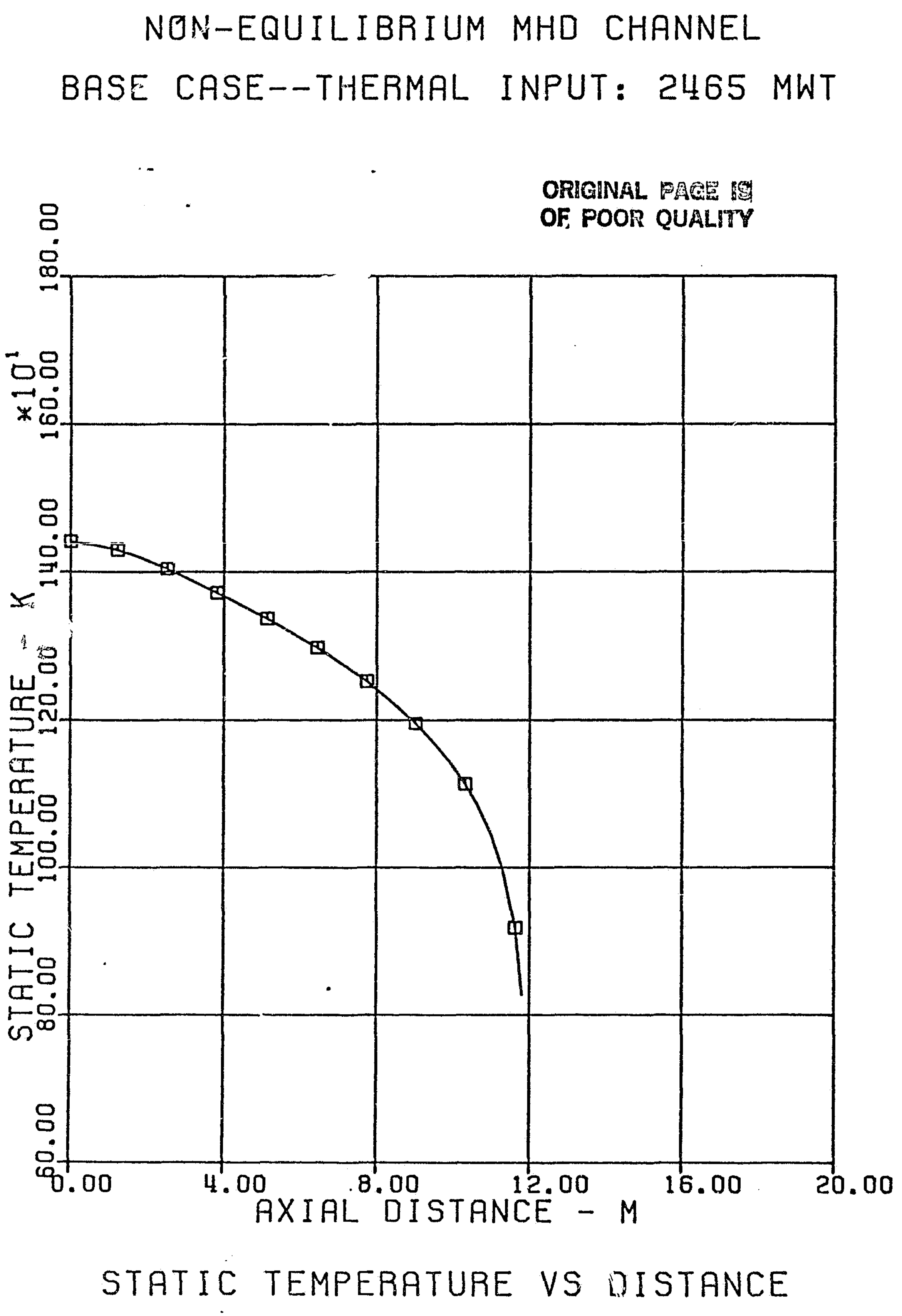


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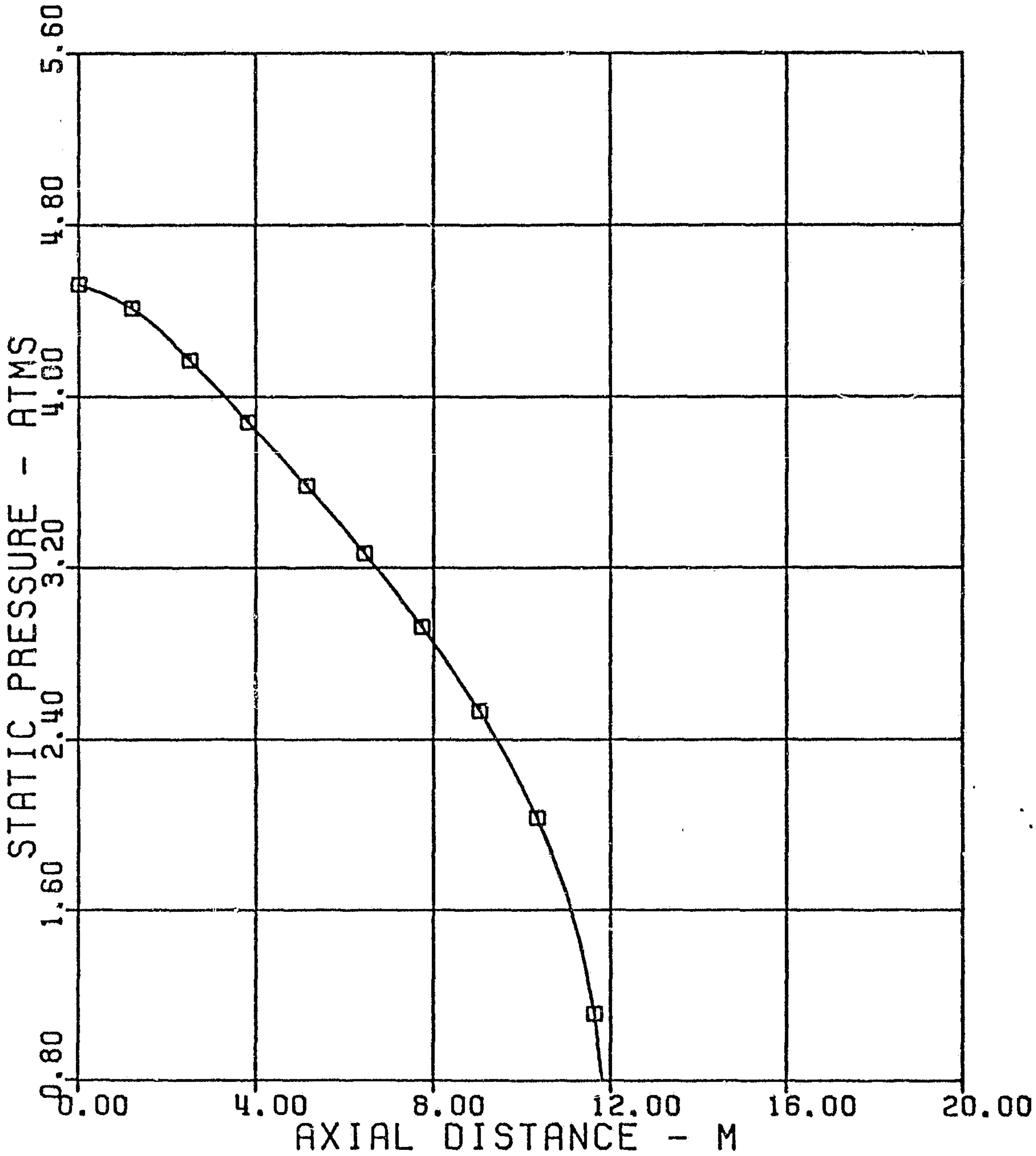
Figure A-4

NØN-EQUILIBRIUM MHD CHANNEL

BASE CASE--THERMAL INPUT: 2465 MWT



STATIC PRESSURE VS DISTANCE

Figure A-5

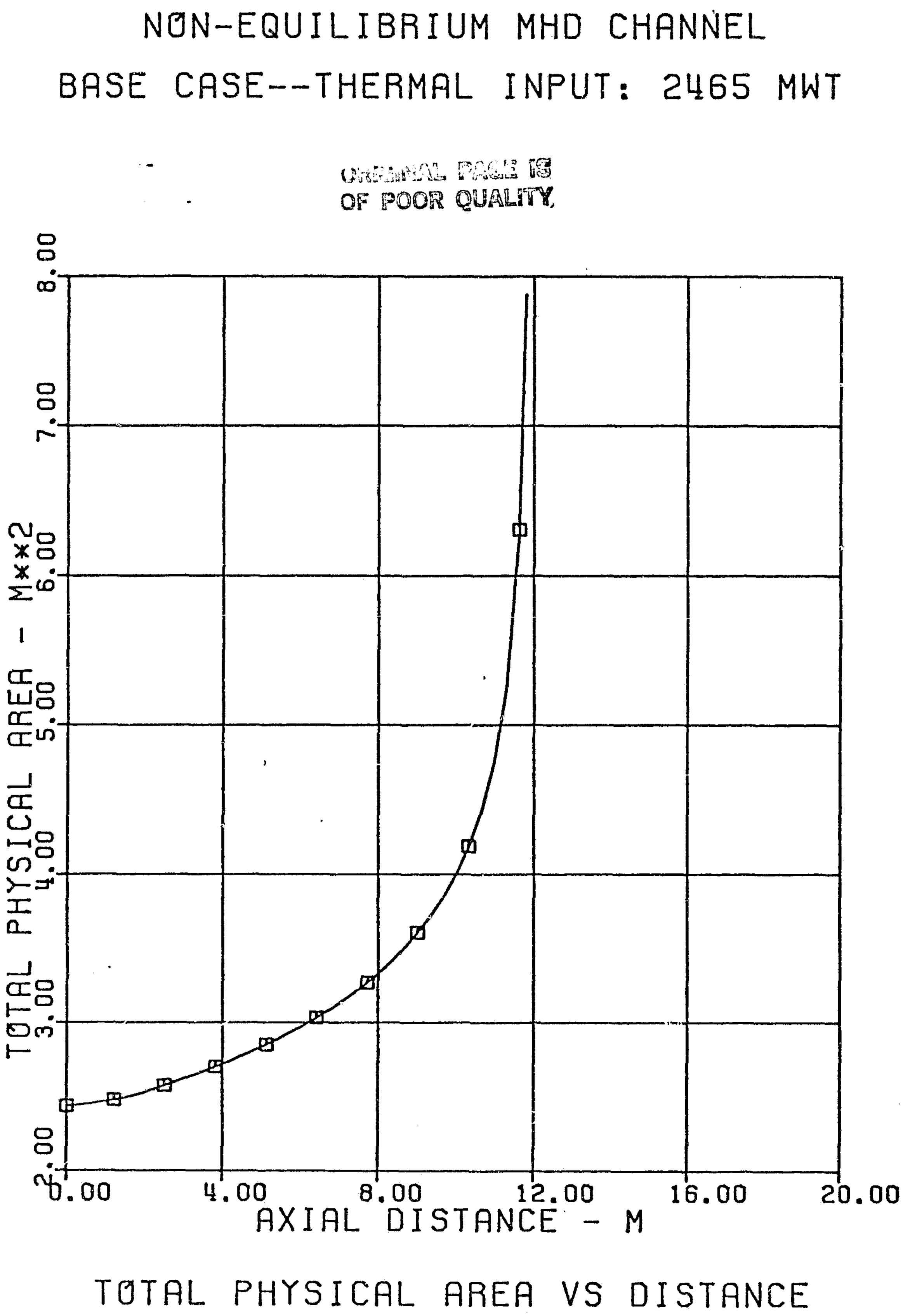
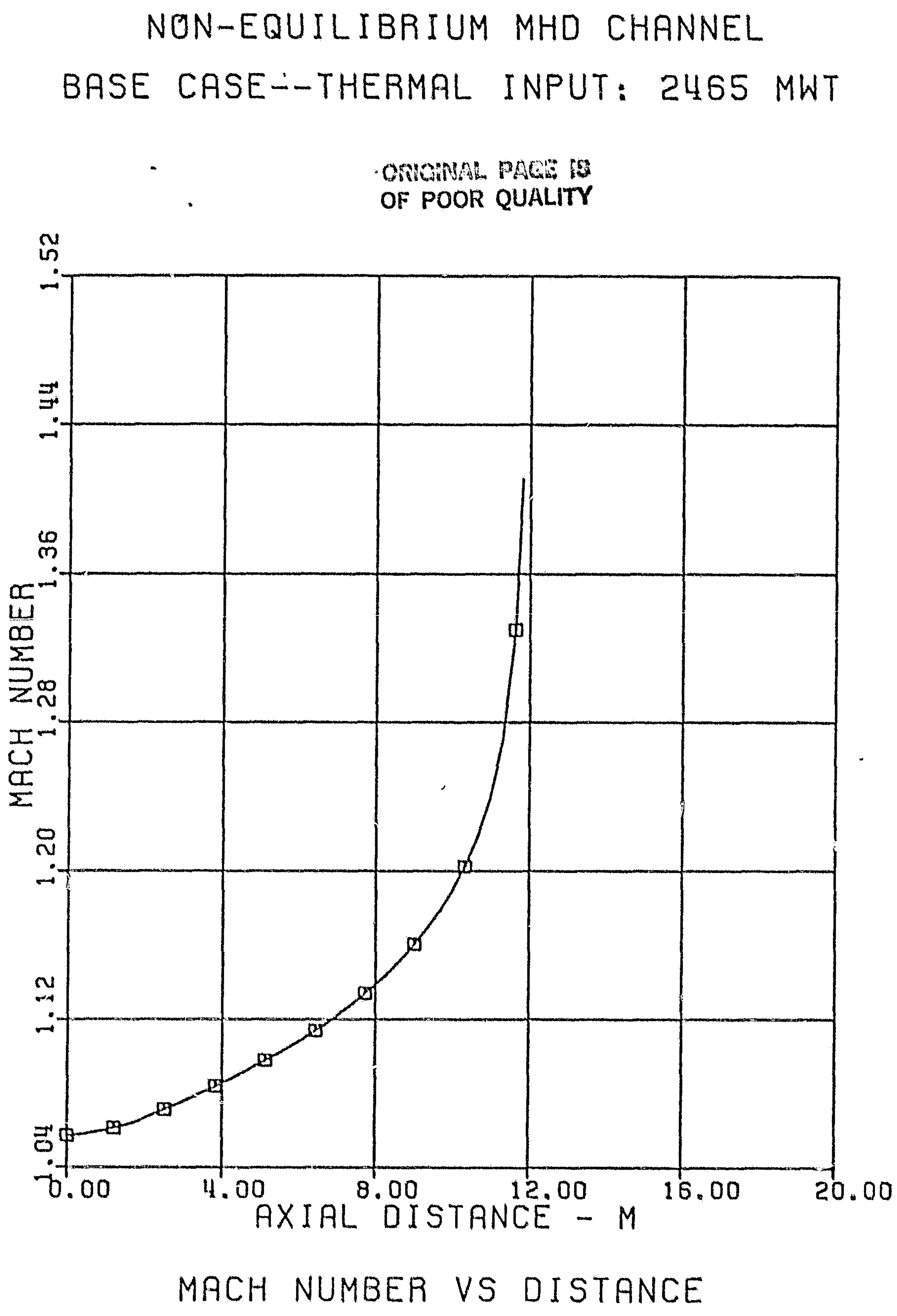

NØN-EQUILIBRIUM MHD CHANNEL
BASE CASE--THERMAL INPUT: 2465 MWT
8.00
7.00
6.00
5.00
4.00
3.00
2.00
TØTAL PHYSICAL AREA - M**2
0.00
4.00
8.00
12.00
16.00
20.00
AXIAL DISTANCE - M
TØTAL PHYSICAL AREA VS DISTANCE

Figure A-6


NON-EQUILIBRIUM MHD CHANNEL
BASE CASE--THERMAL INPUT: 2465 MWT
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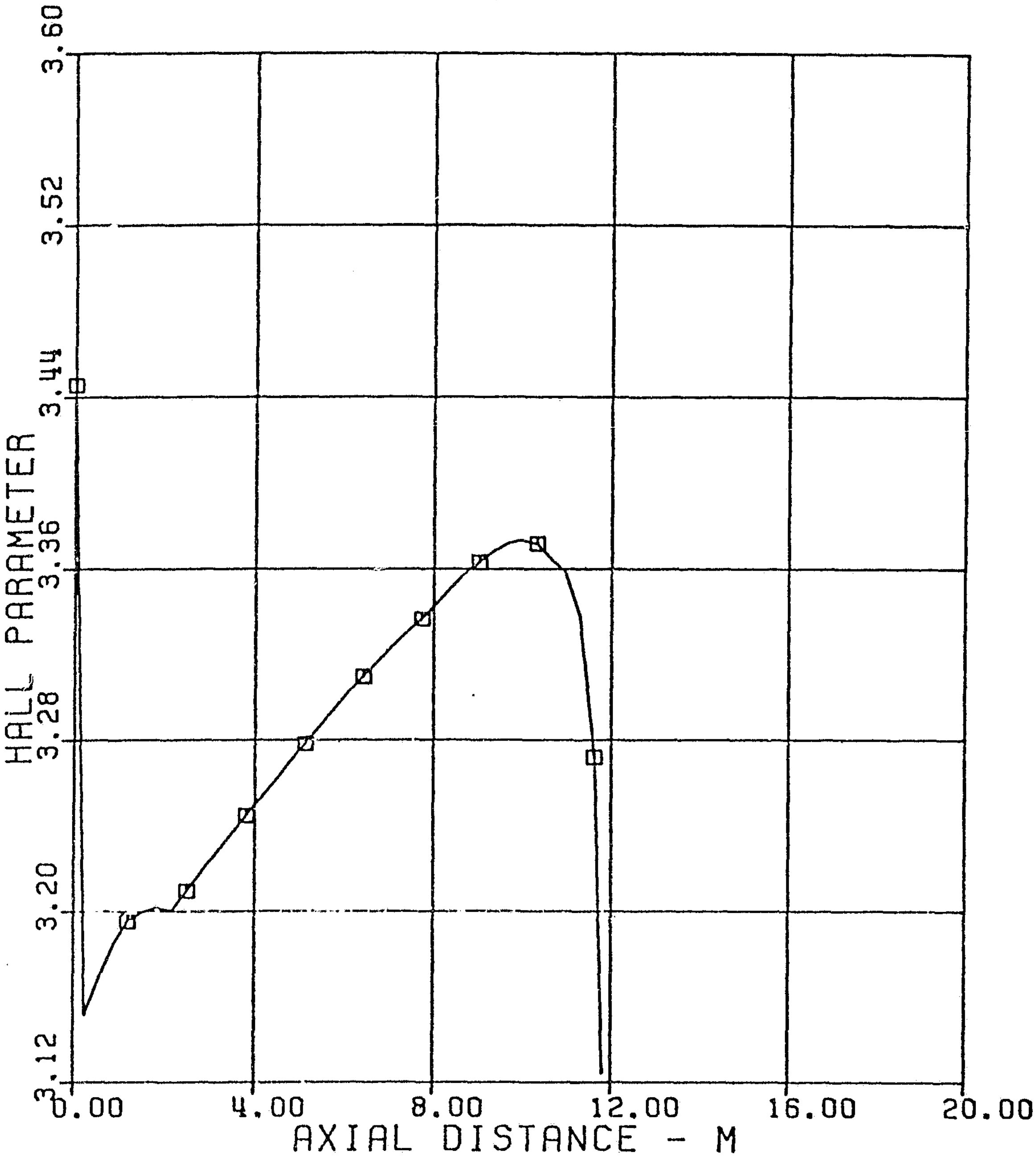
MACH NUMBER
1.04
1.12
1.20
1.28
1.36
1.44
1.52
0.00
4.00
8.00
12.00
16.00
20.00
AXIAL DISTANCE - M
MACH NUMBER VS DISTANCE

Figure A-7

NØN-EQUILIBRIUM MHD CHANNEL

BASE CASE--THERMAL INPUT: 2465 MWT

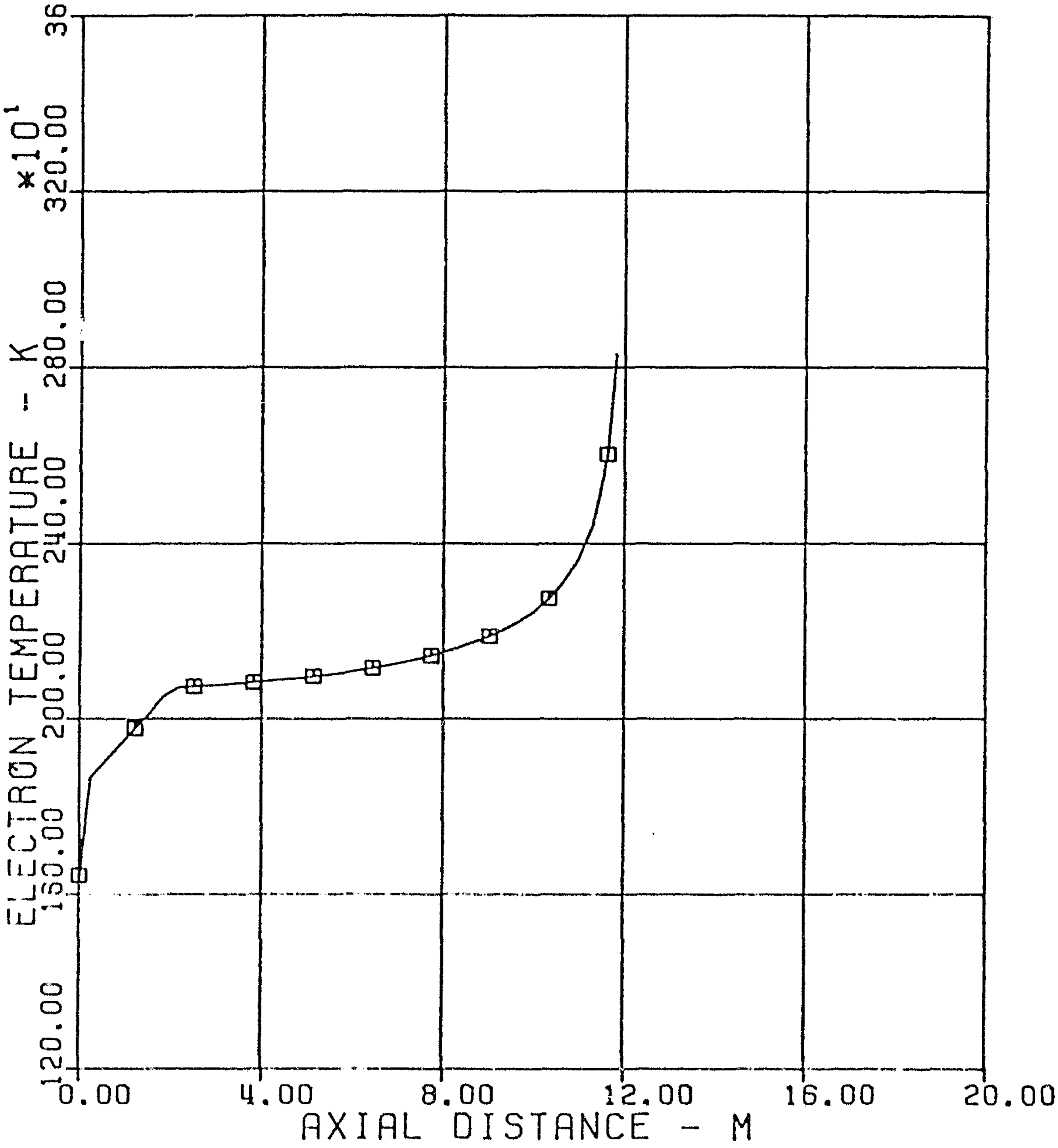


HALL PARAMETER VS DISTANCE

Figure A-8

NØN-EQUILIBRIUM MHD CHANNEL

BASE CASE--THERMAL INPUT: 2465 MWT

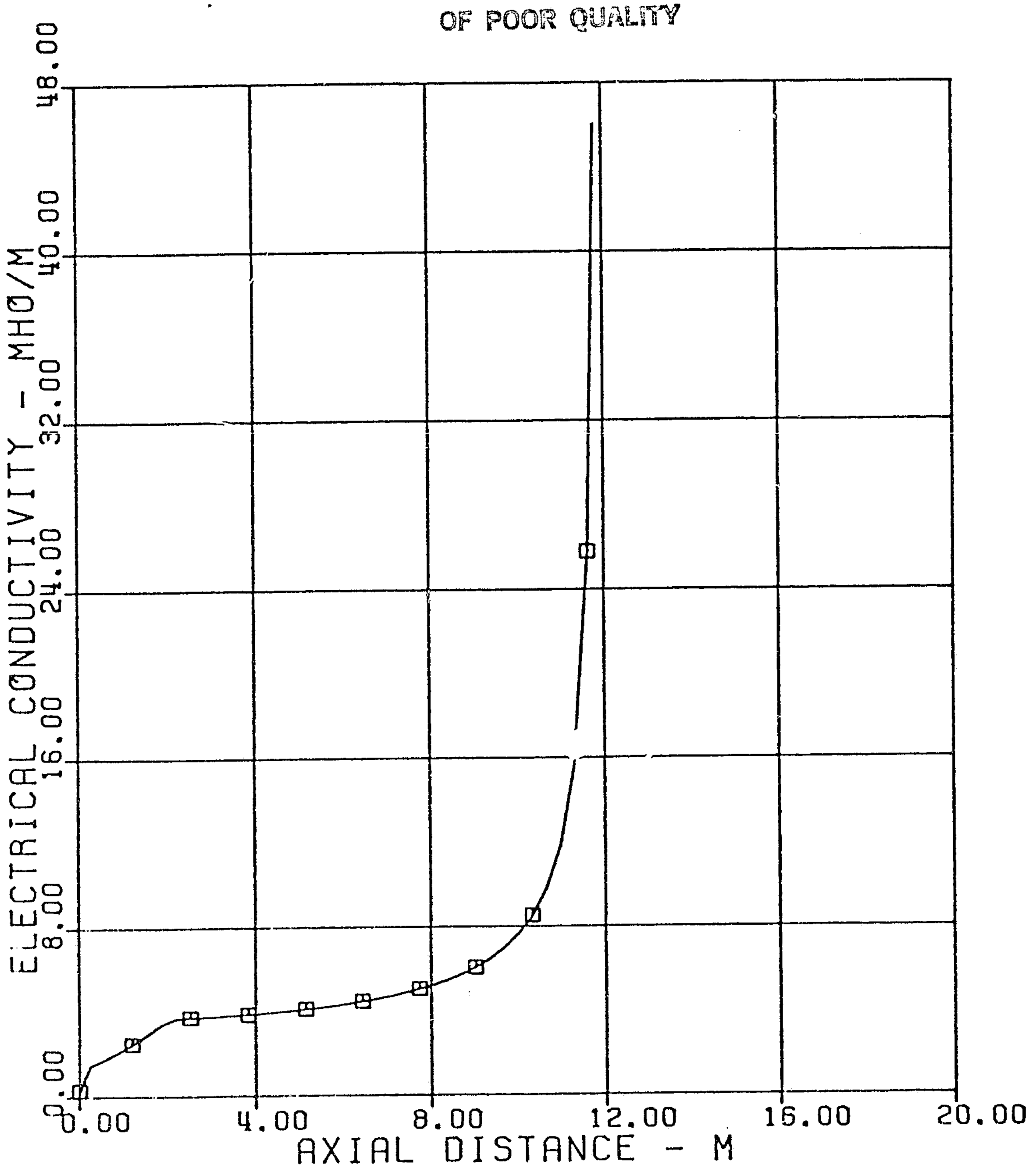


ELECTRØN TEMPERATURE VS DISTANCE

Figure A-9

NØN-EQUILIBRIUM MHD CHANNEL

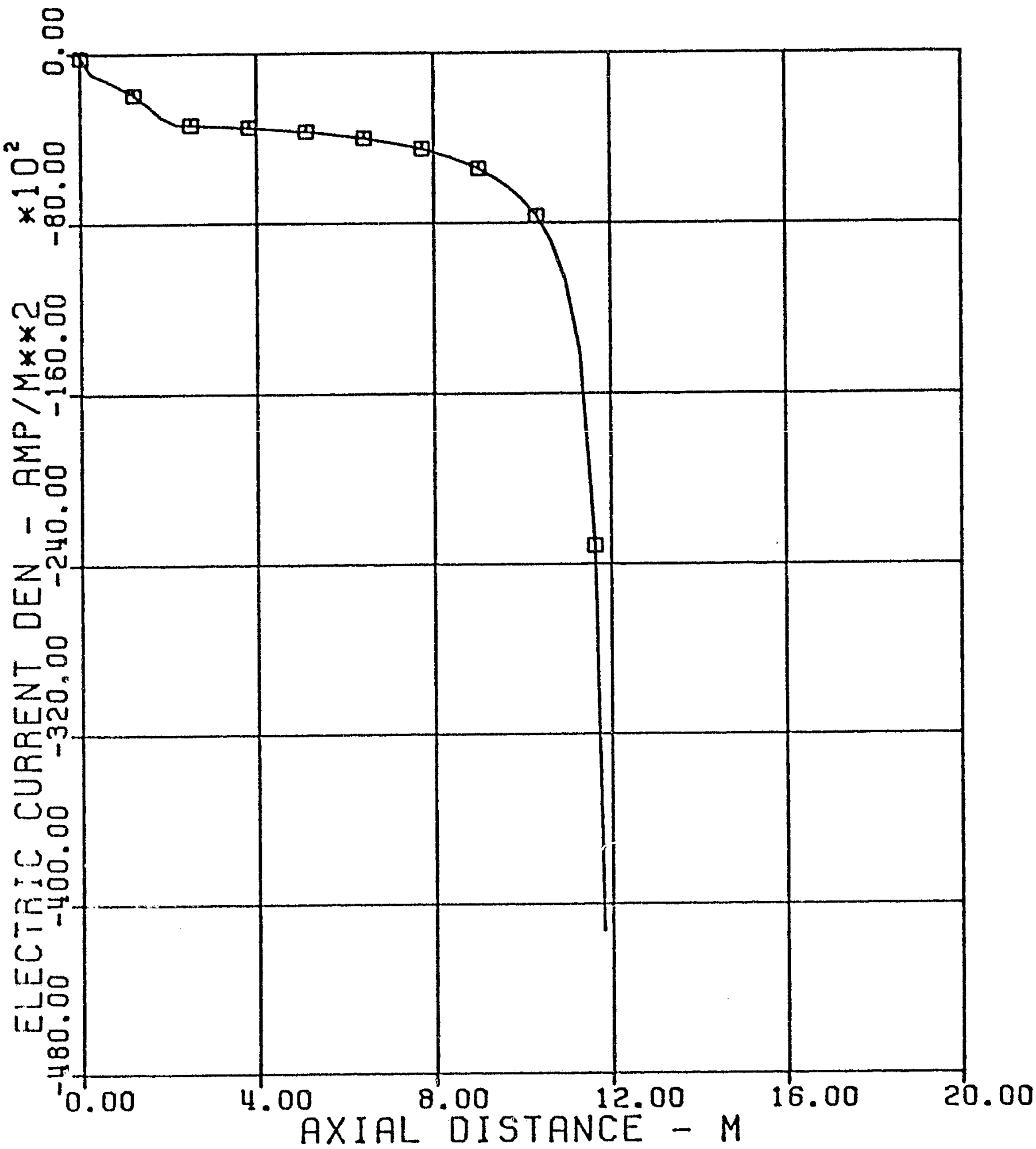
BASE CASE--THERMAL INPUT: 2465 MWT



ELECTRICAL CØNDUCTIVITY VS DISTANCE

Figure A-10

NØN-EQUILIBRIUM MHD CHANNEL



ELECTRIC CURRENT DENSITY VS DISTANCE

Figure A-11

The results of this channel analysis, compared with those from the base case, are summarized in Table A-8. The power output and power extraction from this channel were 875 MWe and 32.7 percent which are not significantly different from those of the supersonic base case channel.

Comparison of the core profiles is presented in Figure A-12 and additional comparisons for this case are given in Figures A-13 through A-19.

Velocity Effect with Constant Electrical Loading Parameter. Two channel runs, one at supersonic velocity (747 m/sec) and the other at subsonic velocity (635 m/sec), but with constant electrical loading parameter (0.77) for each run, were made. The purpose of this investigation was to determine how the channel performance and geometry would vary with velocity.

The results from this investigation are summarized in Table A-9. As the velocity decreased from 747 meters/sec to 635 meter/sec, the channel length increased from 5.5 meters to 17.3 meters. This is because the power density decreases with decreasing velocity and thus results in a reduced rate of change of properties, such as pressure and temperature along the channel.

The electrical output from the subsonic channel is slightly more (1.4%) than that from the supersonic channel. The power extraction and the overall efficiency for each of the channels are also presented in Table A-9.

Effect of Varying Electrical Loading Parameter. The effect of varying the electrical loading parameter (K) on the channel geometry (L/D and L) and performance was studied in this investigation. The channel computation was conducted for each of the K values of 0.77, 0.80, and 0.81, while maintaining the

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TABLE A-8: Velocity Variation with Nearly Constant (L/D)

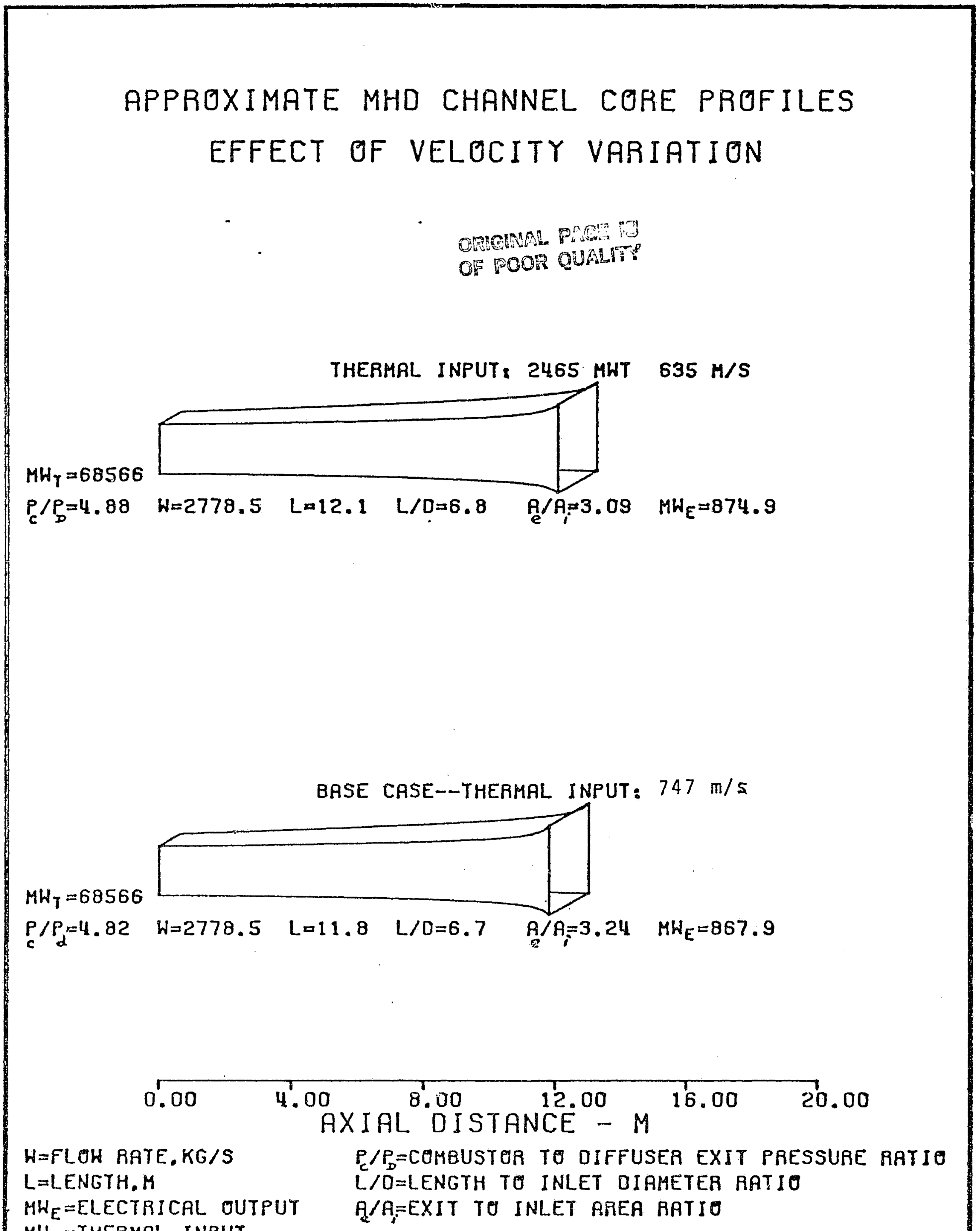
Thermal Input	2465 MW_t
Inlet Stagnation	10 Atm, 3100° F
Exit Stagnation	2 Atm (approx.)

Velocity (M/sec)	L (M)	L/D	K	Mach No. In.	Mach No. Ex.	Power (Mwe)	Power Extraction (%)	Channel Eff. (%)	Overall Eff. (%)
747	11.8	6.7	0.8	1.057	1.411	868	33	87.9	70.7
635	12.1	6.8	0.755	0.855	1.146	875	32.7	81.4	69.6

TABLE A-9: Velocity Variation with Constant Electrical Loading Factor

Thermal Input	2465 MW_t
Inlet Stagnation	10 Atm, 3100° F
Exit Stagnation	2 Atm (approx.)
Electrical Loading Factor	0.77

Velocity (M/sec)	L (M)	L/D	Channel Area In. (M^2)	Channel Area Ex. (M^2)	Power (MW_e)	Power Extraction (%)	Overall Efficiency (%)
747	5.5	3.11	2.43	7.88	870	31.7	68
635	17.3	9.72	2.47	7.62	882	33.4	71


APPROXIMATE MHD CHANNEL CORE PROFILES
EFFECT OF VELOCITY VARIATION
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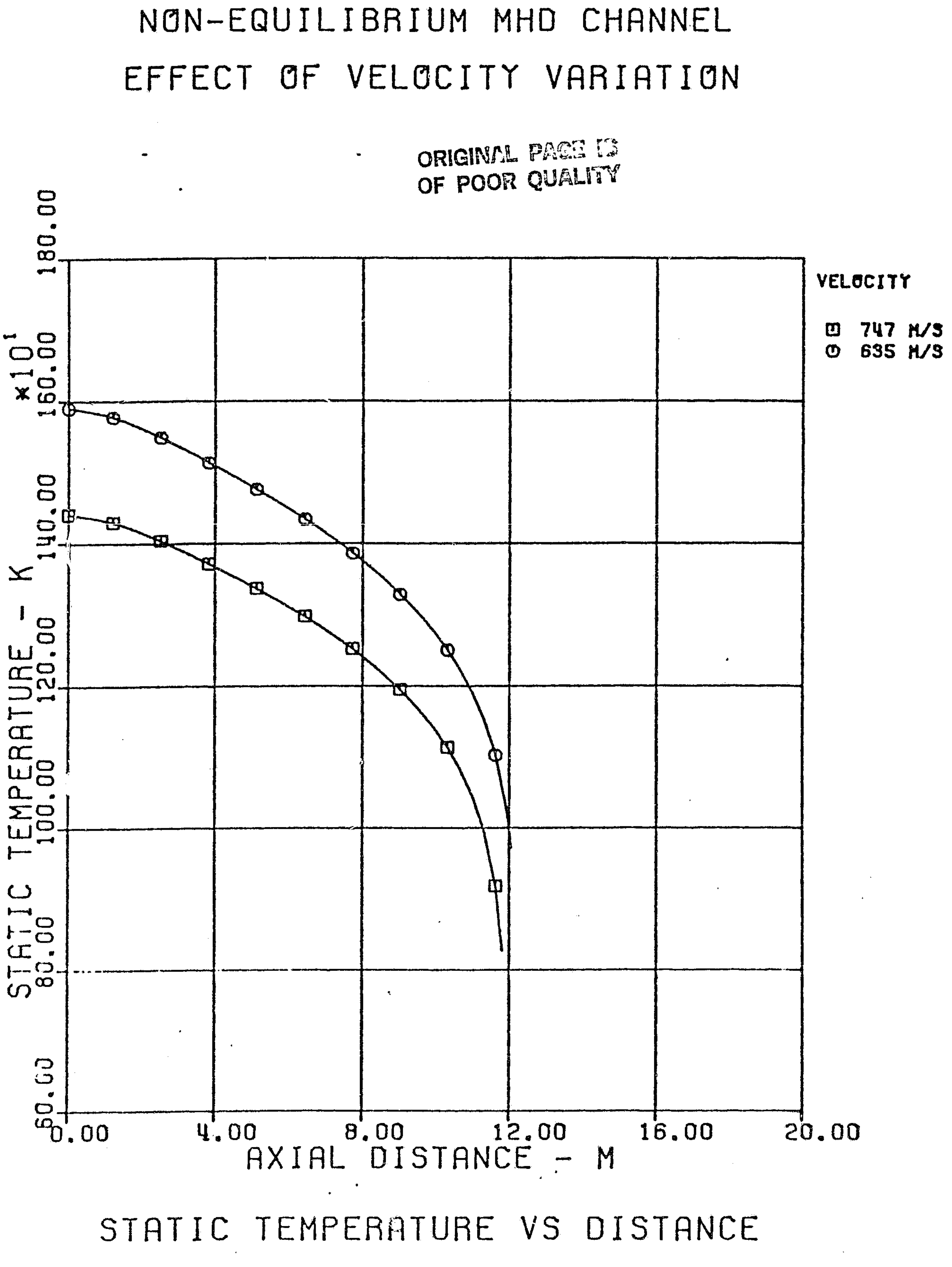
THERMAL INPUT: 2465 MWT 635 M/S
MW_T=68566
P_c/P_D=4.88 W=2778.5 L=12.1 L/D=6.8 A_e/A_i=3.09 MW_E=874.9
BASE CASE--THERMAL INPUT: 747 m/s
MW_T=68566
P_c/P_d=4.82 W=2778.5 L=11.8 L/D=6.7 A_e/A_i=3.24 MW_E=867.9
0.00
4.00
8.00
12.00
16.00
20.00
AXIAL DISTANCE - M
W=FLOW RATE,KG/S
L=LENGTH,M
MW_E=ELECTRICAL OUTPUT
MW_T=THERMAL INPUT
P_c/P_D=COMBUSTOR TO DIFFUSER EXIT PRESSURE RATIO
L/D=LENGTH TO INLET DIAMETER RATIO
A_e/A_i=EXIT TO INLET AREA RATIO

Figure A-12


NON-EQUILIBRIUM MHD CHANNEL
EFFECT OF VELOCITY VARIATION
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STATIC TEMPERATURE - K ×10[1]
180.00
160.00
140.00
120.00
100.00
80.00
60.00
0.00
4.00
8.00
12.00
16.00
20.00
AXIAL DISTANCE - M
VELOCITY
747 M/S
635 M/S
STATIC TEMPERATURE VS DISTANCE

Figure A-13

NON-EQUILIBRIUM MHD CHANNEL

EFFECT OF VELOCITY VARIATION

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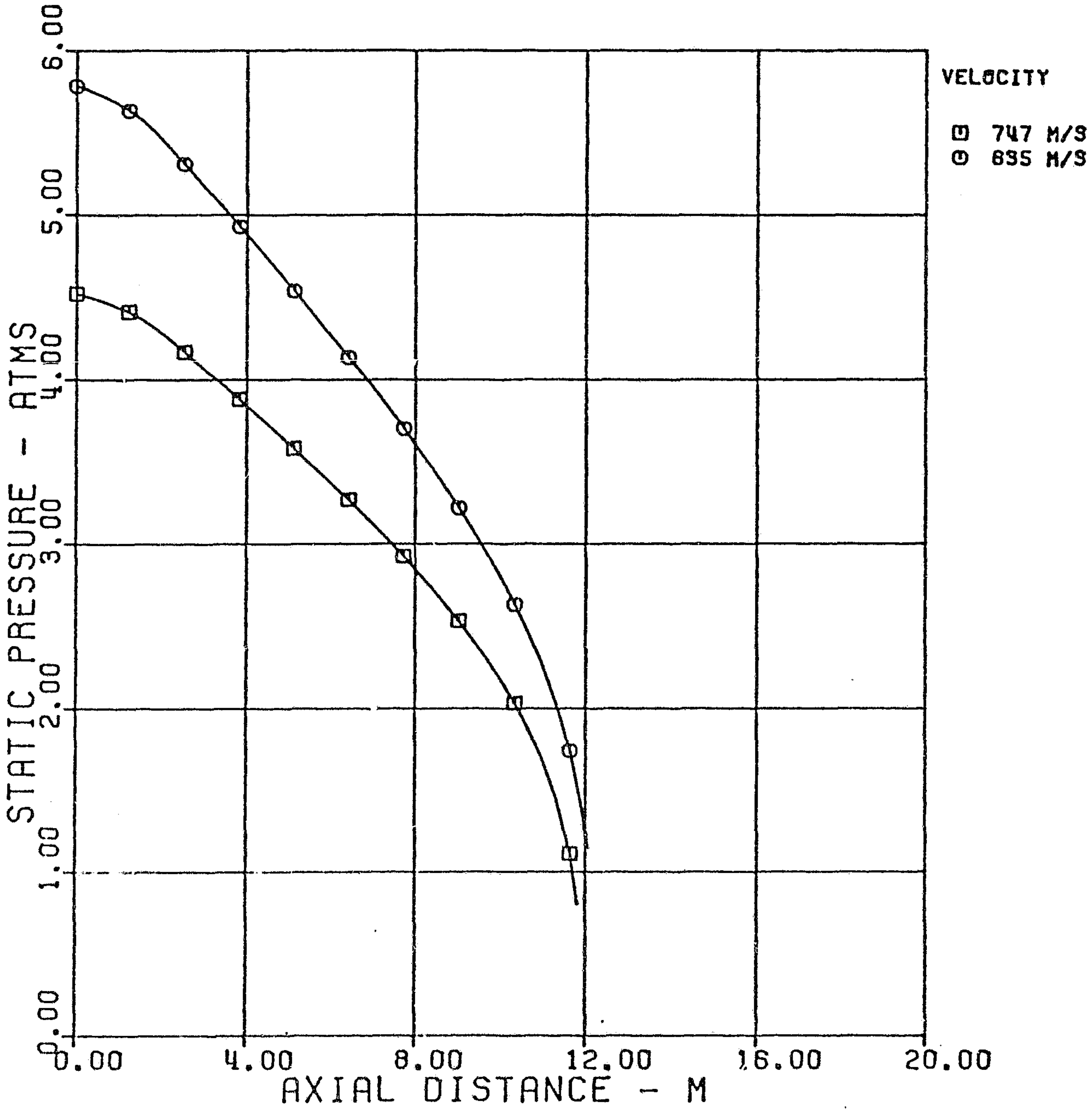
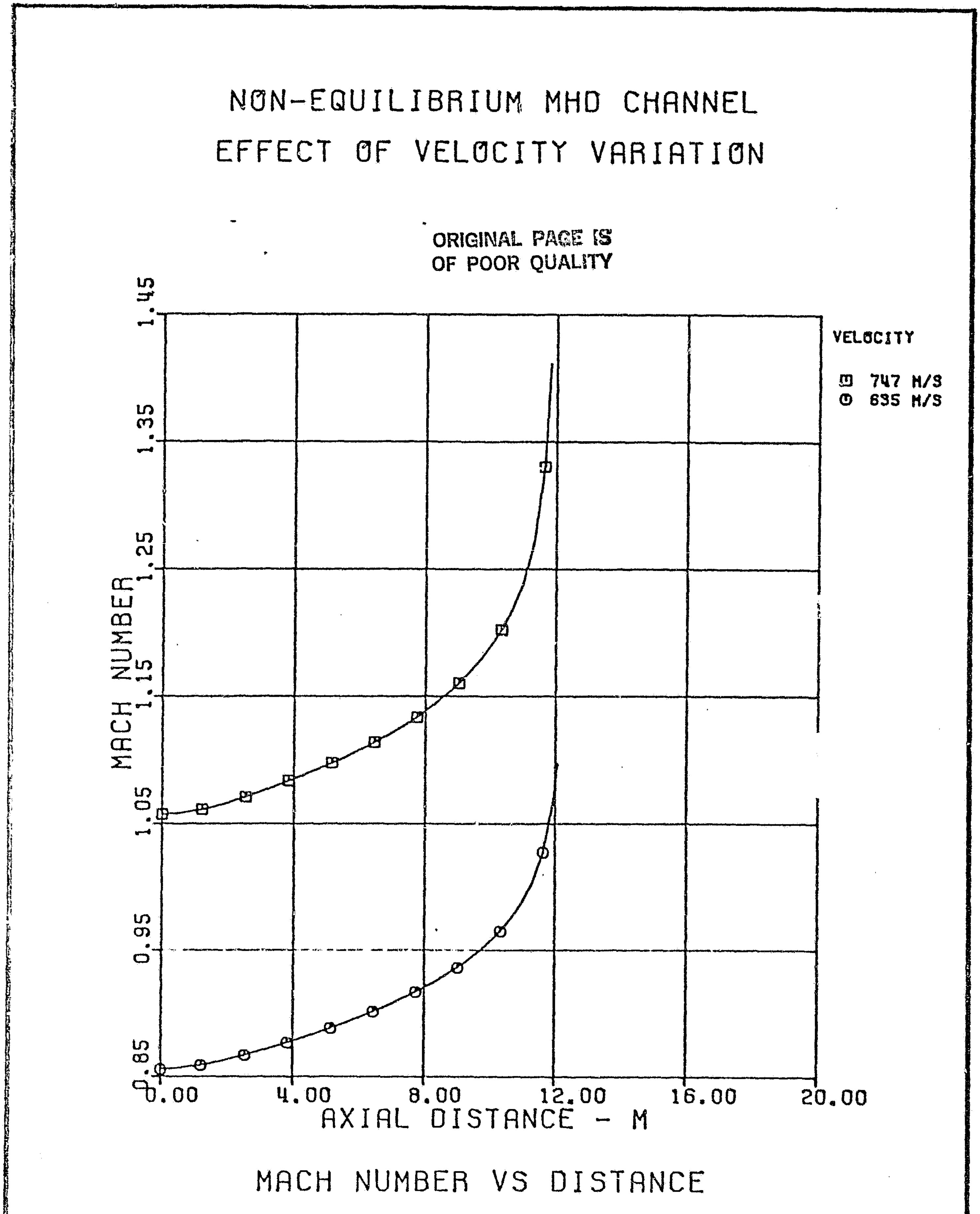


STATIC PRESSURE VS DISTANCE

Figure A-14


NON-EQUILIBRIUM MHD CHANNEL
EFFECT OF VELOCITY VARIATION
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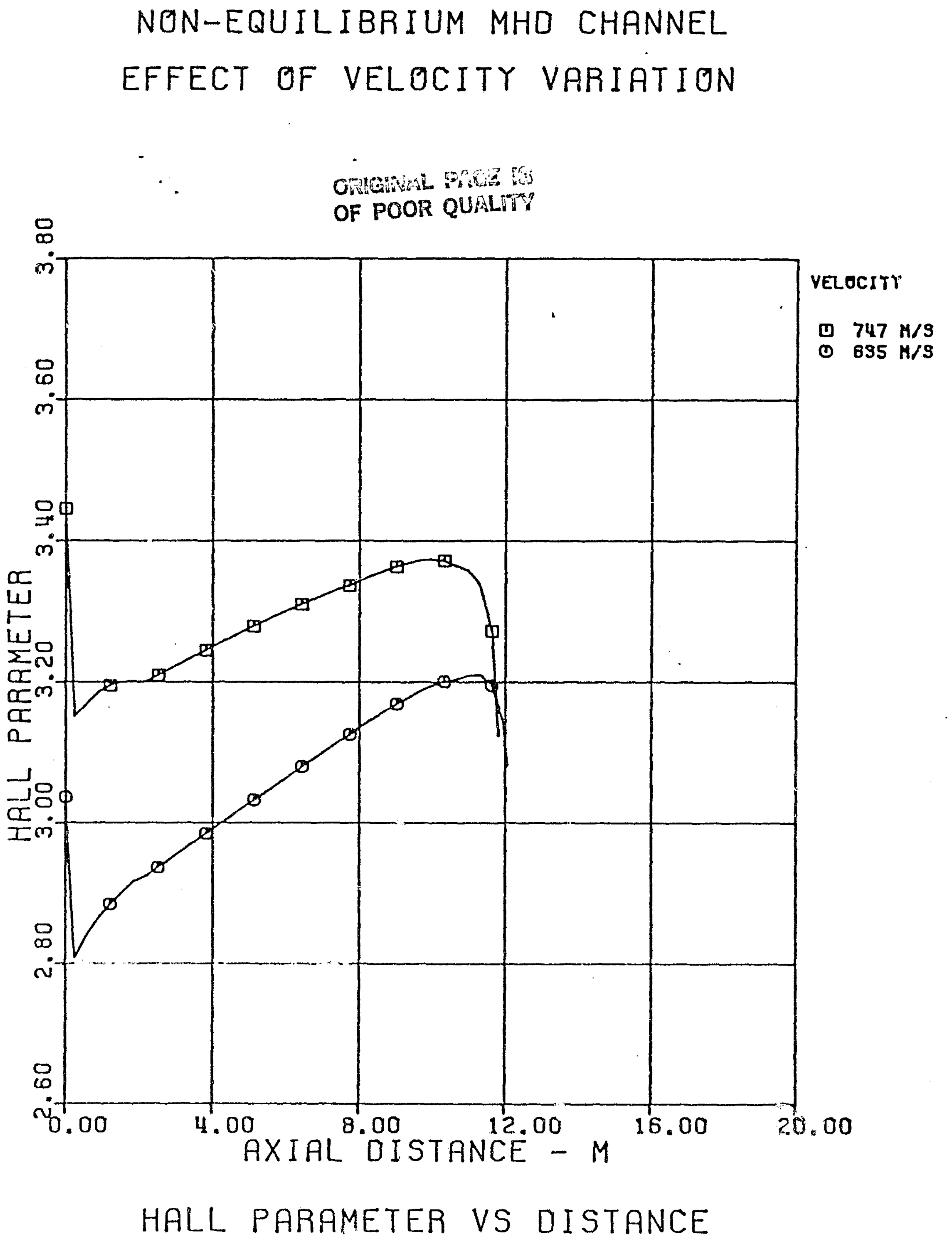
VELOCITY
747 M/S
635 M/S
MACH NUMBER
1.45
1.35
1.25
1.15
1.05
0.95
0.85
0.00
4.00
8.00
12.00
16.00
20.00
AXIAL DISTANCE - M
MACH NUMBER VS DISTANCE

Figure A-15


NON-EQUILIBRIUM MHD CHANNEL
EFFECT OF VELOCITY VARIATION
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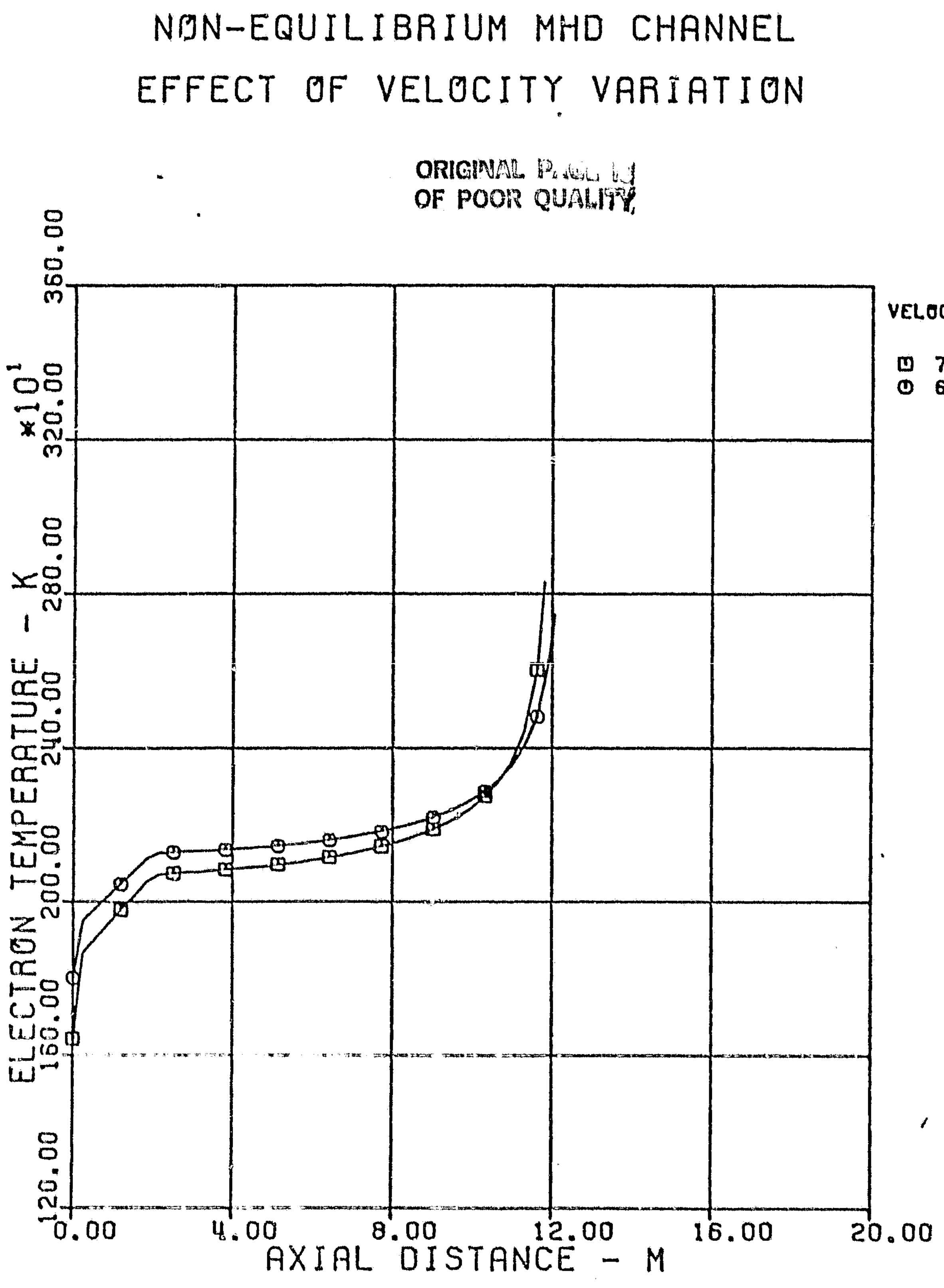
VELOCITY
747 M/S
635 M/S
HALL PARAMETER
2.60
2.80
3.00
3.20
3.40
3.60
3.80
0.00
4.00
8.00
12.00
16.00
20.00
AXIAL DISTANCE - M
HALL PARAMETER VS DISTANCE

Figure A-16



NØN-EQUILIBRIUM MHD CHANNEL
EFFECT ØF VELØCITY VARIATIØN
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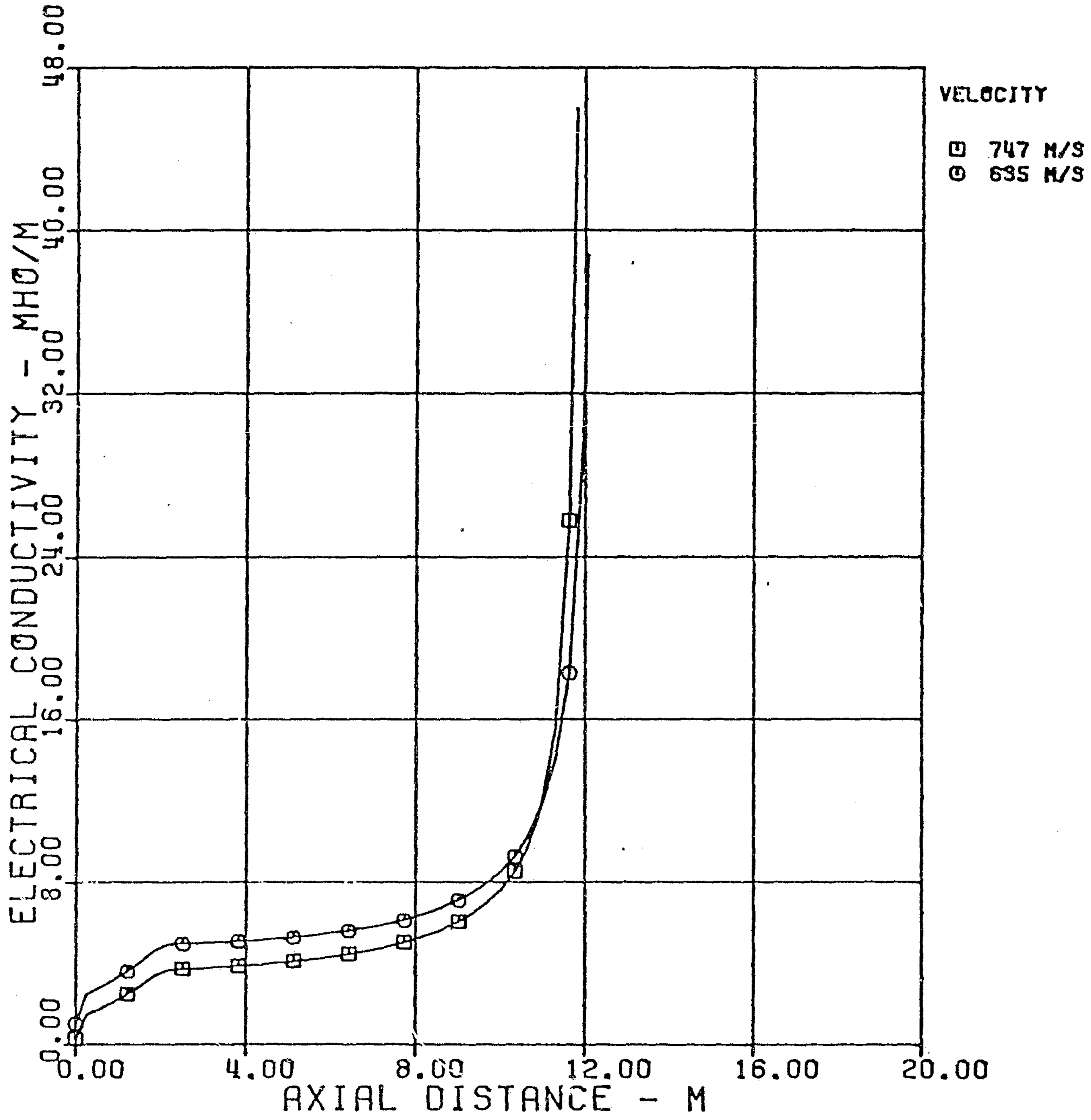
VELØCITY
747 M/S
695 M/S
ELECTRØN TEMPERATURE - K *10[1]
120.00
160.00
200.00
240.00
280.00
320.00
360.00
0.00
4.00
8.00
12.00
16.00
20.00
AXIAL DISTANCE - M
ELECTRØN TEMPERATURE VS DISTANCE

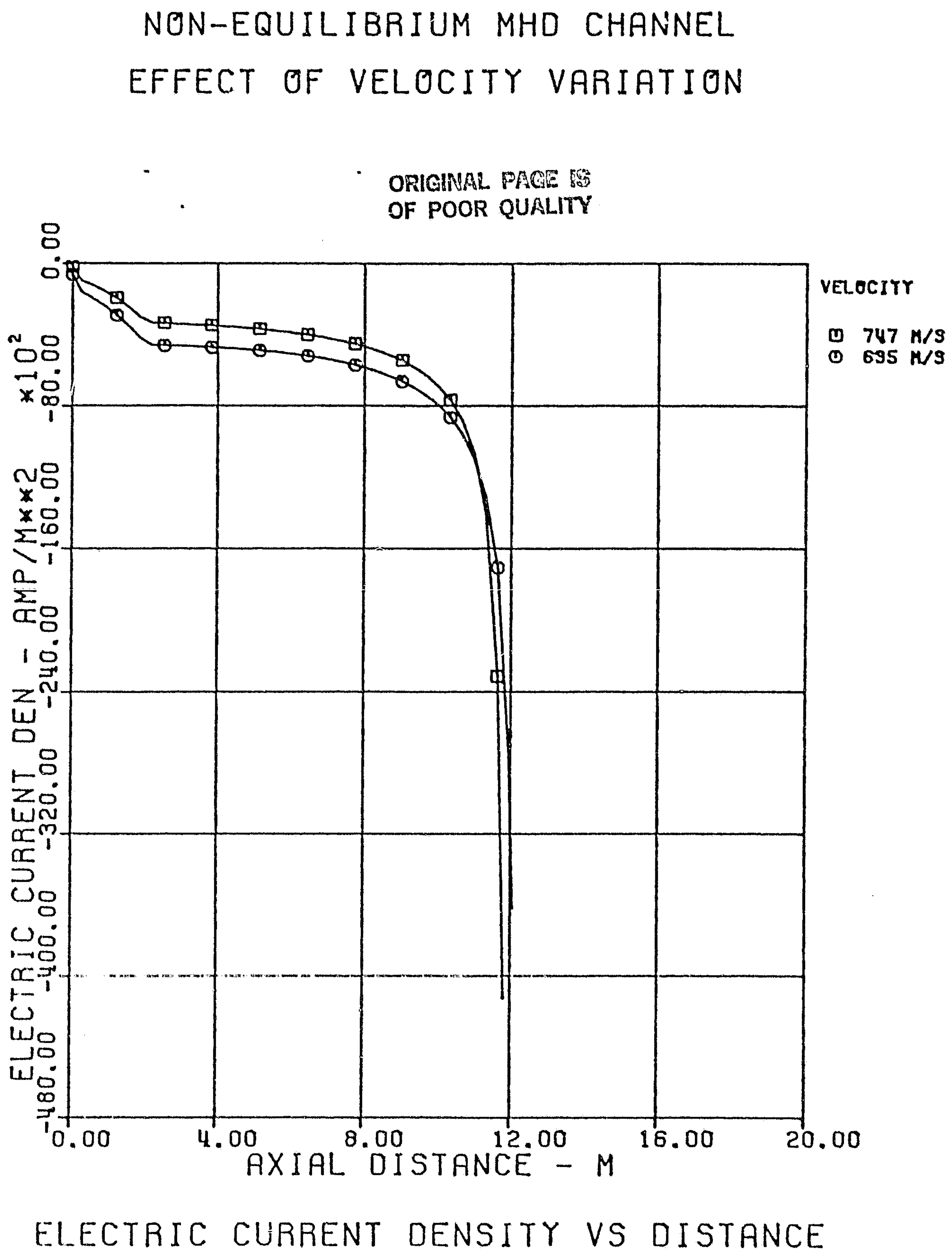
Figure A-17

NON-EQUILIBRIUM MHD CHANNEL
EFFECT OF VELOCITY VARIATION



ELECTRICAL CONDUCTIVITY VS DISTANCE

Figure A-18


NON-EQUILIBRIUM MHD CHANNEL
EFFECT OF VELOCITY VARIATION
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VELOCITY
747 M/S
695 M/S
ELECTRIC CURRENT DEN - AMP/M**2 ×10²
0.00
-80.00
-160.00
-240.00
-320.00
-400.00
-480.00
0.00
4.00
8.00
12.00
16.00
20.00
AXIAL DISTANCE - M
ELECTRIC CURRENT DENSITY VS DISTANCE

Figure A-19

same design gas velocity of 747 meter/sec (supersonic) in each of the channel runs. The other parameters such as the gas flow rate, inlet stagnation pressure, and temperature, and the exit stagnation pressure were held constant in each of the channel computations.

Results from this investigation are summarized in Table A-10. As the electrical loading parameter increased from 0.77 to 0.81, both the channel length L and L/D ratio increased by a factor of nearly 3 (length from 5.5 meters to 16.2 meters, L/D from 3.1 to 9). The electrical power output was nearly the same at 870 MWe. Since the electrical loading parameter is a measure of how effectively the channel is operating, as the K-parameter was increased, both the power extraction and the overall efficiency increased.

A.4.2 Alternate Sizes

The effect of varying the thermal input on the channel performance and size was evaluated in this study. The required change in the thermal input was accomplished by changing the plasma flow rate to the channel entrance. The thermal inputs considered were 2465 MW_t, 1233 MW_t, 616 MW_t, and 247 MW_t. For each of the thermal input, a channel run was made with the following parameters held constant: inlet stagnation pressure, 10 atm; inlet stagnation temperature, 3100° F; exit stagnation pressure, approximately at 2 atm; design gas velocity, 747 meters/sec. In all of these four channel computations, the channel length to diameter ratio was held constant within 10 percent. This was accomplished by varying the electrical loading parameter (from K = 0.8 at 2465 MW_t to K = 0.76 at 247 MW_t).

The results from this alternate size study are summarized in Table A-11. The electrical power output ranged from 868 MWe at 2465 MW_t case to a value of 85 MWe at 247 MW_t case. The overall

TABLE A-10: Effect of K-Factor Variation

Thermal Input	2465 MW_t
Velocity	747 M/sec
Inlet Stagnation	10 Atm, 3100°F
Exit Stagnation	2 Atm (approx.)

Electrical Loading Factor K	L (L)	L/D	Channel Area In. (M^2)	Channel Area Ex. (M^2)	Power (MW_e)	Power Extraction (%)	Overall Eff. (%)
0.77	5.5	3.11	2.43	7.88	870	31.7	67.8
0.80	11.8	6.7	2.43	7.88	868	33.0	70.7
0.81	16.2	9.2	2.43	7.97	872	33.7	71.9

TABLE A-11: Alternate Sizes

Inlet Stagnation	10 Atm, 3100°F
Exit Stagnation	2 Atm (approx.)
Velocity	747 M/sec

Thermal Input (MW_t)	L (M)	L/D	K	Channel Area In. (M^2)	Channel Area Ex. (M^2)	Power (MW_e)	Channel Efficiency (%)	Overall Eff. (%)	Power Extraction (%)
2465	11.80	6.7	0.8	2.43	7.88	868	87.9	70.7	33.0
1233	8.25	6.6	0.788	1.22	4.1	442	87.2	70.2	33.4
616	6.3	7.2	0.778	0.61	2.06	218	86.4	69.4	33.1
247	4.2	7.5	0.76	0.243	0.797	85	84.4	67.2	31.6

efficiency decreased from 70.7 percent for the largest thermal input case to a value of 67.2 percent for the smallest.

A comparison of the approximate channel sizes is illustrated in Figure A-20. Figures A-21 through A-28 represent the comparisons of the important channel parameters (i.e., static temperature, static pressure, cross sectional area and Hall parameter etc.).

A.4.3 Inlet Stagnation Temperature Variation Effect

The effect of varying the inlet stagnation temperature, between 3000°F and 3100°F, on the channel performance was investigated. The computation was performed with the following parameters held constant at the same values as the base channel case: plasma flow rate, inlet stagnation pressure, exit stagnation pressure, electrical loading parameter, and the plasma velocity.

The results from this investigation are summarized in Table A-12. The electrical power output decreases to 826 MWe from the base case value of 868 MWe. However, the overall efficiency and the power extraction remained nearly at the same values as those from the base case.

A.5 COST OF MAGNET AND CHANNEL

Costs for the magnet and channel were estimated using DOE/MHD cost estimation procedure and with the assumption that the construction materials and methods for closed cycle components would be similar to those for the open cycle.

Magnet. The magnet cost was estimated to be $36.8 million in mid-1978 dollars.

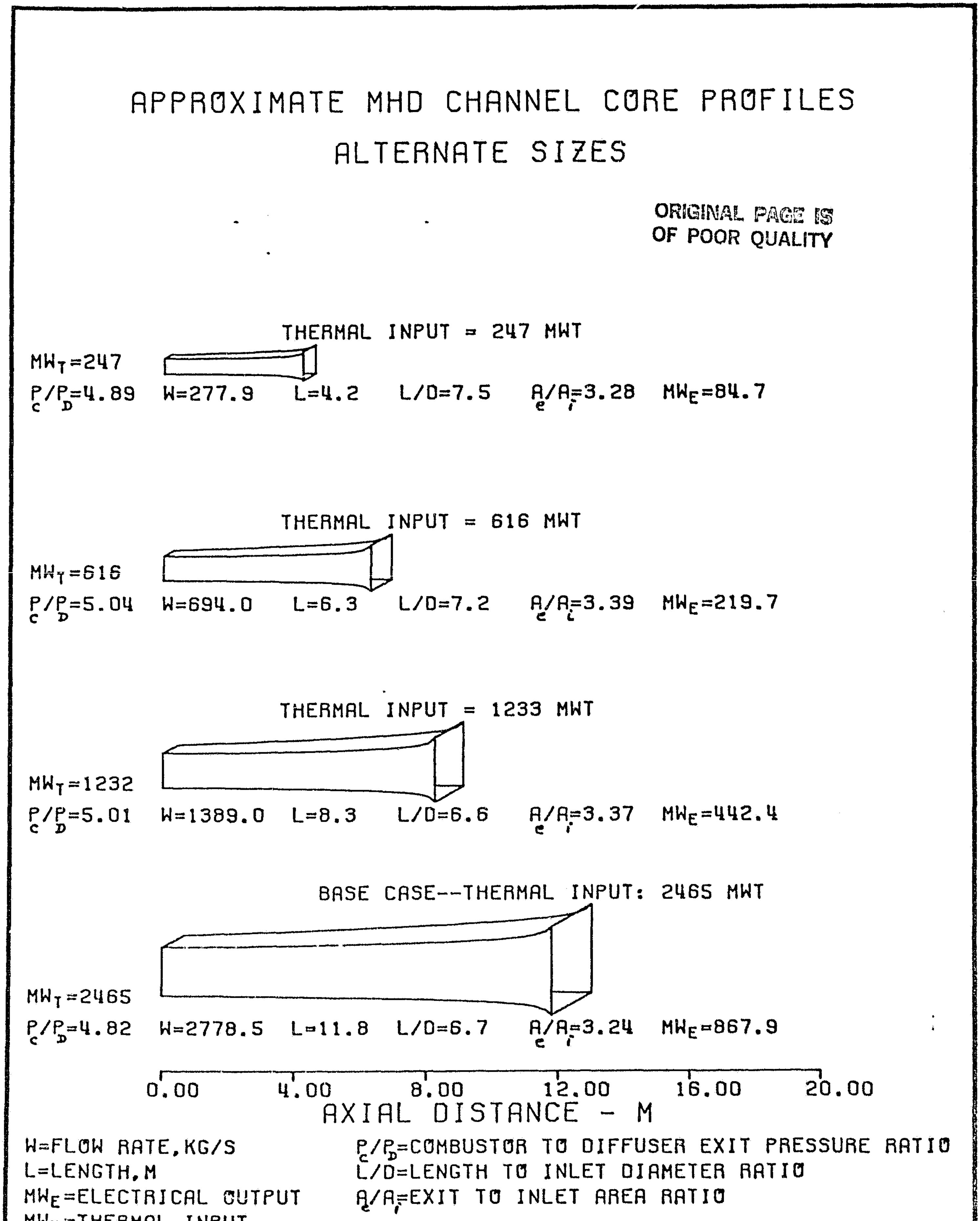
APPROXIMATE MHD CHANNEL CORE PROFILES
ALTERNATE SIZES
THERMAL INPUT = 247 MWT
MW_T=247
P_C/P_D=4.89 W=277.9 L=4.2 L/D=7.5 A_e/A_i=3.28 MW_E=84.7
THERMAL INPUT = 616 MWT
MW_T=616
P_C/P_D=5.04 W=694.0 L=6.3 L/D=7.2 A_e/A_i=3.39 MW_E=219.7
THERMAL INPUT = 1233 MWT
MW_T=1232
P_C/P_D=5.01 W=1389.0 L=8.3 L/D=6.6 A_e/A_i=3.37 MW_E=442.4
BASE CASE--THERMAL INPUT: 2465 MWT
MW_T=2465
P_C/P_D=4.82 W=2778.5 L=11.8 L/D=6.7 A_e/A_i=3.24 MW_E=867.9
0.00 4.00 8.00 12.00 16.00 20.00
AXIAL DISTANCE - M
W=FLOW RATE, KG/S
L=LENGTH, M
MW_E=ELECTRICAL OUTPUT
MW_T=THERMAL INPUT
P_C/P_D=COMBUSTOR TO DIFFUSER EXIT PRESSURE RATIO
L/D=LENGTH TO INLET DIAMETER RATIO
A_e/A_i=EXIT TO INLET AREA RATIO

Figure A-20

NØN-EQUILIBRIUM MHD CHANNEL

ALTERNATE SIZES

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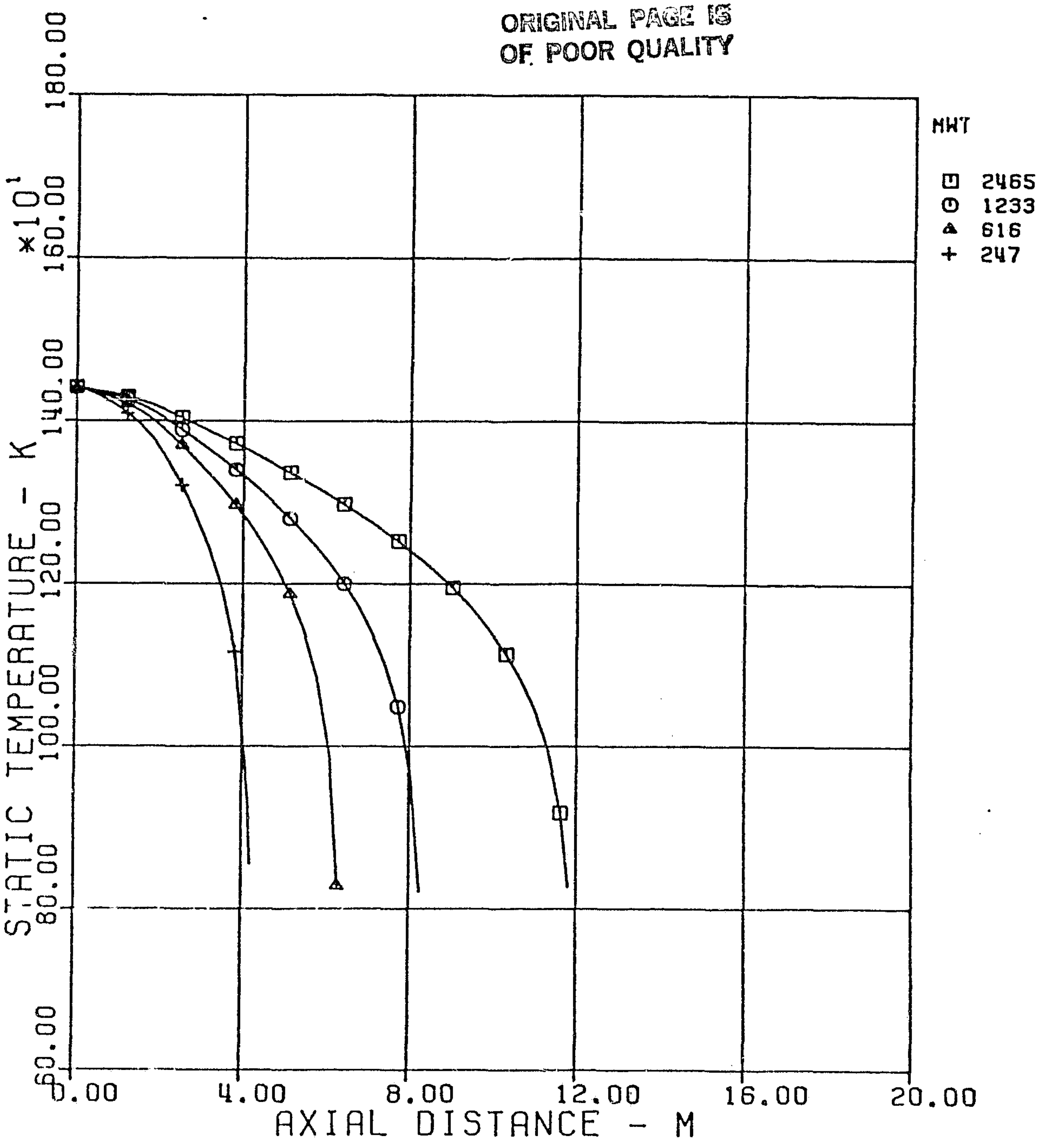
OF POOR QUALITY



STATIC TEMPERATURE VS DISTANCE

Figure A-21

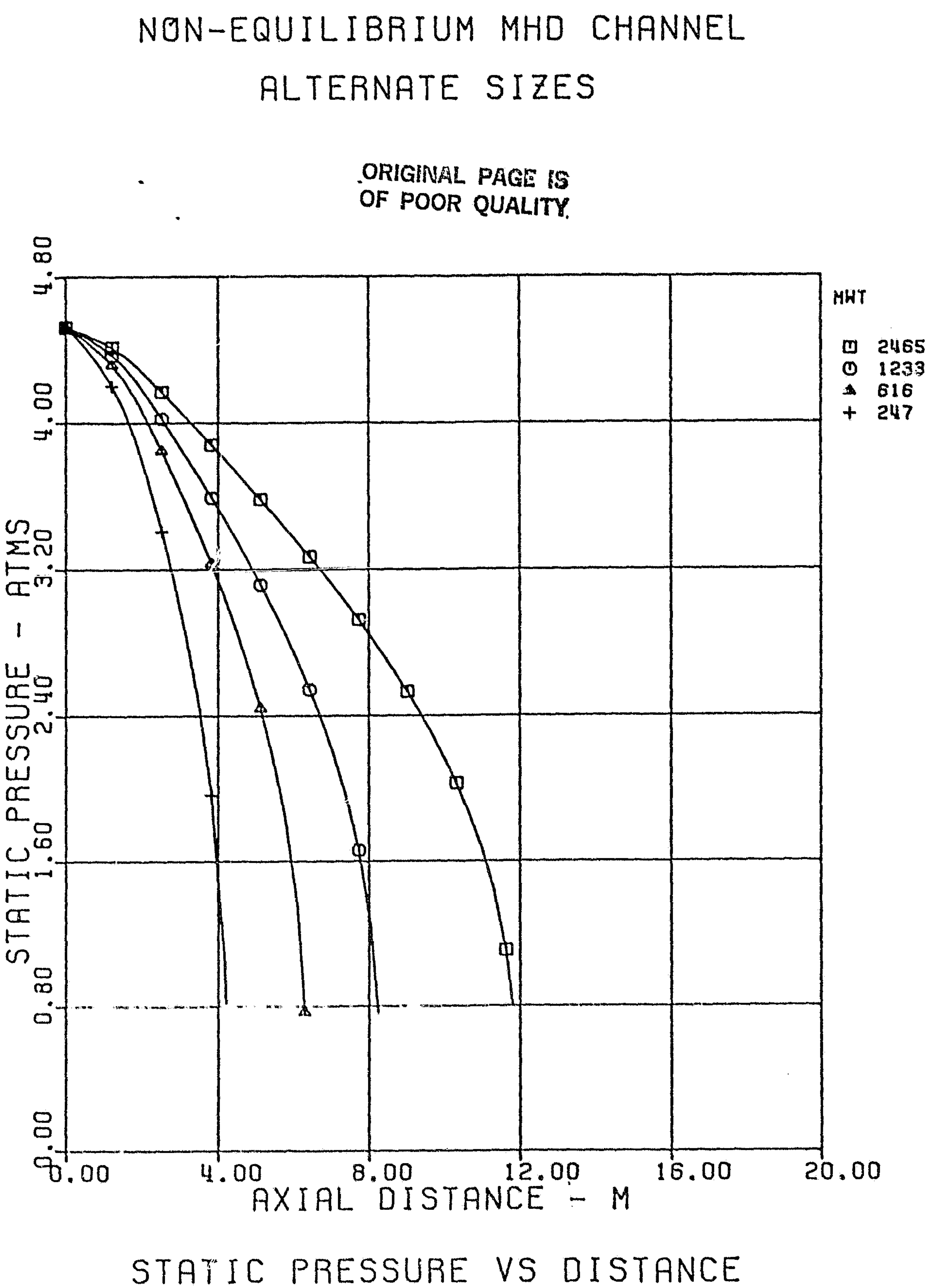
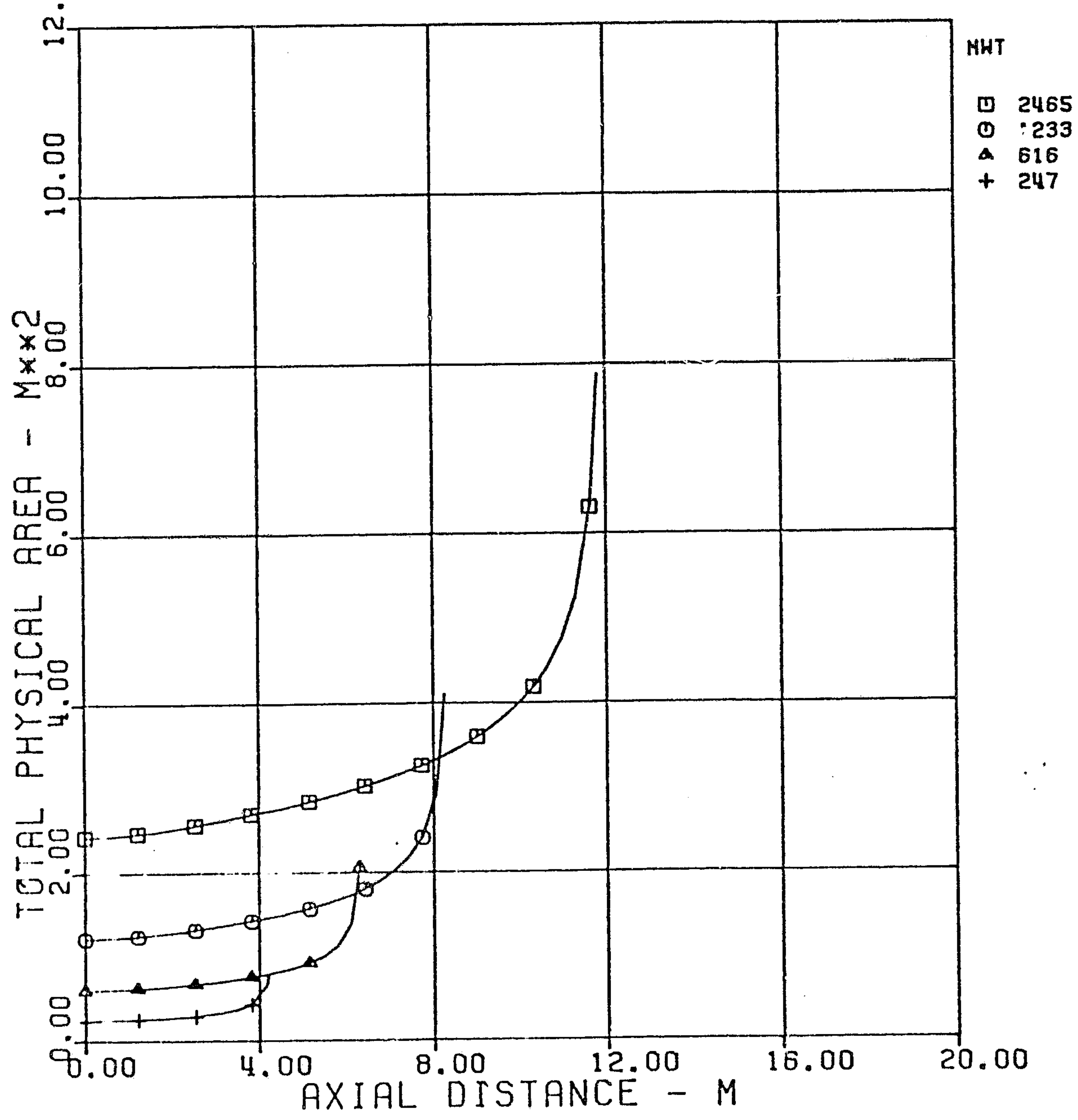

NØN-EQUILIBRIUM MHD CHANNEL
ALTERNATE SIZES
ORIGINAL PAGE IS
OF POOR QUALITY
MWT
2465
1233
616
247
STATIC PRESSURE - ATMS
0.00
0.80
1.60
2.40
3.20
4.00
4.80
0.00
4.00
8.00
12.00
16.00
20.00
AXIAL DISTANCE - M
STATIC PRESSURE VS DISTANCE

Figure A-22

NON-EQUILIBRIUM MHD CHANNEL

ALTERNATE SIZES



TOTAL PHYSICAL AREA VS DISTANCE

Figure A-23

NØN-EQUILIBRIUM MHD CHANNEL
ALTERNATE SIZES

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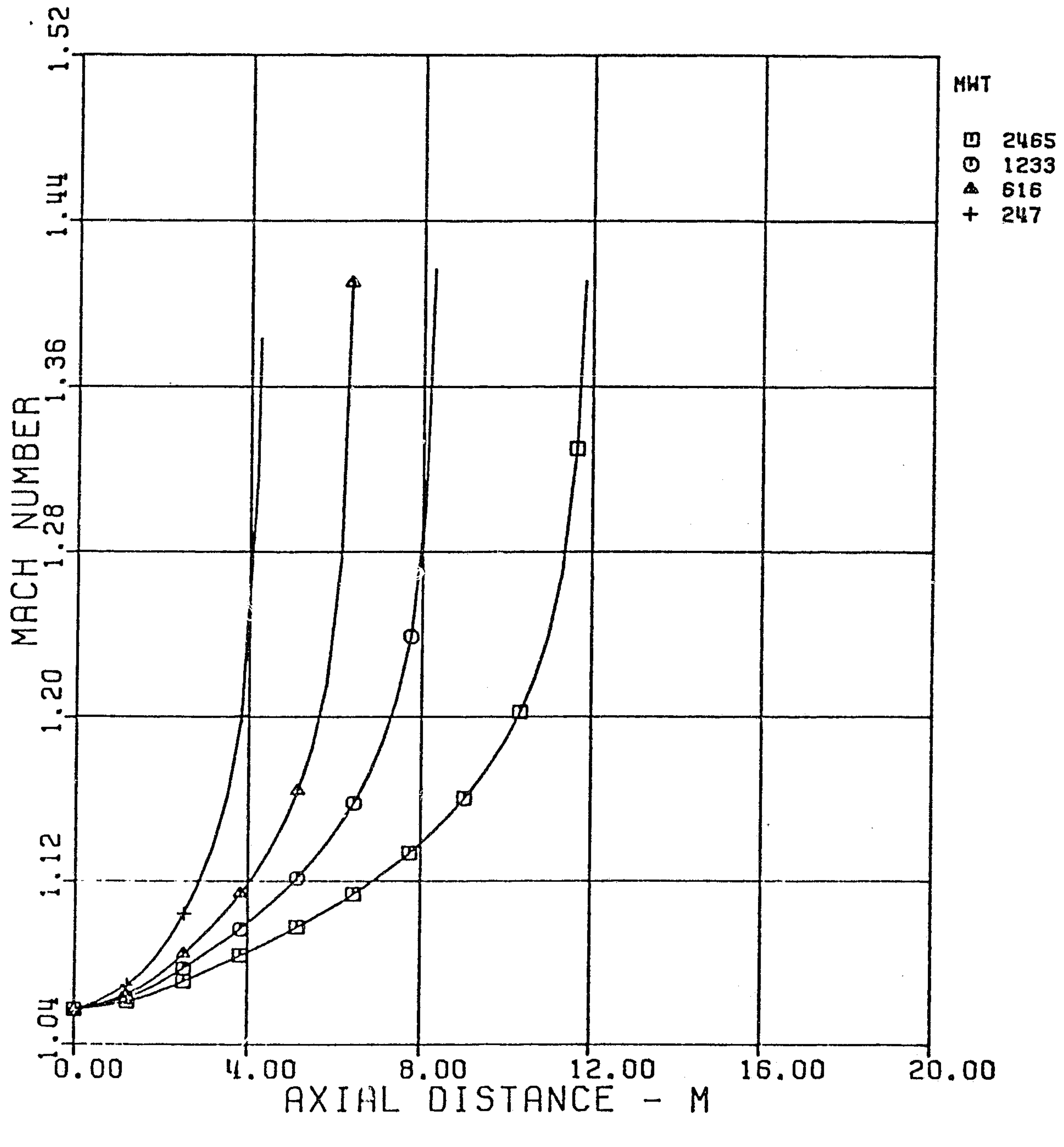
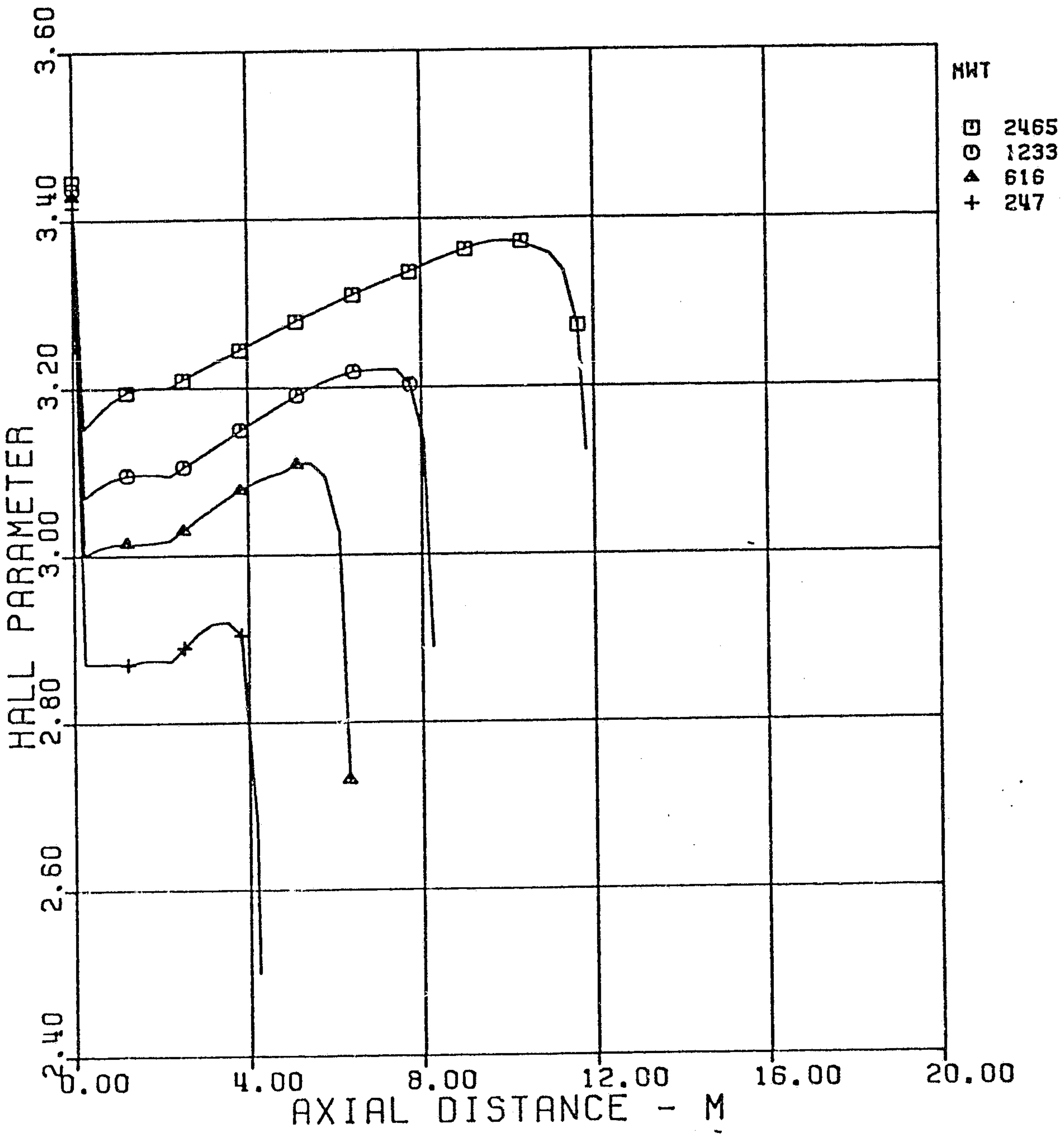


MACH NUMBER VS DISTANCE

Figure A-24

NØN-EQUILIBRIUM MHD CHANNEL

ALTERNATE SIZES



HALL PARAMETER VS DISTANCE

Figure A-25

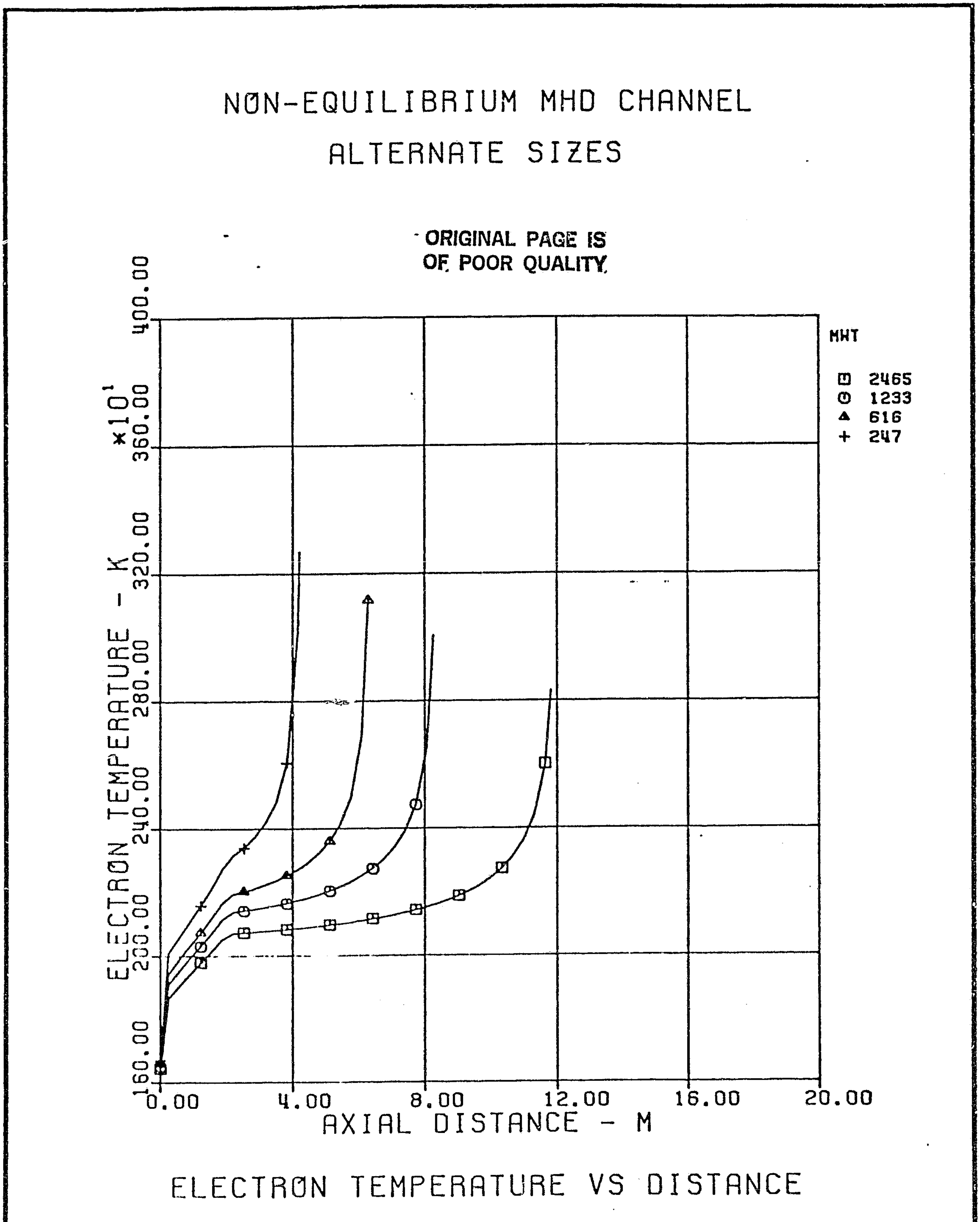

NON-EQUILIBRIUM MHD CHANNEL
ALTERNATE SIZES
ORIGINAL PAGE IS
OF POOR QUALITY
ELECTRON TEMPERATURE - K *10^1
160.00
200.00
240.00
280.00
320.00
360.00
400.00
0.00
4.00
8.00
12.00
16.00
20.00
AXIAL DISTANCE - M
MWT
2465
1233
616
247
ELECTRON TEMPERATURE VS DISTANCE

Figure A-26

NON-EQUILIBRIUM MHD CHANNEL

ALTERNATE SIZES

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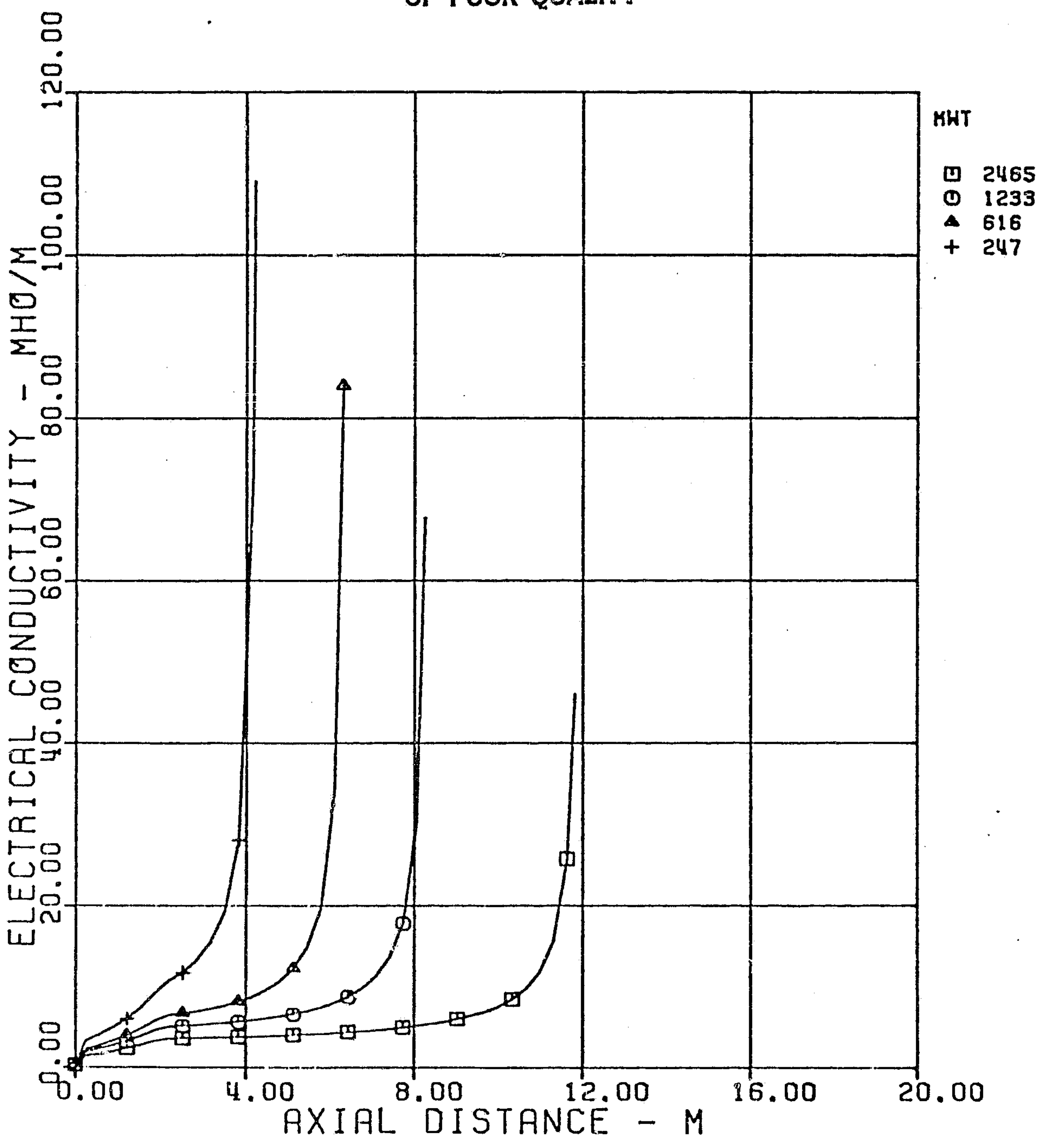

ELECTRICAL CONDUCTIVITY VS DISTANCE

Figure A-27

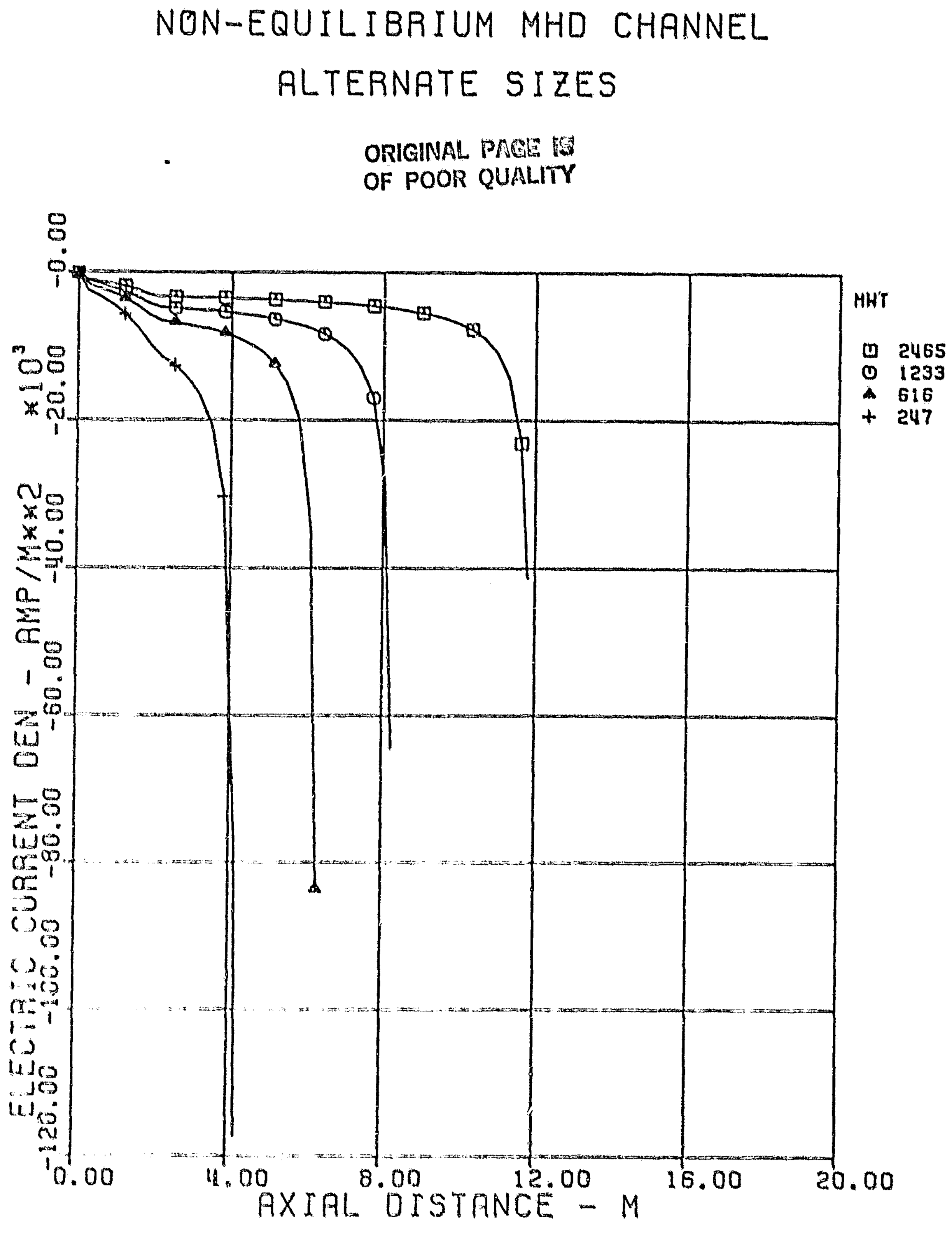
NON-EQUILIBRIUM MHD CHANNEL
ALTERNATE SIZES
ORIGINAL PAGE IS
OF POOR QUALITY
ELECTRIC CURRENT DEN - AMP/M**2 *10³
-0.00
-20.00
-40.00
-60.00
-80.00
-100.00
-120.00
0.00
4.00
8.00
12.00
16.00
20.00
AXIAL DISTANCE - M
MWT
2465
1233
616
247
ELECTRIC CURRENT DENSITY VS DISTANCE

Figure A-28

TABLE A-12: Variation of Inlet Stagnation Temperature

Plasma Flow Rate	2778.5 kg/sec
Inlet Stagnation Pressure	10 Atm
Exit Stagnation Pressure	2 Atm (approx.)
Electrical Loading Factor	0.8
Velocity	747 M/sec

Temperature at Noz. Inlet (°F)	L (M)	L/D	Thermal Input (MW_t)	Mach No. In.	Mach No. Ex.	Power (MW_e)	Power Extr. (%)	Overall Eff. (%)
3100	11.80	6.7	2465	1.057	1.411	868	33.0	70.7
3000	13.93	8.0	2384	1.078	1.444	826	32.7	70.6

The estimated cost was based on a magnet having the following characteristics:

Maximum field strength	=	6T
Magnet volume utilization	=	0.5
Channel inlet area	=	2.47 M^2
Channel exit area	=	7.62 M^2
Channel length	=	12.1 M

Channel. The channel cost was estimated to be $13.2 million in mid-1978 dollars.

The estimated cost was based on the configuration of a subsonic channel (as requested by NASA) with the geometry specified in the Magnet Section.

A.6 CONCLUSIONS

The following conclusions were reached as a result of this investigation:

1) The maximum power generation is approximately 870 MW_e for the base case channel with thermal input of 2465 MW_t. This power output, however, represents 13 percent less than that assumed in the earlier plant studies.

2) The base case channel's power extraction is approximately 33 percent, which is 3 percent less than that assumed in the earlier studies.

3) The base channel is 11.8 meters in length with an L/D ratio of 6.7.

4) The isentropic efficiency for the base channel is 88 percent.

5) The alternate-size-study resulted in the channel power output ranging from 868 MW_e at 2465 MW_t case to a value of 85 MW_e at 247 MW_t case.

6) For the case with an inlet stagnation temperature of 3000°F, the channel produces 826 MW_e of electrical power output, a 5% decrease from the base case.

7) The total cost, for the magnet and channel, was estimated to be $50.0 million in mid-1978 dollars.

A.7 REFERENCES

1. Sutton, G. W., and Sherman, A.: Engineering Magnetohydrodynamics, McGraw Hill Book Company, New York (1965).

2. Cool, T. A., and Zukoski, E. E.: "Recombination, Ionization, and Nonequilibrium Electrical Conductivity in Seeded Plasmas," Phys. Fluids 9, (1966).

3. Zauderer, B., Marston, C., and Cook, C.: "Closed Cycle MHD for Central Station Power with Fossil or Nuclear Stations," GE Report - 73SD231, Space Division, Valley Forge Center, (1973).

Nomenclature

P	Pressure
H	Enthalpy
Q	Heat Transfer
$J \times \bar{B}$	Lorentz Force
ρ	Plasma Density
V	Velocity
$J \cdot E$	Power Density
A	Cross Sectional Area
X	Distance
R	Gas Constant
W	Plasma Flow Rate
Jy	Current Density
σ_{Eff}	Effective Electrical Conductivity
Ne	Electron Density
Te	Electron Temperature
Tg	Gas Temperature
K	Boltzmann's Constant
γ_{eh}	Collision Frequency
m_h	Mass of Heavy Species
μ_{eh}	Electron Mobility
B	Magnetic Field Strength
E_y	Applied Voltage
N_h	Number Density of Gas Species H
N_s	Number Density of Seed
ξ	Turbulence Factor
β	Hall Parameter
$\beta_{Eff} = f(\beta, \xi)$	Effective Hall Parameter
Q_h	Collision Cross Section of the Heavy Species h
e	Electric Charge
N_e	Electron Density

APPENDIX - B

PARAMETRIC ANALYSIS OF CLOSED-CYCLE MHD POWER PLANTS - STUDIES OF NOBLE GAS REGENERATIVE HEATERS

1.0 INTRODUCTION

The NASA Lewis Research Center is supporting a study of closed-cycle MHD power plants. As a part of this effort, FluiDyne Engineering Corporation of Minneapolis, Minnesota has prepared an analysis of noble gas (argon) heaters under contract with Gilbert Associates, Inc. (Fluidyne Report 1223).

The basic requirement is to heat argon at 1069 Pa (155 psia) to 1979K (3100°F). Regenerative, ceramic heat exchangers were selected because of the high temperatures involved. These heat exchangers operate cyclically with so-called "on gas" and "on argon" cycles. When "on gas" the heater beds absorb heat from the reheat gas; when "on argon" the beds release heat to the argon. A system of valves is used for switching flows from "on gas" to "on argon" and vice versa. To provide a steady flow of heated argon requires a number of heaters. In the various systems analyzed, the number of heaters ranged from 5 to 20.

The purpose of the study was to examine the influence on the heater system of the method of firing the heaters and overall plant size. Three methods of firing were considered: (1) coal-fired combustors operating at either 1 atm or 6 atm, (2) gasifier that furnishes clean gas at 10 atm, and (3) gasifiers that furnishes clean gas at 1 atm. With the coal-fired combustors a 10% ash carryover to the heater system was specified. This required use of larger flow passages in the heater beds and special provisions to prevent clogging of the passages with slag.

A variety of cases were examined of which eight were analyzed in detail. The directly coal-fired cases apply to a 1000 MWe plant. The gasifier cases apply to plant sizes of 100, 250, 500, and 1000 MWe. Some cases furnished by Gilbert Associates were very similar and therefore separate detailed calculations were not made for each.

The results presented include a description of the flow conditions and operating sequence, size of components, solution of refractory materials, estimated heater system cost (mid-1979 dollars), and a discussion of the development needs corresponding to each method of firing.

2.0 HEATERS FIRED DIRECTLY WITH COAL

2.1 General Considerations

With direct coal-firing the combustors are attached directly to the heaters (either one per heater or one for the entire heater system). The major significance is that slag carryover from the combustor enters the heaters and must not be allowed to clog the heaters. The melting point of coal slag is in the temperature range over which the beds operate. Thus the slag will condense and tend to clog the flow passages.

Three methods of operation can, in principle, prevent clogging and are being studied in the national MHD program. First, the slag can be allowed to condense and build up over a period of time until the heater pressure drop becomes excessive. Then the heater would be taken off-line and the bottom position heated to a high enough temperature to melt out the deposit. This method is being tested at General Electric Co. and Montana State University. Second, the temperature cycling of the lower part of the bed can be designed so that, in each cycle, the temperature rises high enough to melt the deposit accumulated during each cycle. This method has been tested successfully by FluiDyne Engineering Corporation at subscale for the case where seed (potassium sulfate) and slag are present. In this case the seed appears to flux the slag. Third, a fluxing agent can be added to reduce the melting point and viscosity of the slag. This method could be combined with either of the others.

The specified operating conditions did not permit use of the second method (which would not necessarily have been chosen anyway) and therefore the first method was selected. The hole size for the cored brick beds was chosen as 1.5 inch diameter in order to allow some space for slag accumulation.

The possibility of using a fluxing agent was not considered, because this method has not been explored to any significant degree and could not be used within the scope of this study.

Heater performance depends to a large degree upon the length to diameter ratio of the holes. The hole diameter was selected as 0.5 inch for the heaters fired with clean gaseous fuel. With 1.5 inch holes the direct coal-fired heater beds are much longer than those fired with clean fuel.

In the early part of the study it was necessary to make certain changes in the heater operating conditions from those originally specified. The original specification included a reheat gas flow that was significantly less than the argon flow. This caused a large drop in the reheat gas temperature and required a long (high effectiveness) heater bed. (As noted above, the large hole diameter also increased heater length.) Subsequently, following discussions with Gilbert Associates, the reheat gas flow rate was increased. This change in flow rate and the changes needed to accommodate the slag cleaning methods are noted in Table 1.1.

The resulting heater configuration is not entirely satisfactory. The thermal effectiveness is too high which would make the performance very sensitive to heat traces, flow maldistribution, and other secondary effects. Additional study of the interrelationship between the plant and heater would be needed to improve the configuration. Nevertheless, the results are suitable for a first estimate of performance, size, and cost.

Case 1.1, which was specified, was not analyzed separately. A comparison of the operating conditions indicates that Case 1.0 closely approximates Case 1.1.

2.2 Description of Heater System

Two cases were analyzed, one with an atmospheric pressure combustor and the other with a pressurized combustor. Both cases apply to a 1000 MWe plant. The specified operating conditions are presented in Table 1.1. As discussed previously, changes from the original specifications were made and are noted. Design variables selected by FluiDyne are also shown in Table 1.1. The size of the major components, natural selection, weight, cost, and other pertinent informaton is given in Tables 1.2 through 1.5.

As described in Section 2.1, the beds are very long because of the large hole diameter and the specified flow conditions. The other parameter that fixes the total bed size is the flow per unit bed area. This is limited by either thermal stress or pressure drop. In both cases the pressure drop is controlled. That is, the beds are sized to give the allowable pressure drop as indicated in Table 1.1. At those conditions the thermal stress will be less than the allowable value. However, in making these calculations the effect of accumulated slag in the beds was ignored. There were no special requirements introduced to maintain heater-to-heater flow equality under conditions of slag accumulation. These assumptions are appropriate to the accuracy of the cost estimate.

The diameter of the ducts, manifolds, and valves, is based on the allowable pressure drop or the maximum velocity allowed by erosion considerations (about 200 ft/sec). In Case 1.1 the erosion limit applied; therefore, an increase in allowable pressure drop would not cause a reduction in duct size and cost. On the other hand, for Case 1.4 the duct diameters were limited by pressure drop. In this case an increase in the allowable values would give a reduction in duct size and cost. However, cost of the ducts, manifolds, and valves is only 16% of the total cost (Table 1.4) and the potential reduction is not large.

The number of heater vessels (Table 1.2) depends primarily on the bed diameter selected. In both cases bed diameters were selected that are somewhat less than the maximum used in blast furnace stoves. The bed diameter also fixes the valve diameters (Table 1.3). Again, the valve sizes are approximately equal to the maximum used with blast furnace stoves (up to about 8 ft). Each system includes two "non-flow" heaters, one for flow switching over, and one for cleanout. No standby heaters were included.

The idealized ripple shown in Table 1.2 is an estimate of the argon temperature fluctuation at the exit of the heater system. This estimate is the individual heater temperature droop (Table 1.1) divided by the number of heaters on argon. The values are acceptably low. The corresponding values are also given for the combustion gas. They would be modified slightly by capacitance effects in the ducts and manifolds.

The amount of gas stored in each heater at different times during the cycle is given in Table 1.5. Especially important is the utilization of residual argon. When a heater is switched from argon to combustion gas, the residual argon could be vented. However, the makeup requirements may be too costly and reuse of this argon may be necessary. The possible need for purification and its implications on cost have not been examined.

The ceramic materials for various parts of the system are identified in Table 1.3. High purity alumina was assumed for the beds and the highest temperature regions, and castable materials were used for the insulation. These selections are discussed in Section 2.3.

Estimates of weight and costs are presented in Table 1.4. The costs are in mid-1979 dollars and do not include the contractors overhead and profit. Also, it may be desirable to recover energy from the high pressure combustion gas stream in Case 1.4. This gas stream will be laden with slag particulates. Energy recovery equipment is not included in the cost estimate.

The costs for Case 1.1 and 1.4 are the highest of all cases studied. This results from the large amount of heater bed material. As described earlier, beds are large because of the large holes needed to permit slag accumulation and because of the specified operating requirements. Revisions to this operating requirements would reduce costs.

2.3 Development Needs

Development work is needed in four areas to show technological feasibility and to develop the data base for design. These areas are (1) ceramic materials, (2) operability of the heater system, i.e. preventing clogging of the heater passages with slag, (3) bed support, and (4) valves.

Materials: The heater bed, hot gas inlet ducts and manifolds, upper vessel dome, vessel liner, and lower vessel plenum must be constructed of a material which will resist corrosion/erosion by molten slag at temperatures up to 2130K (3370 F), as well as being cost effective with acceptable mechanical properties. The major problem is the very high temperatures. Operating experience and test data at these conditions is very limited.

Chrome-bearing, fused cast refractories usually have the best slag resistant properties at very high service temperatures. However, they are expensive and have poor thermal shock resistance. Preliminary tests of high alumina materials have shown some swelling due to slag pick-up, but continuing tests at Montana State University indicate that this material has considerable potential. Further tests at subscale and with various slags are needed.

A three-layer castable insulation scheme was selected for this study. Castable materials are becoming more common but have not been applied in comparibly severe conditions. Very limited tests suggest that a material of sufficient slag resistance with similar mechanical properties and cost can be developed. Also needed is a suitable method of anchoring the castable liner to the castable back-up layers and further tests at subscale.

Operability: Slag present in the hot gas stream will accumulate in the argon heaters and associated ducting. A clean-out procedure will be required in which the slag is melted and flows out of the heaters. The frequency and duration of the clean-out cycle must be determined, and will be dependent on many factors including the specific ash characteristics of the coal.

As noted in Section 2.1, preliminary tests with a subscale heater (20 ft high bed) are being done at Montana State University. Additional testing at this scale would be needed, followed by tests at a larger scale in order to progressively move to the commercial plant size.

Bed Support: The foregoing discussion points out the need for a bed support which will endure temperatures up to free flowing slag temperatures while not obstructing the slag flow. The choice of bed support used for this study was a cooled metal grate which has active cooling only during clean-out cycles. During normal operation it is simply a "cold bottom" metallic bed support which reduces operating heat loss. Development work is needed to determine the accumulation of slag on cooled surfaces during the clean-out cycle, materials of construction, and overall structural integrity. Tests at or near full-scale will be needed to verify satisfactory operation during both normal and clean-out thermal cycling. The use of fluxing additives to reduce the slag melting point and viscosity would also need to be tested.

Another choice would be to split the heater matrix into multiple sections. This would reduce the amount of material that is exposed to molten slag. Refractory supported sections (less than 25 feet in height) could be built to limit the support stresses on the ceramic dome used to support the matrix. This design would have the potential of at least partial clean-out during each cycle. The balance of the matrix would only see "dry slag" particles and would have a "cold bottom" and no clean-out cycle requirement. Such a design is complex and was not attempted for this study.

<u>Valves</u>: Six valves are required for the operation of each heater in the regenerative heater system. These are: combustion products inlet and outlet, argon inlet and outlet, and two smaller valves to accommodate fluid changeover. The combustion gas inlet valve has the most severe service. This valve will require some development and the others will require verification testing.

A test of a small scale prototype gas inlet valve has been run at FluiDyne in a seed/slag environment with encouraging results. The test valve was a gate valve in which the gate and a follower ring form an integral structure which slides back and forth in the body of the valve. The follower ring protects the valve seal from fouling by the seed/slag in the hot gas. Both the gate/follower ring and the valve body are water-cooled and refractory-faced to protect them from the harsh service condition at minimum heat loss.

Testing of the other valves is anticipated as part of the air heater development program.

Summary: Each of the four issues described above are under development in the Department of Energy MHD heater development programs. Work being done at FluiDyne includes related seed/slag application work in all four areas. In addition, slag-only operability and experimental materials work is in progress at Montana State University and at General Electric. All of these efforts are coordinated and the participants actively interchange information.

3.0 HEATERS FIRED WITH A HIGH PRESSURE GASIFIER

3.1 General Description

The specified arrangement is a gasifier directly coupled to the heaters and with hot gas clean-up. This very compact matrix was chosen because the working fluids are clean, i.e., there are no constituents to condense out of the bulk stream and adhere to the matrix passages. This allowed the selection of 0.5 inch diameter flow passages which greatly reduces the bed sizes, as compared with the directly-fired cases. For systems of the size needed for MHD applications, matrix compactness is highly desirable and leads to substantial cost reductions. However, as with the directly-fired cases, the gas side flow was increased to avoid excessively high values of heat exchanger effectiveness (see Table 2.1). For all cases the effectiveness is approximately 0.9 and therefore the dimensions are feasible and performance would not be significantly affected by heat losses and other secondary effects.

3.2 Description of Heater System

Three cases were analyzed, for 1000, 500, and 250 MWe. The latter two had flow rates of one half and one quarter of the 1000 MWe case, respectively. Cases 2.3 and 2.6 were not analyzed separately because they were very similar to Case 2.0 after the combustion gas flow had been increased (as described in Section 3.1). The results of the design analyses are presented in Tables 2.1 through 2.5, in the format used for Cases 1.1 and 1.4.

As noted earlier, the clean gases allowed selection of a small hole diameter, 0.5 inch. A web thickness of 0.25 inch was considered but 0.375 inch was selected (Table 2.1) to avoid having a shortened cycle time (Table 2.2).

The designs were not thermal stress limited; therefore, the controlling sizing parameters were allowable pressure loss. The allowable pressure loss was sufficient and did not have a significant size or cost impact. The systems were sized for clean, but roughened matrix flow passages, with a balanced distribution between manifold losses and parallel leg losses.

Smaller bed diameters (Table 2.3) were chosen than in Cases 1.1 and 1.4. Larger diameters could be used to reduce the number of heaters.

However, the cost of the system is relatively insensitive to the total number of heaters. The optimum number will be dependent on the design of the interface manifold between the heater and adjacent components. Only one heater was provided for fluid changeover (or switching); and there is no need for a clean-out heater; and a stand-by heater was not included.

The idealized argon temperature ripple (Table 2.2) is low so that passive control should be acceptable. The ideal ripple does not include manifold capacitance or other secondary effects on system output.

A partial refractory list is given in Table 2.3. With clean combustion products, the choice of materials is much broader. High-density and high-purity alumina can be used for the hot insulation and the hot portion of the bed. The cooler parts of the bed were specified to be of lower cost materials which significantly reduced matrix costs. The specified inlet temperatures are possible, but still on the high side of what is achievable in a long-life, full-scale system. The upper portion of the bed may have limited life due to creep. Available creep data are limited, but suggest that 3200 F might be the upper limit for tall beds.

A castable three-layer insulation option was specified, and this is a development item.

The valves (Table 2.3) are well within the state-of-the-art with respect to dimensions, and the leakage estimates may be conservatively high for a clean environment. The only development question concerns heat loss estimates with a refractory lining compatible with the 3350 F gas inlet temperature.

Costs are presented in Table 2.4. The pressurized gasifier heater system is the smallest of the three systems studied, and therefore has the lowest cost.

In Table 2.5 residual mass data and time averaged flow rates are provided.

3.3 Development Needs

The development needs for clean-fired argon heaters are significantly reduced when compared to the direct coal-fired heaters, the most similar industrial equipment to blast furnace stoves. Significant differences are the higher temperature, higher thermal effectiveness, larger physical size, and operating requirements associated with electric power generation.

Measurements of high temperature creep are needed to assure satisfactory life of the ceramics. Tests of blast furnace valves at high temperatures are needed. Tests of the proposed castable insulation is needed. The castable insulation could be replaced with bricks. This would increase the cost of Case 2.0 (1000 MWe) by about 15 million dollars.

Following small-scale tests, a test of a heater module at sufficient sizes to permit scaling to full plant size would be needed to verify the materials and design.

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4.0 HEATERS FIRED WITH AN ATMOSPHERIC PRESSURE GASIFIER

4.1 General Considerations

The requirements for these systems is very similar to those for the pressurized gasifier systems except that the combustor and associated ducting operates at atmospheric pressure. Again the fuel gas is assumed to be clean, allowing small holes in the heater beds.

The operating conditions and geometric constraints are given in Table 3.1. The exceptions taken to the received operating conditions are noted. As with the other cases, the main exception was to increase combustion product flow rate (see Section 3.1). The hole pattern for the heater beds is identified with that of the pressurized gasifier cases.

4.2 Description of Heater System

Salient system operational characteristics and discussions are provided in Tables 3.2 and 3.3. The bed diameters for Cases 3.0 and 3.3 were chosen to be about that of blast furnace stoves; this fixed the number of heaters. Because of the relatively short cycle time, two heaters were needed to provide the fluid changeover time interval for Case 3.0.

The idealized outlet temperature ripple is suitably low for Case 3.0, but may be too high for Cases 3.3 and 3.7. The ripple could be reduced by use of a passive capacitor or by active control.

Matrix and valve dimensions and a partial refractory list is given in Table 3.3. Ceramic materials are identical to those selected for the pressurized gasifier systems (see Section 3.2).

Estimated system costs are given in Table 3.4 with costs again expressed in mid-1979 dollars.

4.3 Development Needs

The development needs for these systems are identical to those discussed for the pressurized gasifier systems in Section 3.3.

5.0 SUMMARY

A comparison of the eight systems is presented in Table 4. The unit costs provides a basis for approximate scaling to other sizes. The effect of the large holes (1.5 inch) selected for the directly-fired cases, combined with the specified operating conditions, required the largest heater beds and therefore the highest cost. These cases could be optimized to yield a smaller system and lower cost.

TABLE 1.1

Specified Conditions and System Constraints

Direct Coal-Fired Cases

Specified Conditions	Case 1.0		Case 1.4	
Plant Size, MW_e	1000		1000	
Comb. Products Inlet Temperature, K (°F)	2127	(3369)	2129	(3372)
Comb. Products Outlet Temperature, K (°F)	766	(918) (1)	781	(946) (1)
Argon Inlet Temperature, K (°F)	640	(692) (1)	478	(700) (1)
Argon Outlet Temperature, K (°F)	1982	(3107)	1981	(3105)
Comb. Products Inlet Pressure, kPa (psi)	119	(17.3)	605	(87.7)
Argon Inlet Pressure, kPa (psi)	1069	(155)	1069	(155)
Comb. Products Flow Rate, kg/sec, (lbm/sec)	1114	(2447) (1)	1038	(2528) (1)
Argon Flow Rate, kg/sec, (lbm/sec)	2762	(6093)	2770	(6093)
Allowable Comb. Products Pressure Drop, kPa, (psi)	6.2	(0.9)	7.7	(1.12) (1)
Allowable Argon Pressure Drop kPa, (psi)	53	(7.7)	53	(7.7)
System Constraints (set by FluiDyne)				
Allowable Thermal Stress, kPa (psi)	15500	(2250)	15500	(2250)
Hole Diameter, mm (in)	38	(1.5)	38	(1.5)
Web Thickness, mm (in)	19	(0.75)	19	(0.75)
Heater Argon Temperature Droop (2), K (°F)	111	(200)	111	(200)

(1) Droop is the range in delivered air temperature for a single heater.

(2) Changed from original specification.

TABLE 1.2

Operational Characteristics
Direct Coal-Fired Cases

Heater Distribution	Case 1.0		Case 1.4	
On Argon	3		5	
On Combustion Products	11		9	
Changeover (Press/Depress)	1		1	
Cleanout	1		1	
Standby	0		0	
Total	16		16	
Cycle Times				
Full, sec	4030		2650	
Argon, sec	806		883	
Combustion Products, sec	2956		1589	
Valving, sec	105		76	
Fluid Changeover, (Hot-to-Cold), sec	82		50	
Fluid Changeover, (Cold-to-Hot), sec	82		50	
Ripple - Idealized				
Period, sec.	269		177	
Argon Outlet Temperature, K (°F)	±19	(±33)	±11	(±20)
Combustion Products Outlet Temperature Temperature, K (°F)	±4	(±8)	±6	(±10)
Losses (Excluding Changeover and Cleanout)				
Valve Heat Loss, MW_e (10^6 Btu/hr)	18	(60)	10	(34)
Valve Leakage, kg/sec (lbm/sec)	15	(33)	10	(21)
Other Heat Losses, MW_e (10^6 Btu/hr)	41	(140)	30	(94)

TABLE 1.3

Heater Dimensions and Materials

Direct Coal-Fired Cases

Heater Bed	Case 1.0		Case 1.4	
Height m (ft)	37	(122)	41	(133)
Diameter m (ft)	7.8	(25.6)	6.1	(20)
Valve Flow Diameter				
Gas Inlet, m (ft)	3.35	(10.8)	2.1	(7.0)
Gas Outlet, m (ft)	2.3	(7.4)	1.5	(5.0)
Argon Inlet, m (ft)	1.8	(6.0)	1.5	(5.0)
Argon Outlet, m (ft)	2.9	(9.4)	2.3	(7.6)

Materials for Case 1.0 and 1.4

Heater Beds: High Density Alumina

Insulation: Lightweight Castable (F), dense for hot slag exposure

	Layer 1	Layer 2	Layer 3
Comb. Products Inlet	3300	2800	2600
Comb. Products Outlet	2000	-	-
Argon Inlet	2000	-	-
Argon Outlet	3300	2800	2600
Bed Cylinder (Average Location)	2600	2000	-

TABLE 1.4

Estimated Heater System Cost*

Direct Coal-Fired Cases

	Case 1.0				Case 1.4			
	Mass		Cost 10^6\$		Mass		Cost 10^6\$	
Heaters	10^3 kg	(tons)	Material	Labor	10^3 kg	(tons)	Material	Labor
Bed	53600	(59000)	108	30.6	34100	(37500)	68.3	19.4
Insulation	2980	(3280)	1.7	2.7	2160	(2380)	1.1	1.9
Vessel	9850	(10835)	6.5	23.2	6060	(6670)	4.0	14.3
Ducts and Manifolds								
Insulation	4390	(4830)	4.3	2.8	2600	(2860)	2.5	1.61
Steel	1670	(1840)	1.1	3.9	1870	(2053)	1.2	4.9
Flanges	1420	(1560)	3.1		970	(1070)	1.8	
Expansion Joints			1.2				0.8	
Valves	833	(917)	18.4		562	(618)	13.0	
Bed Support	982	(1080)	3.8		461	(507)	1.8	
Structural Steel	11400	(12500)	8.8	23.8	7300	(8030)	5.6	15.3
Instrumentation/Controls			0.68				0.7	
TOTALS	86900	(95800)	157.6	87.0	56000	(61700)	100.8	57.4
TOTAL SYSTEM COST 10^6 \$			244.6				158.2	

* Direct costs are shown, i.e., contractors overhead and profit are not included. Installation is included in the labor component. All costs are in mid-1979 dollars.

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TABLE 1.5

Residual Fluid Mass Per Heater

Direct Coal-Fired Cases

Vessel Condition	Fluid	Case 1.0 Mass kg	Case 1.0 Mass (lbm)	Case 1.4 Mass kg	Case 1.4 Mass (lbm)
Beginning of Comb. Products Flow	Comb. Prod.	490	(1080)	1500	(3300)
End of Comb. Products Flow	Comb. Prod.	470	(1030)	1410	(3100)
Beginning of Argon Flow	Argon	6100	(13500)	3650	(8050)
End of Argon Flow	Argon	6400	(14000)	3870	(8300)

	Flow kg/sec	Flow (lbm/sec)	Flow kg/sec	Flow (lbm/sec)
Time Averaged Argon Exchange Per Cycle	24	(52.0)	21	(47)
Time Averaged Comb. Products Exchange Per Cycle	1.7	(3.8)	8	(17.5)

TABLE 2.1

Specified Conditions and System Constraints

High Pressure Gasifier Fired Argon Heaters

Specified Conditions	Case 2.0		Case 2.4		Case 2.5	
Plant Size, MW_e	1000		500		250	
Gas Inlet Temp., K (°F)	2117	(3350)	2116	(3350)	2116	(3350)
Gas Outlet Temp., K (°F)	787	(957) (1)	786	(955) (1)	787	(957) (1)
Argon Inlet Temp., K (°F)	640	(692)	640	(692)	640	(692)
Argon Outlet Temp., K (°F)	1981	(3105)	1980	(3105)	1980	(3105)
Gas Inlet Pressure, kPa (psi)	993	(144)	993	(144)	993	(144)
Argon Inlet Pressure, kPa (psi)	1069	(155)	1069	(155)	1069	(155)
Comb. Products Flow Rate, kg/sec (lbm/sec)	1057	(2325) (1)	528.2	(1162) (1)	264	(581) (1)
Argon Flow Rate, kg/sec (lbm/sec)	2770	(6093)	1385	(3047)	692	(1523)
Combustion Products Pressure Drop, kPa (psi)	48	(7.0)	48	(7.0)	48	(7.0)
Argon Pressure Drop, kPa (psi)	53	(7.7)	53	(7.7)	53	(7.7)
System Constraints						
Allowable Thermal Stress, kPa	15500	(2250)	15500	(2250)	15500	(2250)
Hole Diameter, mm (in)	12.7	(.5)	12.7	(.50)	12.7	(.5)
Web Thickness, mm (in)	9.5	(.375)	9.5	(.3750)	9.5	(.375)
Heater Air Temperature Droop (2) K (°F)	261	(470)	261	(470)	261	(470)

(1) Changed from original specification

(2) Droop is the range in delivered air temperature for a single heater.

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TABLE 2.2

Operational Characteristics

High Pressure Gasifier-Fired Argon Heaters

	Case 2.0	Case 2.4	Case 2.5
Heater Distribution			
On Argon	10	5	4
On Combustion Products	5	3	2
Changeover (Press/Depress)	1	1	1
Clean Out	0	0	0
Standby	0	0	0
Total	16	9	7
Cycle Times			
Full, sec.	1960	2029	2143
Argon, sec.	1225	1127	1225
Combustion Products, sec.	612	676	612
Valving, sec.	58	56	47
Fluid Changeover (hot to cold), sec.	32	85	130
Fluid Changeover (cold to hot), sec.	32	85	130
Ripple - Idealized			
Period, sec	122.5	225	426
Argon Outlet Temperature, K (°F)	13 (±24)	26 (±47)	33 (±59)
Combustion Products Outlet Temp., K (°F)	28 (±50)	46 (±83)	70 (±125)
Losses (Excluding Changeover & Clearout)			
Valve Heat Loss, MW_e (10^6 Btu/hr)	3.5 (20)	1.8 (10)	1.1 (6.1)
Valve Leakage, kg/sec (lbm/sec)	2.9 (6.4)	1.6 (3.5)	1.1 (2.4)
Other Heat Losses, MW_e (10^6 Btu/hr)	9.0 (50)	4.9 (27)	3.1 (17.4)

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TABLE 2.3

Heater Dimensions and Materials

High Pressure Gasifier-Fired Argon Heaters

	Case 2.0		Case 2.4		Case 2.5	
Heater Bed						
Height, m (ft)	15.6	(51)	15.4	(51)	15.5	(51)
Diameter, m (ft)	5.2	(17)	5.1	(17)	4.2	(13.6)
Valve Flow Diameter						
Comb. Products Inlet, m (ft)	1.7	(5.4)	1.5	(4.9)	1.3	(4.3)
Comb. Products Outlet, m (ft)	1.3	(4.2)	1.2	(3.9)	1.0	(3.4)
Argon Inlet, m (ft)	1.2	(4.0)	1.2	(4.0)	1.0	(3.2)
Argon Outlet, m (ft)	1.6	(5.3)	1.6	(5.3)	1.3	(4.2)

Materials (same for all three cases)

Heater Beds: High density 99% alumina

Insulation: Lightweight castable (°F)

	Layer 1	Layer 2	Layer 3
Comb. Products Inlet	3300	2800	2600
Comb. Products Outlet	2000	-	-
Argon Inlet	2000	-	-
Argon Outlet	3300	2800	2600
Bed Cylinders (Average Location)	2600	2000	-

TABLE 2.4

Estimated Heater System Cost*

High Pressure Gasifier-Fired Argon Heaters

	Case 2.0				Case 2.4				Case 2.5			
	Mass		Cost 10^6 $		Mass		Cost 10^6 $		Mass		Cost 10^6 $	
	10^3 kg	Tons	Materials	Labor	10^3 kg	Tons	Materials	Labor	10^3 kg	Tons	Materials	Labor
Heaters												
Bed	9200	(10200)	11.3	5.8	4800	(5300)	5.9	3.0	2530	(2800)	3.1	1.6
Insulation	1600	(1800)	1.2	1.6	880	(970)	.7	.8	506	(560)	.4	.5
Vessel	2700	(3000)	1.8	6.4	1420	(1600)	.9	3.3	730	(800)	.5	1.7
Ducts and Manifolds												
Insulation	1600	(1800)	1.5	0.8	880	(970)	.8	.4	570	(620)	.5	.4
Steel	1200	(1300)	.8	2.8	640	(700)	.4	1.5	380	(420)	.3	.9
Flanges	560	(620)	1.2		300	(320)	.7		180	(190)	.4	
Expansion Joints			.5				.3				.2	
Valves	400	(440)	9.7		220	(240)	5.2		138	(150)	3.4	
Bed Support	190	(210)	.7		94	(100)	.4		40	(45)	.2	
Structural Steel	2600	(2900)	1.7	5.8	1380	(1500)	1.1	2.9	760	(835)	.6	1.6
Instrumentation/Controls			.7				.4				.3	
TOTALS	20200	(22300)	31.1	23.2	10600	(11700)	16.8	11.9	5820	(6420)	9.9	6.7
TOTAL SYSTEM COST 10^6 $				54.3				28.7				16.6

*Direct costs are shown, i.e. contractors overhead profit are not included. Installation is included in the labor component. All costs are in mid-1979 dollars.

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TABLE 2.5

Residual Fluid Mass Per Heater

High Pressure Gasifier-Fired Argon Heaters

Vessel Condition		Case 2.0 Mass kg	(lbm)	Case 2.5 Mass kg	(lbm)	Case 2.5 Mass kg	(lbm)
Beginning of Comb. Products Flow	Comb. Prod.	1108	(2440)	1020	(224)	670	(1470)
End of Comb. Prod. Flow	Comb. Prod.	990	(2200)	910	(2010)	600	(1330)
Beginning of Argon Flow	Argon	159	(3500)	1467	(3230)	960	(2110)
End of Argon Flow	Argon	170	(3750)	1570	(3450)	1020	(2250)

	Flow kg/sec	(lbm/sec)	Flow kg/sec	(lbm/sec)	Flow kg/sec	(lbm/sec)
Time Averaged Argon Exchange Per Cycle	13.9	(30.6)	6.9	(15.3)	3.3	(7.4)
Time Averaged Comb. Prods. Exchange Per Cycle	8.1	(17.8)	4.1	(8.9)	2.0	(4.3)

TABLE 3.1

Specified Conditions and System Constraints

Low Pressure Gasifier-Fired Argon Heaters

Specified Conditions	Case 3.0		Case 3.3		Case 3.7	
Plant Size, MW_e	1000		500		100	
Comb. Products Inlet Temperature, K (°F)	2123	(3361)	2123	(3361)	2123	(3361)
Comb. Products Outlet Temp., K (°F)	788	(958)(1)	788	(958)(1)	788	(957)(1)
Argon Inlet Temp., K (°F)	640	(692	640	(692	640	(692)
Argon Outlet Temp., K (°F)	1978	(3100)	1978	(3100)	1978	(3100)
Comb. Product Inlet Pressure, kPa (psi)	119	(17.3)	119	(17.3)	119	(17.3)
Argon Inlet Pressure, kPa (psi)	1067	(154.8)	1067	(154.8)	1067	(154.8)
Comb. Products Flow Rate, kg/sec (lbm/sec)	1141	(2510)(1)	570.5	(1255)(1)	141	(251)(1)
Argon Flow Rate, kg/sec (lbm/sec)	2770	(6093)	1385	(3047)	277	(609)
Comb. Products Pressure Drop, kPa (psi)	8.3	(1.2)	8.3	(1.2)	8.3	(1.2)
Pressure Drop, kPa (psi)	52	(7.5)	52	(7.5)	52	(7.5)
System Constraints						
Allowable Thermal Stress, kPa, (psi)	15500	(2250)	15500	(2250)	15500	(2250)
Hole Diameter, mm (in)	12.7	(.5)	12.7	(.5)	12.7	(.5)
Web Thickness, mm (in)	9.5	(.375)	9.5	(.375)	9.5	(.375)
Heater Argon Temperature Droop(2), K (°F)	167	(300)	167	(300)	167	(300)

(1) Changed from original specification

(2) Droop is the range in delivered air temperature for a single heater.

Operational Characteristics

Low Pressure Gasifier-Fired Argon Heaters

	Case 3.0	Case 3.3	Case 3.7
Heater Distribution			
On Argon	4	2	1
On Combustion Products	14	7	3
Changeover (Press/Depress)	2	1	1
Cleanout	0	0	0
Standby	0	0	0
Total	20	10	5
Cycle Times			
Full, sec	1952	1952	2214
Argon, sec	390	391	443
Combustion Products, sec	1367	1367	1328
Valving, sec	84	84	57
Fluid Changeover (Hot to Cold), sec	55.5	55.5	193
Fluid Changeover (Cold to Hot), sec	55.5	55.5	193
Ripple - Idealized			
Period, sec	97.6	195	443
Argon Outlet Tempeature, K (°F)	±21 (±38)	±42 (±75)	±84 (±150)
Combustion Products Outlet Temperature, K (°F)	±6 (±11)	±12 (±22)	±28 (±51)
Losses (Excluding Changeover and Cleanout)			
Valve Heat Loss, MW_e (10^6 Btu)	9.1 (51)	4.5 (25)	1.1 (6.3)
Valve Leakage kg/sec (lbm/sec)	15.1 (33)	7.6 (17)	2.4 (5)
Other Heat Losses, MW_e (10^6 Btu/hr)	16.5 (93)	8.3 (47)	2.4 (13)

TABLE 3.3

HEATER DIMENSIONS AND MATERIALS

LOW PRESSURE GASIFIER-FIRED ARGON HEATER

	Case 3.0		Case 3.3		Case 3.7	
Heater Bed						
Height, m (ft)	10.6	(34.7)	10.6	(34.7)	10.4	(34.1)
Diameter, m (ft)	7.9	(25.8)	7.9	(25.8)	5.3	(17.5)
Valve Flow Diameter						
Comb. Prod. Inlet, m (ft)	2.5	(8.3)	2.5	(8.3)	1.7	(5.6)
Comb. Prod. Outlet, m (ft)	1.9	(6.2)	1.9	(6.2)	1.3	(4.2)
Argon Inlet, m (ft)	1.8	(5.8)	1.8	(5.8)	1.1	(3.7)
Argon Outlet, m (ft)	2.4	(7.9)	2.4	(7.9)	1.5	(5.0)

Materials (same for all three cases)

Heater Beds: High density 99% alumina

Insulation: Lightweight castable (°F)

	Layer 1	Layer 2	Layer 3
Comb. Products Inlet	3300	2800	2600
Comb. Products Outlet	2000	-	-
Argon Inlet	2000	-	-
Argon Outlet	3300	2800	2600
Bed Cylinders (Average location)	2600	2000	-

TABLE 3.4

Estimated Heater System Cost*

Low Pressure Gasifier - Fired Argon Heaters

	Case 3.0				Case 3.3				Case 3.7			
	Mass		Cost 10^6 \$		Mass		Cost 10^6 \$		Mass		Cost 10^6 \$	
	10^3 kg	Tons	Materials	Labor	10^3 kg	Tons	Materials	Labor	10^6 kg	Tons	Materials	Labor
Heaters												
Bed	17300	(19000)	17.9	10.9	8700	(9500)	8.9	5.5	1960	(2160)	2.0	1.2
Insulation	3800	(4200)	3.3	3.3	1900	(2100)	1.7	1.7	510	(560)	.4	.4
Vessel	7000	(7700)	4.6	16.5	3500	(3900)	2.3	8.2	720	(790)	.5	1.7
Ducts and Manifolds												
Insulation	3600	(3900)	3.4	1.8	1800	(2000)	1.7	.9	450	(500)	.4	.2
Steel	1120	(1240)	.7	2.6	560	(620)	.4	1.3	120	(130)	.1	.3
Flanges	1300	(1400)	2.8		640	(710)	1.4		170	(190)	.4	
Expansion Joints			1.2				.6				.2	
Valves	830	(910)	18.9		420	(460)	9.4		125	(140)	3.0	
Structural Steel	5300	(5900)	4.1	11.2	2700	(2900)	2.1	5.6	620	(680)	.5	1.3
Bed Support	640	(700)	2.5	0	320	(350)	1.2		50	(55)	.2	
Instrumentation/Controls			.9				.4				.2	
TOTALS	40800	(44900)	60.3	46.3	20400	(22500)	30.1	23.2	4700	(5200)	7.9	5.1
TOTAL SYSTEM COST 10^6 \$				106.6				53.3				13.0

* Direct costs are shown, i.e. contractors overhead and profit are not included. Installation is included in the labor component. All costs are mid-1979 dollars.

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TABLE 3.5

Residual Fluid Mass Per Heater

Low Pressure Gasifier-Fired Argon Heaters

Vessel Condition	Fluid	Case 3.0 Mass kg	Case 3.0 Mass (lbm)	Case 3.3 Mass kg	Case 3.3 Mass (lbm)	Case 3.7 Mass kg	Case 3.7 Mass (lbm)
Begining of Comb. Products Flow	Comb. Prod.	290	(640)	290	(640)	115.3	(250)
End of Comb. Prod. Flow	Comb. Prod.	270	(590)	310	(590)	107.3	(240)
Begining of Argon Flow	Argon	3620	(7960)	3620	(7960)	1430	(3140)
End of Argon Flow	Argon	3770	(8290)	377	(8300)	1490	(3270)

	Case 3.0 Flow kg/sec	Case 3.0 Flow (lbm/sec)	Case 3.3 Flow kg/sec	Case 3.3 Flow (lbm/sec)	Case 3.7 Flow kg/sec	Case 3.7 Flow (lbm/sec)
Time Averaged Argon Exchange Per Cycle	38.7	(85.1)	19.3	(42.5)	3.4	(7.4)
Time Average Comb. Prod. Exchange for Cycle	2.7	(6.0)	1.4	(3.0)	.2	(.5)

TABLE 4

COMPARISON OF EIGHT SYSTEMS

Case	Type	Power MWe	Weight Tons	Cost 10^6 \$	Cost \$/lb	Cost \$/10^6 B/hr	Cost \$/kWe	Bed Dia. ft	Bed Height ft	No. Heaters
1.1	DF atm	1000	95800	245	1.28	37500	245	25.6	122	16
1.4	DF press	1000	61700	158	1.28	24400	158	20	133	16
2.0	Pr gas	1000	22300	54.3	1.22	8340	54	17	51	16
2.4	Pr gas	500	11700	28.7	1.23	8800	57	17	51	9
2.5	Pr gas	250	6420	16.6	1.29	10200	66	13.6	51	7
3.0	Atm gas	1000	44900	107	1.19	16400	107	25.8	34.7	20
3.3	Atm gas	500	22500	53.3	1.19	16400	107	25.8	34.7	10
3.7	Atm gas	100	5200	13.0	1.25	20000	130	17.5	34.1	5

APPENDIX - C
CLOSED CYCLE EQUIPMENT LIST

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TABLE 1
CCMHD

EQUIPMENT LIST - CASE-1.0

ITEM	QUANTITY	DESCRIPTION
1. MHD Generator	1	Inlet Plasma: 3100°F, 6112.70 lb/sec Length: 57.7 ft. Inlet Area: 26.2 ft.^{2} Outlet Area: 71.7 ft.^{2} Length-to-Diameter Ratio: 10 Enthalpy Extraction: 36.0%
2. Magnet	1	Field: 6 Tesla (Maximum) Length: 82.8 ft. Dewar Outside Diameter: 40 ft.
3. Diffuser	1	Inlet Area: 71.7 ft.^{2} Outlet Area: 407.5 ft.^{2} Length: 115.5 ft. Pressure: 18.0 psia Inlet Temp: 1240°F Outlet Pressure: 30.60 psia Outlet Temp: 1818.5°F
4. Superheater	1	Steam: 820.65 lb/sec In: 3726.1 psia, 721.1°F Out: 3500.0 psia, 1000°F Argon: 6112.7 lb/sec In: 2.083 atm, 1818.5°F Out: 2.045 atm, 1200.0°F Cesium Condensation: 0 lb/sec Height x Width x Length: 75 ft. x 75 ft. x 12 ft. Heat Rating: 494.9 MW_t
5. Reheater	1	Steam: 820.65 lb/sec In: 612.4 psia, 542.6°F Out: 463.0 psia, 1000°F Argon: 6112.7 lb/sec In: 30.073 psia, 1200.0°F Out: 29.782 psia, 904.8°F Height x Width x Length: 75 ft. x 75 ft. x 18 ft. Heat Rating: 236.2 MW_t
6. Boiler	1	Steam: 820.65 lb/sec In: 4025.0 psia, 413.6°F Out: 3765.0 psia, 708.2°F Argon: 6112.7 lb/sec In: 29.782 psia, 904.8°F Out: 29.006 psia, 475.0°F

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TABLE 1 (CONTINUED)
CCMHD

EQUIPMENT LIST - CASE-1.0

ITEM	QUANTITY	DESCRIPTION
		Cesium Condensation: 17.9 lb/sec Height x Width x Length: 75 ft. x 75 ft. x 15 ft. Heat Rating: 343.9 MW_t
7. High Pressure Economizer	1	Steam: 820.65 lb/sec In: 4045.0 psia, 369.3°F Out: 4025.0 psia, 413.6°F Argon: 6094.8 lb/sec In: 29.006 psia, 475.0°F Out: 28.896 psia, 425.0°F Cesium Condensation: 2.0 lb/sec Height x Width x Length: 75 ft. x 75 ft. x 12 ft. Heat Rating: 39.97 MW_t
8. Low Pressure Economizer	1	Steam: 812.07 lb/sec In: 245.0 psia, 108.3°F Out: 150.0 psia, 347.9°F Argon: 6092.8 lb/sec In: 28.896 psia, 425.0°F Out: 28.785 psia, 165.0°F Cesium Condensation: 0.34 lb/sec Height x Width x Length: 75 ft. 75 ft. x 60 ft. Heat Rating: 207.83 MW_t
9. Argon Cooler	1	Argon: 6112.70 lb/hr In: 1.958 atm, 165.0°F Out: 1.954 atm, 100.0°F Cesium Condensation: 0 lb/sec Height x Width x Length: 75 ft. x 75 ft. x 18 ft. Heat Rating: 51.9 MW_t
10. Argon Purifier	1	Argon: 6112.70 lb/sec In: 1.954 atm, 100°F Out: 1.951 atm, 80°F Heat Rating: 16 MW_t
11. Argon Compressor	1	Outlet Pressure: 10.615 atm, 695.7°F Electrical Consumption: 492.2 MW_e AR flow: 6092.46 lb/sec Compressor Pressure Ratio: 5.44 Steam Turbine Driven

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TABLE 1 (CONTINUED)
CCMHD

EQUIPMENT LIST - CASE-1.0

ITEM	QUANTITY	DESCRIPTION
17. Flue Gas to Air Heat Exchanger	1	Air: 1758.8 lb/hr In: 1.278 atm, 100.9°F Out: 1.208 atm, 625.0°F Argon: 2307.7 lb/sec In: 1.093 atm, 762.9°F Out: 1.208 atm, 398.0°F Height x Width x Length: 75 ft. x 75 ft. x 40 ft. Heat Rating: 240.8 MW_t
18. Baghouse (Upstream of the Recirculation Fan)	1	Gas: 362.45 lb/sec In: 15.893 psia, 398.0°F Out: 15.0 psia, 398.0°F Inlet Loading: 0.3348 lb/sec Outlet Loading: 0.048 lb/sec Efficiency: 98.5%
19. Gas Recirculation Fan	1	Outlet Pressure: 18.0 psia, 530°F Electrical Consumption: 0.77 MW_e Gas: 362.45 lb/sec Electrical Motor Driven
20(A) Coal Dryer	1	Gas Flow: 1944.92 lb/sec In: 15.893 psia, 398.0°F Out: 15.575 psia, 246.4°F Coal Flow In: (22.7% mois. by wt.) = 245.6 lb/sec Coal Flow Out: (10.0% mois. by wt.) = 210.93 lb/sec
20(B) Mechanical (Cyclone) Collectors	1	Gas Flow: 1991.7 lb/sec In: 15.575 psia, 246.4°F Out: 15.53 psia, 246.4°F Coal Collected: 11.52 lb/sec Collection Efficiency: 99.9%
20(C) Baghouse	1	Gas Flow: 1980.2 lb/sec In: 15.53 psia, 246.4°F Out: 15.5 psia, 246.4°F Coal Collected: 0.02 lb/sec Collection Efficiency: 99.9%

TABLE 1 (CONTINUED)
CCMHD

EQUIPMENT LIST - CASE-1.0

ITEM	QUANTITY	DESCRIPTION
20(D) Transport Gas Flow	1	Gas Flow (Maximum): 7.0 lb/sec Coal Flow (Maximum): 11.54 lb/sec
21(A) Spray Dryer	1	Gas Flow (Inlet): 1973.18 lb/sec In: 15.5 psia, 246.4°F Gas Flow (Outlet): 1980.2 lb/sec Out: 15.264 psia, 201.8°F
21(B) Baghouse	1	Gas Flow (In) = 1980.2 lb/sec. In: 15.264 psia, 201.8°F Gas Flow (out) = 1973.17 lb/sec. Out: 15.15 psia, 195.0°F

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TABLE 2
CCMHD

EQUIPMENT LIST - CASE-2.0
(WITH IGT PRESSURIZED GASIFIER & HOT GAS CLEAN UP)

ITEM	QUANTITY	DESCRIPTION
1. MHD Generator	1	Inlet Plasma: 3100°F, 6112.70 lb/sec Length: 57.7 ft. Inlet Area: 26.2 ft.^{2} Outlet Area: 71.7 ft.^{2} Length-to-Diameter Ratio: 10 Enthalpy Extraction: 36.0%
2. Magnet	1	Field: 6 Tesla (Maximum) Length: 82.8 ft. Dewar Outside Diameter: 40 ft.
3. Diffuser	1	Inlet Area: 71.7 ft.^{2} Outlet Area: 407.5 ft.^{2} Length: 115.5 ft. Pressure: 18.0 psia Inlet Temp: 1240°F Outlet Pressure: 30.60 psia Outlet Temp: 1818.5°F
4. Superheater	1	Steam: 815.84 lb/sec In: 3726.1 psia, 720.7°F Out: 3500.0 psia, 1000°F Argon: 6112.7 lb/sec In: 2.083 atm, 1818.5°F Out: 2.045 atm, 1200.0°F Cesium Condensation: 0 lb/sec Height x Width x Length: 75 ft. x 75 ft. x 12 ft. Heat Rating: 494.9 MW_t
5. Reheater	1	Steam: 815.84 lb/sec In: 612.3 psia, 542.6°F Out: 463.0 psia, 1000°F Argon: 6112.7 lb/sec In: 2.045 stm, 1200.0°F Out: 2.026 stm, 906.7°F Height x Width x Length: 75 ft. x 75 ft. x 18 ft. Heat Rating: 236.2 MW_t

TABLE 2 (CONTINUED)
CCMHD

EQUIPMENT LIST - CASE-2.0
(WITH IGT PRESSURIZED GASIFIER & HOT GAS CLEAN UP)

ITEM	QUANTITY	DESCRIPTION
6. Boiler	1	Steam: 815.84 lb/sec In: 4025.0 psia, 413.7°F Out: 3765.0 psia, 708.2°F Argon: 6112.7 lb/sec In: 2.026 stm, 906.7°F Out: 1.973 atm, 475.0°F Cesium Condensation: 17.9 lb/sec Height x Width x Length: 75 ft. x 75 ft. x 15 ft. Heat Rating: 343.9 MW_t
7. High Pressure Economizer	1	Steam: 815.84 lb/sec In: 4045.0 psia, 369.3°F Out: 4025.0 psia, 413.6°F Argon: 6094.8 lb/sec In: 1.973 atm, 475.0°F Out: 1.965 atm, 425.0°F Cesium Condensation: 2.0 lb/sec Height x Width x Length: 75 ft. x 75 ft. x 12 ft. Heat Rating: 39.97 MW_t
8. Low Pressure Economizer	1	Steam: 808.67 lb/sec In: 245.0 psia, 108.3°F Out: 150.0 psia, 347.9°F Argon: 6092.8 lb/sec In: 1.965 atm, 425.0°F Out: 1.958 atm, 165.0°F Cesium Condensation: 0.34 lb/sec Height x Width x Length: 75 ft. 75 ft. x 60 ft. Heat Rating: 207.83 MW_t
9. Argon Cooler	1	Argon: 6092.46 lb/hr In: 1.958 atm, 165.0°F Out: 1.954 atm, 100.0°F Cesium Condensation: 0 lb/sec Height x Width x Length: 75 ft. x 75 ft. x 18 ft. Heat Rating: 51.9 MW_t

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TABLE 2 (CONTINUED)
CCMHD

EQUIPMENT LIST - CASE-2.0
(WITH IGT PRESSURIZED GASIFIER & HOT GAS CLEAN UP)

ITEM	QUANTITY	DESCRIPTION
10. Argon Purifier	1	Argon: 6112.70 lb/sec In: 1.954 atm, 100°F Out: 1.951 atm, 80°F Heat Rating: 16 MW_t
11. Argon Compressor	1	Outlet Pressure: 10.615 atm, 695.7°F Electrical Consumption: 492.2 MW_e AR flow: 6092.46 lb/sec Compressor Pressure Ratio: 5.44 Steam Turbine Driven
12. Argon Heat Exchanger	1	Argon: 6092.46 lb/sec In: 10.615 atm, 1100°F Out: 10.020 atm, 3103°F Flue gas: 2681.4 lb/sec In: 144.0 psia, 3350°F Out: 136.8 psia, 773.8°F Heat Rating: 1601.1 MW_t
13. Cesium Injector	1	Cesium: 20.24 lb/sec In: 10.3 atm, 100°F Mixer Out (Ar & CS): 10.00 atm, 3100°F Mixer Flow: 6112.7 lb/sec
14. Pressurized Gasifier	6	Type: IGT Coal: 10.0 Moist, 242.2 lb/sec. 10,404.0 Btu/lb Oxidant: Air 604°F, 147.0 psia 604.45 lb/sec. Steam: 247.0 psia, 400.0°F 126.24 lb/sec. Slag: 24.0 lb/sec. 1735.0°F Fuel Gas: 135.6 psia, 1335.0°F 926.7 lb/sec. Inner diameter x Length = 22 ft. x x 30 ft. Overall Length = 35 ft. Construction Material = Carbon Steel Refractory Lined

TABLE 1 (CONTINUED)
CCMHD

EQIUPMENT LIST - CASE-1.0

ITEM	QUANTITY	DESCRIPTION
12. Flue Gas to Argon Preheater	1	Argon: 2307.7 lb/sec In: 10.615 atm, 695.7 ^{o}F Out: 10.530 atm, 1100 ^{o}F Flue gas: 2652.8 lb/sec In: 16.400 psia, 1238.8 ^{o}F Out: 16.073 psia, 888.0 ^{o}F Height x Width x Length: 75 ft. x 75 ft. x 40 ft. Heat Rating: 323.14 MW_t
13. Argon Heat Exchanger	1	Argon: 2307.7 lb/sec In: 10.530 atm, 1100 ^{o}F Out: 10.020 atm, 3103 ^{o}F Flue gas: 2652.8 lb/sec In: 17.26 psia, 3350 ^{o}F Out: 16.4 psia, 1227.3 ^{o}F Heat Rating: 1601.1 MW_t
14. Cesium Injector	1	Cesium: 20.24 lb/sec In: 10.3 atm, 100 ^{o}F Mixer Out (Ar & CS): 10.00 atm, 3100 ^{o}F Mixer Flow: 6112.7 lb/sec
15. Main Combustor	3	Pressure: 17.50 psia Coal: 10.0 moist, 210.93 lb/sec 10,404.0 Btu/lb Air: 625.0 ^{o}F, 1758.8 lb/sec 1.05% Stochiometric air flow Slag: 3100 ^{o}F, 24.52 lb/sec Flue Gas: 3350 ^{o}F, 2652.8 lb/sec Inner Diameter x Length = 17 ft. x 67 ft. Overall Length: 75 ft.
16. Air Compressor	1	Outlet Pressure: 18.79 psia, 100.9 ^{o}F Electrical Consumption: 19.87 MW_e Air Flow: 1758.8 lb/sec Electrical Motor Driven

TABLE 2 (CONTINUED)
CCMHD

EQUIPMENT LIST - CASE-2.0
(WITH IGT PRESSURIZED GASIFIER & HOT GAS CLEAN UP)

ITEM	QUANTITY	DESCRIPTION
15. Single Stage Combustor	3	Pressure: 144.0 psia Fuel Gas: 135.6 psia, 1335.0°F 926.7 lbs/sec Air: 604.0°F, 1197.8 lb/sec 1.05% Stochiometric air flow Flue Gas: 3350°F, 2681.4 lb/sec Inner Diameter x Length = 10 ft. x 67 ft. Overall Length: 75 ft.
16. Air Compressor	1	Outlet Pressure: 147.0 psia, 604.0°F Electrical Consumption: 179.0 MW_e Air Flow: 1197.8 lb/sec Electrical Motor Driven
17. Gas Recirculation Fan	1	Outlet Pressure: 144.1 psia, 790.9°F Electrical Consumption: 0.27 MW_e Gas: 50.27 lb/sec Electrical Motor Driven
18 (A) Coal Dryer	1	Gas Flow: 2124.6 lb/sec In: 16.1 psia, 355.6°F Out: 15.78 psia, 200.0°F Coal Flow In: (22.7% mois. by wt.) = 273.7 lb/sec Coal Flow Out: (10.0% mois. by wt.) = 222.03 lb/sec
18 (B) Mechanical (Cyclone) Collectors	1	Gas Flow: 2176.6 lb/sec In: 15.78 psia, 200°F Out: 15.65 psia, 200°F Coal Collected: 12.10 lb/sec Collection Efficiency: 99.9%
18 (C) Baghouse	1	Gas Flow: 2163.9 lb/sec In: 15.65 psia, 200.0°F Out: 15.50 psia, 200.0°F Coal Collected: 0.07 lb/sec Collection Efficiency: 99.9%