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ENERGY CONVERSION ALTERNATIVES STUDY

-ECAS-

GENERAL ELECTRIC PHASE I FINAL REPORT

VOLUME III, ENERGY CONVERSION SUBSYSTEMS AND COMPONENTS
Part 2, Primary Heat Input Systems and Heat Exchangers

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16. Abstract

A parametric study was performed to assist in the development of a data base for the comparison of advanced energy conversion systems for utility applications using coal or coal-derived fuels. Estimates of power plant performance (efficiency), capital cost, cost of electricity, natural resource requirements, and environmental intrusion characteristics were made for ten advanced conversion systems. Over 300 parametric points were analyzed to estimate the potential of these systems. Emphasis of the study was on the energy conversion system in the context of a base loaded utility power plant. Although cases employing transported coal-derived fuels were included in the study, the fuel processing step of converting coal to clean fuels was not investigated except for cases where a low-Btu gasifier was integrated with the power plant. All power plant concepts were premised on meeting emission standards requirements. The investigative approach focused on achieving consistency and comparability in the analysis of the various conversion systems. Recognized advocate organizations were employed to analyze their respective cycles and to present their analyses for power plant integration by the GE systems evaluation team. Wherever possible, common subsystems and components for the various systems were treated on a uniform basis. A steam power plant (3500 psig, 1000 F, 1000 F) with a conventional coal-burning furnace-boiler was analyzed as a basis for comparison. Combined cycle gas/steam turbine system results indicated competitive efficiency and a lower cost of electricity compared to the reference steam plant. The Open-Cycle MHD system results indicated the potential for significantly higher efficiency than the reference steam plant but with a higher cost of electricity. The information contained in this report constitutes results from the first phase of a two phase effort. In Phase II, a limited number of concepts will be investigated in more detail through preparation of conceptual designs and an implementation assessment including preparation of R&D plans estimating the resources and time required to bring the systems to commercial fruition.

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FOREWORD

The work described in this report is a part of the Energy Conversion Alternatives Study (ECAS)—a cooperative effort of the Energy Research and Development Administration, the National Science Foundation, and the National Aeronautics and Space Administration.

This General Electric contractor report for ECAS Phase I is contained in three volumes:

Volume I - Executive Summary

Volume II - Advanced Energy Conversion Systems

- Part 1 - Open-Cycle Gas Turbines
- Part 2 - Closed Turbine Cycles
- Part 3 - Direct Energy Conversion Cycles

Volume III - Energy Conversion and Subsystems and Components

- Part 1 - Bottoming Cycles and Materials of Construction
- Part 2 - Primary Heat Input Systems and Heat Exchangers
- Part 3 - Gasification, Process Fuels, and Balance of Plant

In addition to the principal authors listed, members of the technical staffs of the following subcontractor organizations developed information for the Phase I data base:

General Electric Company
- Advanced Energy Programs/Space Systems Department
- Direct Energy Conversion Programs
- Electric Utility Systems Engineering Department
- Gas Turbine Division
- Large Steam Turbine-Generator Department
- Medium Steam Turbine Department
- Projects Engineering Operation/I&SE Engineering Operation
- Space Sciences Laboratory

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Avco Everett Research Laboratory, Incorporated

Bechtel Corporation

Foster Wheeler Energy Corporation

Thermo Electron Corporation

This General Electric contractor report is one of a series of three reports discussing ECAS Phase I results. The other two reports are the following: Energy Conversion Alternatives Study (ECAS) Westinghouse Phase I Final Report (NASA CR-134941), and NASA Report (NASA TMX-71855).

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Summary

ENERGY CONVERSION SUBSYSTEMS AND COMPONENTS

The objective of Phase I of the Energy Conversion Alternatives Study (ECAS) for coal or coal-derived fuels was to develop a technical-economic information base on the ten energy conversion systems specified for investigation. Over 300 parametric variations were studied in an attempt to identify system and cycle conditions which indicate the best potential of the energy conversion concept. This information base provided a foundation for selection of energy conversion systems for more in-depth investigation in the conceptual design portion of the ECAS study. The systems for continued study were specified by the ECAS Interagency Steering Committee.

The major emphasis of this study was the evaluation of the prime cycle portion of the energy conversion system. The energy conversion subsystems and auxiliary systems are coupled to the prime cycle to produce a complete power plant. These subsystems were applied to each of the prime cycles on a consistent basis. Each of the subsystems, e.g., furnaces, bottoming cycles, balance of plant, was analyzed by its respective independent study team for each specific application to an energy conversion system.

The furnace systems included both direct combustion of coal and combustion of process fuels derived from coal. The furnaces with direct coal combustion employing fluidized beds with in-bed sulfur capture appear to be the most attractive options for the closed-cycle advanced energy conversion systems.

Both organic and steam cycles were studied for bottoming many of the prime cycles. The characteristics of the organic cycles made them most attractive in ratings up to 100 MWe and peak organic cycle temperature less than 500 F (533 K). Although the addition of an organic bottoming cycle to a prime cycle showed an efficiency improvement, a relatively high capital cost addition for the organic bottoming cycle and its related balance of plant was estimated. A steam bottoming cycle was an essential requirement for use with many of the prime cycles; e.g., Combined Cycle Gas Turbine, Liquid Metal Topping Cycle, MHD Systems, and High-Temperature Fuel Cells. The steam bottoming cycles were all analyzed by the same study team to assure a uniform assessment. Steam throttle conditions and feedwater heating chains were varied, however, to accommodate specific prime cycle requirements for improvement of the system efficiency.

In energy conversion systems which could utilize coal directly, the employment of clean fuels produced from coal did not appear to be economically attractive. In systems which require a fuel processing step, e.g., open-cycle gas turbines, the semi-clean liquid fuels produced from coal appeared to be an attractive alternative and were close to an economic standoff with the low-Btu integrated gasifier technique for producing an acceptable gas turbine fuel.

Introduction

ENERGY CONVERSION SUBSYSTEMS AND COMPONENTS

Many advanced energy conversion techniques which can use coal or coal-derived fuels have been advocated for power generation applications. Conversion systems advocated have included open- and closed-cycle gas turbine systems (including combined gas turbine-steam turbine systems), supercritical CO_2 cycle, liquid metal Rankine topping cycles, magnetohydrodynamics (MHD), and fuel cells. Advances have also been proposed for the steam systems which now form the backbone of our electric power industry. These advances include the use of new furnace concepts and higher steam turbine inlet temperatures and pressures. Integration of a power conversion system with a coal processing plant producing a clean low-Btu gas for use in the power plant is still another approach advocated for energy conserving, economical production of electric power. Studies of all these energy conversion techniques have been performed in the past. However, new studies performed on a common basis and in light of new national goals and current conditions are required to permit an assessment of the relative merits of these techniques and potential benefits to the nation.

The purpose of this contract is to assist in the development of an information base necessary for an assessment of various advanced energy conversion systems and for definition of the research and development required to bring these systems to fruition. Estimates of the performance, economics, natural resource requirements and environmental intrusion characteristics of these systems are being made on as comparable and consistent a basis as possible leading to an assessment of the commercial acceptability of the conversion systems and the research and development required to bring the systems to commercial reality. This is being accomplished in the following tasks:

Task I	Parametric Analysis	(Phase I)
Task II	Conceptual Designs	(Phase II)
Task III	Implementation Assessment	(Phase II)

This investigation is being conducted under the Energy Conversion Alternatives Study (ECAS) under the sponsorship of Energy Research and Development Administration (ERDA), National Science Foundation (NSF), and National Aeronautics and Space Administration (NASA). The control of the program is under the direction of an Interagency Steering Committee with participation of the supporting agencies. The NASA Lewis Research Center is responsible for project management of this study.

The information presented in this report describes the results produced in the Task I portion of this study. The emphasis in this task was placed upon developing an information base upon which comparisons of Advanced Energy Conversion Techniques using coal or coal-derived fuels can be made. The Task I portion of the study was directed at a parametric variation of the ten advanced energy conversion systems under investigation. The wide-ranging parametric study was performed in order to provide data for selection by the Interagency Steering Committee of the systems and specific configurations most appropriate for Task II and III studies.

The Task II effort will involve a more detailed evaluation of seven advanced energy conversion systems and result in a conceptual design of the major components and power plant layout. The Task III effort will produce the research and development plans which would be necessary to bring each of the seven Task II systems to a state of commercial reality and then to assess their potential for commercial acceptability.

A prime objective of this study was to produce results which had a cycle-to-cycle consistency. In order to accomplish this objective and still ensure that each system was properly advocated, an organization which is or had been a proponent of the prime cycle was selected to advocate the energy conversion system and to analyze the performance and economics of the prime cycle portion of the energy conversion system, i.e., the parts of the system which were novel or unique to the system. The remaining subsystems, e.g., fuel processing, furnaces, bottoming cycles, balance of plant, were analyzed by technology specialist organizations which presently have responsibility for supplying these subsystems for utility applications. The final plant configuration and performance were produced by the General Electric Corporate Research and Development study team and this group performed the critical integration of the final plant concept. This methodology was used to provide a system-to-system consistency while maintaining the influence of a cycle advocate.

The energy conversion subsystems and components which were applied on a common basis to each of the advanced energy conversion systems are described in this Volume. The discussion and results for each of the advanced systems is given in Volume II.

Bottoming Cycles are applied to most of the advanced energy conversion systems. To the maximum extent possible, the bottoming cycles were assumed to be composed of state-of-the-art components. Steam bottoming cycles are utilized for "high-temperature" applications bottoming with steam conditions being limited to 1000 F (811 K). Organic fluid bottoming cycles are employed for the low-temperature applications (temperatures less than 600 F [589 K]).

The Materials of Construction are defined for each of the energy conversion systems. This includes both the identification of the materials and the assumptions which were made with respect to design criteria.

Primary Heat Input Systems were employed for all closed-cycle applications. The heat exchanger equipment provides for the transport process to introduce thermal energy into the cycle working fluids. Advanced furnace techniques for direct combustion of coal and combustion of clean fuels were considered. The atmospheric fluidized bed with direct coal was utilized as a reference furnace for the closed-cycle parametric variations.

Heat Exchangers were employed in all advanced energy conversion systems. This fluid-to-fluid exchange equipment provided for transport processes within the cycles, e.g., the regeneration of thermal energy, heat rejection precoolers, and low temperature air preheaters.

Gasification and Process Fuels derived from coal were employed as clean fuel sources for combustion systems. The low-Btu gasifier employed for integrated plants was the fixed bed gasifier with low-temperature cleanup. The process fuels were considered as delivered to the plant boundary. The cost and conversion efficiency for these clean fuel production processes were directly related to the fixed bed gasifier. This gave a basis for cost comparison between the use of process fuels and integrated gasifier systems.

The Balance of Plant for the advanced energy conversion concepts considered the installation of the specific components of the energy conversion cycle and primary heat input heat exchangers and the supply and installation of the auxiliary plant equipment. The fuel supply and storage system and the heat rejection system were two of the major elements evaluated as balance-of-plant items.

Section 6

PRIMARY HEAT INPUT SYSTEMS

INTRODUCTION

The primary heat-input heat exchangers were employed to supply thermal energy to the working fluid of a closed power cycle. The thermal energy initially came from the combustion of a fuel, and in many cases the combustion process was an integral part of the primary heat-input heat exchanger. However, in some energy conversion systems (for example, open-cycle MHD) the combustion process was remote from the heat exchange equipment and the primary transport mode was convection and radiation from high-temperature fluids to the heat transfer tubes that contain the cycle working fluid. The fuels considered were coal or coal derived liquid and gaseous fuels.

Primary heat input heat exchangers were evaluated for eight energy conversion systems:

1. Advanced steam
2. Supercritical CO_2
3. Closed-cycle helium gas turbine
4. Liquid metal topping
5. Closed-cycle liquid metal MHD
6. Closed-cycle inert gas MHD
7. Open-cycle MHD
8. High-temperature fuel cell

The following furnaces were considered:

1. Conventional furnace (coals, semiclean liquid fuel [SRC])
2. Atmospheric fluidized bed (coals with limestone)
3. Pressurized fluidized bed (coals with dolomite)
4. Pressurized furnace (high-Btu and low-Btu gases)

Each energy conversion system also included primary, intermediate, and heat recovery exchangers as required and auxiliary combustion support systems.

All of these furnaces are in different stages of development. A discussion of these furnaces for the advanced cycles under consideration in this study follows:

- <u>CF (Conventional Furnaces) and CH (Combustion Chambers)</u>

  Conventional furnace technology is well developed through 1100 F (866.7 K) working-fluid exit temperatures. Firing slagging fuels for high-temperature cycles presents serious metals problems due to coal ash corrosion in the 1000 F to 1400 F (811 to 1033.3K) outside metal temperature range. Resistant materials must be developed in this range if this technique is to be applied, permitting surface metal temperatures to 1600 F (1144.4 K).

- <u>AFB (Atmospheric Fluidized Beds)</u>

  Engineering development work is continuing and is required in the area of peripheral solids systems to minimize costs, complexity, and solids handling problems. The regeneration of spent acceptor requires a considerable development effort and was not considered in this study.

  For certain high-temperature cycles were heat transfer surface requirements control fluidized bed size and volume, extended surfaces were not used in this study, as they require further testing and development work.

- <u>PFB (Pressurized Fluidized Beds)</u>

  Engineering development work is continuing in the area of peripheral solids systems to minimize costs, complexity, and solids handling problems. These problems, typical to fluidized bed combustion processes, are compounded at pressure. The presence of alkali compounds in the exhaust gases at high temperature present serious problems in dust collection and gas turbine design.

- <u>PF (Pressurized Furnaces)</u>

  Engineering development work is required in the area of large minimum volume diffusion flame burners which produce low-NO_X emissions. In order to take advantage of the high transfer rates and gas temperatures typical to pressurized furnaces, high metal temperatures are expected. The need for development of high-temperature metals is greatest with this type of furnace. A weldable, readily formed, wrought alloy is required which exhibits reasonable design stress limits in the 1500 F to 1600 F (1088.9 to 1144.4 K) temperature range. The only material that has this characteristic at present is a proprietary centrifugally cast material. The use of this material is unproven in high-pressure service and it is difficult to weld.

In reviewing the results of this study for combustors and primary heat exchangers the following qualifications should be applied:

1. The power cycles studied are not optimized with respect to capital cost, thermal efficiency, or cost of elec-

tricity. The base cases and parametric variations were selected to indicate the potential of and sensitivity to varying approaches, furnace types, and conditions for the cycles. The objective was an approximate exploratory effort to reveal the main characteristics of the power cycles studied.

2. This study represents different levels of familiarity with and conceptual development for the various combustion systems investigated. It is apparent that all furnace/cycle combinations have not been thought through to an equal level of development. Table 6-1 gives an engineering evaluation of the characteristics and operating limitations of these furnace concepts.

PRIMARY COMBUSTOR AND HEAT EXCHANGER ARRANGEMENT

The following code list was used to represent furnaces and equipment used in this study:

CF = Conventional fired units

CH = Combustion chambers

AFB = Atmospheric fluidized bed units

PFB = Pressurized fluidized bed units

PF = Pressurized furnace units

OC-MHD = Open-Cycle MHD, secondary combustion units

IHX = Indirect Intermediate Heat Exchangers

HTAH = High-temperature air heaters

AUX = Auxiliary equipment

Of the furnace types considered in this study for the advanced applications, the conventional furnace applications are the nearest to commercialization. Where the conventional furnace would be extended for advanced power cycles with high temperatures, development work would be required. The pressurized or supercharged furnace units firing clean fuels are considered the next in order of commercial availability. Where these furnace types are applied to advanced power cycles with high working-fluid temperatures and it is desired to take advantage of the very high heat absorption rates that are possible, a major effort would be required to develop high-temperature metals, particularly when used to heat gases and superheated or reheated steam.

The atmospheric fluidized bed furnace and pressurized fluidized bed furnaces, respectively, are considered next in order of commercial availability. Economic and efficient operation is projected from smaller-scale test units currently in operation. The open-cycle MHD radiant furnace, intermediate-temperature air heater, and heat recovery steam generator, although very similar in operational duty to conventional furnaces,

Table 6-1

FURNACE CHARACTERISTICS

Combustion System	Fuels Fired	Level of Development for Advanced Cycle Conditions Increasing Severity → Std*	HP*	HT*	HP+HT*	VHT	Maximum Furnace Temperatures	High-Temp. Mate-rials	Main Limitations Ash-Seed Corrosion	Velocity, Bed Volume
CF	Coals	1**	2	4	4	5	3500 F	x	x	
	SRC	1	2	3	3	5	3500 F	x		
	Gases	1	1	2	3	5	3500 F	x		
AFB - 1-stage	Coals	2	2	3	4	5	1400-1600 F			x
2-stage	Coals	4	4	5	5	5	1600-2000 F	x		x
PFB - 1-stage	Coals	3	3	4	4	5	1600-1800 F			x
2-stage	Coals	5	5	5	5	5	1800-2000 F	x		x
PF	Gases	1	2	3	4	5	2200-2800 F	x		

*Cycle Conditions

Std	1000 F at 3500 psi	-	Standard
HP	1000 F at 4000 psi	-	High pressure
HT	1200 F+ at 1000 psi	-	High temperature
HP+HT	1200 F+ at 4000 psi	-	High pressure + temperature
VHT	1400 F+ at 1000 psi	-	Very high temperature

**Level	Rank	Engineering Stage
1	Established	Complete
2	Proven	Advanced
3	Certain	Intermediate
4	Probable	Preliminary
5	Feasible	Conceptual

require special attention and consideration because of unknowns surrounding the combination of coal-ash corrosion, seed material corrosion, and slagging. Combustion conditions in the open-cycle MHD furnace are controlled near stoichiometric and are substoichiometric in the major part of the radiant furnace. The correct combination of outside tube metal temperature limits, protective tube coatings or materials, and furnace design must be found for the real environment of MHD channel exhaust gases. The closed-cycle inert gas MHD furnaces or combustion chambers must be large in order to control slag and ash pickup and carryover when burning substoichiometrically in the main part of the furnace. This is necessary for control of nitrogen oxide emissions.

Indirect intermediate heat exchangers and heat recovery steam boilers used with the advanced cycles of this study require additional attention to ensure economic designs.

The high-temperature air heaters (for furnace applications other than open-cycle MHD) required with these advanced power cycles similarly require additional attention to ensure economic design. Tubular type air heaters were utilized in all high-temperature applications. Where the working fluids and gas streams handled are relatively clean, compact recuperative heat exchanger surfaces could be considered in future studies.

The auxiliary equipment considered in this study to support conventional furnaces and combustion chambers, atmospheric fluidized beds, pressurized fluidized beds, and pressurized furnaces were developed from the standpoint of minimum process requirements. The solids feeding and disposal systems for the fluidized bed units require process development, optimization, and integration into the overall power-plant cycle to ensure economic and reliable operation.

These furnaces and equipment are described in this subsection. The characteristics, scope of equipment included, development requirements, main advantages, major problems, limitations, and special design considerations are indicated.

CONVENTIONAL FURNACE

The main characteristics of conventional furnaces firing coal, the auxiliary equipment included in this study, and indicated development requirements are summarized on Table 6-2. The main advantages, major problem areas, limitations, and special design considerations are summarized in Table 6-3. Coal-ash corrosion is a major problem when firing coal in a conventional furnace. The projected corrosion rates due to coal-ash corrosion over a temperature range from 600 to 2000 F (588.9 to 1366.7 K) are indicated approximately in Figures 6-1 and 6-2. Figure 6-2 indicates a zone of outside metal temperatures from 1350 to 1550 F (1005.6 to 1116.7 K) in which high-temperature cycles could possibly operate with reduced corrosion rates. However, operation in this temperature range is seen to be extremely risky when the

Table 6-2

CONVENTIONAL FURNACE ARRANGEMENT
(Reference Furnace)

Characteristics

- Large furnace combustion chamber (required for combustion efficiency and to avoid slagging, ash corrosion).
- Pulverized coal fired (70% < 200 mesh for combustion efficiency).
- Sootblowers required, wall blowers required (to maintain furnace heat transfer effectiveness).
- No SO_2 control, higher cold-end sulfuric acid dew points.
- Difficult NO_x control with high adiabatic flame temperatures; must consider off-stoichiometric firing, gas recirculation, lower-temperature air preheater (AP), low-turbulence burners.
- Requires control of particulate emissions (electrostatic precipitators or high-efficiency wet scrubbing with gas reheat).

Auxiliary Equipment Included

- Feeders and pulverizers (coal firing)
- Air preheaters (recuperative type)
- Sootblowers (coal and oil firing)
- Forced draft (FD), Induced draft (ID), and primary air fans (coal firing)
- Combustion and burner controls
- Valves

Development Requirements

- None for fluid temperatures to 1000 F
- Materials subject to liquid slag corrosion from 950 to 1350 F and above 1500 F fluid temperatures.
- Present high-temperature alloys offer limited suitability in furnace for high-temperature cycles.

Table 6-3

GENERAL CONSIDERATIONS
(Conventional Furnace)

Main Advantages

Technology proven for combustion control, sarety, etc.

High availability for fluid temperatures to 1000 F maximum.

Major Problems

NO_X, SO_X, dust emissions require cleanup with dirty fuels

Slagging, ash corrosion, erosion

Ash disposal

High-temperature cycle—materials

Fabrication, cost, availability, expansion, stress anslysis, etc.

Limitations

Largest water-walled panel sections shipped about 12 ft wide by 100 ft long.

Maximum furnace widths about 100 ft maximum as limited by staying and expansion problems.

Large furnace size and weight set by combustion efficiency, slag control.

Gas recirculation for required NO_X control in certain extreme cases.

Staged firing for NO_X control preferred, but sufficient residence time at high temperature must be maintained in the furnace volume to ensure complete combustion after injection of the balance of combustion air.

Special Design Considerations

Complex—long path

Circulation analysis

Furnace cleaning

Sootblowing

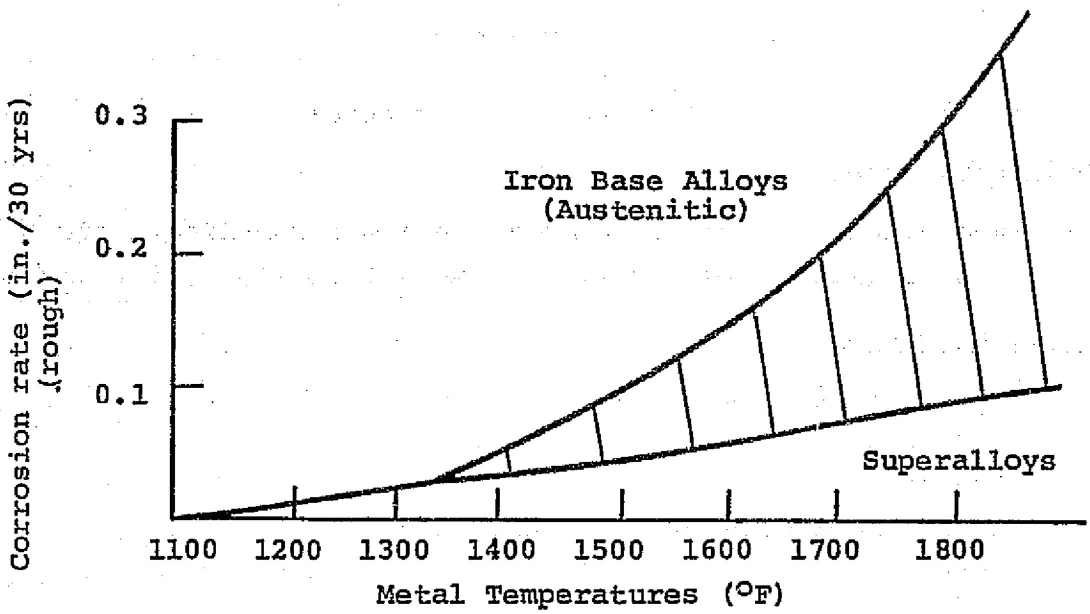


Figure 6-1. Example of Oxidation versus Metal Temperature

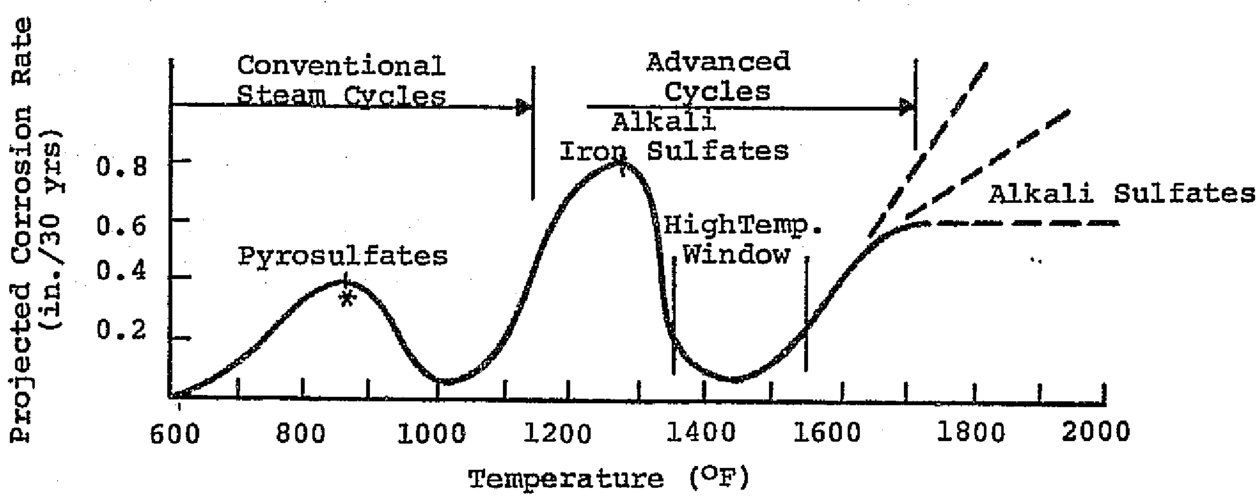


Figure 6-2. Example of Corrosion from Alkali-Sulfur Compounds

unknown factors influencing coal-ash corrosion and oxidation are considered. Every time a unit starts up and shuts down it must transition through the 1000 to 1400 F (811 to 1033.3 K) range where accelerated coal-ash corrosion occurs. Starting up and shutting down with auxiliary fuels or clean fuels might be a possible approach, but this solution is not very attractive economically.

In applications for the advanced steam power cycle firing coal, metal temperatures fall in the range from 1350 to 1550 F (1005.6 to 1116.7 K), or in this high temperature window. However, suitable materials and/or protective coating may still have to be considered, to permit actual operation. Hastelloy X tubing was utilized for this cycle from the standpoint of stress, corrosion, and oxidation resistance. The creation of this high-temperature window presents limitations on operating flexibility of the boiler, as metal temperatures must be maintained in this range with little room for error. Where metal temperatures will be higher than 1400 to 1600 F (1033.3 to 1144.4 K), i.e., above the vapor point of most corrosive liquid-phase alkali-iron-sulphate complexes, metal depletion allowances will have to be made. The assumption is also made that the units will 1) be base load units and will not experience more than one to two shutdowns per year and 2) be maintained at a sufficiently high load level to maintain metal temperatures above this critical ash corrosion temperature. Above 1600 F (1144.4 K) coal-ash corrosion rates accelerate for most ferrous tube materials while ash is still molten.

Furnace design policy, permitting the use of economic materials, indicates that the gas should be chilled below the ash solidification temperatures (1900 to 2400 F [1311.1 to 1588.9 K]) before passing over high-temperature surfaces operating above 1100 F (866.7 K) metal temperatures. This philosophy indicates that cooling in the radiant furnace enclosure should be accomplished with cycle working-fluid temperatures lower than 950 F (783.3 K).

In all cases where coals and high-sulfur fuels are fired directly, sulfur removal equipment is required. In this study these cleanup processes were considered to be a part of the balance of plant. The cost of these cleanup systems is not reflected in the study described in this section.

Figure 6-3 presents an example of the determination of nitrogen oxide emission; the effect of off-stoichiometric firing is indicated in Figure 6-4. Present furnace design policy for firing coals prefers the use of off-stoichiometric firing as the main means of NO_X limitation and control. With this control system, fuel is burned substoichiometrically in lower stages of the furnace, with the balance of combustion air added at the top of the furnace after the gas has a chance to cool below accelerated formation temperatures for nitrogen oxide. In addition to

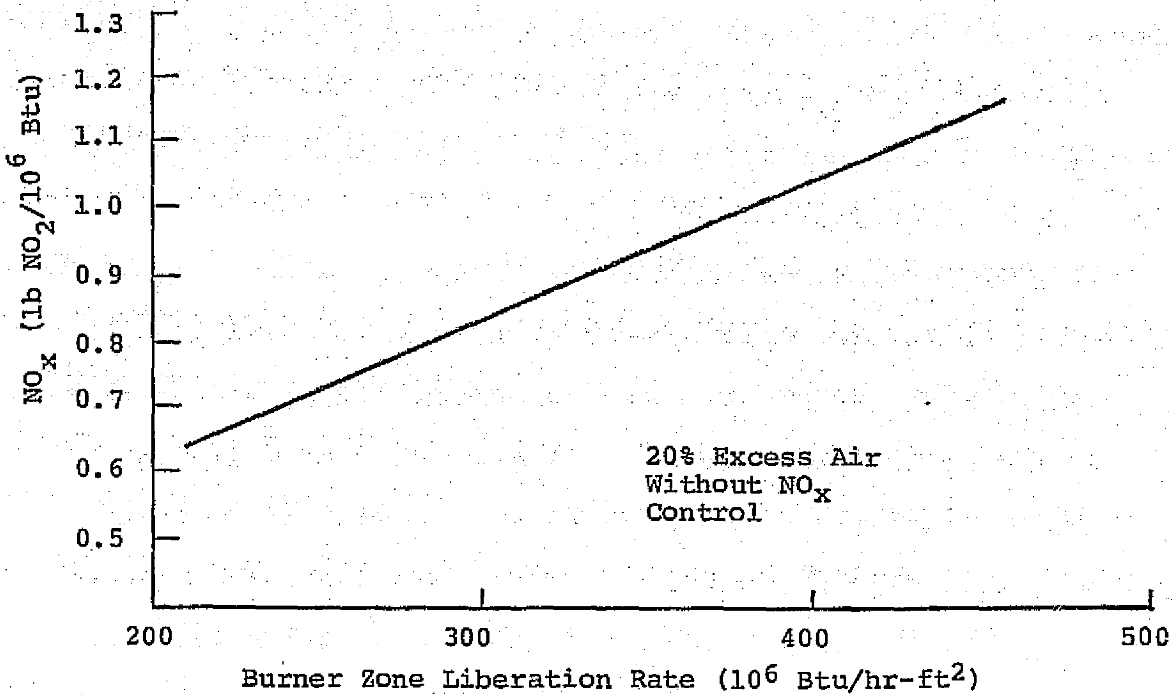


Figure 6-3. NO_x Emissions

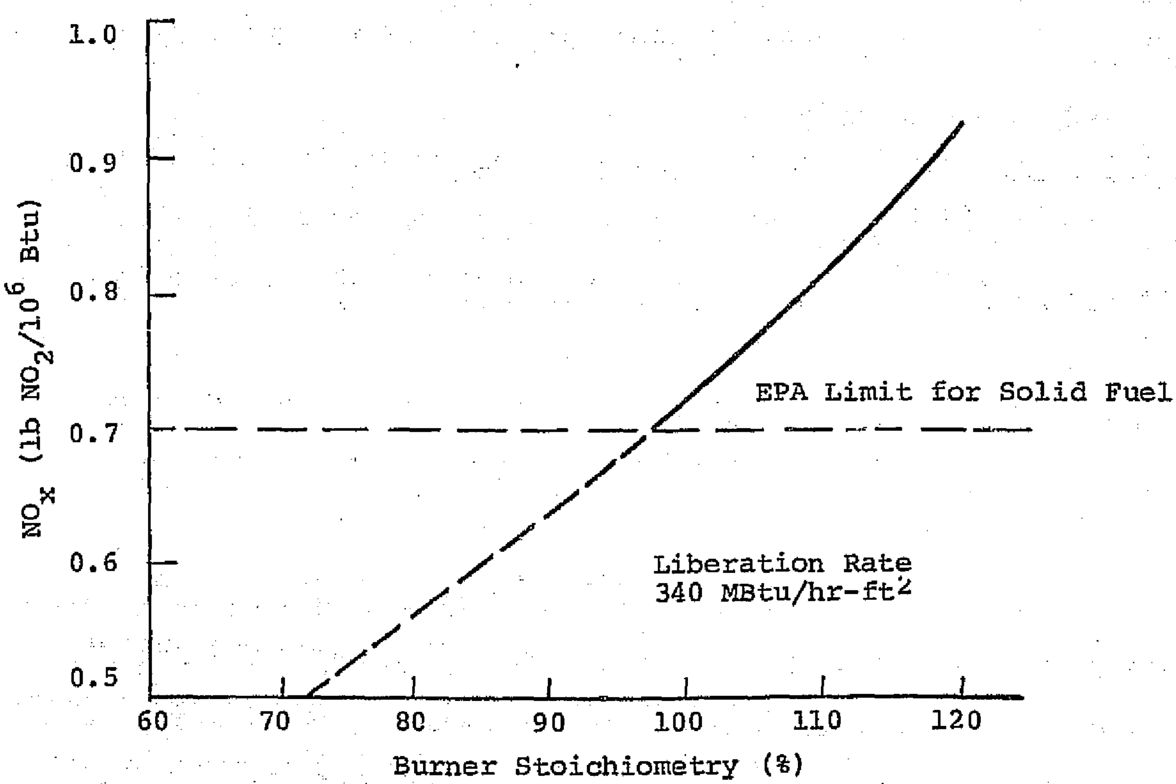


Figure 6-4. Effect of Burner Stoichiometry on NO_x Emission

recirculation of off-stoichiometric firing flue gas, low-temperature air heat and low-turbulence burners may also be used to depress nitrogen oxide formation in the furnace.

The effect of sulfur content in coals fired in a conventional furnace on fly-ash resistivity is indicated in Figure 6-5. With the firing of low-sulfur coal, there is a tendency for resistivity to increase considerable over the average experienced for sulfur-content coals. As precipitator temperature is raised, resistivity tends to be reduced to a more desireable range. Higher precipitation temperatures may therefore be indicated for more economical precipitator design.

The effects of furnace chemistry and iron on sodium content in coal fly ash on ash softening temperatures are indicated in Figures 6-6 and 6-7. As a furnace design policy it is desirable to cool combustion gases below these ash softening temperatures before passing gases over high-temperature convection surfaces. The impact of this condition for coals with low ash softening temperatures is to increase the radiant furnace size in proportion to the rest of the boiler. Molten slag on metallic surfaces should be avoided in regions where alkali, iron and nickel sulfates, and pyrosulfates might form. (Temperature range is from 1100 to 1400 F [866.7 to 1033.3 K]).

The conventional furnace was considered only for the advanced steam energy conversion system. This furnace configuration is shown in Figure 6-8.

ATMOSPHERIC FLUIDIZED BED FURNACE

The atmospheric fluidized bed (AFB) was employed for the base cases in the closed-cycle energy conversion systems. The characteristics (auxiliary equipment included) and development requirements for this type of furnace are presented in Table 6-4. Its main advantages, major problem areas, limitations, and special design considerations are given in Table 6-5.

Figure 6-9 is a general-flow schematic diagram for an AFB steam power plant. An example of the AFB modules are shown in Figures 6-10 and 6-11 for two applications: advanced steam cycle and closed helium gas-turbine cycle, respectively.

The sulfur dioxide emissions reduction attainable in atmospheric fluidized beds is shown in Figure 6-12. There is a temperature range in which acceptable SO_2 reduction can be achieved; for this application, the temperature range is 1480 to 1560 F (1077.8 to 1122.2 K). Figure 6-12 represents the SO_2 reduction attainable with a lime-to-sulfur stoichiometric ratio of approximately 2 to 1 and fluidizing conditions typical of those being considered in this study. This characteristic presents a very real limitation on the application of fluidized beds to very high-temperature cycles where sulfur must be removed and captured.

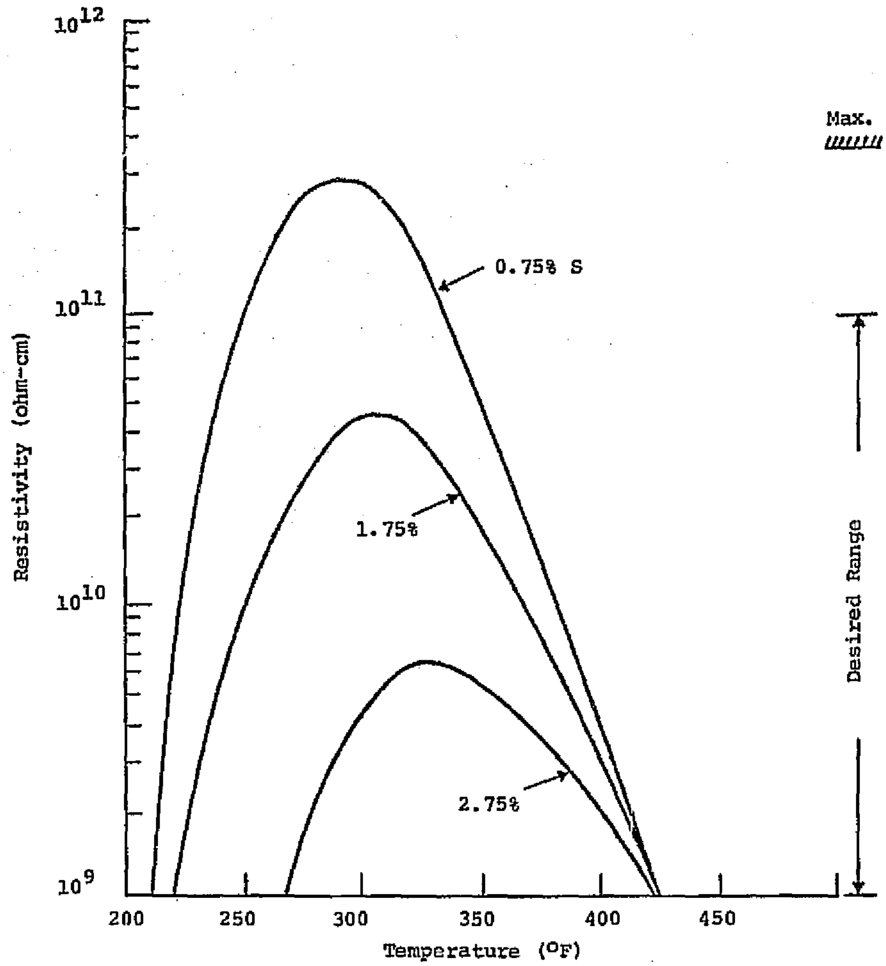


Figure 6-5. Fly-ash Resistivity Versus Changes in Temperature and Percentage Sulfur in Coal (∿ 5% Conversion of SO_2 to SO_3)

Carbon burnup cells are employed to complete combustion of particles that escape the main bed and are removed from the exhaust gas stream by the cyclone separators. The normal carbon burnup cells provide only about 10 percent of the total thermal output of the bed system; they operate at about 2000 F (1366.7 K) and accomplish very little sulfur removal. This characteristic suggests the development of a two-stage furnace bed system in which a lower stage operating at elevated temperature would exhaust gas into an upper-stage bed operating under optimum sulfur capture conditions.

An example of NO_x emissions from atmospheric fluidized beds is presented on Figure 6-13. This curve indicates the effect of fuel nitrogen content and bed operating temperatures on NO_x emissions. Practically all of the nitrogen oxides produced in fluid bed combustion are derived from fuel-based nitrogen, with

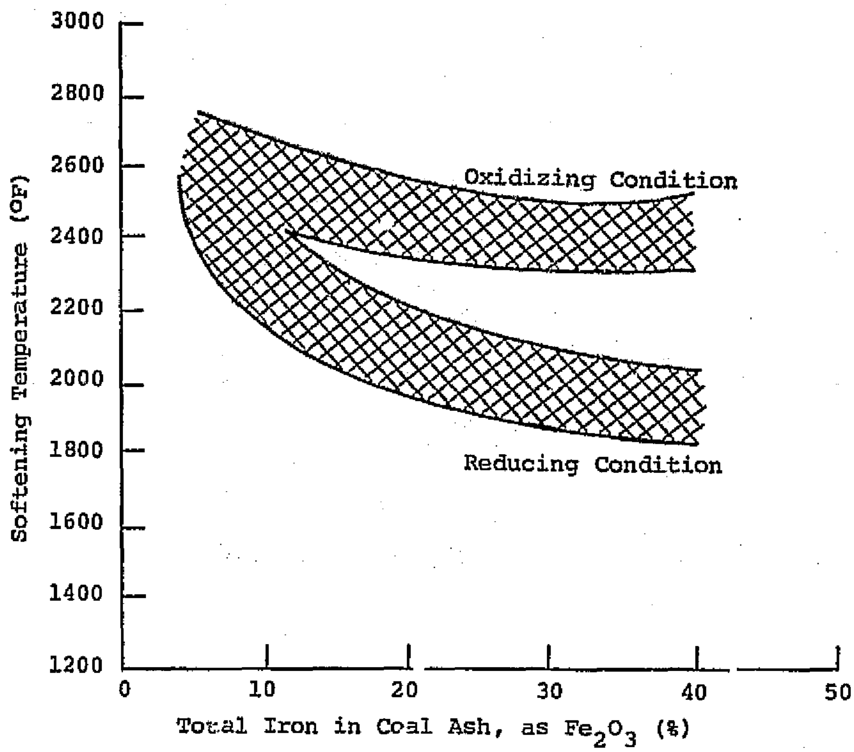


Figure 6-6. Effect of Stoichiometry on Softening Temperature

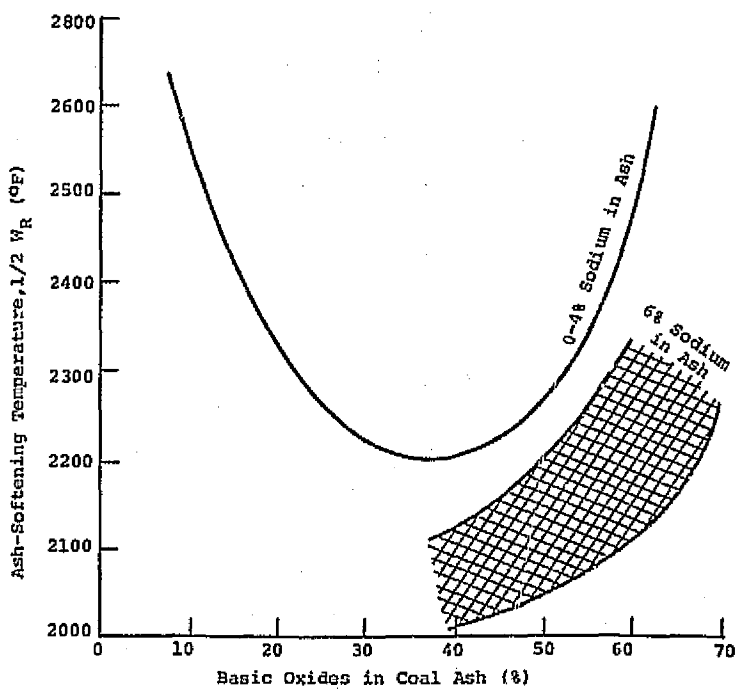


Figure 6-7. Sodium Effect on Ash Softening Temperature

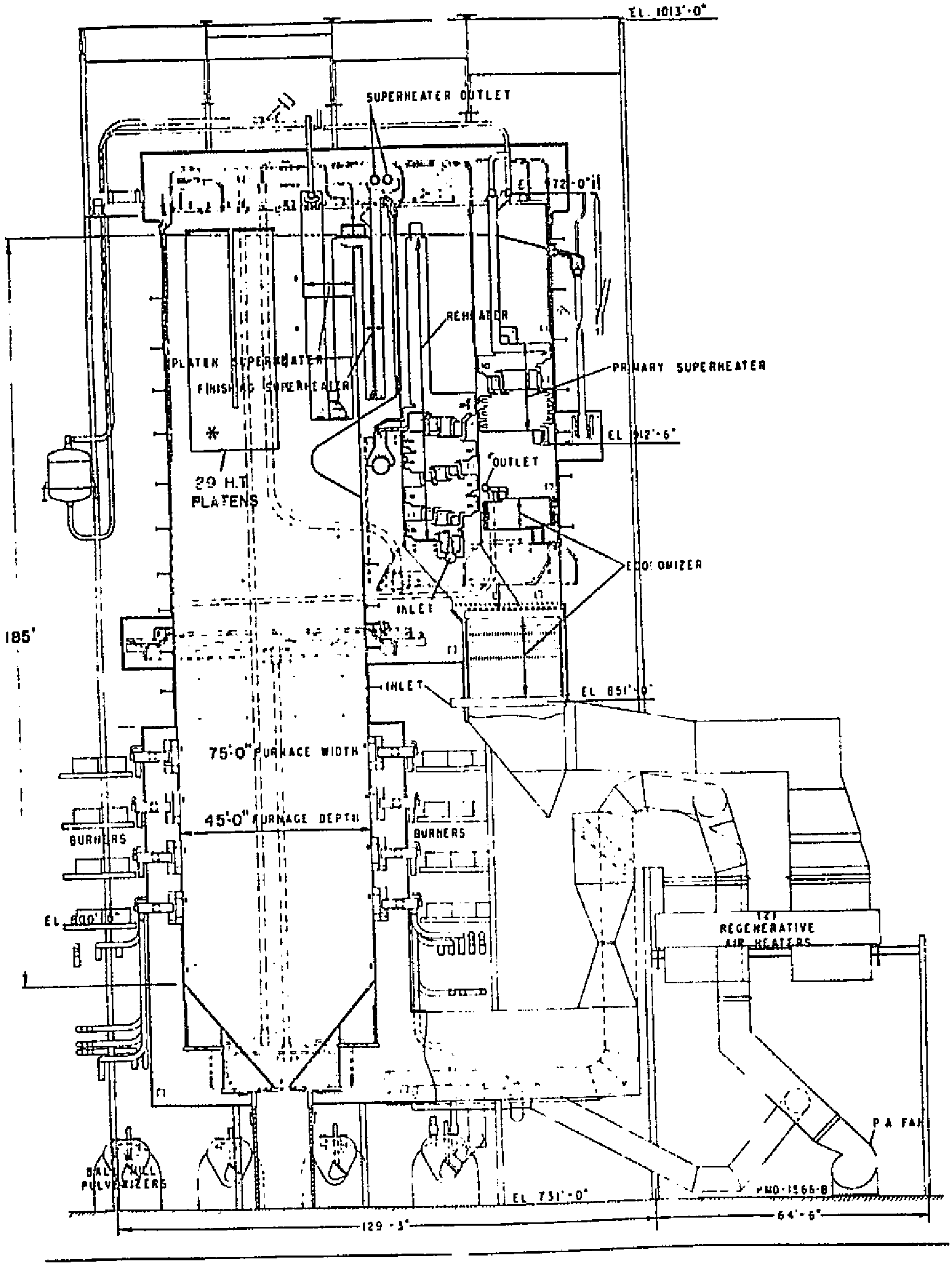


Figure 6-8. Conventional Furnace for Advanced Steam Cycle (3500 lb/1200 F/1000 F)

Table 6-4

ATMOSPHERIC FLUIDIZED BED ARRANGEMENT
(Basic Furnace—Most Cycles)

Characteristics

- Reduced size (about 40% plan area of conventional furnace)
- Crushed coal fired (1/4 inch - 0)
- Limestone bed material for SO_2 control (1/8 inch - 0)
- Low NO_x emissions (half specified limit)
- Requires control of particulate emissions
- Low cold-end dewpoints
- Good heat transfer characteristics
- No coal-ash corrosion

Auxiliary Equipment Included

- Coal transport, storage, drying, crushing, feeding, and injection system
- Limestone transport, storage, drying, crushing, feeding, and injection system
- Bed material extraction, transport, and cooling system
- Mechanical and electrostatic dust collectors
- Air preheaters
- Forced draft, induced draft, primary air and exhaust fans
- Combustion control and safety interlock system
- Valves

Development Requirements

- Alloys for high-temperature cycles
- Process development for solids handling
- High-temperature-solid transport equipment
- Mechanical development for tower construction
- Multi-tower (multibed) integrated steam and combustion
- Control system development

Table 6-5

ATMOSPHERIC FLUIDIZED BED—GENERAL CONSIDERATIONS

Main Advantages

- Reduced furnace size and weight
- No coal-ash corrosion
- No-SO_X tailgas cleanup
- No molten-slag-related problems
- Maximum modular shop fabrication
- Burn any solid fuel
- Reduced coal grinding and drying
- Low NO_X emissions

Major Problems

- High dust emissions require cleanup (mechanical and electrostatic collectors)
- High-temperature tube sulfiding, corrosion-erosion
- Solids handling, higher solids heat losses
- Increased alkaline spent solids disposal (dry)
- CO emission control, higher carbon losses
- High-temperature cycle materials
- Fabrication, cost, availability, expansion, stress analysis, etc.

Limitations

- Maximum surface packing about 10 ft^2/ft^3 (bed volume) with 1 inch o.d. tubes 1400-1600 F maximum bed temperature set by SO_X removal at 1550 F maximum but below ash softening temperature
- Minimum bed ignition temperature 1000 F
- Minimum stable operation 1300 F
- Burn coals only in this study

Special Design Considerations

- Tube vibration and support
- Corrosion and erosion

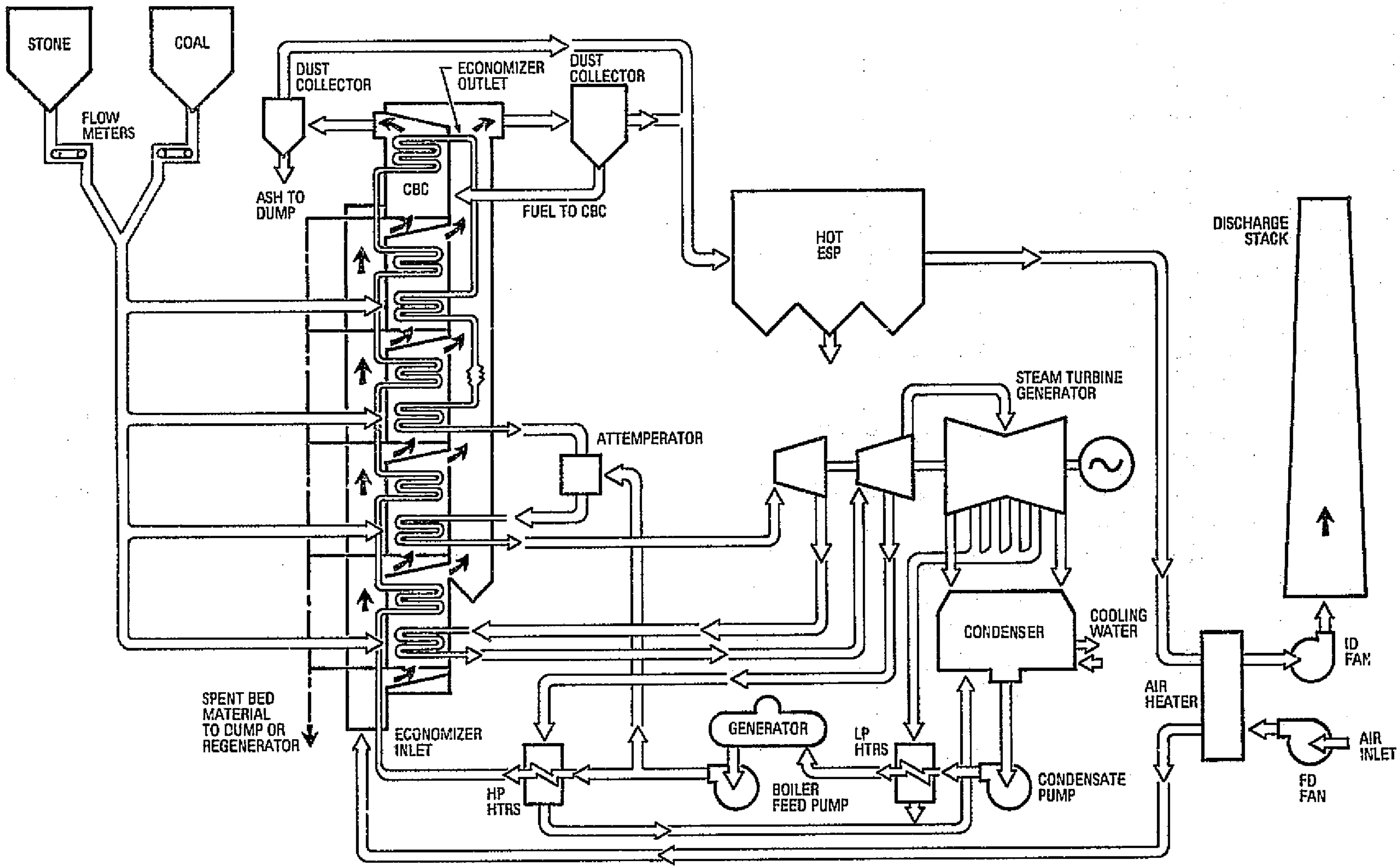


Figure 6-9. AFB Power Plant—General-Flow Schematic Diagram

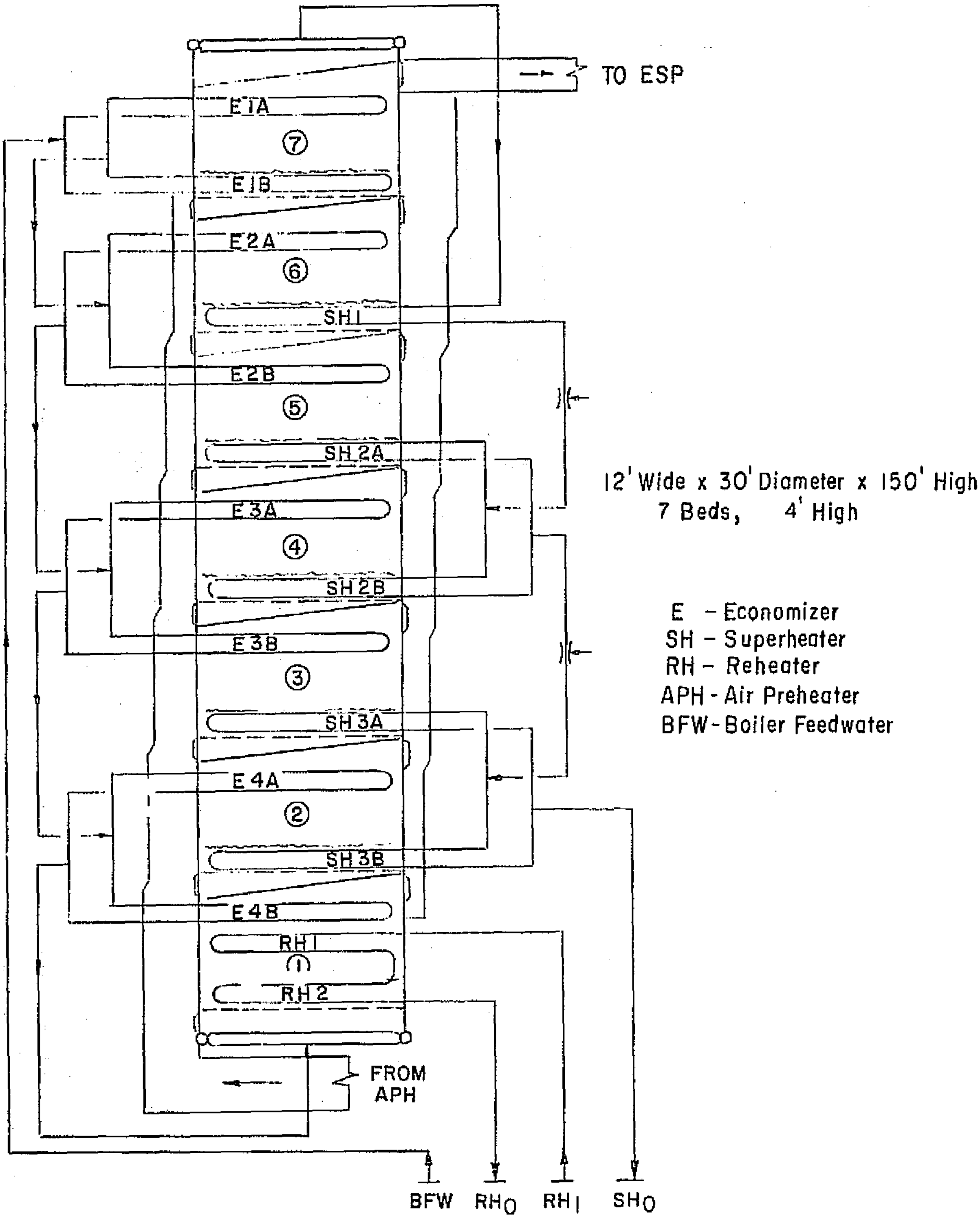


Figure 6-10. Advanced Steam Cycle in AFB (200 MWe) Tower

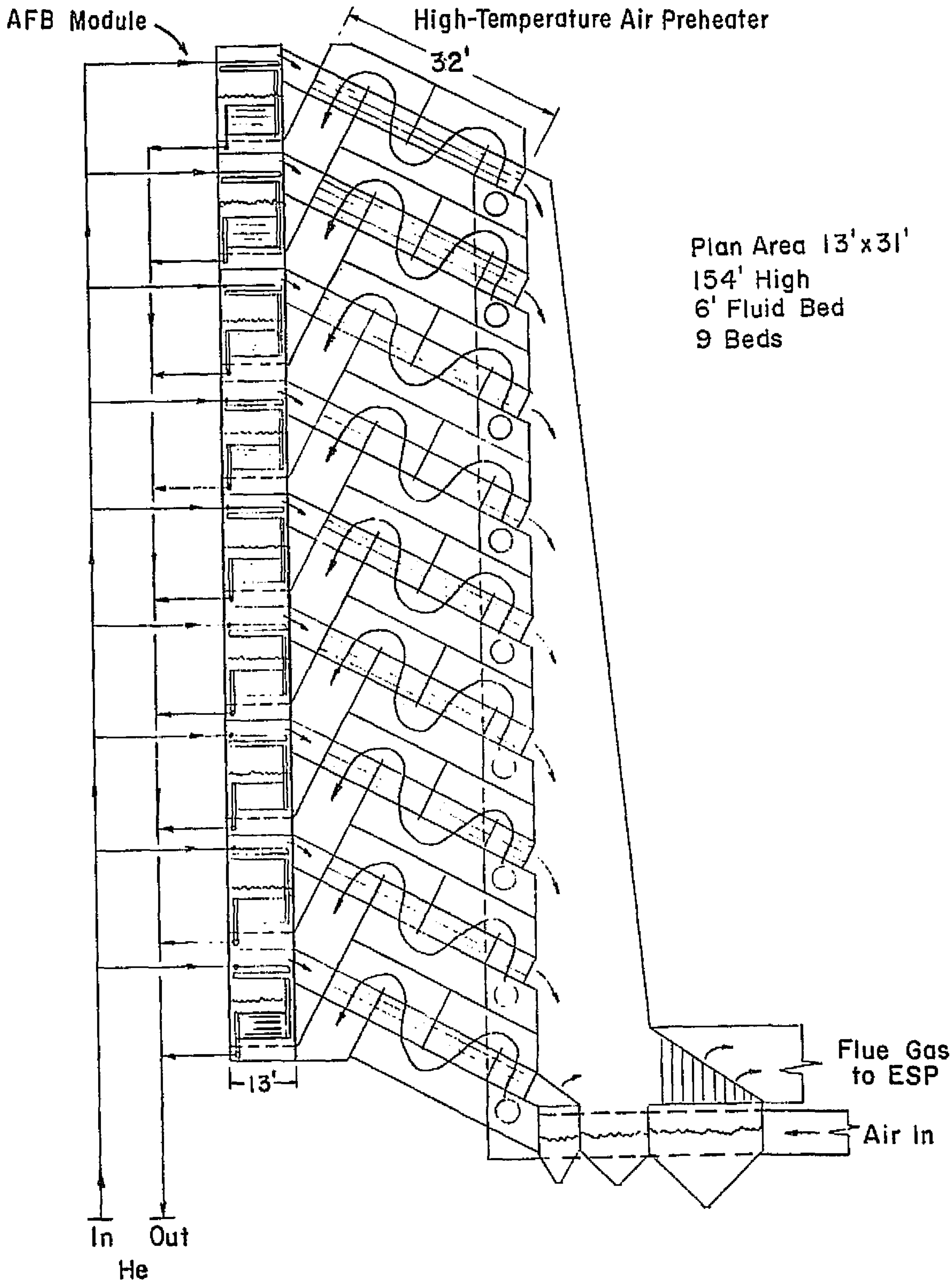


Figure 6-11. Helium Gas-Turbine Cycle in AFB Tower

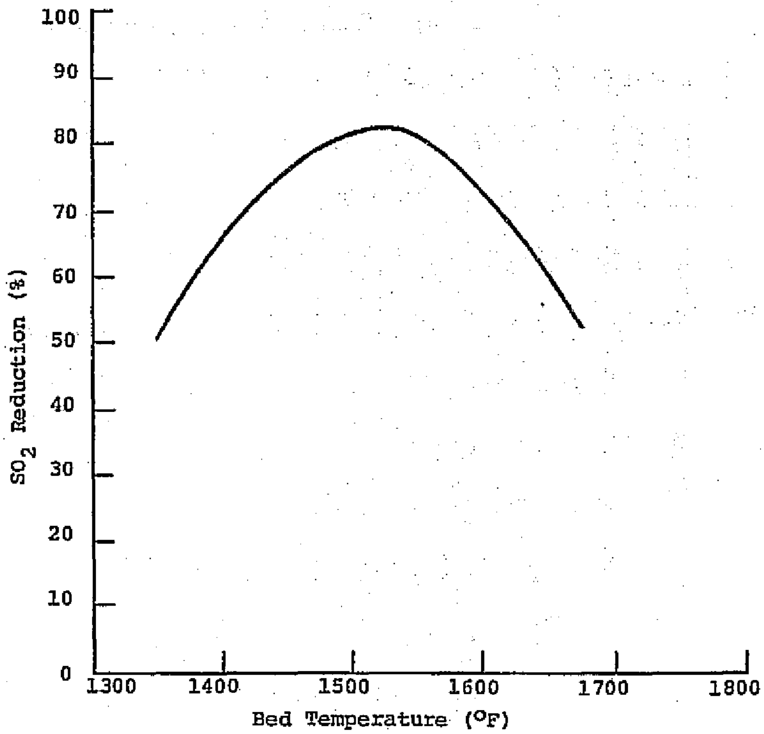


Figure 6-12. SO_2 Reduction from Atmospheric Fluidized Beds

very little thermal NO_x produced from the reaction of gas-phase nitrogen and oxygen. This is a result of mean bed temperatures, which are unfavorable to promoting this reaction.

The resistivity of fly ash from fluidized bed combustors as it varies with temperature is shown in Figure 6-14. The low acidity and free sulfur of dust emitted by fluidized beds results in high resistivity relative to conventional fly ash, and higher precipitation temperatures are definitely indicated.

Typical parameters for fluidized bed arrangement for atmospheric units operating at one atmosphere are presented in Figure 6-15 and Table 6-6. It was not possible to maintain these conditions for every energy conversion system, because of very real limits on the ability to install required heat transfer surface in the fluidized beds and still maintain a size that provided a shippable module configuration. The maximum surface density or packing achievable with 1 inch (2.54 cm) tubes is about 10 ft^2 (32.75 m^2) of heat transfer surface per cubic foot (m^3) of active bed volume.

The fluidizing conditions given in Table 6-6 are indicated for a conventional steam boiler with the heat transfer surface requirements required by that cycle when contained in the given

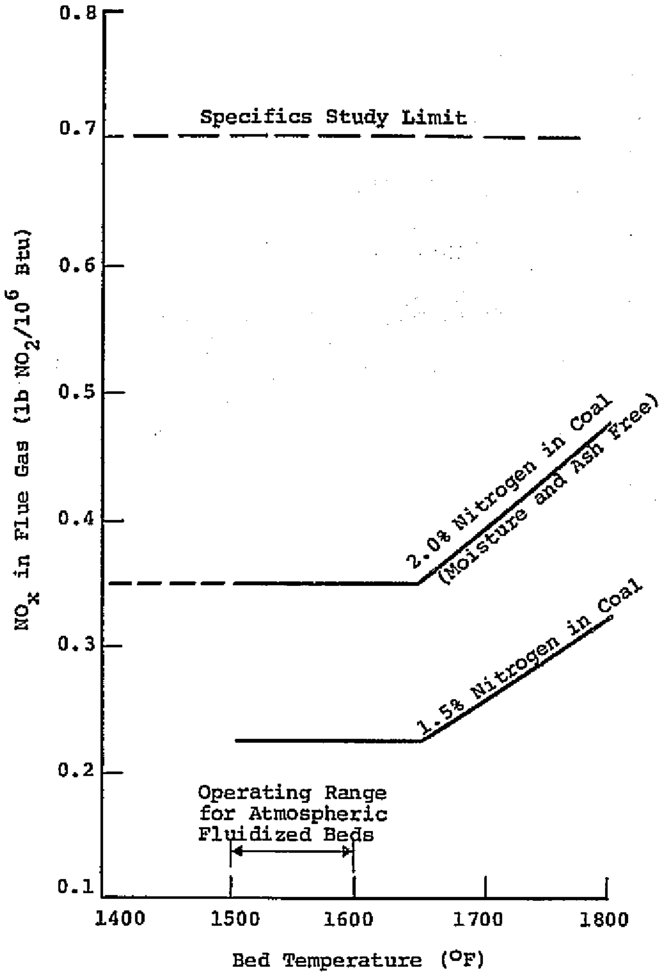


Figure 6-13. NO_X Emissions from Atmospheric Fluidized Beds (Ca O/S ∿ 2/1)

bed configuration. For high-temperature cycles in which the temperature differences for heat transfer are small (relative to conventional steam conditions), a problem is experienced in attempting to fit the required heat transfer surface into the bed volumes indicated in this table. This led to a compromise in operational conditions. Wherever more than seven beds of the indicated size were used in this study, the fluidizing velocities had to be reduced to provide sufficient volume in the beds proper for the required heat transfer surface. In certain cases the bed depth had to be increased, resulting in increased fan power consumption for the units in which there was a limit on the fluidizing velocities.

Atmospheric fluidized bed units were not designed for fluidizing velocities lower than about 2 ft/s (0.61 m/s). With a low range of fluidizing velocities, a smaller mean particle size results in the bed. This causes improved heat transfer and

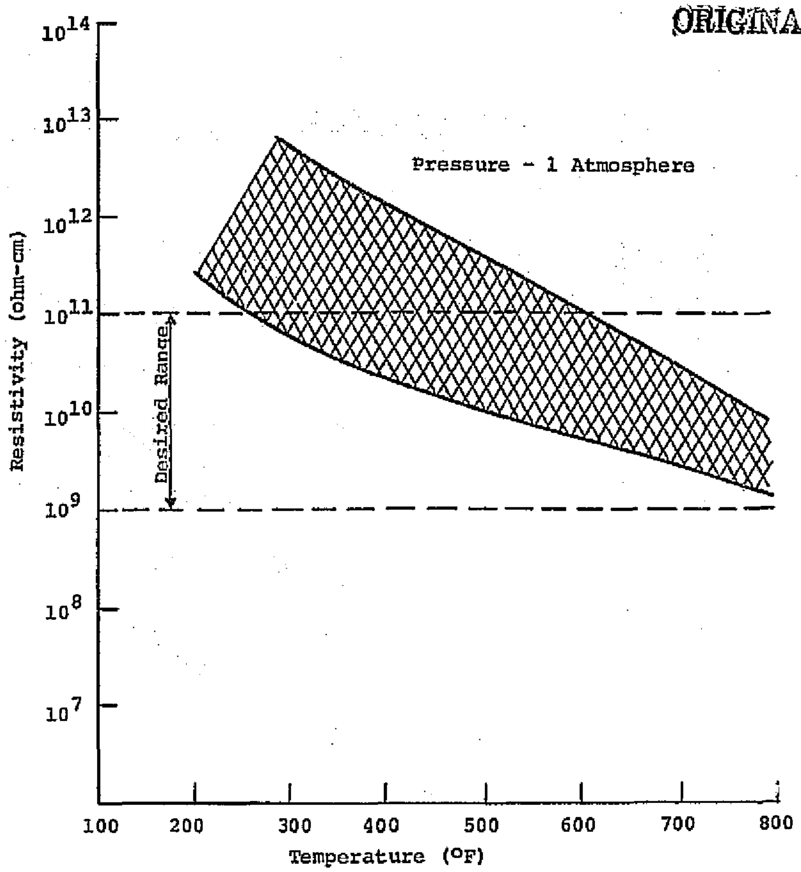


Figure 6-14. Resistivity of Particulates from Fluidized Bed Combustors

combustion efficiency. As fluidizing velocities are reduced to low levels, however, bed agitation is reduced and a tendency of coal material to conglomerate and stick to bed material is to be expected.

The following are general characteristics of the AFB combustion systems as employed in this study:

1. Modular construction reduces field labor considerably.

2. Reduced size—the AFB furnace utilizes approximately 40 percent of the furnace plan area occupied by a con-conventional furnace.

3. Crushed coal is fired. The crushed coal in an atmospheric fluidized bed boiler is sized 1/8 inch (0.318 cm) to 0, as compared to 70 percent less than 200 mesh for a conventional furnace. This sizing requires considerably less grinding power.

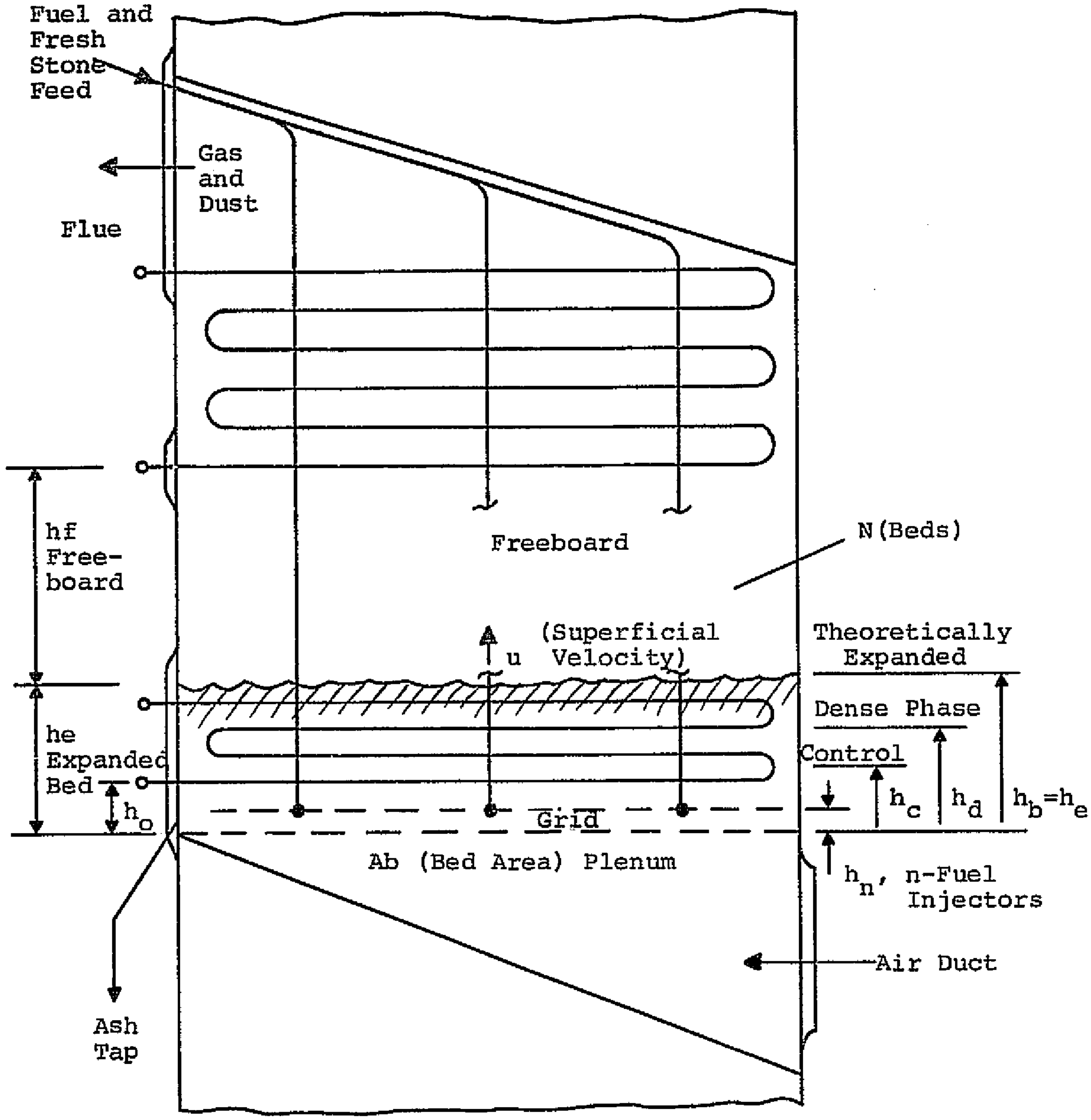


Figure 6-15. Fluidized Bed Parameters

Table 6-6

TYPICAL PARAMETERS FOR FLUIDIZED BED ARRANGEMENT
(Atmospheric Units)

CONDITIONS	MAIN BEDS	CARSON BURNUP CELL (CBC)
Bed Temperature (T_B), °F	1550 F	2000 F
Excess air (XSA), %	20	20
Particle Size Fed*	1/8 in.-0	1/8 in.-0
<u>Velocities</u>		
U_{mf}, ft/s** (min. fluidizing)	4.7	5.2
Uc, (control)	5.0	5.3
Ud, (design)	10.0	8.0
Ut, (terminal)	26.0	32.0
<u>Transfer coefficient</u>		
U_o, $\frac{Btu}{hr\text{-}ft^2\ F}$	Tube 40 Wall 50	40 50
<u>Heights</u>		
h_s (static), ft	2.0	2.2
h_c (control), ft	2.5	2.4
h_d (dense phase), ft	3.8	3.5
h_b (design bed), ft	4.0	4.0
<u>Bed densities</u>		
ρ_B (bulk), lb/ft^3	89	89
ρ_P (particulate), lb/ft^3	164	164
<u>Geometry</u>		
h_n, ft	0.25	0.25
h_o, ft	1.5	(1.5)
h_e - h_b**, ft	4.0	4.0
h_f, ft	6	6
Ab/n, (ft^2/injector)	9	3
P_B (in. H_2O) bed	34+2=36	36
P_G (in. H_2O) grid	12	12
P_S (in. H_2O) total	48	48
P_S (in. (psi) total	1.73	1.73
<u>Example</u>		
200 MW N (beds)	6	1
Tower		
Fluid W (width) ft	12	12
Bed		
Size L (height) ft	30-	30-
Ab (area) ft^2	360	360
n injectors	40	120

* Fluidizing condition estimates are based on a single 0.05 inch mean particle size. U_{mf} will vary roughly with mean particle size to the 3/2 power.

** Certain high-temperature cycles are surface-controlled and fluidizing velocities decreased, with bed heights increased to fit required heat transfer surfaces into beds. Maximum surface packing is about 10 ft^2 per cubic foot of bed volume when 1 inch o.d. tube, is used.

† Parameters defined in Figure 6-15.

4. Dry limestone bed material was used for SO_2 control. This material is sized 1/8 inch (0.318 cm) to 0 and requires only nominal grinding power and drying to render it suitable for use in an AFB boiler. Limestone selection is BCR case 1359 or its equal and limestone consumption was estimated using a 2/1 lime-to-sulfur stoichiometric ratio to achieve an 85 percent sulfur removal efficiency. SO_2 emissions were controlled within the study limits.

5. Low NO_X emissions are typical in the atmospheric fluidized bed. Approximately one-half of the allowable limit is emitted. Practically all of the nitrogen oxides formed are derived from fuel-based nitrogen. Very little thermal NO_X is formed in the combustion process, because of low combustion temperatures. The main fluidized beds operate at 1500 F (1116.7 K) and the carbon burnup cell operates at 2000 F (1366.7 K), and the carbon burnup cell operates at 2000 F (1366.7 K).

6. The atmospheric fluidized bed requires greater control of particulate emissions than does the conventional furnace. In the combustion process approximately 0.26 pound (0.117 kg) of limestone per pound (0.45 kg) of fuel is required. In addition to the stone reaction products that are formed, the coal fly ash must be collected and removed. Mechanical collectors from the main combustion cells and the carbon burnup cell are required before the gas is emitted to the electrostatic precipitator. Approximately 0.15 pound (0.07 kg) of solids per pound (0.45 kg) of fuel is tapped from the unit as granular bed material. Approximately 0.16 pound (0.07 kg) of dust per pound (0.45 kg) of fuel is collected from the carbon burnup cyclone and approximately 0.02 pound (0.009 kg) of dust per pound (0.45 kg) of fuel is collected at the electrostatic precipitators. The total solids produced by the atmospheric fluidized bed are approximately 0.33 pound (0.15 kg) per pound (0.45 kg) of fuel. This material is alkaline and should be handled in a dry state.

7. The AFB operates in an alkaline environment at reduced temperature level, and very low sulfuric acid cold-end dew points are attainable. This realistically allows for the reduction in stack gas temperature and consequent improvement in efficiency.

8. The AFB achieves on average improved heat transfer coefficients, as compared with a conventional furnace, resulting in lower installed heat transfer surface and weight.

9. There is no coal-ash corrosion problem. The ash products produced are soft and will not adhere to heat transfer surfaces. With proper combustion control, the classic coal-ash corrosion problem will not occur in fluid bed combustion processes.

The fuel diagram for the AFB furnace is shown as Figure 6-16. The auxiliary equipment required with a fluidized bed unit, which is discussed in later sections, comprises the following items:

1. Coal handling, drying, crushing, and feeding system.

2. Limestone handling, drying, crushing, and feeding system.

3. Hot-bed-material handling and cooling system. In this system, heat is recovered from the spent bed material with an airstream and is used to accomplish the coal and limestone drying duty.

4. Mechanical and electrostatic dust collectors.

5. Recuperative air heaters, where required for low-level heat recovery, were considered with a leakage rate of approximately 7 percent. This leakage rate is believed to be somewhat optimistic, but obtainable with advanced design. If air leakage rates of less than 10 percent cannot be economically obtained in recuperative air heaters, the use of an extended surface tubular air heater of advanced design would be indicated.

6. Forced draft, induced draft, primary air, and drying air exhaust fans are required. The power consumption of the fans on a fluid bed unit is typically higher than a conventional furnace. This is offset to some extent by reduced grinding power consumption.

7. A combustion control and safety interlock system is required. Because of the multiplicity of heat transfer circuits and combustion stages or beds required with this system, a more complex startup system, control valve arrangement, and safety interlock system are required.

Development Requirements

The AFB is an advanced furnace concept. Development work is required in the following areas to reduce the cost and complexity of the AFB:

1. A major effort is indicated in establishing requirements for the solids handling system associated with the fluid bed boiler. Much progress has been made in this area, but optimization relative to the whole system and its implied availability is not complete.

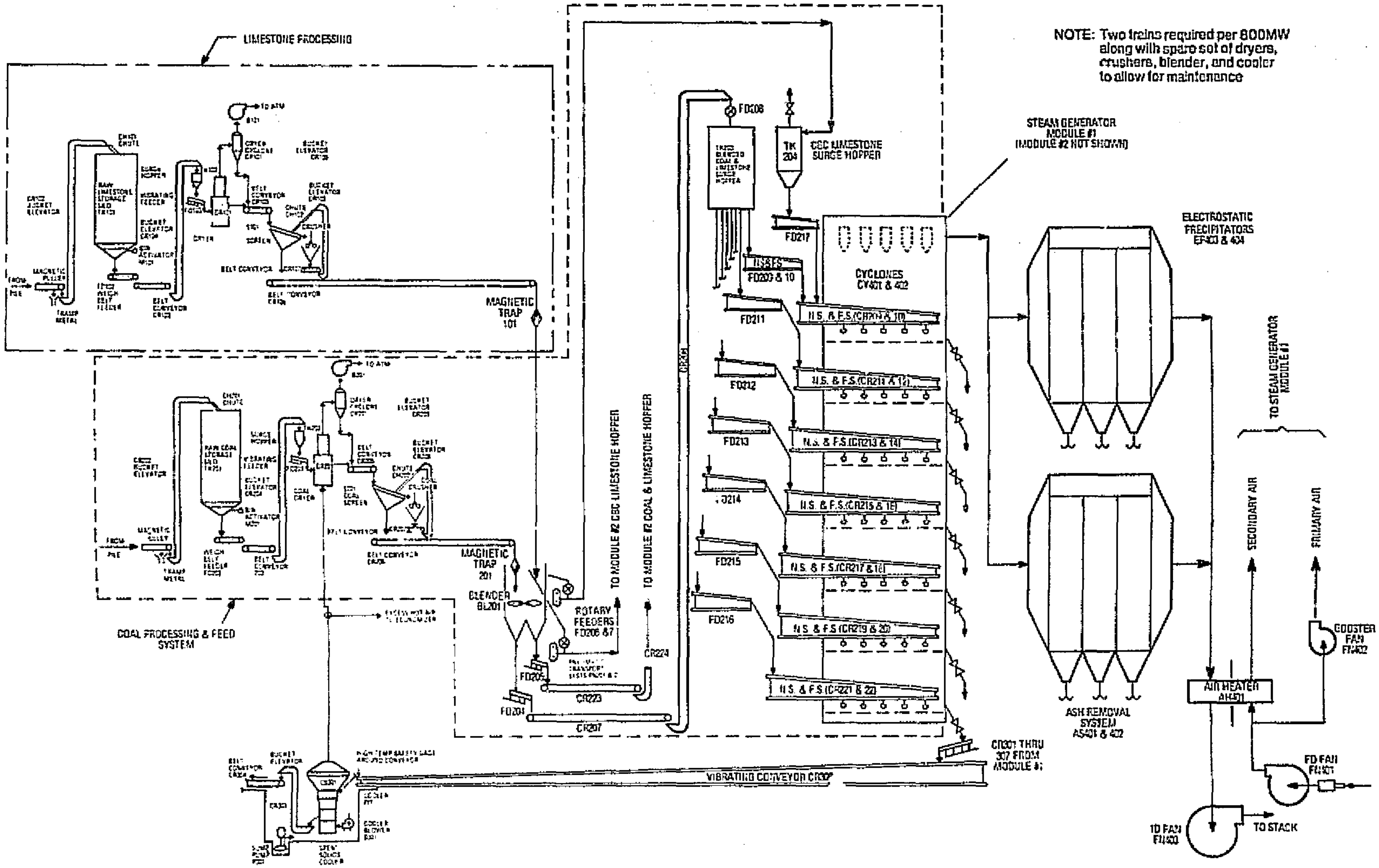


Figure 6-16. Solids Handling and Hot Gas Cleanup Train for AFB Furnace Design

2. Further evaluation of heat transfer surfacing of the individual beds in the unit relative to circulation requirements, pressure drop limitations, and integration into the overall unit control system is indicated.

3. Mechanical development areas:

   - Shutter type dampers for air inlet control.
   - Partial dampers under fluidization grids to permit segmental fluidization of beds.
   - Mechanical support of long tube bundles.
   - Header covers for headers in hot gas streams and in fluidized beds.
   - Solids drop lines and valves.
   - Bed startup and fuel injection systems.

Boiler Efficiency and Performance

Performance analysis indicates the need for further improvement in thermal efficiency, with additional solids cooling and heat recovery for coal drying. There is additional room for improvement in recovery of heat from solids and in reducing unburned combustible losses for the system. One of the more important uncertainties in fluid-bed boiler design is combustion loss and heat release patterns in the main combustion cells.

Depending upon bed operating condition and the type of fuel fired in the main combustion cells, the amount of heat released in the combustion bed itself and the amount of heat released in the freeboard above the bed can vary significantly. Of the heat released in the main bed, 90 percent occurs in the bed and 10 percent in the freeboard. This was assured by assuming enough residence time and sufficient oxygen at high temperature to burn off most of the hydrocarbon, carbon monoxide, and methane being emitted from the bed.

Estimated Emissions

In these analyses, a lime-to-sulfur stoichiometric ratio of 2 to 1 was provided. No addition of salt was employed to activate the bed material. With data available, this appears to be a reasonable estimate for the expected sulfur capture of 85 percent, but a slightly higher lime-to-sulfur stoichiometric ratio may be required in practice to permit some flexibility in bed temperature control. The effect of the additon of salt catalyst may permit the use of lime-to-sulfur stoichiometric ratios of less than 2 to 1, but this requires further development testing. The effect of salt and resulting HCl partial pressure in the bed and their effect on highly stressed stainless steel tubing are not fully known at this time. The estimated emissions from the atmospheric fluid bed are within the specifications of the study with NO_x emissions approximately one-half of the level of that permitted.

PRESSURIZED FLUIDIZED BED

The main characteristics, auxiliary equipment included, and development requirements for pressurized fluidized bed (PFB) furnaces are summarized in Table 6-7. The main advantages, major problem areas, limitations, and special design considerations for pressurized fluid bed units are summarized in Table 6-8. A schematic diagram for a PFB steam power plant is shown as Figure 6-17.

Examples of this furnace module are shown in Figure 6-18, for the advanced steam cycle, and in Figure 6-19, for the closed-cycle gas turbine. The advanced steam example is a tower module with internal waterwalls which would contain the fluidized beds. The closed-cycle gas turbine example is a barrel type with internal refractory lining containing the fluidized beds. In the latter arrangement, post feeders and risers with stick-type tubes are used to limit gas-side pressure drops for the working fluid. This unit operates at lower fluidizing velocities than does the advanced steam module and permits the installation of more heat transfer surface in a given bed. This reduces the number of bed control systems and working-fluid control circuits. Within a 13 1/2 foot (4.12 m) outside diameter for transportability, the maximum bed height achieved with this type of configuration is about 6 feet (1.83 m). Larger sizes would require considerable field fabrication of the containment shell with refractory lining.

When heat transfer surface requirements could not be achieved with this type of construction, the vertical tower type was employed, but the bed depths were increased at the expense of many more beds. An increase in barrel diameter (to greater than 13 feet) to achieve more bed height means field fabrication of shells and was not considered in this study.

An air, gas, and solids diagram in support of the PFB combustion system is presented in Figure 6-20.

An example of SO_2 reduction from pressurized fluidized beds is indicated in Figure 6-21. Dolomite appears to be a more efficient acceptor of SO_2 than limestone for a lime-to-sulfur stoichiometric ratio of about 2 to 1. This curve does not reflect the increased dolomite elutriation or explosion losses that may be experienced in pressurized fluidized beds. With dolomite injection, acceptable sulfur recoveries are achievable over a wider temperature range (up to ∿1900 F [1311.1 K]).

Under PFB operating conditions, a sufficiently high bed temperature must be maintained to calcine the calcium carbonate

Table 6-7

PRESSURIZED FLUIDIZED BED ARRANGEMENT

Characteristics

- Reduced size (about 10 percent plan area of conventional furnace)
- Crushed-coal-fired (1/4 inch to 0)
- Dolomite bed material for SO_2 control (1/4 inch to 0)
- Reduced NO_x emissions (1/3 specified limit)
- Significant particulate removal control required
- Low cold-end dew points
- Good heat transfer characteristics
- No coal-combustor ash corrosion

Auxiliary Equipment Included

- Coal transport, storage, drying, crushing, feeding, and pressure injection system
- Dolomite transport, storage drying, crushing, feeding, and pressure injection system
- Bed material extraction, pressure letdown, transport, and cooling system
- Particulate removal equipment (mechanical and sand bed filters)
- Primary air system, cooling air, and exhaust fans
- Combustion control and safety interlock system
- Valves

Development Requirements

- Alloys for high-temperature cycles
- Process development for solids handling
- High-temperature, high-efficiency particulate removal system
- Mechanical development for tower construction
- Solids feeding and injection systems at pressure
- High-temperature solids transport equipment
- Multi-tower (multibed) integrated gas turbine, steam turbine, and combustion control system development

Table 6-8

PRESSURIZED FLUID BED—GENERAL CONSIDERATIONS

Main Advantages

- Compact furnace—reduced size and weight
- No coal-ash corrosion
- No SO_X tail gas cleanup
- No molten slag problems
- Maximum modular shop fabrication
- Burns any solid fuel
- Reduced coal grinding and drying

Major Problems

- Startup and control of furnace and gas and steam turbines
- High alkali metal content in gases at high temperatures
- High dust emissions require extreme high-temperature gas cleanup for gas turbine; high dolomite elutriation losses, $CaCO_3$ recombination
- High-temperature—tube sulfiding, corrosion - erosion
- Solids handling at high-pressure—injection, extraction
- Increased alkaline spent solids disposal (dry)
- High-temperature cycle—materials fabrication, cost, availability, expansion, stress analysis, etc.

Limitations

- Maximum surface packing, 10 ft^2/ft^3 bed volume with 1 inch o.d. tube
- 1600-1800 F maximum bed temperature set by dolomite calcination and by SO_X removal at 1750 F maximum; below ash softening temperature
- Burns coal only, in this study
- Turbine exhaust heat recovery required
- 1000 F minimum bed ignition temperature; minimum stable operation 1300 F

Special Design Considerations

- Tube vibration support, corrosion - erosion
- Differential thermal expansion
- Startup (auxiliary gas-turbine combustor)
- Unit access

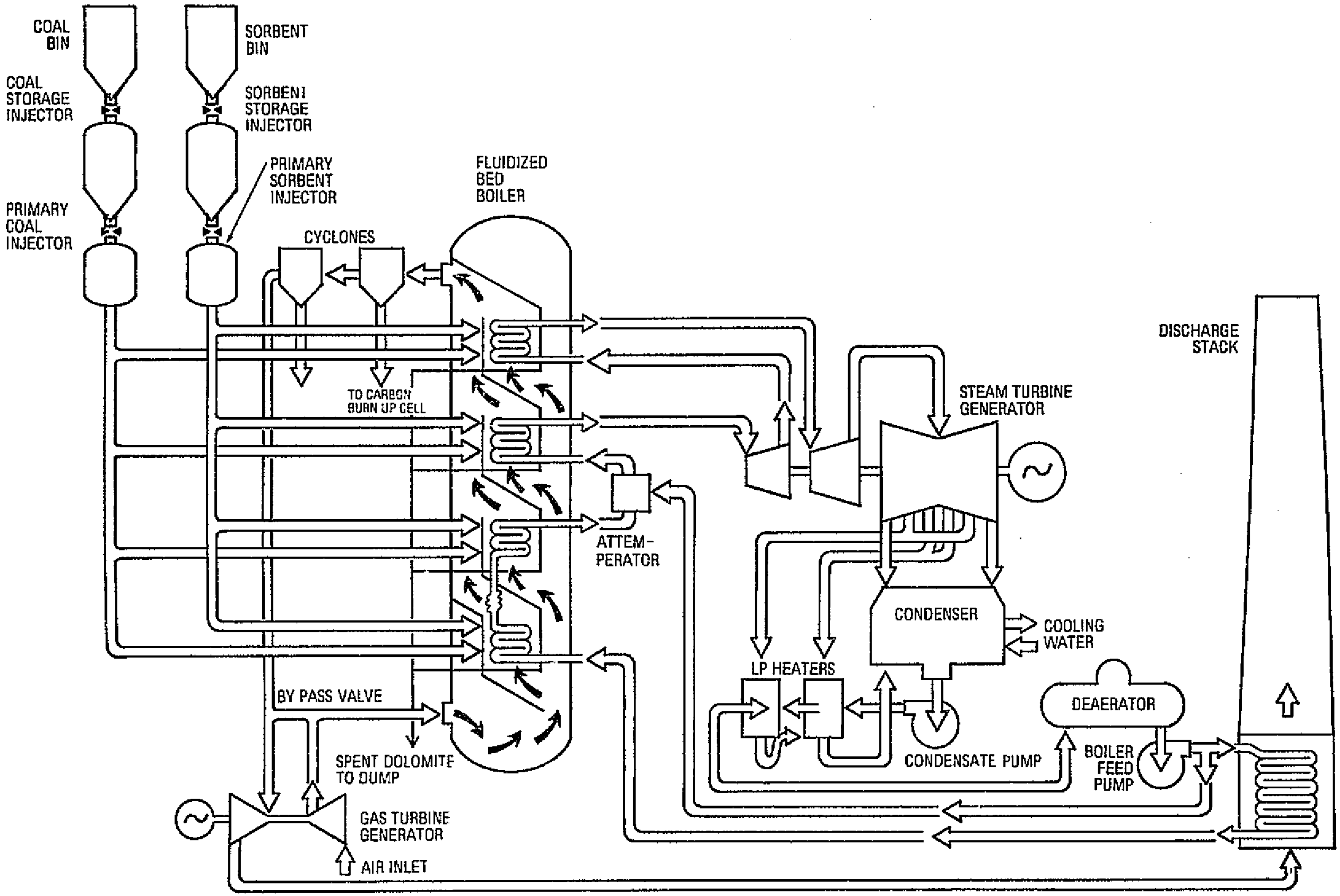


Figure 6-17. PFB Steam Power Plant—Schematic Diagram

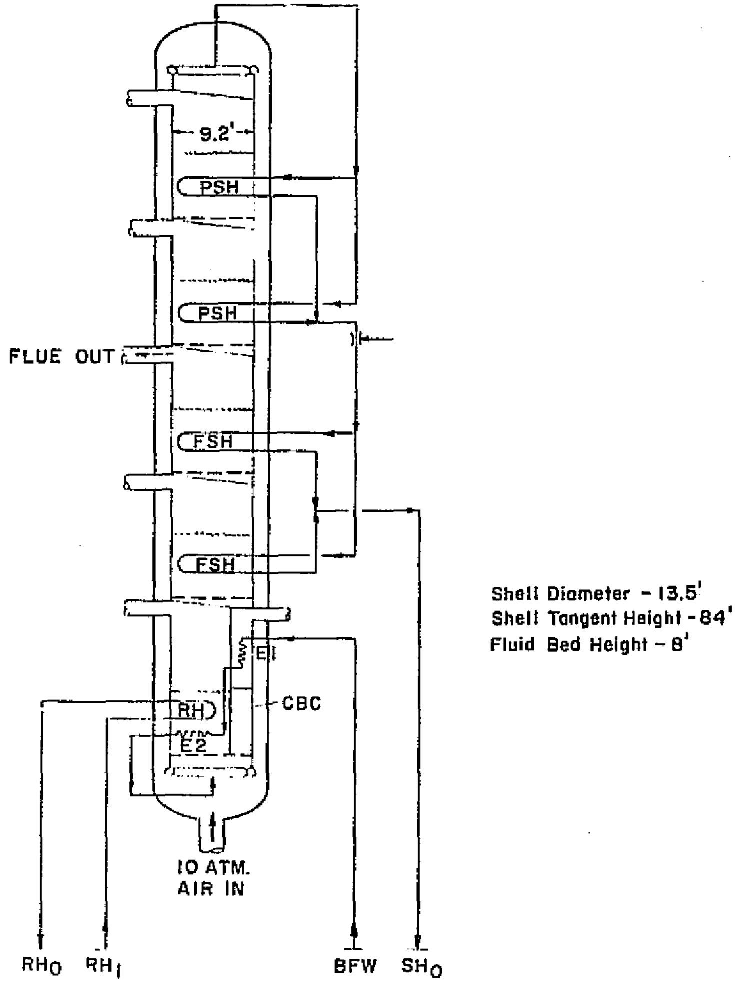


Figure 6-18. Advanced Steam Cycle in 200 MWe PFB Tower

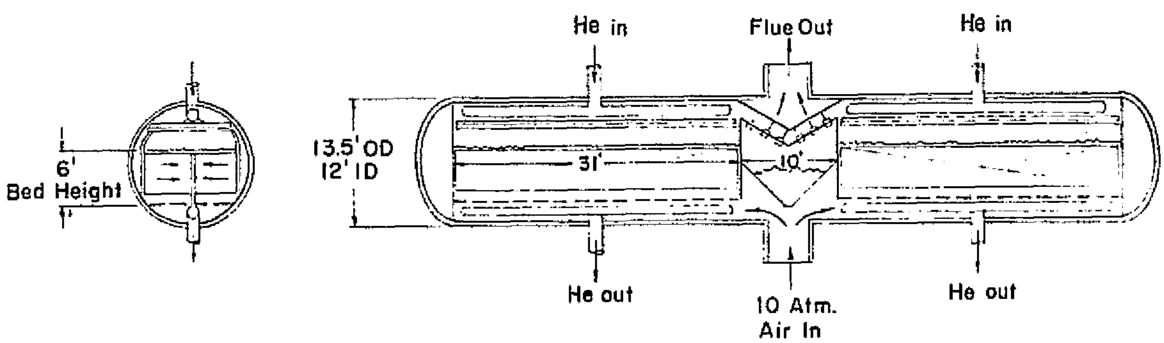
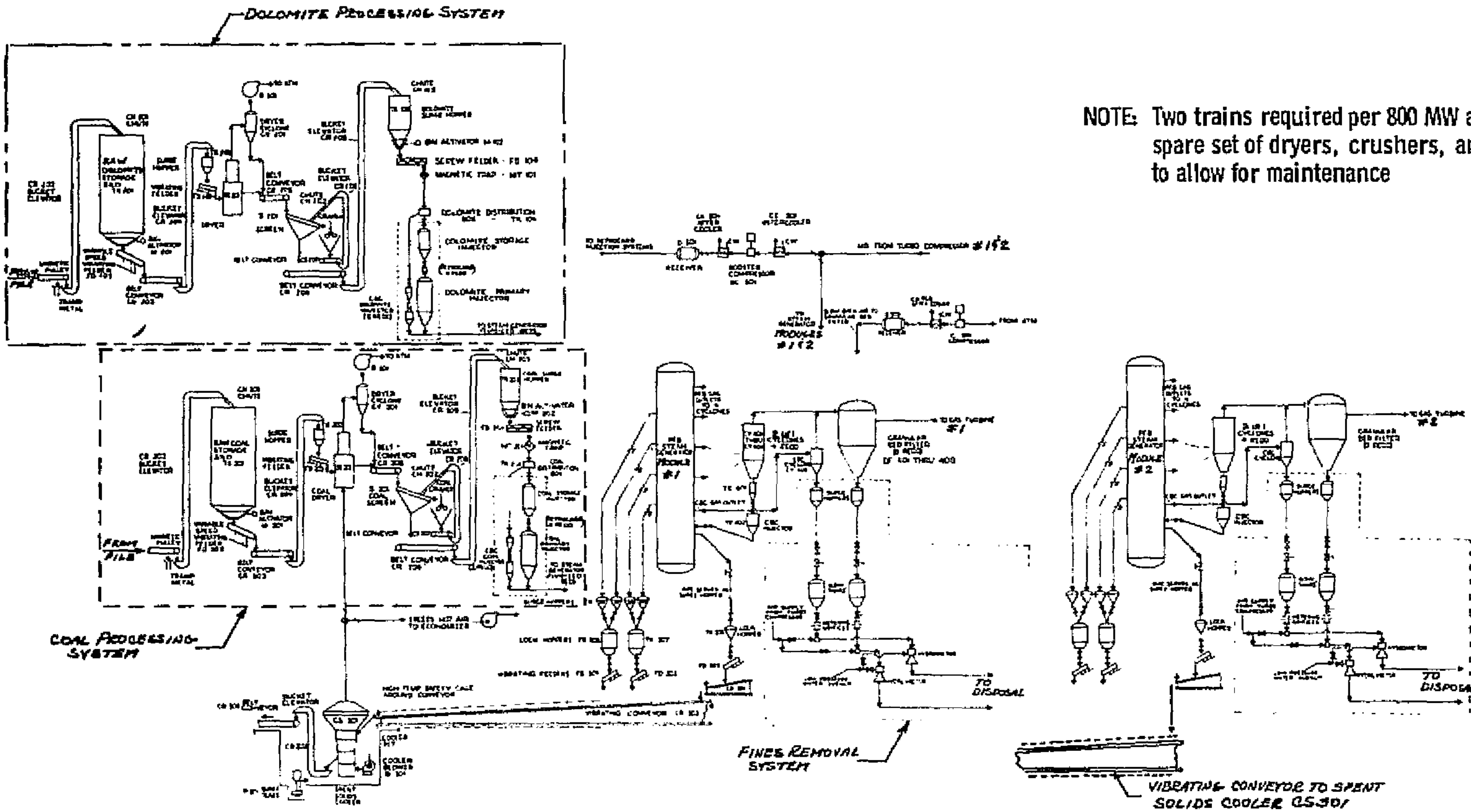


Figure 6-19. Closed Gas-Turbine Cycle in PFB Barrel



NOTE: Two trains required per 800 MW along with spare set of dryers, crushers, and cooler to allow for maintenance

Figure 6-20. Solids Handling and Hot Gas Cleanup Process Train for PFB Furnace Design

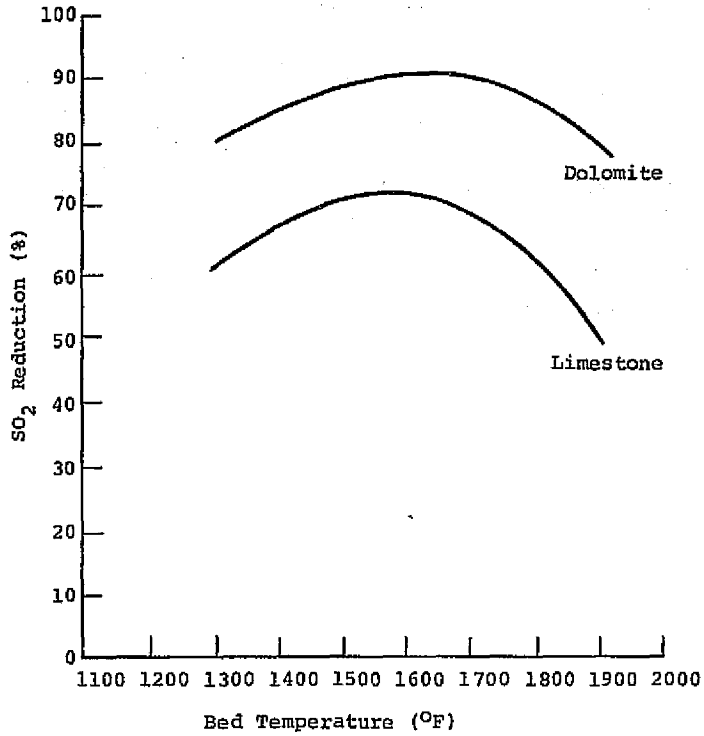


Figure 6-21. SO_2 Reduction from Pressurized Fluidized Beds

present in the feed material. This feature is indicated on the equilibrium decomposition diagram of calcium carbonate in Figure 6-22. For the conditions selected in this study, bed temperature of 1650 F (1172.2 K) is indicated to assure satisfactory calcination and reaction with excess air in the range of 20 to 300 percent.

As the bed operating temperature is raised, alkali vapors would increase in the gas. This will result in corrosion problems for the pressurizing gas turbine and may tend to be a practical limit.

In order to ensure a gas-turbine-quality gas, considerable research and development work must be exerted in the removal of high-temperature particulates from the exhaust gas stream from these units. Additives and modified combustion conditions must be experimented with to reduce the alkali compounds coming off in this exhaust gas stream. Equilibrium analyses of typical pressurized fluidized bed effluences are definitely required to indicate how effluent chemistry might be altered by modified combustion conditions.

An example of nitrogen oxide emissions from pressurized fluidized beds is indicated in Figure 6-23. The NO_X emissions also tend to increase as the lime-to-sulfur stoichiometric ratio increases, and as bed operating temperature rises. The effect of higher lime-to-sulfur stoichiometric ratios on increased NO_x emissions is not fully understood but is thought to be related to the increased CO_2 partial pressure present from the products of calcination of the feed, dolomites, and limestones tested. Higher acceptor circulation rates also diminish the exposed coal-ash species which adhere to the surface of the bed material and are thought to have a catalytic effect on this process.

Typical parameters for fluidized bed arrangement for pressurized units operating at 10 atmospheres (6.89×10^4 N/m^2) are presented in Table 6-9. These conditions reflect desirable

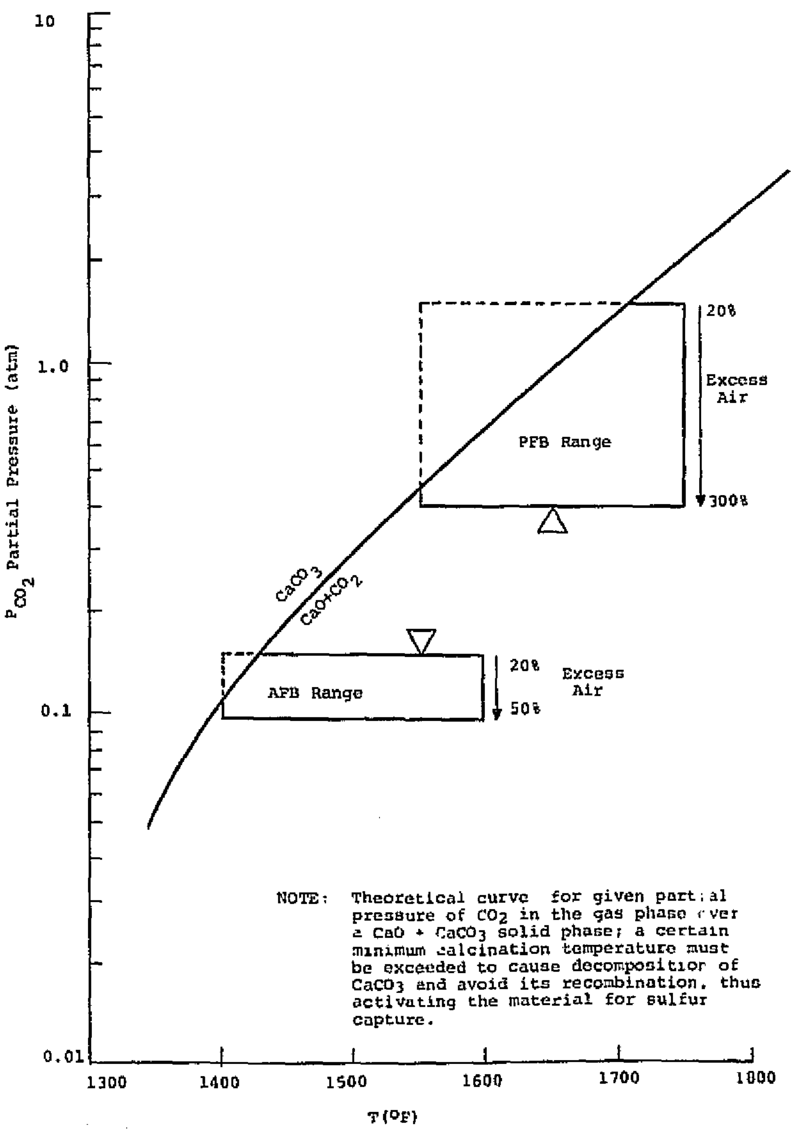


Figure 6-22. Equilibrium Decomposition of $CaCO_3$ in Fluidized Beds

operating conditions for a steam boiler module. As with the atmospheric fluidized bed, high-temperature advanced power cycles with lower log mean temperature differences than exit with the steam cycle require that considerably more heat transfer surfaces be installed in the bed combustion systems. The barrel type of pressurized fluidized bed, previously discussed, permitted the installation of more heat transfer surface in a given bed.

The following are general statements of characteristics of the PFB combustion system studied:

1. Compact modular construction minimizes field labor.

2. Reduced size; the pressurized fluid bed furnace is approximately 13 percent of the furnace plan area occupied by a conventional furnace.

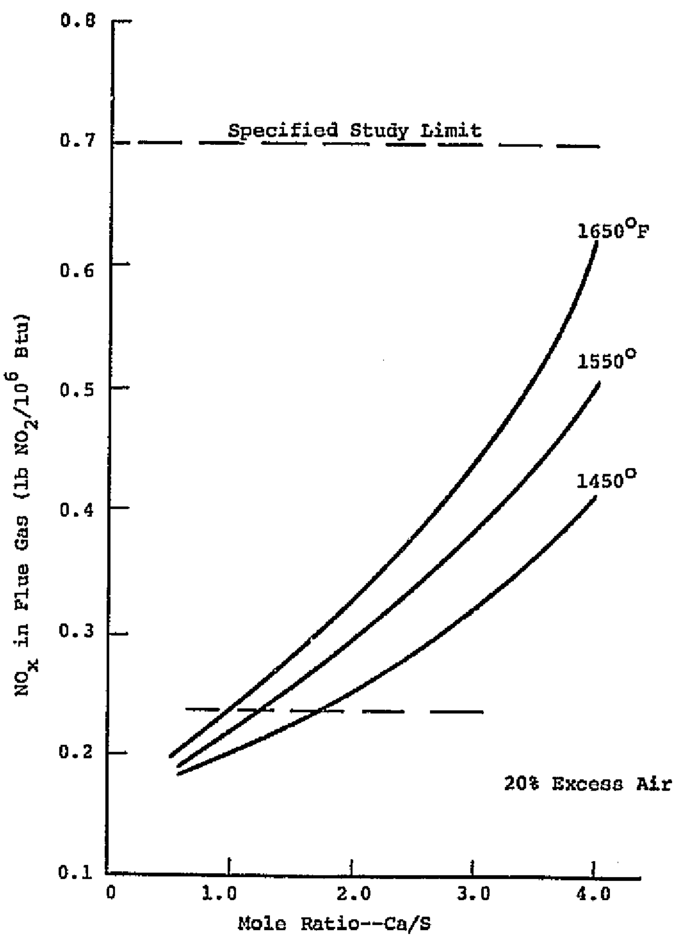


Figure 6-23. NO_X Emissions from Pressurized Fluidized Beds

Table 6-9

TYPICAL PARAMETERS FOR FLUIDIZED BED ARRANGEMENT PRESSURIZED UNITS - 10 ATA

CONDITIONS		MAIN BEDS	CARBON BURNUP CELL BEDS
Bed temperature (T_B), °F		1650°	2000 F
Excess air, %		20	20
Particle size feed		1/4 inch - 0	1/4 inch - 0
Velocities			
Minimum Fluidizing (U_{mf}), ft/s		1.3	1.4*
Control (U_c), ft/s		2.0	1.5
Design (U_d), ft/s		7(4.0)**	4.8(3.0)**
Terminal ($\bar{U}_t$), ft/s		8.0	8.7
Transfer Coefficient			
U_o, $\frac{Btu}{hr\text{-}ft^2\ F}$	Tube	50	50
	Wall	60	60
Heights			
Static (h_s), ft		6.0*	6.2*
Control (h_c), ft		6.5	6.4
Dense Phase (h_d), ft		7.8	7.5
Design Bed (h_b), ft		8.0**	8.0**
Bed Densities			
ρ_B - lb/ft^3 - Bulk		66	66
ρ_p - lb/ft^3 - Particulate		120	120
Geometry***			
h_n, ft		.50	.50
h_o, ft		1.5	1.5
$h_e = h_b$, ft		8.0*	8.0*
$A_{b/n}$, ft^2/injector		9	3
Pressure Drops			
Bed (ΔP_B), in. H_2O		84	84
Grid (ΔP_G), in. H_2O		28	28
Total ΔP_S, in. H_2O		112	112
Total ΔP_S, psi		4.04	4.04
Example			
200 MW_e	Beds**	4	1
Tower	diameter, ft	13.5	13.5
Fluid	width, ft	9.5	9.5
Bed size	depth, ft	9.5	9.5
	area, ft^2	90	90
No. of Injectors, min.		9	9

* Fluidizing condition estimates based on a single .05 inch mean particle size. U_{mf} will vary roughly with mean particle size to the 3/2 power.

** Certain high-temperature cycles are surface-controlled, fluidizing velocities decreased, and bed heights increased to fit required heat transfer surfaces into beds. Maximum surface packing about 10 ft^2/ft^3 of bed volume, when 1 inch o.d. tubes are used.

*** The parameters are defined in Figure 6-15 for AFB and PFB.

3. Crushed coal is fired. The crushed coal in a PFB unit is sized 1/8 inch (0.318 cm) to 0, as compared to 70 percent less than 200 mesh for a conventional furnace. This sizing requires considerably less grinding power but must be injected at high pressure. Injection air supply runs about 100 psi (28.69×10^6 N/m^2) above module pressure.

4. Dry dolomite bed material was used for SO_2 control. This material is sized 1/4 inch (0.635 cm) by 28 mesh and requires only nominal grinding power and drying to render it suitable for use in a pressurized fluidized bed boiler. Dolomite selection was BCR case 1337 or its equal, and consumption was estimated by a 2:1 lime-to-sulfur stoichiometric ratio, achieving a 90 percent sulfur removal efficiency. Dolomite must be injected at pressure similar to the coal.

5. Low NO_x emissions are typical to the pressurized fluidized bed. Approximately one-half of the allowable study limitation is emitted. Practically all of the nitrogen oxides formed are derived from fuel-based nitrogen. Very little thermal NO_x is formed in the combustion process, because of low combustion temperatures. The main fluidized beds operate at 1650 F (1172. 2 K) and the carbon burnup cell operates at 2000 F (1366.7 K).

6. The pressurized fluidized bed requires greater control of particulate emissions than does the conventional furnace. In the combustion process approximately 0.46 pounds (0.21 kg) of dolomite per pound (0.45 kg) of fuel is required. In addition to the stone reaction products that are formed, the coal fly ash must be collected and removed. Mechanical collectors are required following the main combustion cells and the carbon burnup cell before the gas is emitted to the granular bed filters. Approximately 0.19 pound (0.86 kg) of solids per pound (0.45 kg) of fuel is tapped from the unit as granular bed material. Approximately 0.22 pound (0.1 kg) of dust per pound (0.45 kg) of fuel is collected from the carbon burnup cyclone, and approximately 0.02 pound (0.009 kg) of dust per pound (0.45 kg) of fuel is collected at the granular bed filters. The total solids produced by the pressurized fluidized bed is approximately 0.43 pound (0.2 kg) per pound (0.45 kg) of fuel. This material is alkaline and should be handled in a dry state. Recuperation of spent bed material, which was not considered in this study, requires further development.

7. The pressurized fluidized bed operates in an alkaline environment at reduced temperature level, and very low sulfuric acid cold-end dew points are attainable. This realistically allows for the reduction of stack gas temperature and improvement of efficiency.

8. The pressurized fluidized bed achieves much improved heat transfer coefficients on the average, as compared to a conventional furnace. This permits lower installed heat transfer surface and weight.

9. There is virtually no coal-ash corrosion problem within the pressurized combustor typical to fluidized bed combustion. The ash products produced are soft and will not adhere to heat transfer surfaces, as with conventional firing techniques, at very high temperature. Corrosion of pressurizing gas turbine and particulate removal equipment requires further study.

The auxiliary equipment required with a fluid bed boiler, which will be discussed in a later section, comprises the following:

1. Coal handling, drying, crushing, and pressurized feeding system.

2. Dolomite handling, drying, crushing, and pressurized feeding system.

3. Hot-bed-material handling and cooling system. In this system, heat is recovered from the spent bed material with an airstream and is used to accomplish the coal and limestone drying duty.

4. Mechanical dust collectors. These mechanical collectors must be highly efficient to produce gas-turbine-quality gases at high temperature and pressure. Gas-turbine dust limitations are severe, and performance of commercial dust removal equipment must be demonstrated before major development proceeds. Combustion at pressure results in higher alkali metal compositions than in atmospheric units and turbine blade corrosion-erosion is potentially a serious problem. Further experiments with modified combustion conditions are indicated.

5. Gas-turbine air compressor, primary air recompressor, and drying air forced and exhaust fans are required. The power consumption of the compressors and fans on a pressurized unit is much higher than in a conventional furnace, but is offset somewhat by reduced grinding power consumption.

6. A combustion control and safety interlock system is required. Because of the multiplicity of heat transfer circuits and combustion stages or beds required in this system, a more complex startup system and control valve arrangement is required than for a conventional furnace.

Development Requirements

The PFB is an advanced furnace concept. Development work is required in the following areas, to reduce the cost and complexity of the PFB combustion systems:

1. Need for a major effort is indicated in establishing requirements for the solids handling system associated with the fluid bed unit. Much progress has been made in this area, but optimization relative to the whole system and its implied availability is not complete.

2. Further review of the surfacing of the individual beds in the unit relative to circulation requirements, pressure drop limitations, and integration into the overall unit control system is indicated.

3. The exhaust gas cleanup system which will produce a gas-turbine-quality gas is required.

4. Mechanical Development Areas:

   - Shutter type dampers for air inlet control
   - Mechanical support of tube bundles
   - Header covers for headers in hot gas streams and in fluidized beds
   - Outlet nozzles for high-temperature gas, allowing for differential thermal expansions
   - Solids drop lines and valves
   - Bed startup and fuel injection systems

Boiler Efficiency and Performance

Performance analyses indicated the need for further improvements in thermal efficiency, with additional solids cooling and heat recovery for coal drying. There is room for improvement in recovery of heat from solids and in reducing unburned combustible losses in the system. One of the more important uncertainties in fluid-bed boiler design at this point is combustion loss and heat release patterns in the main combustion beds of the boiler. Depending upon bed operating depth and the fuel injection pattern in the main combustion cells, the amount of heat released in the lower bed and the amount of heat released in the upper bed can vary significantly, resulting in temperature gradients. With deeper beds, higher temperatures, and higher oxygen partial pressures, combustion is expected to be more complete than in atmospheric units. Very low unburned hydrocarbon and carbon monoxide emissions are expected. Development testing may indicate the need for multilevel fuel injection and distribution rather than simple injection at the base of the bed.

Estimated Emissions

The estimated emissions from the pressurized fluid bed are within specified study limitations, with NO_X emissions approximately one-third of the permitted level. In this study a lime-to-sulfur stoichiometric ratio of 2 to 1 was provided. No salt was added to activate the bed material.

Present experiment data indicate that this is a reasonable arrangement for achieving a 90 percent sulfur capture. A slightly higher lime-to-sulfur stoichiometric ratio may be required in practice, to allow some flexibility in bed temperature control. The effect of the addition of a salt catalyst may permit the use of lime-to-sulfur stoichiometric ratios of less than 2 to 1, but this requires further development testing. The effect of salt and resulting HCℓ partial pressure in the bed and their effect on highly stressed stainless steel tubing are not fully known at this time.

PRESSURIZED FURNACE ARRANGEMENT

The characteristics, auxiliary equipment included, and development requirements for pressurized (or supercharged) furnaces are summarized in Table 6-10. The main advantages, major problem areas, limitations and special design considerations for pressurized or supercharged furnaces are summarized in Table 6-11.

An example of the pressurized furnace for one energy conversion system application, advanced steam cycle, is presented in Figure 6-24. In this type of furnace conventional type steam bundles were used with internal refractory lining and support of tube bundles from inlet and outlet headers. The pressurized furnace (Figure 6-24) was used for low-Btu gas firing and featured three stages of firing to control maximum gas temperatures and velocities within the containment shell to limit absorption rates and metal temperatures resulting in the heat transfer circuits. The pressurized furnace shown in Figure 6-25 is an example of another application to the closed-cycle gas turbine. This shows the post and stick type tubing arrangement on the working fluid side to control pressure drops. The pressurized furnace shown in Figure 6-26 fires high-Btu fuel in the closed gas-turbine cycle. In this pressurized furnace, 6 stages of firing were used to control gas temperatures and velocities within the module and hence absorption rates in the tubing.

A very real problem is experienced in these furnaces in the inability to take advantage of the maximum absorption rates, because of serious metal temperature limitations.

Table 6-10

PRESSURIZED FURNACE ORIENTATION

Characteristics

Much reduced size

High heat transfer rates attainable

Clean gas fuel only (no particulates or sulfur)

NO_X control by staged firing

Maximum shop assembly

Fast control response

Auxiliary Equipment Included

Valves

Combustion and burner controls

Development Requirements

Diffusion burners

Serious materials limitations for high-temperature cycles, limiting absorption rates

For high-temperature advanced power cycles, the need to develop metals with 1500 to 1800 F (1088.9 to 1255.6 K) operating temperature and 6000 to 8000 psi (4.14×10^7 to 5.52×10^7 N/m^2) allowable design stress is definitely indicated. Figure 6-27 indicates the general effect that pressure will have on heat transfer surface requirements in pressurized or supercharged furnaces.

OPEN-CYCLE MHD RADIANT FURNACE

In support of the open-cycle MHD process the following units were studied:

- Radiant furnace section
- Heat recovery boiler, including a superheated steam section and a reheat steam section
- Intermediate-temperature air heater
- Heat recovery economizer

Only the radiant furnace is discussed in this subsection. Although the main characteristics of the radiant furnace section are similar to conventional furnace types discussed earlier, it also includes special design features. A two-second holding time

Table 6-11

PRESSURIZED FURNACE—GENERAL CONSIDERATIONS

Main Advantages

- Much reduced furnace size and weight
- No coal-ash corrosion
- Technology in hand
- Good controllability
- Fast response
- Maximum modular shop fabrication
- No gas cleanup

Major Problems

- NO_X control
- Combustion control
- High-temperature cycle—materials
- Fabrication, cost, availability, expansion, stress analysis, etc.

Limitations

- Clean fuels only
- Low sulfur, no ash
- AFT 2200-2800 for NO_X-control staged firing, and absorption rate control
- Turbine exhaust heat recovery required

Special Design Considerations

- Tube vibration and support
- Startup (auxiliary gas-turbine combustor)
- Unit access

must be provided for the gases leaving the MHD channel, to permit reactions started at the exit channel to go to completion. As indicated in Section 2.8, this two-second holding time reduces the nitrogen oxides and allows time for seed reactions. Upon completion of this holding time, the secondary air necessary to complete combustion is introduced.

The secondary air must be introduced at the top of the furnace in such a fashion that it mixes more or less thoroughly before the combustion products are ducted through the regenerative

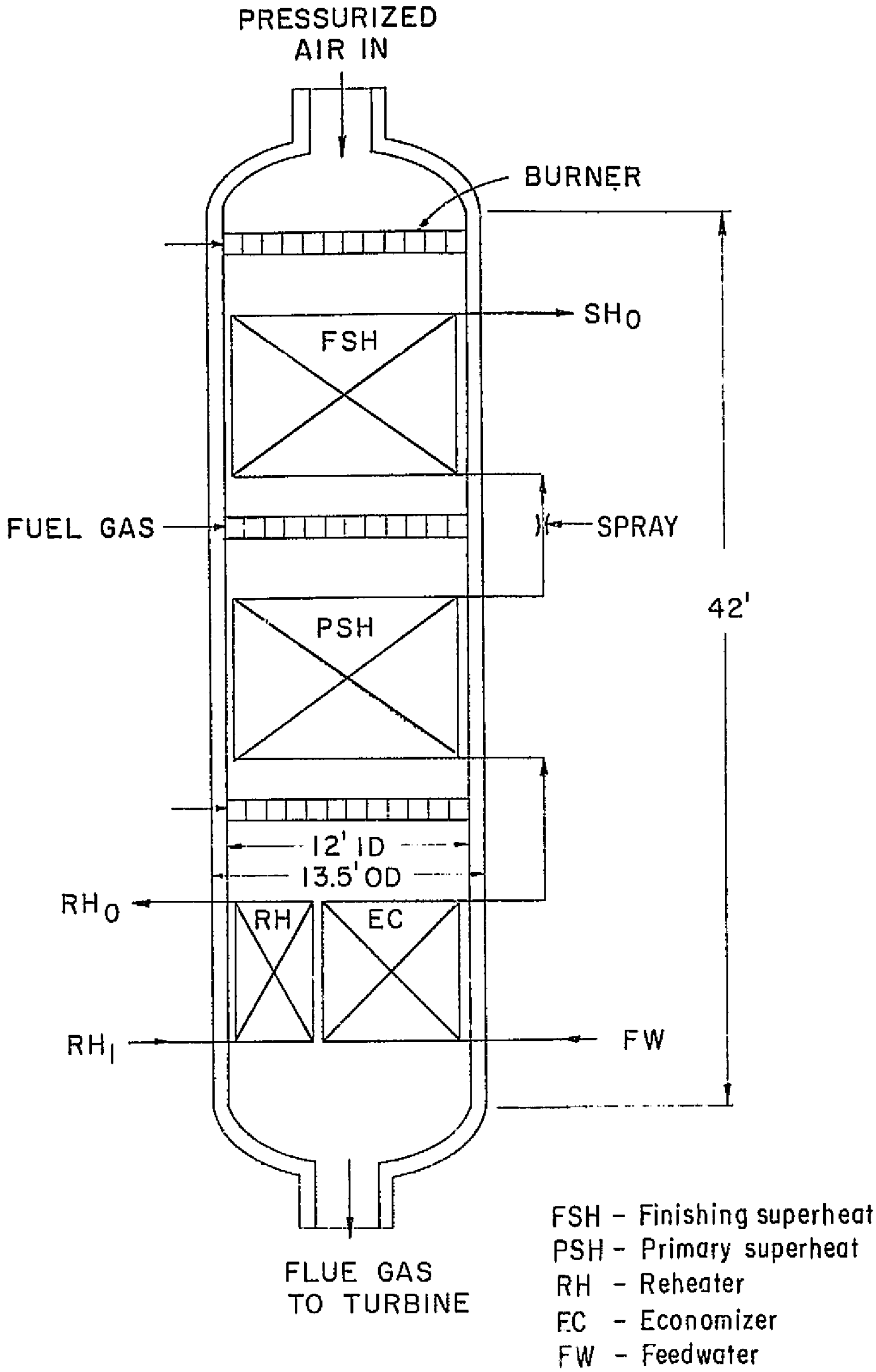


Figure 6-24. Advanced Steam Cycle in Pressurized Furnace

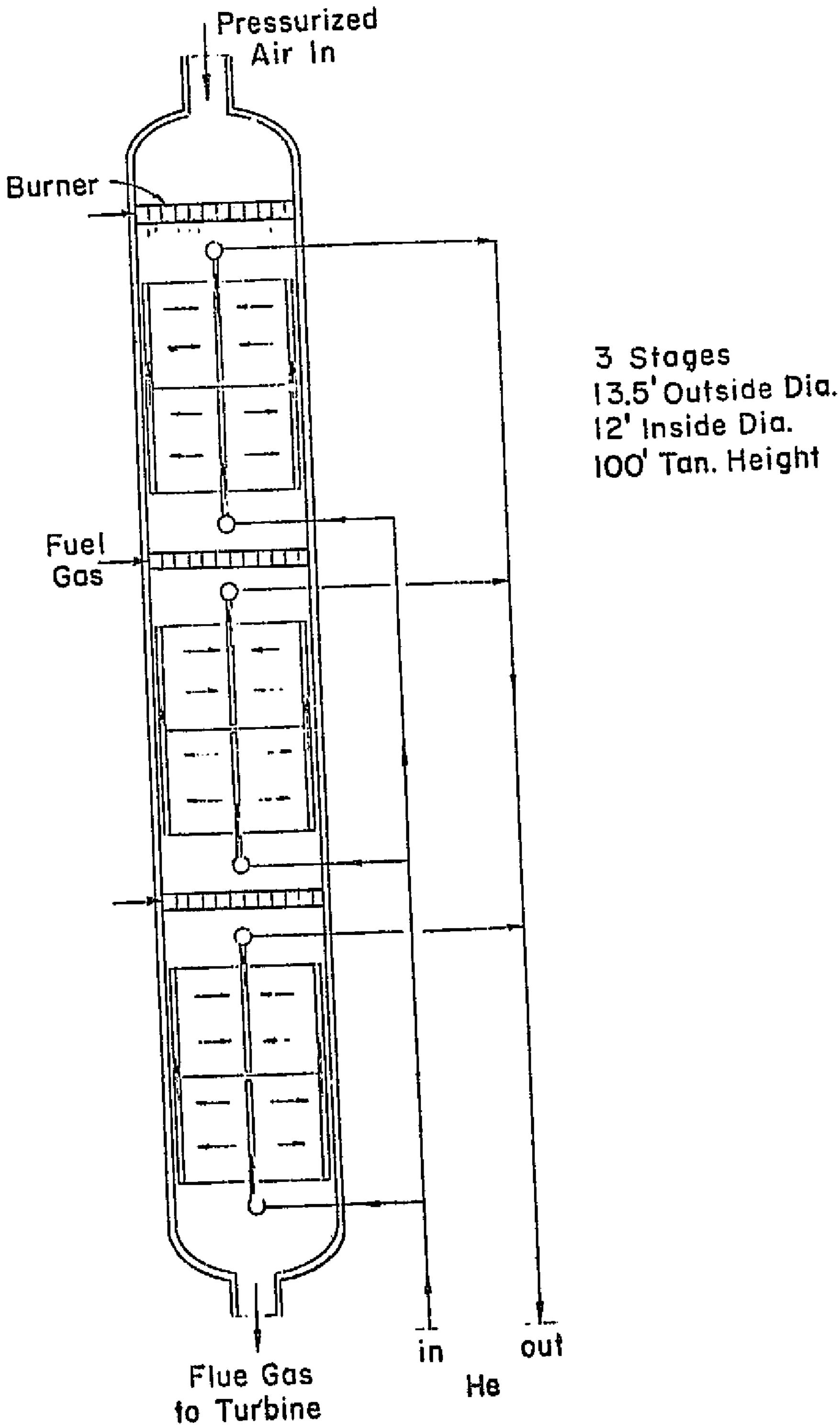


Figure 6-25. Closed Gas-Turbine Cycle in Pressurized Furnace — Low-Btu Fuel

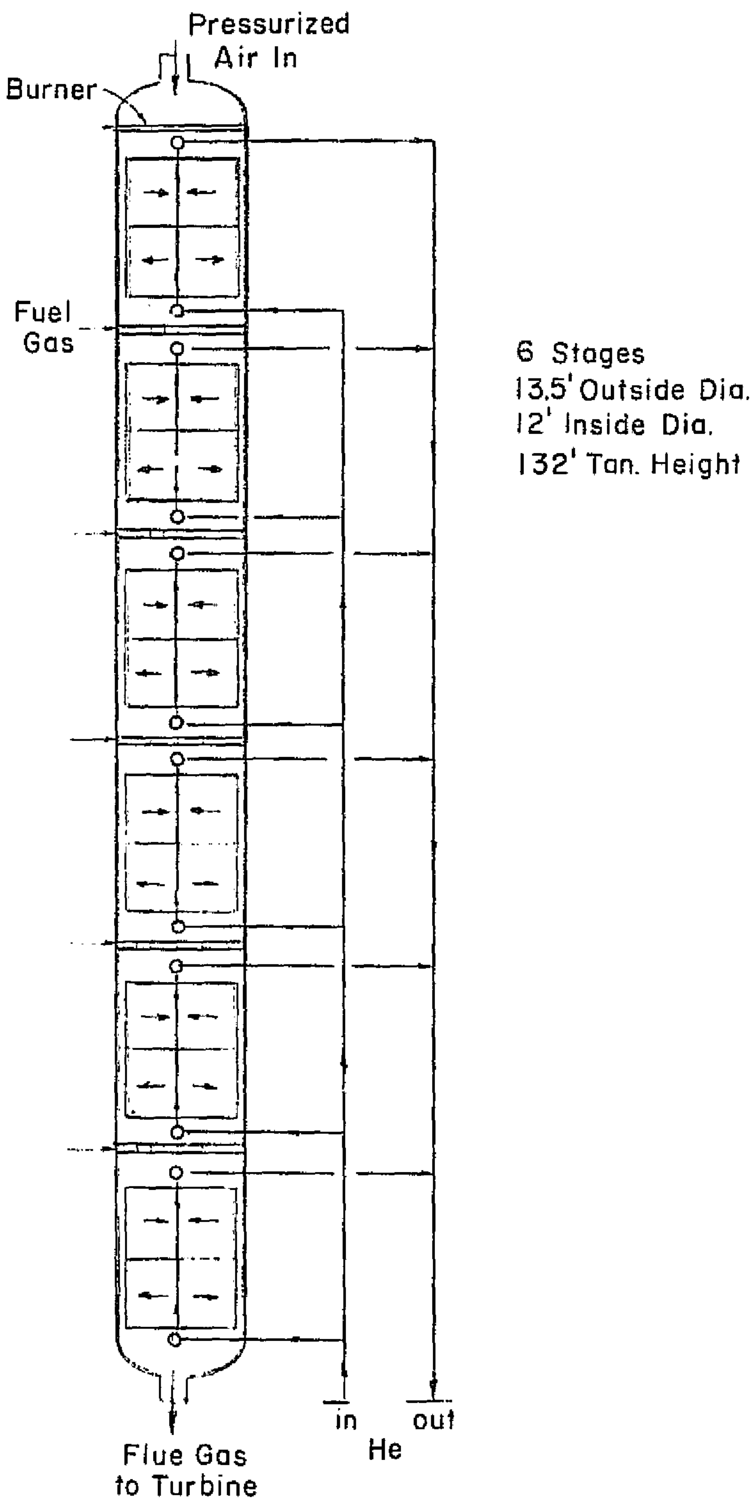


Figure 6-26. Closed Gas-Turbine Cycle in Pressurized Furnace—High-Btu Fuel

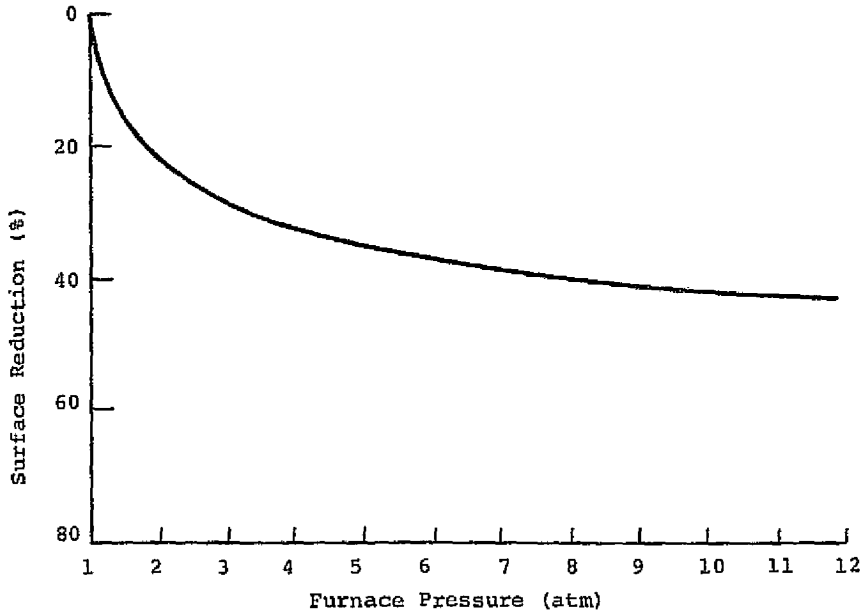


Figure 6-27. Effect of Combustion Pressure on Heat Transfer Surface Requirements

high-temperature air heaters, and an additional mixing volume providing additional residence time must be provided for this purpose. Conditions in the open-cycle MHD combustion gas stream are maintained near or slightly above stoichiometric conditions; that is, oxygen content not exceeding about one-half percent. This is required at the higher temperatures to prevent the formation of potassium pyrosulfates, which in a running liquid condition can be highly corrosive to ferrous and nickel bearing materials.

Special coal-ash and seed corrosion conditions are expected to exist in the radiant furnace section. The environment in the lower furnace is a reducing atmosphere, and slag deposition on the tubes in the radiant furnace is expected to occur. Sootblowers and wall blowers must be provided in the radiant furnace section to allow for cleaning and maintenance of heat transfer surface. It is important that metal temperatures in the radiant furnace section not exceed the 1000 to 1100 F (811 to 866.7 K) outside metal temperatures in the furnace environment.

The main advantages, disadvantages, limitations, and special design considerations for the open-cycle MHD radiant furnace are indicated in Table 6-12. Figure 6-28 is a sketch indicating the design requirements of the radiant furnace.

Table 6-12

OPEN-CYCLE MHD RADIANT FURNACE ARRANGEMENT

Main Advantages

Simple in principle

Uses conventional materials in most of furnace

Disadvantages and Problems

Large furnace size and weight

Difficult NO_x control; substoichiometric combustion

Seed corrosion-erosion, slagging

Coal-ash corrosion-erosion, slagging

Ash disposal, seed recovery required

SO_x, dust emissions require cleanup with dirty fuels

High-temperature cycle-materials, expansion, stress analysis

Limitations

Air injection after holding gases 2 seconds for NO_x reduction

Control gas chemistry below or near stoichiometric at all times

Special Design Considerations

Complex circulation analysis; difficult furnace heat balancing

Furnace cleaning, sootblowing

Slag, seed, and fly-ash removal and separation

NO_x control by staged firing

Secondary combustion stability, furnace mixing in large scale

Corrosion, erosion, refractory selection, and life

Arrangement to suit special air preheating requirements

Development Requirements

Seed and coal-ash corrosion resistance

Materials at very high temperatures, velocities

- K_2SO_4 melts at 1970 F
- $K_2S_2O_7$ melts at 450 F

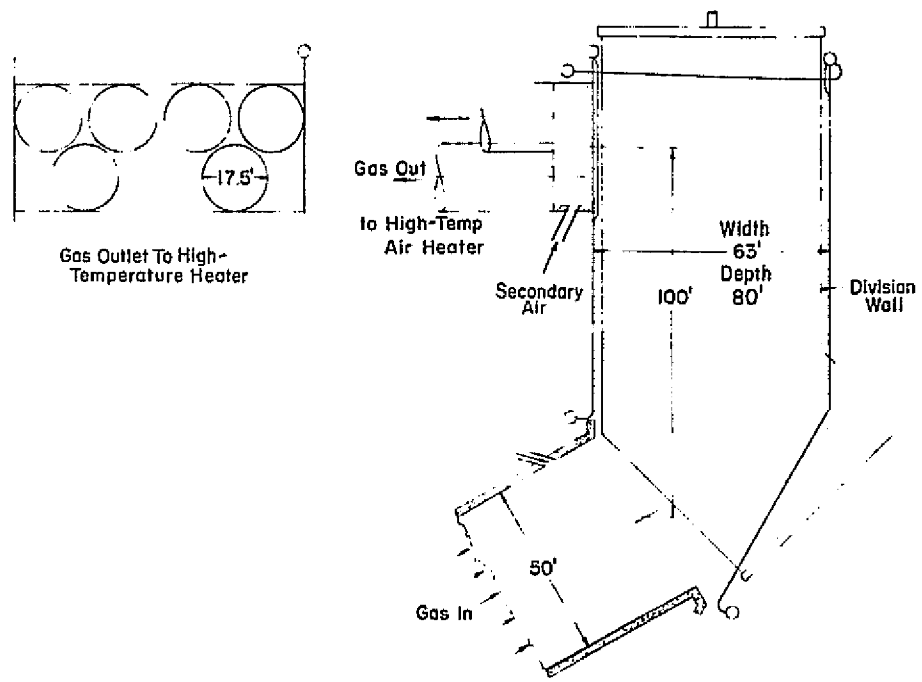


Figure 6-28. Open-Cycle MHD in Radiant Furnace

INTERMEDIATE INDIRECT HEAT EXCHANGERS AND HEAT RECOVERY BOILER

The heat exchange between a prime-cycle working fluid and a secondary-cycle working fluid is defined as an intermediate heat exchange. When heat exchange occurs between products of combustion and a secondary-cycle working fluid, the heat exchanger is defined as a heat recovery boiler. No combustion occurs in either heat exchange device. Table 6-13 summarizes the main advantages and characteristics, major problems, limitations, and special design considerations for these types of exchangers. The intermediate heat exchangers were evaluated for the following components relating to steam generation:

- Condenser for the liquid-metal topping cycle
- Argon cooler for the inert-gas MHD cycle
- Helium cooler for liquid-metal MHD cycle

The heat recovery boilers were considered for:

- Insert-gas MHD cycle (after combustion in parallel cycle)
- Open-cycle MHD (after high-temperature air preheating)
- High-temperature fuel cell cycle.

Figure 6-29 demonstrates an intermediate heat exchanger which was employed for the liquid-metal topping cycle. Figure 6-30 demonstrates a heat recovery boiler for the liquid-metal

MHD cycle. Sketches of other equipment considered in this study of the various cycles are shown in the individual cycle summary parts of this section.

Table 6-13

INTERMEDIATE INDIRECT HEAT EXCHANGER ARRANGEMENT

Main Advantages and Characteristics

Compact, leak-tight construction for gases, liquids

High transfer rates

Reduced size and weight

Modular design, maximum shop fabrication

Major Problems

Control difficulties, control systems

Very high heat fluxes, metal temperatures, and differential temperatures

Cyclic thermal stresses, flow stability, liquid-metal mass transport

High temperature cycle—materials

Fabrication cost, availability, expansion, stress analysis

Limitations

Unit geometry and arrangement limited by complex structural problems

Special Design Considerations

Air in leakage in vacuum units

Tube vibration and support

Structural analysis

Unit access

Cycles Requiring Intermediate Heat Exchange

- Working fluid/steam
  - Liquid-metal topping—condenser
  - Inert-gas MHD—argon cooler
  - Liquid-metal MHD—helium cooler
- Combustion gases/steam
  - Inert-gas MHD
  - Open-cycle MHD
  - High-temperature fuel cells

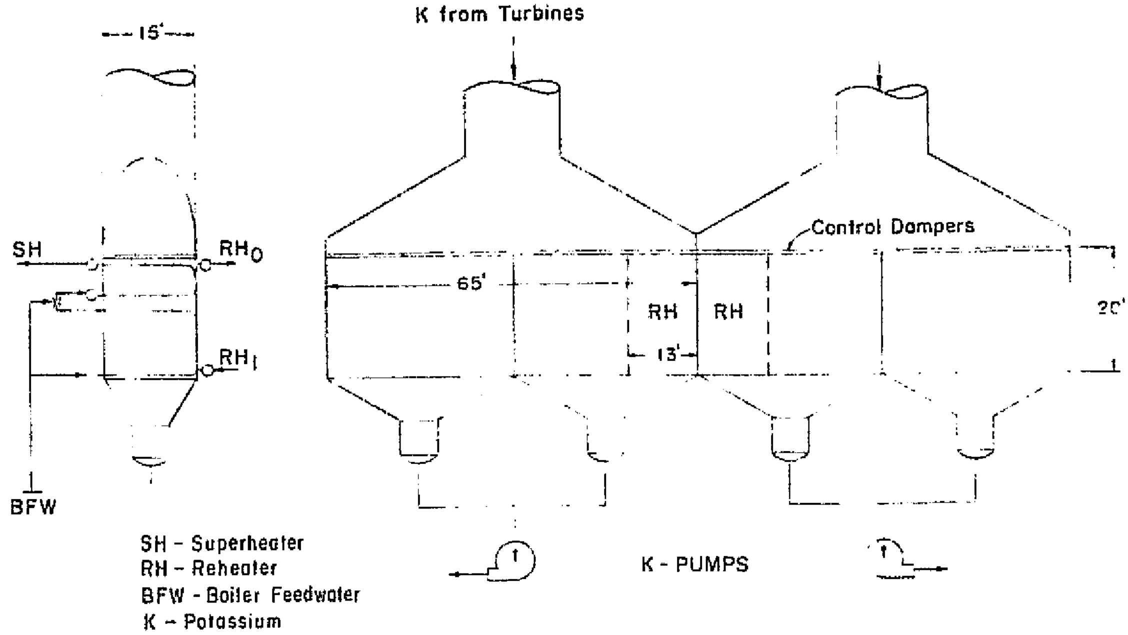


Figure 6-29. Potassium Condenser for Liquid-Metal Topping Cycle

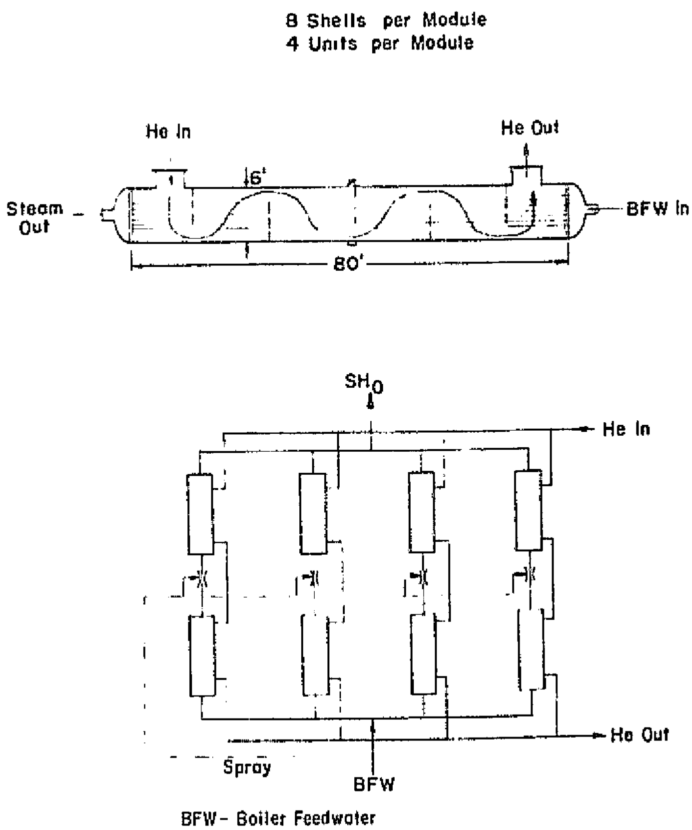


Figure 6-30. Heat Recovery Boiler for Liquid-Metal MHD Cycle

HIGH-TEMPERATURE AIR PREHEATER

High-temperature air preheaters were employed for preheating combustion air. These units were recuperative heat exchangers of metal, tubular construction. The temperature ranges were from 700 F (644 K) to 1500 F (1089 K). The higher temperature ceramic air preheaters employed for final preheat in the open-cycle MHD system are not discussed in this section. The major advantages, disadvantages, limitations, and special design considerations for this type of equipment is summarized in Table 6-14.

Table 6-14

HIGH-TEMPERATURE AIR HEATER ORIENTATION

Major Advantages and Characteristics

Improved cycle thermal efficiency

Available technology

Disadvantages

Large size and weight

Inefficient heat transfer (tubular designs)

Considerable field fabrication (large units)

Limitations

Material strengths and properties at high temperature

Module size limited at higher pressure by shell design considerations

Special Design Considerations

Low heat transfer rates with low allowable pressure drops

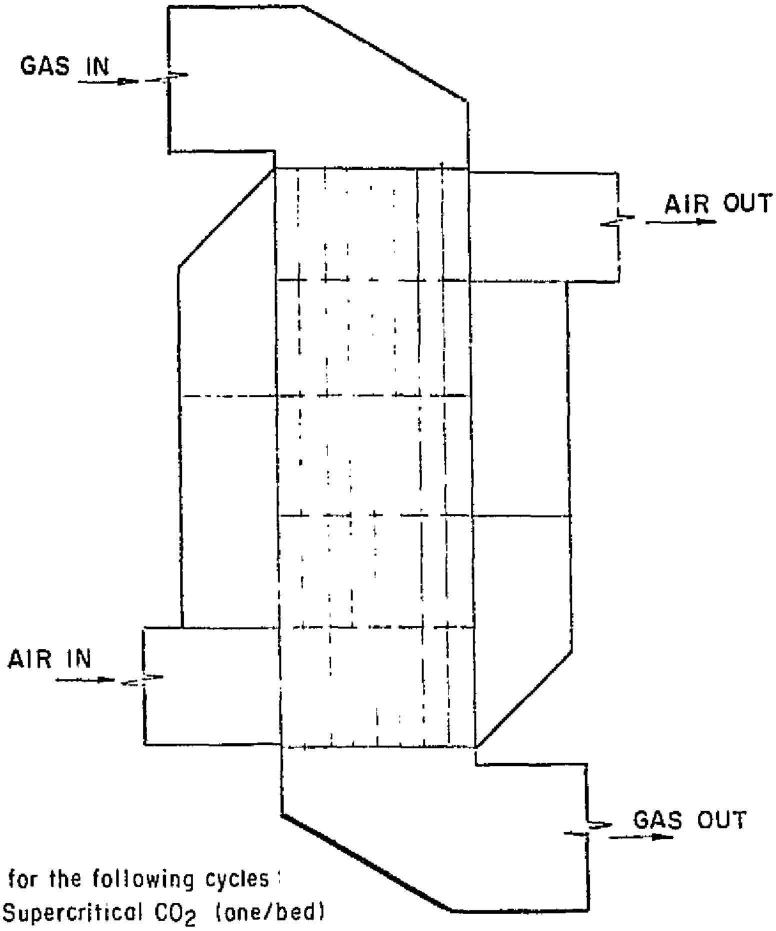
Tube vibration

Gas-side safety; sootblowing

Maintenance accessibility

Startup requirements

Sketches indicating the application of these air heaters to process cycles where required are indicated in Figures 6-31 through 6-34. Where high-pressure air must be heated by lower-pressure gases coming from a direct combustion process, the dirtier gas stream flowed downward into the tube bundles. This allows for a periodic cleaning of the inside of the tubes and is a preferred arrangement for this type of exchanger. Where gases are relatively clean they may flow on the shell side of an exchanger of this type, as indicated in Figure 6-33.



Used for the following cycles:
Supercritical CO_2 (one/bed)
Closed gas turbine (one/bed)
Liquid-metal topping (one/bed)

Figure 6-31. Low-Pressure/High-Temperature Air Preheater—Typical Gas Inlet Tubes

Process requirements for the inert-gas MHD cycle indicated that a special type of air heater/gas heater must be used. In this unit, recirculated gas was blended with preheated air in the air heater itself to allow for lower exit temperatures on the unit and maintenance of the required combustor outlet temperature for the process.

Figure 6-34 indicates an application of higher temperature air heaters to the open-cycle MHD case. This heat exchanger develops running liquid slag through the tube bundle which, it was hoped, would be compensated for by sootblowing as the seed solidified. The potassium sulfate seed in combination with coal ash will be highly corrosive in the running liquid condition, tending to attack iron and nickel constituents of the tubes used in construction. In future efforts this air heater should be redesigned to chill the gas stream and solidify these seed and ash

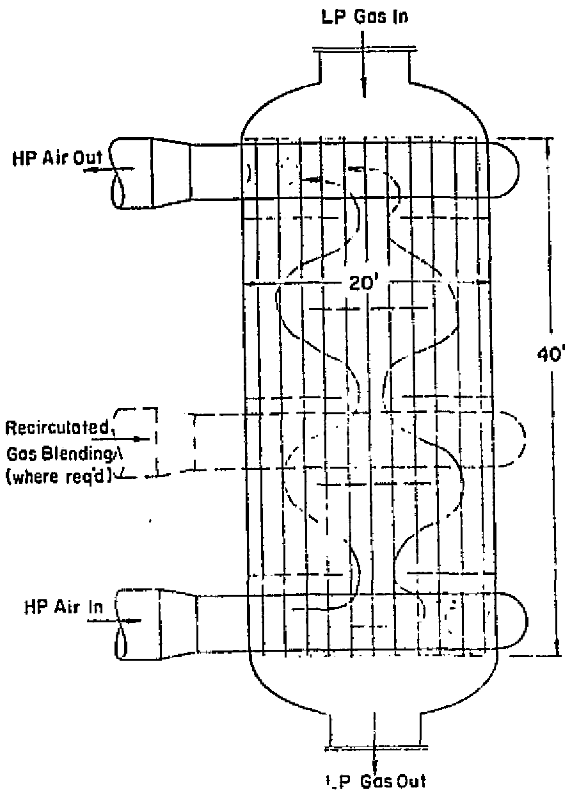


Figure 6-32. High-Pressure/High-Temperature–Typical Gas Inlet Tubes

constituents before they pass over convection surfaces at high temperatures.

AUXILIARY EQUIPMENT

The auxiliary equipment required to support the different combustion and heat exchange systems studied is summarized in Table 6-15. The two furnace systems requiring the most extensive auxiliary equipment system are the atmospheric fluidized bed and the pressurized fluidized bed. These are described here in more detail.

Solids Handling and Hot Gas Cleanup Process Train for AFB Furnaces

In the AFB process, coal and limestone are injected into a fluidized bed on a continuous basis to provide steady combustion conditions and environmentally acceptable SO_2 emissions. SO_2 absorption and bed operating levels are maintained by continuously

Table 6-15

AUXILIARY EQUIPMENT

<u>Conventional Furnaces and Combustion Chambers</u>

Regenerative airheaters (low-temperatures ≤ 850 F [727.8 K] gas inlet)

Steam coil airheaters (high-sulfur coal and semiclean liquid firing)

Fans

- Forced draft
- Primary air (coal firing)
- Induced draft
- Gas recirculation (inert-gas MHD coal combustors)

Burners, ignitors, burner front piping, and valves

Combustion control and safety interlock systems (including control, safety, miscellaneous valves and sensors, regulator and devices)

Pulverizers and feeders (coal firing)

Oil pumping and heating sets (semiclean liquid firing)

Sootblowers (coal, semiclean liquid firing)

Wall blowers and deslaggers (coal firing)

<u>Open-Cycle MHD Combustors and Furnaces</u>

- Combustors
  - Pulverizers and feeders
  - Pressurized coal injection system
  - Booster compressor for coal injection
  - Coal processing equipment
- Heat recovery furnaces
  - Combustion control and safety interlock systems
  - Steam temperature, pressure, and flow control systems
  - Soot and wall blowers

<u>Pressurized Furnaces</u>

- Combustion control and safety interlock systems
- Fluid temperature, pressure, and flow control systems

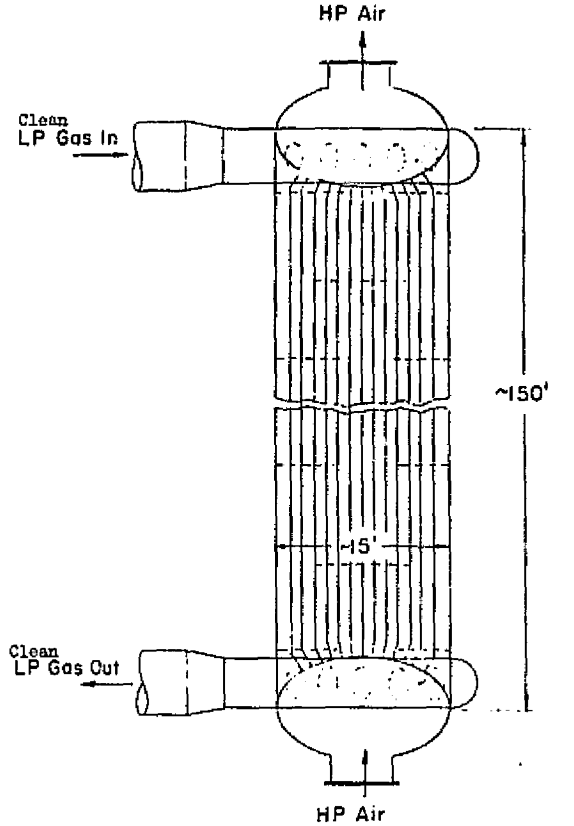
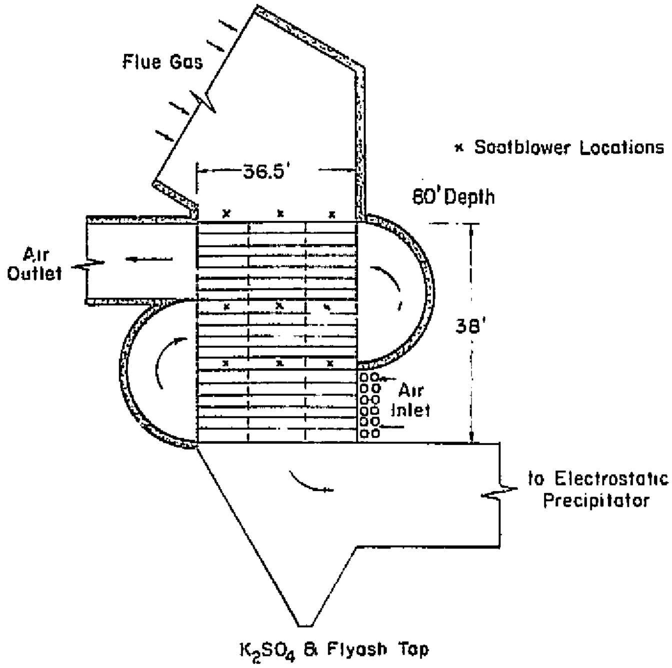


Figure 6-33. High-Pressure/High-Temperature Air Preheater–Typical Air Inlet Tubes

withdrawing spent sorbent materials from the AFB, whereas elutriated bed materials are captured in the hot gas cleanup equipment.

Limestone Handling System. Limestone will be delivered to the plant site by rail, unloaded, transferred to, stored in, and reclaimed from a storage pile for drying and crushing prior to AFB injection (a balance-of-plant responsibility). The limestone processing is similar to the coal handling process.

Coal and Blend Handling and Processing System. Coal will be withdrawn on a continuous basis from a day storage silo and dried to provide proper screening and blend transport properties. Once the AFB's solids removal system has reached steady-state operation, the hot exhaust gases from the spent solids cooler is employed for the coal and limestone driers. The 2 inches (5.08 cm) × 0 dried coal will then be crushed to 1/8 inch (0.318 cm) × 0 by a Gundlach Cage-Paktor crusher with prescreening and recirculation used to minimize the amount of fines. It has been estimated that with the above arrangement the amount of coal less



Note: Some K_2SO_4 will Condense in the High Temperature Air Heater and in the flue to the Low Temperature Air Heater.

Figure 6-34. Low-Temperature Air Heater with Open-Cycle MHD

than 500 microns or 28 mesh in size can be reduced to 20 percent or less.

After being crushed to size, the coal and limestone flows are mixed in a blender and transported by belt conveyors and bucket elevators to 4-hour blend holding hoppers positioned above each steam generator module. The blended material is withdrawn from the holding hoppers by vibrating feeders, which discharge to vibrating conveyor/tables, positioned on each side of a module and containing multiple outlet pipes that feed the module's fluidized beds. Primary air is injected into the vibrating tables to pneumatically assist the blend bed injection process. In addition to the above described blend-feed arrangement a separate limestone holding hopper and feeder is provided for the carbon burnup cell (CBC). This arrangement permits the latter to be operated without coal injection in the event of high carbon carryover from the other beds.

Spent Sorbent Handling System. Spent sorbent material will be withdrawn from the AFB process by the use of a butterfly valve and a high-temperature vibrating feeder. After being withdrawn from the module the spent material is transported via a high-temperature vibrating conveyor to the spent-solids cooler. In the latter unit the hot bed material is cooled to approximately

300 F (422.2 K) in an air fluidized bed. The hot exhaust gases from the solids cooler will be ducted to the dryer units to dry the coal and limestone.

Hot Gas Cleanup Equipment. Solid material elutriated from the AFB beds will be removed from the gas stream in two stages. The first stage of gas cleanup equipment will consist of multitube cyclones, whereas electrostatic precipitators will be provided as the second and final stage cleanup device. Solid material captured in the cyclones will be returned to the AFB process by injecting such into the 2000 F (1366.7 K) CBC for combustion of previously unburned carbon.

The gases exiting from the CBC will pass through multitube cyclones before mixing with the module's main gas stream, upstream of the electrostatic precipitators. Solid material captured in the CBC cyclones will be transported to the solids cooler and removed from the process after cooling. High-temperature, tubular air preheaters will be provided as required to limit cyclone/precipitator gas inlet temperatures to 730 F (661 K). Low-temperature recuperative air preheaters will be provided as required to minimize stack temperature losses.

Forced draft, induced draft, and booster fans are included to supply all necessary primary and secondary air streams.

Solids Handling and Hot Gas Cleanup Process for PFB Furnaces

In the PFB process, coal and dolomite are injected into a fluidized bed on a continuous basis to provide steady combustion conditions and environmentally acceptable SO_2 emissions. The SO_2 absorption and bed operating levels are maintained by continuously withdrawing spent sorbent materials from the PFB; elutriated bed materials are captured in the hot gas cleanup equipment.

Dolomite Handling System. Dolomite will be delivered to the plant site by rail; it will then be unloaded, transferred to, stored in, and reclaimed from a storage pile for drying and crushing prior to PFB injection (a balance-of-plant responsibility). Dolomite processing is similar to the coal handling process, which is next described in detail.

Coal Handling and Processing System. Coal will be withdrawn on a continuous basis from a day storage silo and dried to a surface moisture of less than 1 percent, as required by the Petrocarb injection system. Once the PFB's solids removal system has reached steady-state operation, the hot exhaust gases from the spent-solids cooler will be used for drying. The 2 inch (5.08 cm) × 0 dried coal will then be crushed to 1/4 inch (0.635 cm) × 0 by a Gundlach Cage-Paktor crusher, with prescreening and recirculation used to minimize the amount of fines. Gundlach has estimated that with the above arrangement the amount of coal less than 500 microns or 28 mesh in size can be reduced to 20 percent or less.

After being crushed to size, the coal will be transported to a holding hopper located above the system. Although coal is withdrawn from this hopper in batch amounts, the coal drying and crushing is done on a continuous basis. The coal storage in the intermediate hopper will vary from a minimum of 2 hours to a maximum of 4 hours. Level indicators in this hopper will signal the coal processing equipment to speed up or slow down to match withdrawal rates, while still maintaining at least 2 hours of storage at all times. Before entering the lock hoppers, the coal will be put through a final stage of cleanup consisting of screening and magnetic scalping.

Spent Sorbent Handling System. Spent PFB sorbent materials will be withdrawn from the process and cooled to approximately 300 F (422.2 K) by the use of surge hoppers, lock hoppers, high-temperature vibrating feeders and conveyors, and a fluidized bed cooler. The spent material will drain through a refractory lined pipeline provided at the bottom of the beds and collect in a pressurized lock hopper. A butterfly valve will be provided upstream of the surge hopper to control the amount of solids draining out of the PFB by gravity. When a predetermined solids level is reached in the lock hopper, the butterfly valve will close momentarily, to stop the flow of solids as the upstream lock hopper valve is closed to isolate such from process conditions.

Having completed the isolation, the butterfly valve will be opened to its previous position to resume solids draining but with such now collecting in the surge hopper. The lock hopper will be depressurized, the outlet valve opened, and collected material fed onto a high-temperature vibrating conveyor by a vibrating feeder. The lock hopper will fill during the first 30 minutes of every hour and empty during the last 30 minutes of the hour. When the lock hopper's fill cycle starts up again, the material collected in the surge hopper during the latter 30 minute interval will discharge into the lock hopper. Depressurized solids material will be transported to an air-fluidized solids cooler, where the unburned coal (PFB may contain up to 1 percent coal by weight at any point in time) will be combusted and the material cooled to 300 F (422.2 K). To avoid loss of the heat content of the solids, the exhaust gas from the solids cooler will be ducted to the coal dryer.

Although this could be considered a conventional approach for depressurizing hot solids, it does require the use of leak-tight valving at the lock hoppers. The valving selected and costed for this preliminary design possesses a water-cooled stainless steel body and utilizes gas blast cleaning of the seats and a seat wiping movement of the plug.

Hot Gas Cleanup Equipment. The hot gas cleanup equipment selected and costed consists of Aerodyne two-stage cyclone collectors followed by Ducon granular bed filters. The PFB and CBC exhaust gases pass through the cyclone separators and into the granular bed filters through refractory lined carbon steel piping. Solid material captured in the PFB combustor cyclones will drain through trickle valves into a collecting hopper that feeds by gravity to the CBC injector vessel. A butterfly valve provided in the interconnecting piping will control the hopper draining and, when closed, will permit the CBC injector vessel to be pressurized and fluidized for fines injection into the carbon burnup cell.

Solid material elutriated from the CBC and escaping from the PFB combustor cyclones is expected to contain a negligible amount of unburned carbon. As a result, fines captured in the CBC cyclone and granular bed filters will not be injected into the CBC, but instead will be depressurized in surge hoppers and lock hoppers. Since these particles have already been elutriated once, it is expected that an attempt to cool this material and recover its heat by introducing it into the spent-sorbent-solids cooler would result in re-elutriation and pneumatic transport to the coal dryer. Solids elutriated from the coal dryer will be captured in the dryer's cyclone system and returned to the coal stream, going on to the Petrocarb unit and into the PFB combustor. Rather than recirculate the CBC cyclone and granular bed filter fines, these materials will be pneumatically transported to a water slurry vessel for slurrying and transport to the spent-ash pond.

FURNACE ANALYSIS CONSIDERATIONS

An analysis procedure was established for each of the furnace types previously described. Their purpose was to develop the furnace performance, size and weight, auxiliary energy requirements, and capital costs. Table 6-16 is an outline of the analytical procedures that were followed.

FURNACE SIZING, HEAT AND MASS BALANCES

The module sizes of the various types of furnaces investigated depended on various considerations. The arrangement and construction technique which is employed for each energy conversion system is identified in a later subsection, where each energy conversion system is individually discussed.

The sizes of the atmospheric fluidized bed towers were set by the maximum shippable tube bundle module. This was not more than about 12 to 13 feet (3.66 to 3.97 meters) wide by 25 to 35 feet (7.63 to 10.68 meters) deep. The height of a tower unit (consisting of multiple beds) was set by the number of beds, the depth of the individual bed, the freeboard allowance, and the height of the convection sections installed in each cell of the

Table 6-16

SEQUENTIAL ANALYTICAL PROCEDURE OUTLINE FOR COMBUSTORS AND PRIMARY HEAT EXCHANGERS

I. Combustion Analysis (per unit of fuel basis)

A. Select excess air for good combustion.

Atmospheric fluidized bed — Coal	20%
Pressurized fluidized bed — Coal	20%
Pressurized furnace — LBtu	15%
Pressurized furnace — HBtu	10%
Conventional furnace — Coal	20%
Conventional furnace — SRC	20%
Combustion chamber — Coal	20%
Combustion chamber — IBtu	15%

Perform combustion analyses

B. Determine mass balance.

Air required, lb/lb of fuel

Gas produced, lb/lb of fuel

Solids out, lb/lb of fuel

C. Set unit exit temperature for gas and solids streams.

D. Calculate total sensible heat input.

Net heat released, Btu/lb of fuel

Heat content of air above 80 F, Btu/lb of fuel

E. Calculate: Furnace heat absorption or output to working fluids, Btu/lb of fuel ≐ D. - C.

F. Establish emissions.

SO_X, $lb/10^6$ Btu

NO_X, $lb/10^6$ Btu

CO, $lb/10^6$ Btu

Dust, lb/lb fuel

Percent solids as bottom ash, lb/lb of fuel

G. Calculate: Furnace efficiency.

$Q_{out}/Q_{in} \doteq$ E./D.

Table 6-16 (Cont'd)

H. Select number of firing stages to control NO_X or heat absorption rates or module size.

II. Determination of Flow Rates

A. Set: Output required to prime cycle and/or output per standard module

B. Calculate: Fuel input required, lb/hr

C. Calculate: Flow rates of air, gas, and solids

III. Furnace Sizing

A. Select furnace gas velocities for good combustion, absorption rates, NO_X control, fluidization.

Atmospheric fluidized bed	5 to 10 ft/s
Pressurized fluidized bed	2 to 7 ft/s
Pressurized furnace	30 to 40 ft/s
Conventional furnace	30 to 50 ft/s
Combustion chamber	30 to 35 ft/s

B. Determine furnace dimensions for gas passage at reasonable velocities.

C. Select containment shell type, casing type.

Heat transfer monowall

Refractory lined casing

Refractory lined casing with air gap

D. Calculate weights and select material compositions for casings per foot of height.

IV. Heat Transfer Surface Requirements

A. Define heat transfer services and determine duty of each.

B. Select location of each service in gas passage.

C. Calculate hot fluid temperatures, determine log mean temperature differentials, and set overall heat transfer coefficients.

D. Calculate surface area requirements.

Table 6-16 (Cont'd)

E. Determine volume required for heat transfer surface.

F. Set dimensions of heat transfer surface and containment shell or walls, allowing free space for access, staged firing, sootblowing, etc.

V. Materials Selection and Weights

A. Determine or judge peak absorption rates and estimate inside fluid film and metal conductances for each heat transfer circuit.

B. Determine or judge mean metal temperatures.

C. Select material to give practical tube wall thickness and corrosion resistance for design pressures and temperatures.

D. Determine heat transfer surface weights and add 5 to 30 percent for tube stubs, headers, piping, etc., depending upon complexity.

VI. Cost Estimating

A. Combustors and heat exchangers:

+ surface weight × cost rate

+ casing, shell, etc. weight × cost rate

+ allowance for flues, ducts, valves, buckstays, hanger, insulation, lagging, etc.

+ allowance for burners, grids, needles

Heat exchanger costs

B. Auxiliary equipment:

Vendor costs for basic furnace systems

Proration of costs for cycle conditions

boiler. Additional allowance must be made for inlet and gas outlet areas in each cell.

The size of the pressurized fluidized bed depended upon a maximum shippable module that was 13.5 feet (4.12 meters) in outside diameter. Within this containment shell configuration, certain sizes of beds were permitted. A 9.5 × 9.5 foot (2.9 × 2.9 meter) square bed with waterwall-type enclosures was permitted for a vertical tower arrangement. With refractory lining a similar type construction was permitted. The height of the pressurized fluidized bed tower depended upon bed height, allowable freeboard height, space for convection tubes, and gas outlet and air inlet areas of the unit.

A barrel type of design was also considered for the pressurized fluidized bed. The actual bed area installed in each barrel is limited by maximum allowable gas outlet velocities in the outlet sector at the top of the bed. These units were limited to a bed height of 6 feet (1.83 meters) and a reduced freeboard section considered, with the lower fluidizing velocities used in these designs. For a given quantity of fuel fired and gas generated at these lower fluidizing velocities, this type of fluidized bed permits the installation of more heat transfer surface than the tower type with an equal number of beds. The tower type of design permitted maximum bed heights of 8 to 10 feet (2.44 to 3.05 meters), as compared with the 6 foot (1.83 meter) maximum in the barrel design.

The pressurized furnace sizing was based on the maximum shippable diameter, with deductions for internal refractory lining and/or air-gap cooling in combination with refractory linings. The height of the pressurized furnace was mainly determined by the space required for installed heat transfer surfaces, burner staging, and access areas.

The conventional furnace sizing depended on conventional practice, the allowable input per unit of plan area and furnace surfacing considerations governing height. With the conventional furnace type of construction, waterwall panels wider than 100 feet (30.5 meters) are to be avoided because of thermal expansion and the difficulty in staying and bracing so long a span on the boiler. The height of the conventional furnace is determined by the amount of heat transfer surface that can be installed and by surrounding waterwalls, partition walls, and division walls in the design, with allowance for burner zones within the furnace.

The combustion chambers considered for the inert-gas MHD cycle were of special design concept. In these furnaces fuel is fired around the circumference, to create centrifugal forces that will throw slag toward the refractory surfaces of the unit. It was assumed that a running liquid slag layer on a partially frozen slag layer would be maintained within this type of unit. The process conditions required a 90 percent slag rejection in the combustion chamber before passage of gases into the high-temperature working fluid heaters. After this tangential injection in the main furnace, gases recombine and drift upward at lower velocities, reducing the entrainment of slag, fog, and liquid droplets to the upper zone of the furnace. The height of this unit is set by the burner space required in the unit.

A detailed estimate of NO_X emissions from the various furnaces was not made. It is felt, however, that such emissions are within the capacity of the combustion systems described. The basic weight, cost, and performance of the combustion systems given are not expected to be affected by redesign which might be required for NO_X reduction from the evaluated design concepts.

REVIEW OF FUELS AND COMBUSTION

A combustion analysis was performed for all the fuels and appropriate excess air rates under consideration. These analyses were used to estimate enthalpies of the combustion gases.

To estimate the quantity of gases for AFB and PFB processes a conventional coal combustion analysis was conducted which assumes complete combustion of all species. The gas weight and molecular weights were adjusted to compensate for the products of limestone-sulfur trioxide reaction. For the AFB and PFB cases, the moisture removed in the coal and limestone drying processes was considered to be fired with the fuel in the boiler. This is an approximation, the net effect being to raise the gas quantity and percentage of moisture in the exhaust gases. Table 6-17 lists the fuels that were analyzed. Table 6-18 summarizes the results of the combustion analyses that were performed.

Table 6-17

FUELS FIRED

Coals

- Illinois No. 6
- Montana Subbituminous
- North Dakota Lignite
- Semiclean fuel (solvent refined coal)

Low-Btu Gases (from all three coals)

Intermediate-Btu Gases (from all three coals)

High-Btu Gases (from Illinois No. 6)

Fuel Cell Waste Gas

Table 6-18

SUMMARY COMBUSTION RESULTS

Coals	Illinois No. 6			Montana Subbituminous			North Dakota Lignite			Semiclean Fuel	
HHV, Btu/lb fuel	10788			8944			6890			15602	
Furnaces	CF	AFB	PFB	CF	AFB	PFB	CF	AFB	PFB	CF	CH
Lb combn air/lb fuel	9.9515	9.9521	9.9518	8.1755	8.1755	8.1756	6.3258	6.3262	6.3258+	14.3611	
Excess air, %	20	20+	20+	20	20	20	20	20	20	20	
Lb gas/lb fuel	10.8555	10.8738	10.9729	9.1005	9.0993	9.1222	7.2638	7.2637	7.2832	15.3211	
W/W, % H_2O	6.89	6.1	6.03	7.14	7.1	7.14	9.61	9.6	9.61	6.81	
Lb ash/lb fuel	0.096	0.096	0.096	0.075	0.075	0.075	0.062	0.062	0.062	0.001	
Lb stone products/lb fuel	—	0.233	0.3301	—	0.0519	0.0701	—	0.0445	0.0610	—	
Total solids, lb/lb fuel	0.096	0.3290	0.4261	0.075	0.1269	0.1451	0.062	0.1065	0.1230	0.01	
Acceptor		Stone	Dolomite		Stone	Dolomite		Stone	Dolomite		
Lb/lb fuel		0.2513	0.4475		0.0512	0.5451		0.0451	0.6166		
Fuel Gases	Low-Btu Gas									High Temperature Fuel Cell	
	Illinois No. 6			Montana Subbituminous			North Dakota Lignite				
HHV, Btu/lb fuel		2340			2148			2047		1402	
Furnaces		PF			PF			PF			
Lb combn air/lb fuel		2.8132			2.3371			2.1978		2.1596	
Excess air, %		15			15			15		—	
Lb/gas/lb fuel		3.8132			3.3371			3.1978		3.1596	
W/W % H_2O		15.14			16.53			16.87		—	
Fuel Gases	Intermediate-Btu Gas									High-Btu Gas	
	Illinois No. 6			Montana Subbituminous			North Dakota Lignite				
HHV, Btu/lb fuel		6873			6534			6333		22651	22651
LB combn air/lb fuel		5.0171			4.7264			4.5202		17.8949	18.2263
Excess air, %		15			15			15		8	10
LB gas/lb fuel		6.0171			5.7264			5.5202		18.8949	19.2263
W/W, % H_2O		11.15			11.34			11.35		12.52	12.33

Table 6-19

HEAT TRANSFER PARAMETERS FOR PRIMARY AND
INTERMEDIATE AIR PREHEATER HEAT EXCHANGERS

	Uo(Btu/hr-ft^2-°F)		T_1-T_2(°F) Gas Temperature
	Range	Utilized	Ranges
AFB			
Bed	35 - 45	40	1550
Convection	12 - 15	13	1550 - 730
PFB			
Bed	45 - 55	50	1650
Convection	15 - 20	17	1650 - 1600
PF			
Convection	50 - 60	55	2800 - 1800
			2800 - 1150
CF			
Radiation	15 - 25	20	4000 - 2000
Convection	12 - 15	13	2000 - 700
HTAH			
LP Convection	4 - 6	5	850 - 250
HP Convection	6 - 9	7	1700 - 730
VHP Convection	12 - 15	13	2200 - 700

Cycle	Service	Uo Range	Utilized	T_1-T_2(°F) Ranges
IHX				
LM-MHD	He - Cooler	55 - 65	60	1300 - 200
K TOP	K - Cond.	210 - 230	220	1100 - 1000
IG-MHD	Ar - Cooler	20 - 30	25	1800 - 260
WHB				
IG-MHD	Gas Cooler	15 - 25	20	1930 - 540
OC-MHD	Economizer	12 - 15	13	700 - 300

NOTE:

HTAH = High-temperature air preheater
IHX = Intermediate heat exchanger
LP = Low pressure
HP = High pressure
VHP = Very high pressure

Table 6-20

SUPER ALLOYS AND ALLOYS

	Rough Chemicals			Density (lb/in^3)
	Chromium	Nickel	Other	
Carbon steel + low alloy	0-5	—	—	0.286
Austenitic stainless steel	18	8	—	0.290
Hastelloy X	22	45	19 Fe, 2 Co, 9 Mo	0.297
Mo-Re 2	34	48	16 W, 0.75 Si 0.3 Mn, 0.2 C	0.330
Inconel 601	23	61	1.5 Al, 1 Cu	0.291

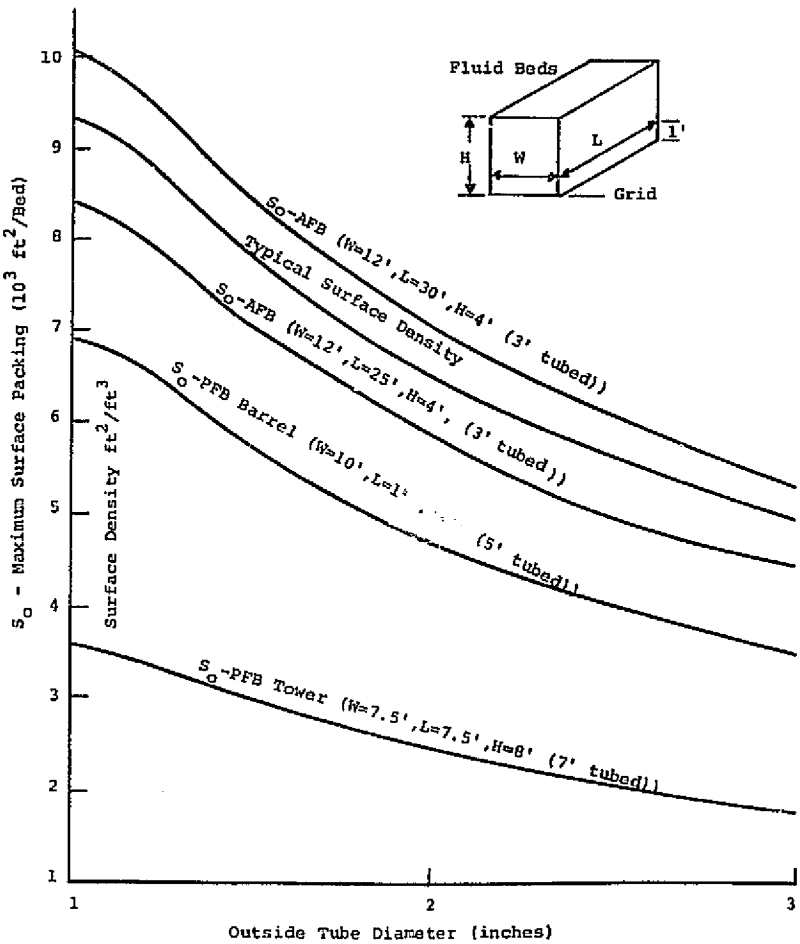


Figure 6-35. Surfacing Arrangements for Fluidized Bed Cases

HEAT TRANSFER AND SURFACING

A detailed determination of heat transfer coefficients for each energy conversion system was not made. Judgments were made as to the magnitude of heat transfer coefficients in individual cases. As an example, the overall average coefficient throughout the convection pass on a large utility boiler at atmospheric pressure will be approximately 12 to 15. These judgment values were employed for heat exchanger analysis. Boiling heat transfer coefficients were not the controlling thermal resistance and were neglected for most determinations. Likewise, where gases heat liquids the inside film resistance is not controlling and was neglected. The metal selections were not optimized and practical judgments based on experience were made.

Table 6-19 indicates the overall coefficients used for the various energy conversion systems in this study. Shortcuts taken in estimating temperature differences or log-mean temperature differences early in the study resulted in eventual changes when time was available for reviewing the method of surfacing and temperature difference calculation.

Where parametric variations affected only capacity, heat transfer element weights and costs were prorated on a capacity ratio. Where parametric variations affected only allowable pressure drops through the banks, prorations of heat transfer surface were made for changes in tube diameters, and resulting weights were required to stay within new pressure drop limits.

Curves are presented in Figure 6-35 indicating surface packings and surface densities attainable in fluidized bed arrangements. Where fluidized bed designs for heat transfer were limited, these curves were employed to establish the number of beds and bed volume required to accomplish the required heat transfer duty for the cycles.

METAL SELECTION AND PRESSURE PART WEIGHT

A brief procedure for estimating weights of heat transfer surfaces and pressure parts includes the following steps:

1. Determine average absorption rates, considering overall heat transfer coefficients and linear variation of fluid temperatures, at hot end, middle, and end of surface.

   Increase absorption rates by a factor to determine spot or local rates for metals selection:

Furnace	Typical Factors
AFB Bed	1.4 - 1.6
AFB Convection	1.4 - 1.6

Furnace	Typical Factors
PFB Bed	1.4 - 1.6
PFB Convection	1.4 - 1.6
PF Convection	1.4 - 1.6
CF Radiation	1.8 - 2.2
CF Convection	1.4 - 1.6

2. Estimate inside fluid film and metal conductances.

3. Determine mean metal temperatures.

4. Select practical metal wall thicknesses for required design pressures and temperatures, and pressure drop limits.

5. Determine surface weights and material compositions.

6. Determine pressure part weights, adding 5 to 30 percent to surface weights for manifolding, headers, piping, etc., reflecting circuit complexity.

The composition and densities of the materials utilized for these parts is given in Table 6-20.

The curves shown in Figure 6-36 indicate the allowable design stresses considered for the materials of construction in this study. Figures 6-37 through 6-39 show the curves that were used in estimating required ratios of average wall thickness to diameter as a function of design temperature and pressure. Some representative minimum wall limits are also indicated on these diagrams. The maximum practical ratio of average wall thickness to diameter considered in this study was about 0.375. The curve presented in Figure 6-40 indicates the weight, in tons per million square feet, of outside heat transfer surface from the wall-thickness selection procedure and the internal flow area for 1000 tubes of the sizes indicated and average wall thickness required (see Table 6-21). No analysis of startup requirements and load changing requirements were performed in this study. In the heat exchangers studied, no more than one metals change was considered within one tube bank.

Table 6-21

HEAT EXCHANGER SURFACE WEIGHT AND FLOW AREA

Surface Weight		Internal Flow Area
$\gamma = 20592\ \alpha(1-\alpha)d_o^2C\ \frac{\text{tons}}{\text{million ft}^2}$	$C = 1.000$ CS	$A_i = \frac{\pi(1-2\alpha)^2 d_o^2}{4(144)} \times 1000\ \text{ft}^2/1000\ \text{tubes}$
$\gamma = \gamma' C$	$= 1.014$ SS	
	$= 1.063$ I601	
$\alpha = \frac{AW}{d_o}$	$= 1.039$ HX	
	$= 1.154$ MO-RE2	

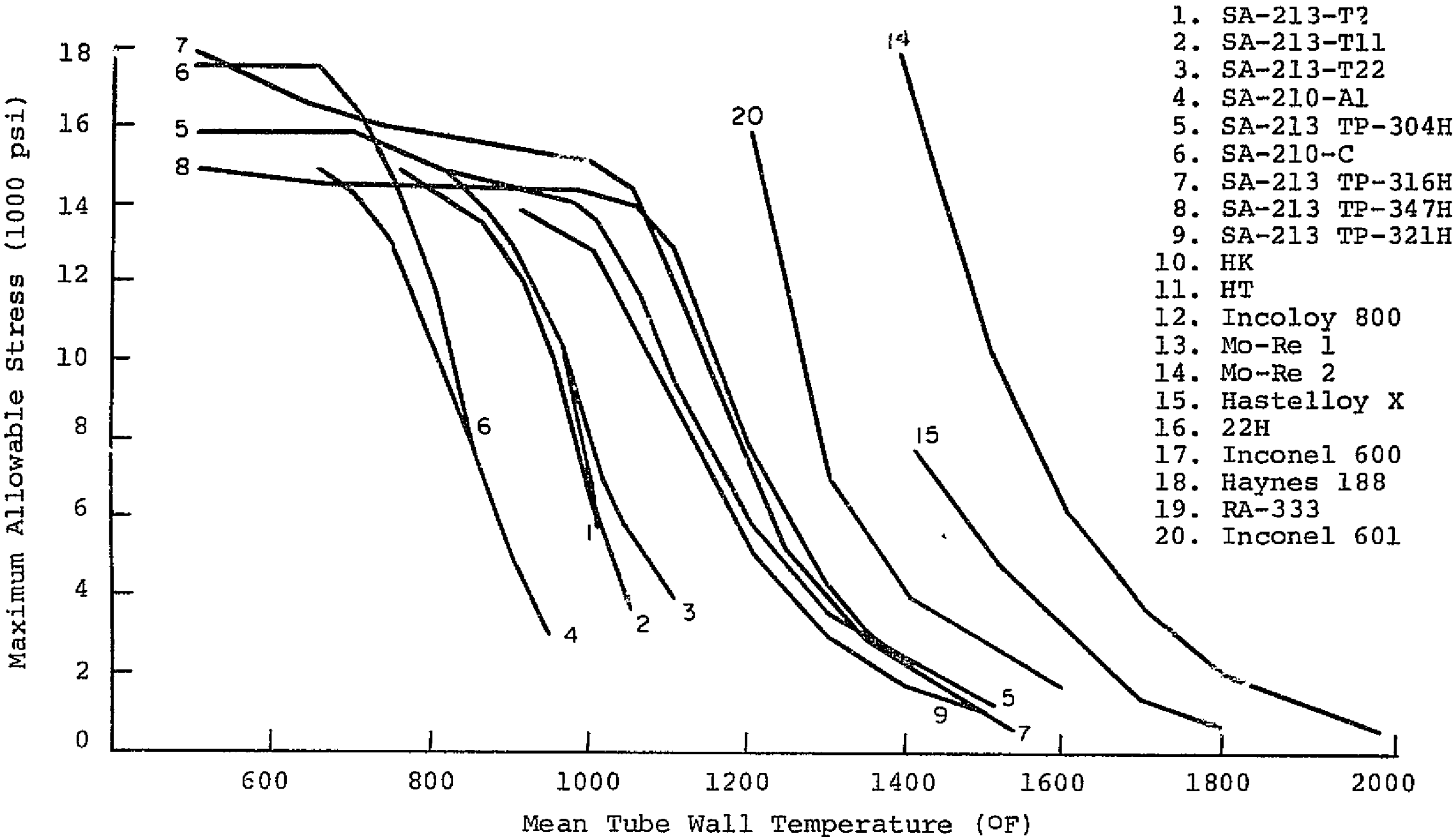


Figure 6-36. Primary Combustor Materials and Allowable Design Stress

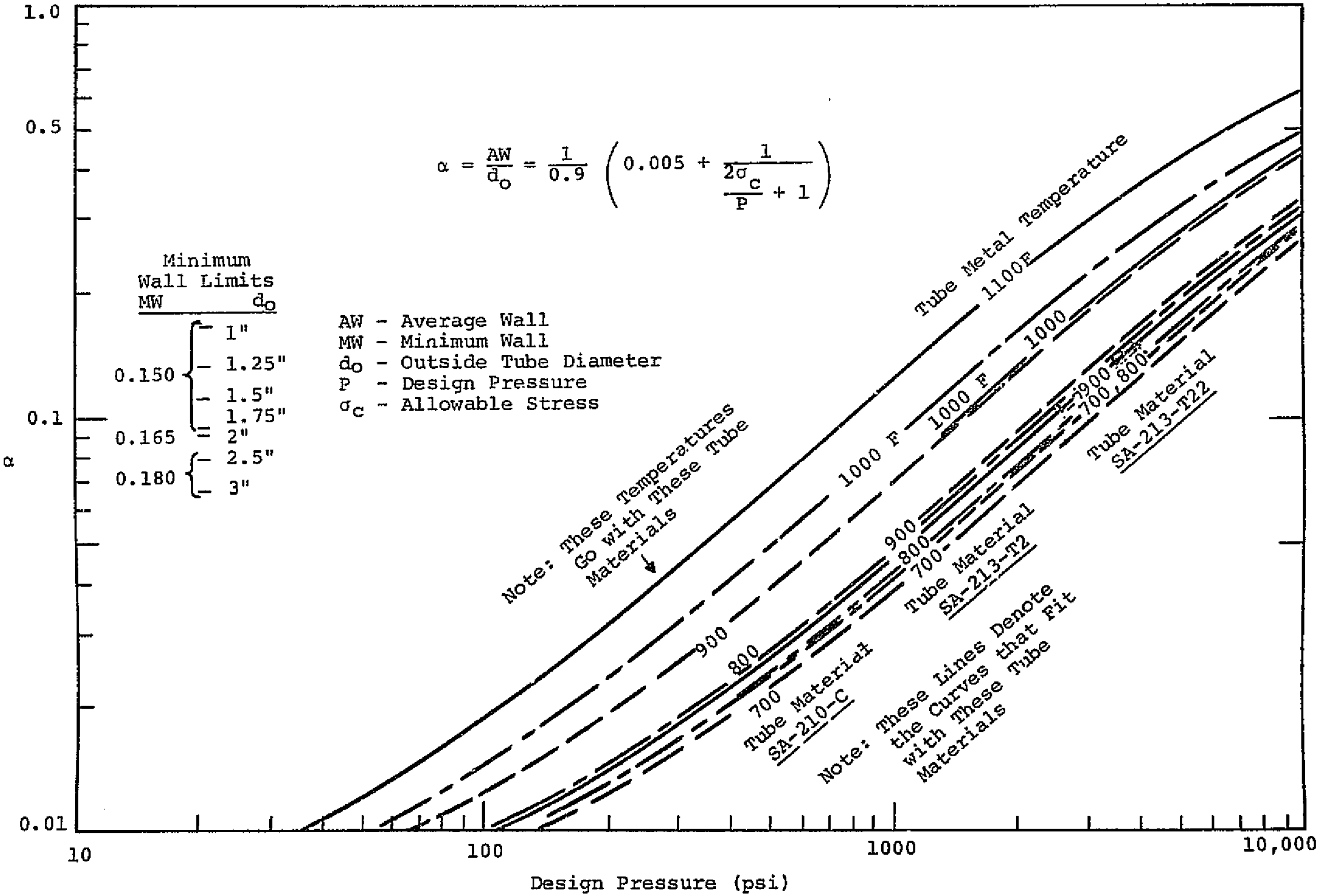


Figure 6-37. Design Stress Versus Internal Pressure and Metal Temperatures

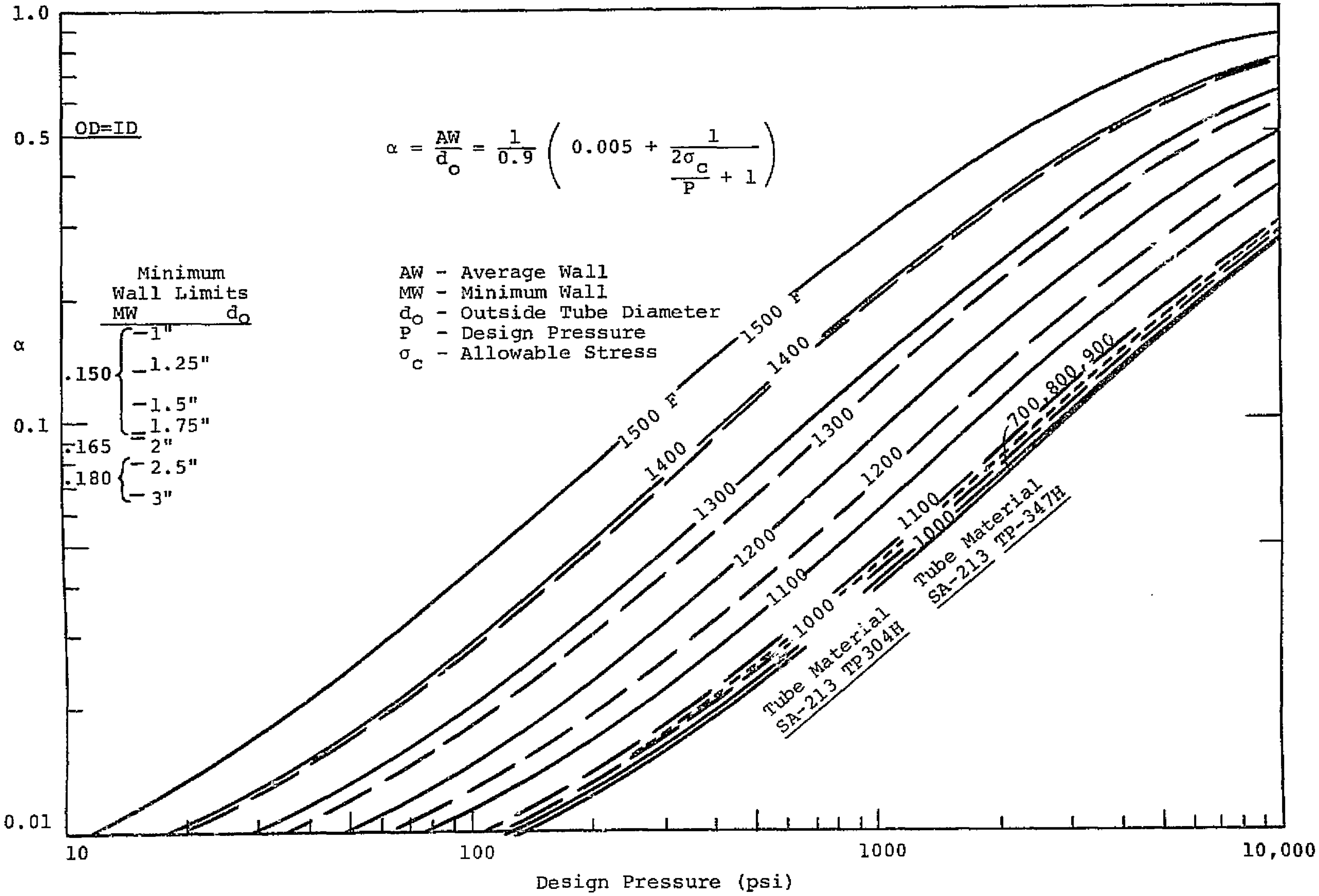


Figure 6-38. Design Stress Versus Internal Pressure and Metal Temperatures

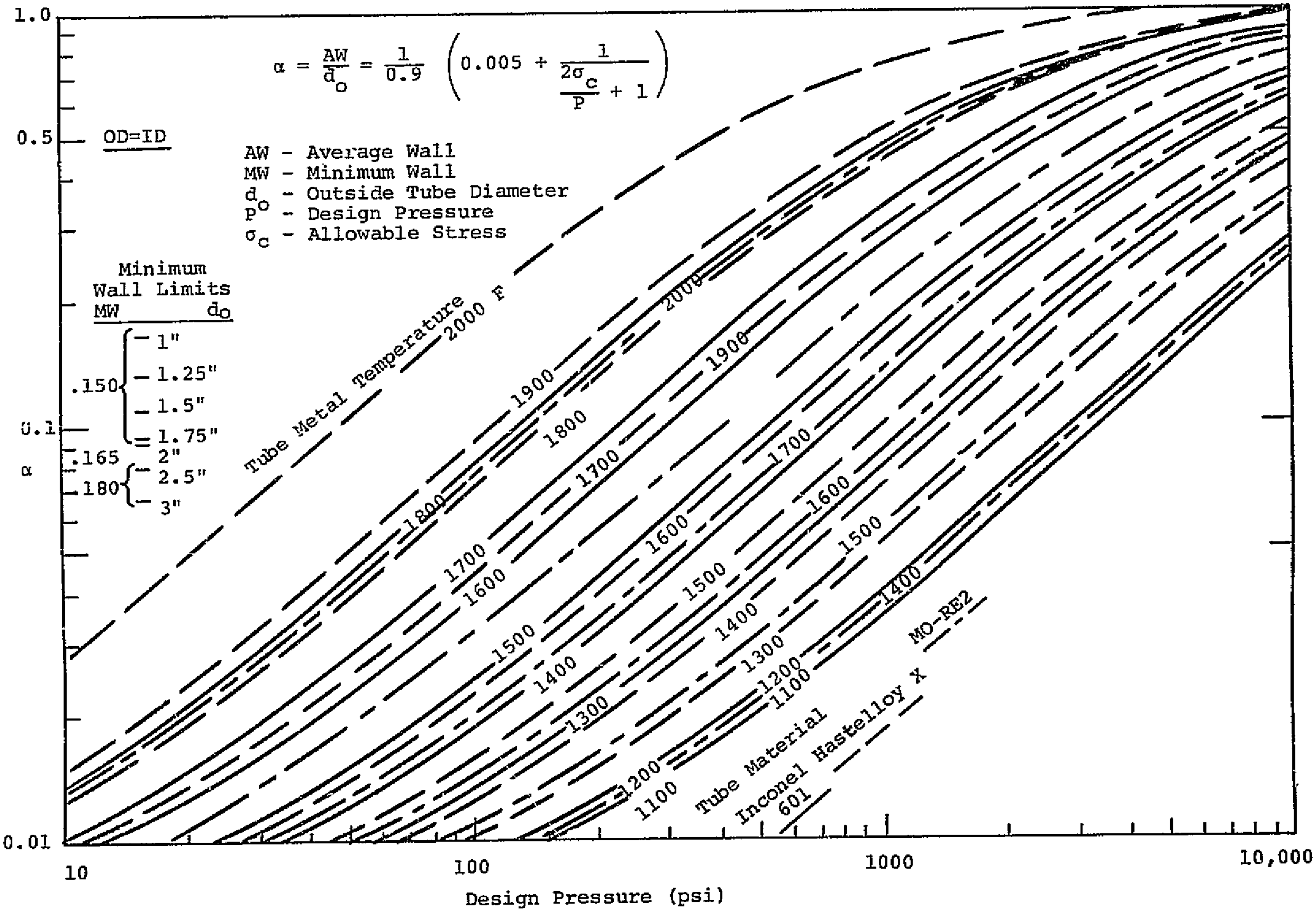


Figure 6-39. Design Stress Versus Internal Pressure and Metal Temperatures

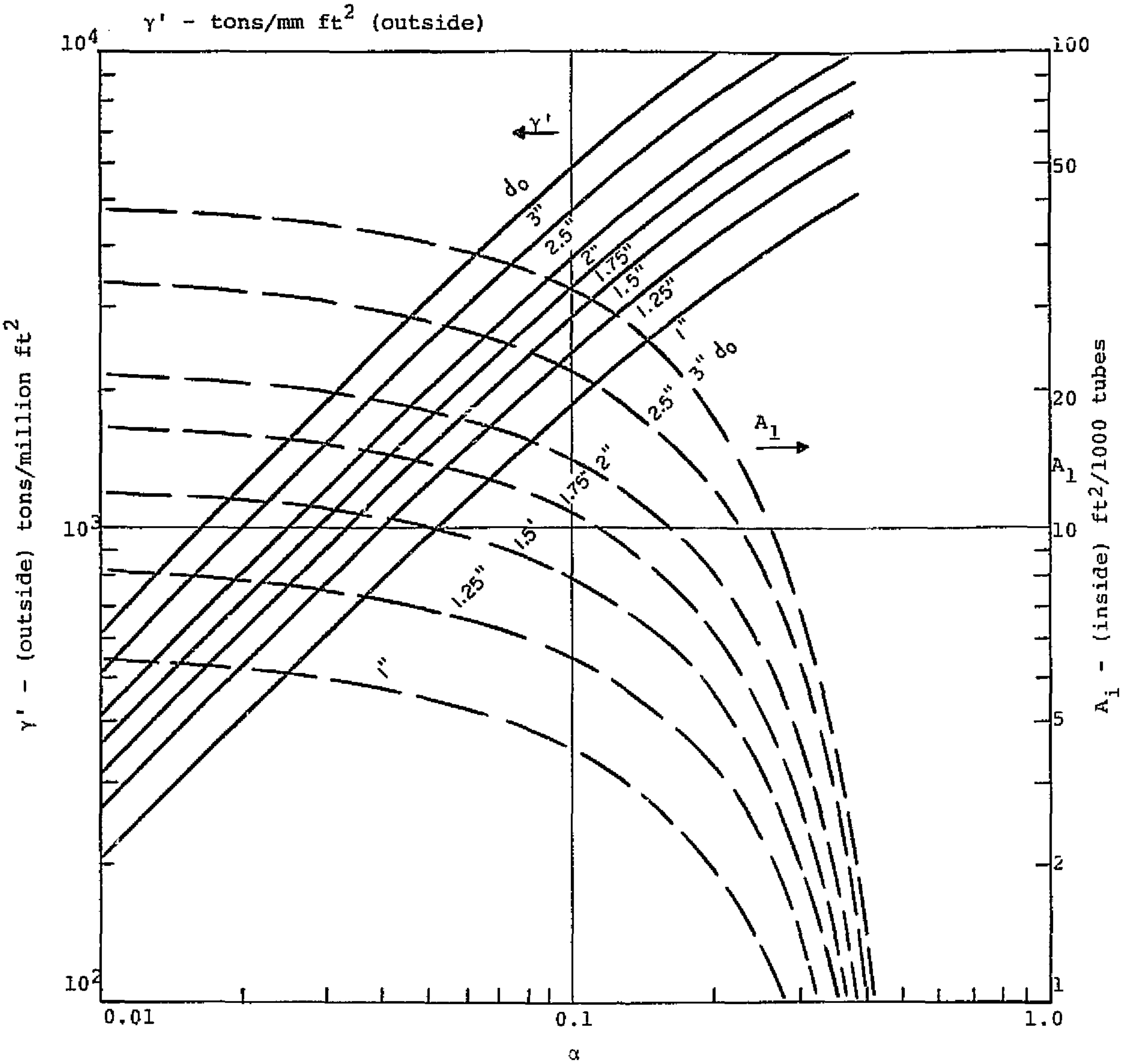


Figure 6-40. Weight of Heat Transfer Surface

COST ESTIMATING

The cost estimate procedure in this study was integrated with the design evaluations. An outline for this procedure is given in Figure 6-41. Table 6-22 summarizes the basic materials and welding costs. It was assumed that one weld would be required every 22 feet (6.8 meters). This permitted the generation of a cost value per foot of welded pipe. When cost estimates were obtained early in 1975, these were prorated back to mid-1974 conditions.

Table 6-22

PRIMARY COMBUSTOR MATERIAL COSTS

In 1974 Dollars

Material	$/Lb	$/Weld*	Remarks
SA-210 Al	0.40	14.20	Conventional
SA-213 T2	0.51	15.45	Conventional
SA-213 T12	0.54	15.45	Conventional
SA-213 T22	0.69	15.45	Conventional
SA-213 TP304H	3.12	14.90	Conventional
SA-213 TP347H	4.15	14.90	Conventional
Inconel 601	3.48	16.40	Heat to bend
Hastelloy-X	6.90	17.90	Can cold-bend
Mo-Re-2	7.50	25.00	Welding not recommended Subject to cracks
Refractory	0.65	—	Clips, ties, barriers included

*Cost for 2-1/2-inch diameter, 1/4-inch wall tube or equivalent.

The selling price for a heat exchange device was derived from the required heat transfer surface area and subsequent material weight. A process cost between 30 and 120 man-hours per ton of material, depending upon the difficulty and complexity of the manufacturing process required for the material, was added. A multiplication factor of 1.27 was utilized to cover product engineering and management, general overhead and administration, and profit.

Some typical examples of prime heat transfer surface costs are given in Table 6-23.

Cost Analysis of Auxiliary Systems

In setting up the solids processing trains for the atmospheric fluidized beds and pressurized fluidized bed systems the

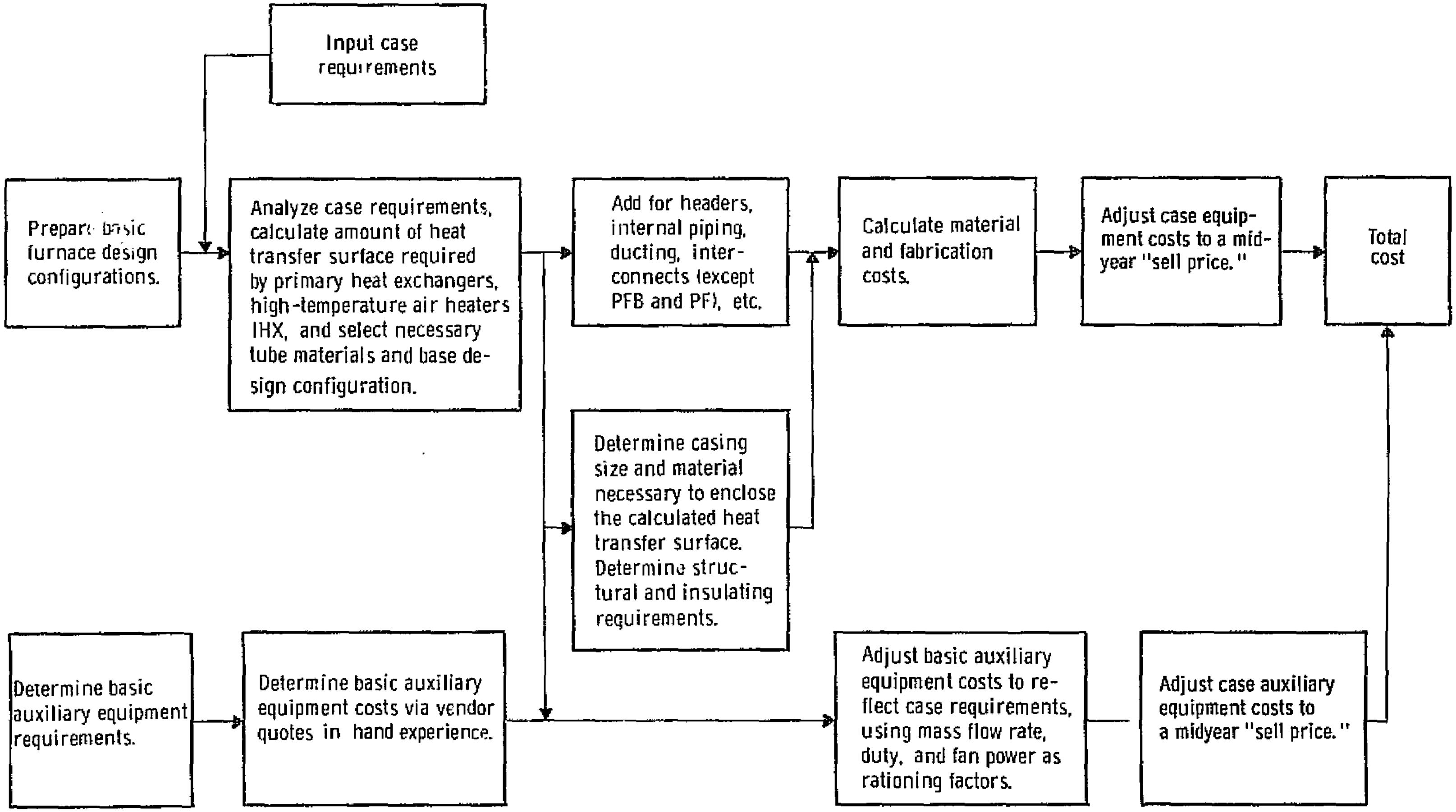


Figure 6-41. Cost Estimating Procedure for Heat Exchanger Systems

Table 6-23

TYPICAL PRIME TRANSFER SURFACE COSTS

Material	$/Ton
Carbon steel	2800
Low alloy T2	3200
Low alloy T22	4000
Stainless steel TP304H	9432
Stainless steel TP347H	12050
Inconel 601	12050
Hastelloy X	19050
Mo-Re 2	19050
High-temperature air heaters (tubular)	2781 - 3250
Low-temperature air heaters (recuperative)	1842 - 1972

approach was taken to build in surplus capacity in on-line equipment, surge hoppers, and conveying equipment, to provide for the worst coal and limestone and solids disposal conditions. Thus, when hoppers and bunkers were operating at full capacity, reasonable down times could be permitted for equipment repair and servicing. Where repair times were long and components subject to failure, full-capacity spare replacements were provided for such components as crushers and dryers. A completely full train of solid processing equipment was not installed.

AFB Furnace. Assuming coal and limestone flow rates of 158 and 40 TPH to a total of two modules, vendor pricing was obtained for all major components, whereas on smaller items such as air coolers and belt conveyors, where vendor pricing was already available for similar process conditions, the referenced pricing was scaled up or down to reflect new conditions in accordance with capacity factors recommended by Guthrie (ref. 1). For the most part the vendor prices obtained were of a budget type, as opposed to firm price quotations. The major equipment costs together with overall system costs are tabulated in Table 6-24.

To account for variations in process flow rates, the compiled equipment costs were varied in accordance with the following assumptions:

- Air fan costs were assumed to vary linearly with power consumption.

- Low-temperature air heater costs were assumed to vary linearly with heat transfer duty.

Table 6-24

MAJOR EQUIPMENT COSTS TOGETHER WITH
OVERALL SYSTEM COSTS
FOR AFB FURNACE

The major equipment cost items proved to be:

Limestone dryer system per train	$	296,700
Coal dryer system per train		679,000
Coal crusher per train		90,700
Spent-solids cooler per train		268,600
Blend bucket elevator per module		122,100
Blend distributor tables per module		393,200
Cyclone collectors per module		196,300
Electrostatic precipitators per module		977,000
Low-temperature air heater per module		718,000
Forced draft, induced draft and booster fans per module		696,100
Control system and valving per module		572,000

System equipment costs per two-module train:

Limestone processing system	$	414,200
Coal processing and feed system		2,316,100
Spent solids handling system		791,600
Hot gas cleanup and air supply system		5,658,700
Boiler control system and valving		1,144,000
Coal dryer system (spares)		339,500
Solids cooler system (spares)		67,200
	Total Cost	$10,761,300
		($5,365,700 per Module)

- Steam generator control system and valving costs were assumed insensitive to process conditions and flow rates.
- All other equipment costs were assumed to vary linearly with process flow rates.

PFB Furnace. Assuming coal and dolomite flow rates of 138 and 62 tons per hour (34.77 and 15.62 kg/s) to a total of two modules, the costs of the depicted equipment was compiled by the same techniques utilized for the AFB. The major equipment costs together with overall system costs are tabulated in Table 6-25.

Table 6-25

MAJOR EQUIPMENT COSTS TOGETHER WITH
OVERALL SYSTEM COSTS
FOR PFB FURNACE

The major equipment cost items proved to be:

Dolomite dryer system per train	$ 342,300
Petrocarb dolomite injection system per train	1,337,800
Coal dryer system per train	599,100
Coal crusher system per train	90,700
Petrocarb coal injection system per train	2,500,000
Spent-solids cooler per train	308,100
Aerodyne 2-in-1 cyclones per module	1,528,100
Ducon granular bed filters per module	4,004,900
Fines removal system per module	305,300
Surge and lock hoppers per module	236,900
Booster air compressor for Petrocarb per module	103,500
Control system and valving per module	381,000

System equipment costs per two-module train:

Dolomite processing system	1,856,000
Coal processing system	3,340,700
Spent solids handling system	1,989,600
Hot gas cleanup system	12,084,100
High-pressure air system	384,900
Boiler control system and valving	762,000
Coal dryer system (spares)	299,500
Solids cooler (spares)	77,000
Total Cost	$20,793,800
	($10,396,900 per Module)

To account for variations in process flow rates, the compiled equipment costs were varied in accordance with the following assumptions:

- Since the coal and dolomite processing system costs are dominated by the Petrocarb injectors, the system costs were assumed proportional to mass flow rate raised to the 0.6 power, the latter cost relationship being recommended by the vendor.

- Control system and valving costs were assumed insensitive to process condition and flow rates.
- All other equipment costs were assumed to vary linearly with process flow rate.

Conventional and Semiclean Liquid (SRC) Fueled Furnaces. The costs of the auxiliary equipment associated with the 800 MWe furnaces were obtained from Foster Wheeler Energy Corporation (FWEC) marketing curve averages for pulverized coal fired units. The costs used were as follows:

Air heater	$ 2,525,000
Forced draft and induced draft fans	1,525,000
Primary air fan	300,000
Combustion controls	643,800
Burner controls	406,300
Pulverizers and feeders	3,056,300
Valves	275,000
Soot and wall blowers	1,056,300
Preheater coils	106,300
Total	$ 9,894,000

For the case of SRC fuel, the above pulverizer/feeder cost was replaced by the cost of a No. 6 fuel oil pumping and heating set, budget priced at $456,500.

Pressurized Furnaces. The auxiliaries costed for the pressurized furnaces were burner controls, combustion controls, and valving. Using FWEC marketing curve averages for pulverized coal fired units, it was assumed that the above would cost approximately half of that required for pulverized units; a cost of $191,000 was used and such was assumed insensitive to flow rate.

Closed-Cycle, Inert-Gas MHD. The auxiliaries included with the closed-cycle inert-gas MHD process were the same as those listed for the conventional and SRC furnaces. For costing purposes, air fan prices were obtained by scaling up 800 MWe conventional furnace fan costs by the ratio of the cycle fan power required to that of the conventional furnace. Pulverizer and oil pumping and heating set costs were similarly varied, but with fuel flow rates as the ratio factor.

Open-Cycle MHD. The coal processing system proposed for the subject cycle was identical with that proposed for the PFB process, with the exception that the Gundlach crushers were replaced by FWEC ball-mill type pulverizers to provide the 70 percent minus 200 mesh coal that would be pneumatically transported to the holding hoppers positioned above the Petrocarb injection system. Budget estimates were obtained from Petrocarb, Inc. for a 160 ton/hour (40.31 kg/s) pulverized coal injector system for both 8 and 20 atmospheres (8.1×10^5 and 2.02×10^6 N/m^2) operating pressures. For other pressures, injector costs were determined by interpolation/extrapolation as recommended by the vendor. A 160 T/hr (40.31 kg/s) flow rate injector is not currently available but is envisioned as being a near term possibility as improved techniques are developed. In order to achieve the larger mass flow rates required by the open-cycle MHD process, additional injector units must be provided. As a result 160 T/hr (40.31 kg/s) coal processing and injector unit costs were scaled up, assuming direct proportionality between cost and mass flow rate. The subject cycle auxiliary costs also include allowances for steam control valving and sootblowers associated with the indirect-heated steam generators. The following costs were used:

FWEC ball mill pulverizers for 160 T/hr (40.31 kg/s)	$ 1,896,200
Petrocarb 160 T/hr (40.31 kg/s) pulverized coal injector for 8 atmospheres	2,282,200 *
Booster compressor for pulverized coal transport air	45,000
Cost of balance of plant coal processing equipment	1,294,300
	$ 5,517,700

*For 20 atmospheres (2.02×10^6 N/m^2) coal injectors, cost is $3,423,400.

SUMMARY—PRIMARY HEAT INPUT SYSTEMS

The primary heat input systems were evaluated for eight energy conversion systems. The furnaces were designed for the base cases and critical parametric case; extrapolations were employed for the other parametric points. The performance and cost data for these systems are given in Appendices A through C of this part of Volume III. Four basic furnace types were employed for the closed-cycle systems. The conventional furnace was utilized only for the advanced steam case.

CONVENTIONAL FURNACE

The conventional furnace has a temperature of 3500 to 730 F (2200 to 661.1 K). In this type of furnace approximately 55 percent of the heat transfer is accomplished before the gas temperature falls below 2100 F (1422.2 K), the approximate ash solidification temperature when firing Illinois No. 6 coal. In the gas temperature region of 2100 F (1422.2 K), an attempt is made to keep the outside metal temperatures well below 1100 F (866.7 K). After the gases are chilled below the ash solidification temperature an attempt is made to match the high-temperature services to the gas temperature.

In the advanced steam power cycle, however, heat balance limits on the boiler resulted in employment of high-temperature platens in the furnace. In these platens steam temperatures are raised from approximately 1000 to 1200 F (811 to 922.2 K), indicating that metal surfaces might fall within the coal-ash corrosion range and Hastelloy X surfacing was used for these platens.

ATMOSPHERIC FLUIDIZED BED

The atmospheric fluidized bed firing coal with limestone has a temperature potential of 1550 to 2000 F (1116.7 to 1366.7 K) in the carbon burnup cell. The hot precipitation of the alkaline dust produced in the process was accomplished at 730 F (661.1 K). Where the exit temperature from the bed is greater than 730 F (661.1 K), heat was recovered in high-temperature air heaters following the furnaces, to bring the exhaust gas temperature to 730 F (661.1 K). With a 730 F (661.1 K) end temperature, approximately 65 percent of the heat transfer is done in the fluidized beds and the balance in the convection space. When 1100 F (866.7 K) exit temperatures are indicated, approximately 77 percent of the heat transfer occurs in the beds and 23 percent in the convection space. When 1420 F (1044.4 K) exit temperatures are indicated, as with the liquid metal topping cycle, approximately 91 percent of the duty is done in the beds and 9 percent in the convection space. With 730 F (661.1 K) exit temperatures, approximately 5.8 percent of the heat transfer occurs in the 2000 F (1366.7 K) CBC bed. This changes to approximately 7.1 percent at an exit temperature of 1420 F (1044.4 K). Therefore in all cases for the atmospheric fluidized bed not more than about 8 percent of the heat transfer can occur above the 1550 F (1116.7 K) temperature level which occurs in the main combustion beds.

PRESSURIZED FLUIDIZED BED

The pressurized fluidized bed operates with a temperature of 1650 F (1172.2 K) in the main beds and 2000 F (1366.7 K) in the CBC cell. With this furnace concept, no convection space heat transfer is normally utilized. For the pressurized fluidized bed not more than about 3 percent of the duty is available in the CBC and only about 2 percent or less available above the 1650 F (1172.2 K) main bed temperature.

PRESSURIZED FURNACE

The pressurized furnaces firing only clean gases from coal were considered. The operational furnaces were 2200 to 1150 F

(1677.8 to 894.4 K). The maximum combustion gas temperatures in this system are limited, to control the heat transfer rates to the tubes in the pressurized furnaces and thus limit metal temperatures and to stay within the study specified limit on NO_X production. For high-temperature cycles heat transfer rates are primarily limited by metals temperatures and secondarily by circulation requirements of boiling liquids.

HEAT EXCHANGER DESIGN CONSIDERATIONS

Tube sizes and wall thicknesses were set by pressure, temperature, and corrosion criteria, with materials selected to give reasonable wall thicknesses and corrosion resistance. Flow pressure drops were evaluated for the basic types of furnaces. A more detailed review of circulation requirements was not performed. This may indicate that different tube sizes should be used and other surfacing arrangements considered. The designs are considered capable of meeting the individual cycle pressure drop limitations, and any further changes are not considered to significantly affect the weight, cost, or approximate size of the equipment studied.

The Advanced Steam Cycle had rather liberal pressure drop allowances for the supercritical steam surface and reheated steam surface when going to higher temperatures. Heavy wall tubing and headers result at supercritical pressures when advanced steam conditions are required. This heavier wall tubing results in increased costs and pressure drops in flow circuits. For the lower pressure reheat surfaces at higher temperatures, increased heat transfer surface area is required, resulting in higher costs and somewhat higher pressure drops. It appears to be economic from a furnace standpoint to reheat steam to higher temperatures at moderate design pressures (up to 1500 psi [10.34×10^6 N/m^2]).

There appears to be no major limitation in producing reheated steam at 1200 F (922.2 K) temperature levels in fluidized bed combustion processes. In conventional furnace designs this also appears to be technically feasible, but with some increased difficulty in circuiting and surfacing the furnaces to avoid coal-ash corrosion.

Pressure drop considerations for the Supercritical CO_2 Cycle were strict for the type of furnaces studied. When heating gases such as CO_2, pressure drop limits limit the maximum working fluid velocities, and therefore heat transfer coefficient, inside the tubes. This subsequently controls the operating mean metal temperatures of heat transfer surfaces. When the allowable pressure drop was reduced by 30 percent the heat exchanger cost increased by 20 percent. When the allowable pressure drop was doubled, the cost reduction was 10 percent. In the pressurized furnace cases an attempt was made to control heat transfer rates, hence metal temperatures, by limiting maximum gas temperature and velocity through the tube bundles by staged firing techniques.

The general type of circuiting in the supercritical CO_2 cycle to control pressure drop was a "stick" tube with "posted" feeders from inlet headers. This type of construction requires considerable welding.

The Closed Cycle Helium Gas Turbine combustors and heat exchangers studied required similar circuiting and metals temperature control as with the supercritical CO_2 cycle. When heating helium to 1700 F (1200 K) in a pressurized furnace, extensive use of Mo-Re-2 material was required. This material is difficult to weld and basically unproven in high-pressure service. Heating helium to 1500 F (1088.9 K) in atmospheric fluidized beds resulted in low log mean temperature difference and required disproportionately large heat transfer surfaces. For economic heating of gases as process working fluids, a 200 F (366.7 K) temperature difference should be allowed between the working fluid and the mean operating temperature of a fluidized bed in order to limit the required heat transfer area.

Circulation requirements for the Liquid-Metal Topping Cycle were perhaps the most critical. It is necessary with liquid metal boilers to control the pressure drop from the boiling tube elements to the discharge tube separators to as low a value as possible. Potassium, for example, experiences roughly a 10 F (K) increase for every 1 psi (6894.7 N/m^2) of pressure drop above the discharge pressure. Because of the limited temperature differences available in atmospheric fluidized bed furnaces, and to a lesser degree in pressurized fluidized beds, the amount of heat transfer surface required varies directly with the boiling-to-bed temperature difference. Circulating pumps were indicated to ensure positive circulation in all circuits.

Stable flow regimes were required in order to prevent the formulation of a vapor blanket and cause overheating of the tube. Considering elevation effects, pressure drop in headers and manifolds and tube bundles, and pressure drops in risers, separators, and outlet headers and manifolds, a pressure drop of 25 psi (1.72×10^5 N/m^2) was allowed for on each boiling circuit. The pressure drop in the liquid-metal circuit varies very strongly with the vapor specific volume. The lower the liquid-metal boiling temperature and pressure the higher the vapor specific volume is. As a result, very high circulation ratios were required at the lower boiling pressures and temperatures. As the boiling temperature was raised, the required circulation ratio decreased. This resulted in lower auxiliary power consumption and cost for these cases.

Circulation requirements for the Liquid-Metal MHD cycle are complicated because of the necessity to provide heating of three separate fluids in the primary heat exchangers; heating helium and sodium to the 1500 F (1088.9 K) and reheating steam to 1000 F (811 K). This arrangement of multiple heat transfer duties would be very difficult to start up and control, and a detailed circulation and startup review was not performed on this system. Heating helium to 1500 F (1088.9 K) in an atmospheric fluidized bed operating at 1550 F (1116.7 K) represents a rather severe temperature pinch, and consequently raises the heat transfer requirements significantly.

Circulation requirements for the Inert-Gas MHD cycle are similar to conventional practice and place no severe restrictions on the process. The argon gas coolers required were not reviewed for mass transport effects of cesium present in the argon being cooled, although some condensation of cesium would be expected. Slag carryover in the argon heaters could be a serious design limitation.

The Open-Cycle MHD cases studied represent rather serious circulation considerations for furnace design. The steam circuiting of these furnaces must provide a low-temperature fluid sink with which to cool the combustion gases in the radiant furnace, intermediate-temperature air heater, and heat recovery boiler below the seed solidification temperature of 1950 F (1338.9 K) before passing the gases over compact convection surfaces. It is desirable to avoid "running liquid" seed and slag mixtures on heat transfer surfaces because of the affinity of these liquids to iron and nickel constituents of the materials of construction used in surfacing the unit. With the series circuiting suggested, it was difficult to accomplish this goal in the units and still maintain outside metal temperatures below 1100 F (866.7 K).

The circulation of steam and water in the High Temperature Fuel Cell heat recovery boiler is more conventional. The surface arrangement and circuitry of this heat recovery boiler follows more conventional designs for box type units firing natural gas fuels.

REFERENCE

1. Guthrie, K.M., "Data and Techniques for Preliminary Capital Cost Estimating," Chemical Engineering, Vol. 76, March 24, 1969, pp. 114-142.

Section 7

HEAT EXCHANGERS

Three types of heat exchangers were used in the cycles analyzed in Task I: conventional tube/shell designs, plate/fin design, and refractory designs. The refractory exchangers were employed when temperatures exceeded 1250 F (950 K), where the use of conventional units was either impossible or, because of the need to employ superalloys in their construction, their cost was prohibitive. The plate/fin designs were used only for the open-cycle gas turbine regenerator and are described in Section 2.1.

The refractory units were arranged as cylindrical columns with checker-brick matrices of alumina for moderate temperatures or zirconia for high temperatures and were designed and costed for the specific system by the cycle advocate. They were used in the open MHD cycle, the inert gas closed MHD cycle, and the high-temperature fuel cell, and are discussed in the sections of Volume II which describe these cycles.

The conventional tube/shell designs include regenerators, precoolers and compressor intercoolers as prime-cycle components and vapor generators, condensers, and regenerators as organic bottoming cycle components. The organic bottoming cycle heat exchangers were designed by Thermo Electron Corporation, Waltham, Massachusetts. The prime cycle heat exchangers were designed by General Electric (Heat Transfer Products Department), who, for consistency, estimated the cost of all units.

Because of the large number of parametric variations studied, it was not feasible to develop a design for each case. Consequently, only a few specific configurations were actually selected for detailed thermodynamic analysis and costing. The selection was, however, organized so that the parameters which influence heat exchanger design (i.e., pressure drop, temperature difference, and temperature and pressure levels) were varied over a sufficiently broad range to enable cost estimates of all other cases to be made by interpolation.

[illegible] heat exchangers in prime cycles were designed as cross-count[illegible] flow units. Configurations included rectangular tube bundles in a round shell, fixed tube sheets, conventional U-tube and, when thermal expansion was expected to create problems during startup, U-tube/U-shell designs. A schematic of the latter is shown in Figure 7-1, and the component configurations for the various cycles are listed in Table 7-1.

Shell size was dictated either by shipping limitations (maximum 12 feet diameter [3.7 m]) or tube sheet thickness (14 inches [0.36 m] maximum). Tube spacing was arranged in a staggered (triangular) array and tube thickness was selected to suit the pressure differential and 18 gauge material was adopted as a

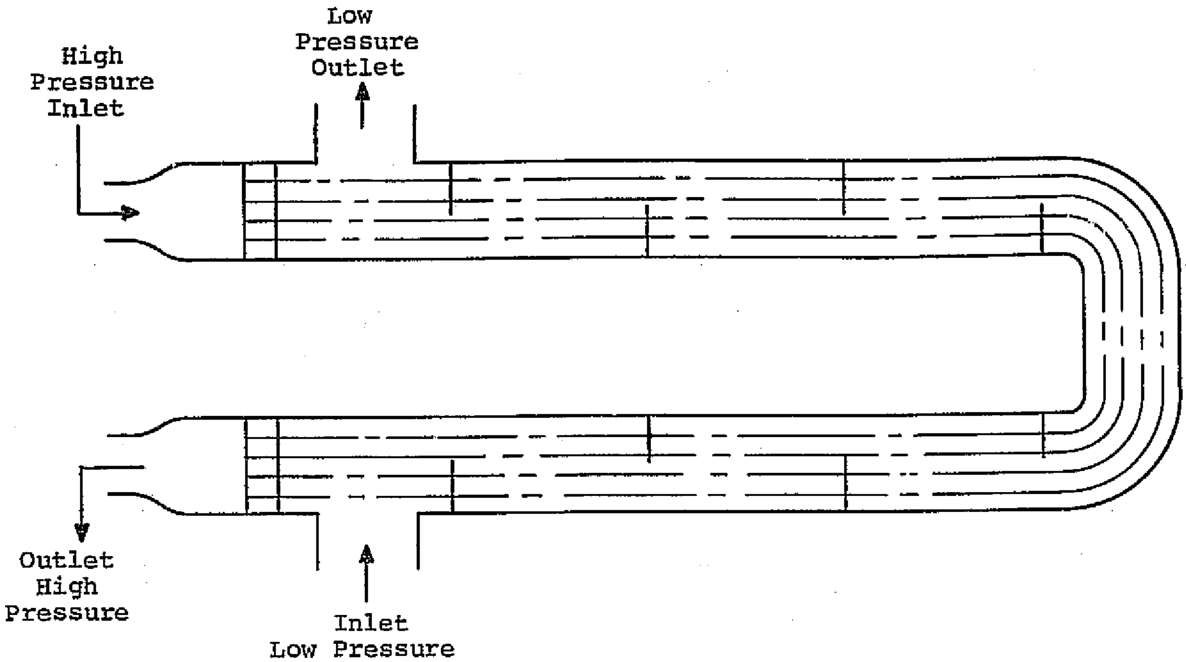


Figure 7-1. U-Tube, U-Shell Heat Exchanger

minimum thickness because of mechanical design and construction limitations. Generally the tube and shell sides accommodated high- and low-pressure fluids, respectively. However, in the case of the closed-cycle helium gas turbine regenerator, it was found that better overall cycle performance could be obtained with this situation reversed. Nominally the pressure loss was split in the ratio 2/1 between the low-pressure (high-temperature) and the high-pressure (low-temperature) sides, respectively. However, this ratio was intended merely as a suggestion to cycle advocates, who ultimately modified its value to give the best cycle performance. Operating pressures and temperatures were within the limits specified for the selected materials by the ASME Section 8 Code. Fouling factors of 0.0003 and 0.001 hr ft^2 F/Btu (5.29 x 10^{-5} to 1.76 x 10^{-4} ft^2 K/watt) were assumed throughout for the process side and water side, respectively. Standard heat transfer and fluid flow correlations were used to predict heat transfer coefficients and determine heat exchanger configurations from given input values of pressure drop. These correlations, which were modified to reflect manufacturing practice, are proprietary information of the heat exchanger designer.

Cost estimates were based primarily on the total tube length with allowances factored in for shells, channels, and tube sheets. Estimates of heat exchanger costs for selected base case cycles are given along with a physical description of the heat exchangers in Table 7-2. The material selection for the component parts of all tube shell heat exchangers used in this study is given in Table 7-3. Clearly there are large incremental changes in total

Table 7-1

HEAT EXCHANGER CONFIGURATIONS

Cycle/Component	Heat-Exchanger Configuration
• Closed-cycle gas turbine/Helium regenerator (case 40)	Conventional U-tube (straight shell)
• Closed-cycle gas turbine/Helium precoolers and intercoolers (cases 10-13) • CO_2-cycle/precoolers and intercoolers	Fixed tube sheets (straight tubes, straight shell)
• Closed-cycle liquid metal MHD/Helium regenerator (case 17) • CO_2-cycle/high-temp & low-temp regenerators • Closed-cycle gas turbine/Helium regenerators <u>except</u> cases 10-13	U-tube/U-shell
• Closed-cycle liquid metal MHD/Helium precooler (case 17) • Closed-cycle inert gas MHD/argon precoolers • Open-cycle air-cooled gas turbine combined cycle/air intercooler (case 20)	Rectangular tube bundle in round shell

cost due to the different materials selected for different operating temperatures. Carbon steel was selected for temperatures below 800 F (700 K). Between 800 F (700 K) and 1100 F (867 K) stainless steels were used. Above this range, inconel alloys were necessary. The precoolers and intercoolers used in each cycle were all fabricated with SB111 tubes and carbon steel shells, channels, and tube sheets at a cost of between \$3.3/ft (\$10.7/m) and \$4.5/ft (\$14.6/m) for unfinned tubes and about \$7/ft (\$22.7/m) for finned tubes (finning on outside of tubes only). Price variations are due to different tube densities and tube sizes within shells. Regenerators of carbon steel throughout cost between \$3.46/ft (\$11.25/m; CO_2 cycle) and \$5.02/ft (\$16.3/m; liquid metal MHD cycle). When stainless steel tubes were combined with carbon steel shells, regenerator costs increased to \$12/ft (\$39/m), and to over \$16/ft (\$52/m) when stainless steel was used throughout, as in the high-temperature CO_2 regenerator. In this task no attempt was made to estimate the cost of composite regenerators, i.e., single regenerators made up of both high- and low-temperature alloys. If gas temperatures warranted the use of high-temperature alloys, then these alloys were used throughout.

Table 7-2

PHYSICAL DESCRIPTION OF HEAT EXCHANGERS

Unit	Parametric Case Number	Size Each (ft) (HxWxL)	Wt. each (lb)	No. Units	Tube Count Per Unit	Tube Length (in.)	Tube Dia. (in.)	Area/Unit (ft^2)	Total Cost $ x 10^{-6}	Cost** $/ft
Supercritical CO_2										
CO_2 regenerator (high temp)	1	4x14x34	170,000	160	1342	582	0.5	9050	166.4	15.8
CO_2 regenerator (low temp)	1	7x23x42	300,000	32	4821	720	0.5	37800	32.0	3.46
CO_2 precooler	1	8x8x60	170,000	10	6522	554	0.5	39400	9.9	3.29
Closed Cycle & Gas Turbine										
He regenerator/closed cycle gas turbine	1	7x24x45	200,000	20	1052	703	1.0	16100	15.2	12.33
He precooler/closed cycle gas turbine	1	6x18x28	45,000	10	1734	360	0.75	10206	2.3	4.42
Organic condenser/closed cycle gas turbine	37	12x12x42	400,000	4	16379	384	0.625	849087*	6.8	3.24
Liquid Metal MHD										
He regenerator/liquid metal MHD	17	20x6x36	125,000	40	1992	480	0.75	14600	16.0	5.02
He precooler/liquid metal MHD	17	12x12x40	210,000	4	3844	324	1.0	70369*	3.0	7.23
Open Cycle Gas Turbine Combined										
Air intercooler/open-cycle air-cooled gas turbine combined cycle	20	10x10x40	125,000	1	2700	312	1.0	47600*	0.5	7.12
Closed Cycle Inert Gas MHD										
Argon precooler/inert gas MHD	1	12x12x85	500,000	2	3696	840	1.0	351000*	3.6	6.95
Open Cycle Gas Turbine-Recuperative										
Organic boiler/open cycle gas turbine	34	11x8x13	160,000	4	3120	276	1.0	72000*	0.8	2.78
Organic regenerator/open cycle gas turbine	34	7x7x45	200,000	2	783	480	1.0	96000*	0.5	7.98

*Finned tubes (outside only)
**Average cost per linear foot, of heat exchanger tube

Table 7-3

MATERIAL SELECTION FOR HEAT EXCHANGER COMPONENT PARTS

Heat Exchanger	Tubes	Shell Channel & Tube Sheets
* All precoolers and intercoolers	SB111 (90/10CuNi)	Carbon Steel
* <u>Supercritical CO_2 Cycle</u>		
• High-temperature regenerator	321HSS	321HSS
Case 10	Inconel 800	Inconel 800
• Low-temperature regenerator	Carbon Steel	Carbon Steel
* <u>Closed-Cycle Helium Gas Turbine Cycle</u>		
• Regenerator	304SS	Shells 1C-1/2MO Channel & tube sheet 1/2Cr-1/2MO
Cases 12,13,25,28,31 & 32	304SS	Shells 2-1/2Cr-1MO Channel & tube sheet 304SS
• Organic boiler & condenser	Carbon Steel	Carbon Steel
* <u>Liquid Metal MHD Cycle</u>		
• Regenerator	Carbon Steel	Carbon Steel
* <u>Open-Cycle Gas Turbine</u>		
• Organic regenerator, boiler and condenser	Carbon Steel	Carbon Steel

All organic heat exchangers employed finned tubes and were constructed of carbon steel. In the closed gas turbine cycle, the organic fluid was contained within the tubes of the organic boiler. The tubes were externally finned with a pitch of 2 fins per inch (78 fins/m) and a fin height of 0.125 inches (0.0032 m). The tube arrangement was staggered with a pitch of 1.375 inches (0.035 m). The external flow was multipass cross-counter flow.

The organic boiler for the open gas turbine cycle again contained the organic fluid within the tubes, with exhaust gas flow from the gas turbine on the shell side. The tubes were externally finned (6 fins/inch [234 fins/m], fin height 0.375 inches [0.0096 m]) and the tubes were arranged in a staggered formation on a 2-inch (0.05 m) pitch.

In the organic condensers, the condensation process occurred on the shell side. Again tubes were staggered on a 0.9375-inch pitch (0.024 m). Fin spacing was 19 fins/inch (741 fins/m) and the fin height was 0.0625 inches (0.0016 m). The costs of the organic boiler and the organic condenser were approximately the same at about $3/ft ($9.75/m) of tube length.

An organic regenerator was used in the open gas turbine regenerative cycle at a cost of about $8/ft ($26/m).

APPENDIX A

DESIGN DATA AND EQUIPMENT OUTLINES FOR PRIMARY HEAT INPUT SYSTEMS

An example is given here of the design data for each energy conversion system under study. Line drawings denote the furnace configurations that were evaluated.

LIST OF TABLES

LIST OF ILLUSTRATIONS

Table A-1

BASIC FURNACE SIZING AND SURFACING CONSIDERATIONS
ADVANCED STEAM CYCLE

	AFB Coal Tower	PFB Coal Tower	PF Lo Btu	CF coal	APH	
Surfacing						
Cycle Point	1	24	21	17	1	
Gas - Hot Fluid Flow (10^6#/hr)	1.749	1.606	1.859	7.115	-	
Sectional Area (ft^2)	2520	423.2	113	3427	-	
Furnace Velocity (ft/sec)	9.21	5.29	30.11	33.05	-	
Gas T_1/T_2 (°F)	1550/730	1650/1600	2200/1800	1950/730	730/252	
Working Fluid t_1/t_2 (°F)	1200/510				655/86	
LMTD - ΔT) Avg. (°F)	~ 332	~ 647	937 Avg.	798 Avg.	115	
Uo (Btu/hr-ft^2 °F)	~ 25	~ 45	60	13	5e	
q"o (Btu/hr-ft^2)	8280	29102	56228	10370	575	
So (10^6 -ft^2)	.186	.0334	.0114	.594	.371	
Qo (10^6 -Btu/hr)	1540	972	641	6160	213	
Surface Wt. (tons/10^6-ft^2)	4269	6587	3509		-	
Tube size OD (in)	1¼" & 1-3/4"	1¼" & 1-3/4"	1¼" & 1-3/4"	1¼"→ 2½"	Req.	
Weights						
Pressure Part Wt. (tons)	913	286	52	-	397	
Casing, etc. Wt. (tons)	287	134	168	-		
Total Wt. (tons)	1200	420	220	6000	397	
Costs						
Unit cost \$ x 10^6	5.624	2.796	1.006	25.881	.731	
Cost rate \$/ton	4687	6657	4573	4314	1841	
Cost rate \$/ft^2	30.24	83.71	88.25	43.57	1.97	
Cost rate \$/10^6 Btu/hr	3652	2877	1570	4202	3432	

Pressure Part Composition	AFB Coal Tower		PFB Coal Tower		PF Lo Btu		CF coal		APH	
Service (Wt.-Tons)	SH(173)	RH(37)	SH(150)	RH(10)	SH(38)	RH(5)	SH(450)	RH(450)		
Fluid Temperature Exit	1200	1000	1200	1000	1200	1000	1200	1000		
Wt. % MO-RE 2										
Wt. % Hastelloy X	-	-	-	-	-	-	46	-		
Wt. % Inconel 601	50	-	35	-	35	-	-	-		
Wt. % TP 347H	32	-	25	-	25	-	48	-		
Wt. % TP 304H	-	71	-	50	-	60	-	30		
Wt. % T-22	18	-	20	-	20	40	6	30		
Wt. % T-2	-	29	20	50	20	-	-	20		
Wt. % CS	-	-	-	-	-	-	-	-	100	

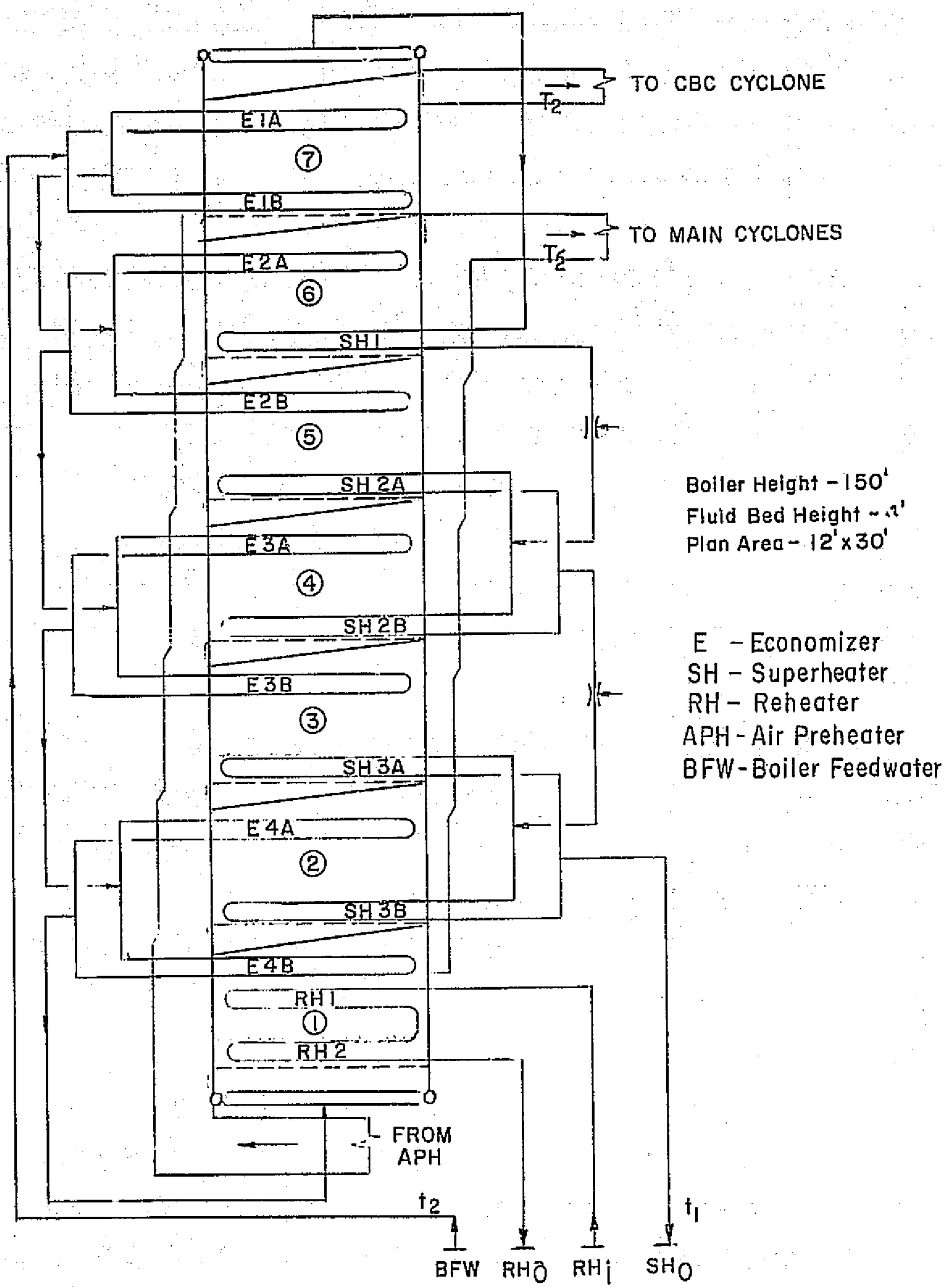


Figure A-1. Advanced Steam Cycle in Atmospheric Fluid Bed—Case 1, 200 MWe Tower

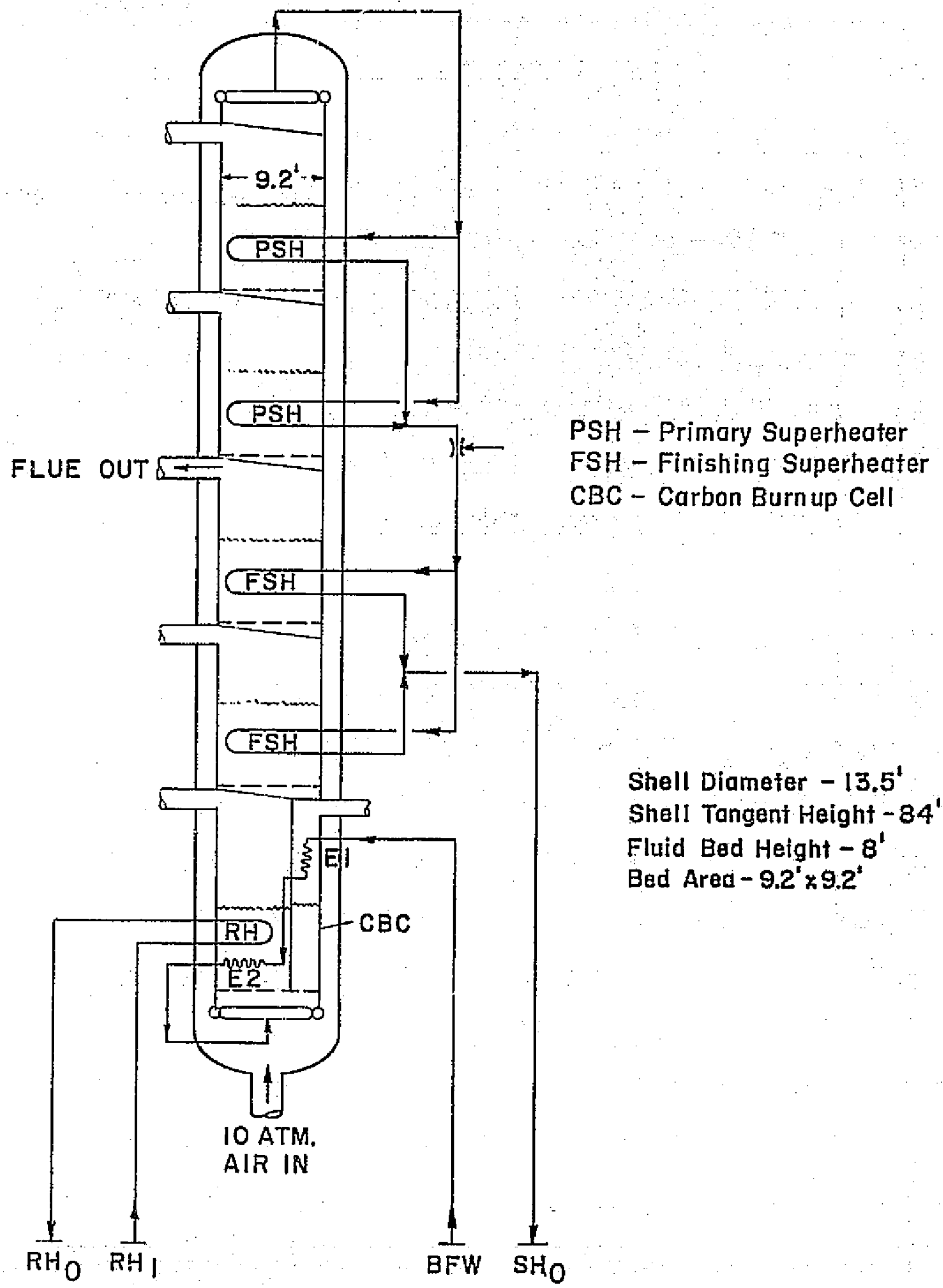
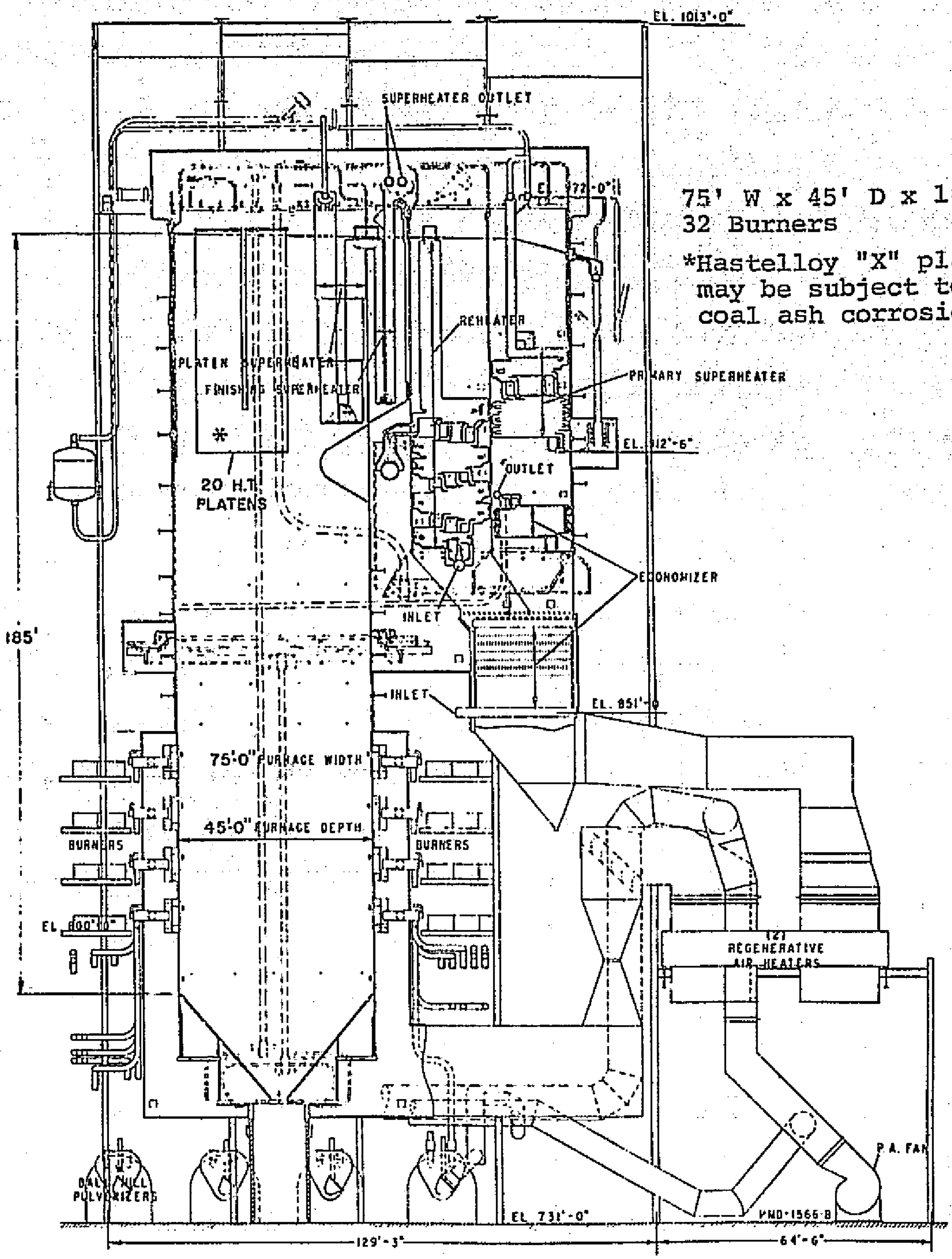


Figure A-2. Advanced Steam Cycle in Pressurized Fluid Bed—Case 24, 200 MWe Tower



75' W x 45' D x 185' High
32 Burners
*Hastelloy "X" platens
may be subject to
coal ash corrosion

Figure A-3. Advanced Steam Cycle in Conventional Furnace—
Case 17, 800 MWe Unit

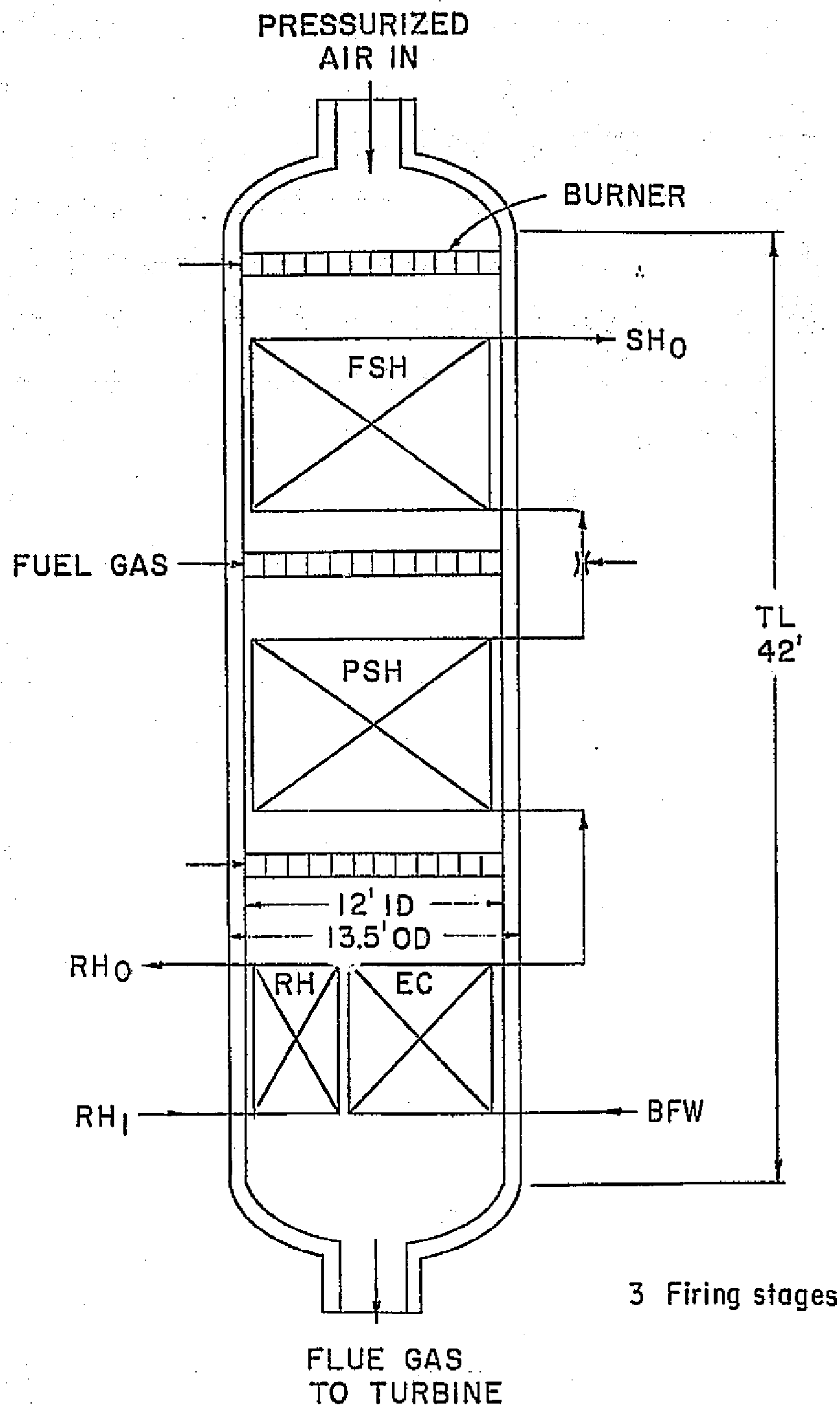


Figure A-4. Advanced Steam Cycle in Pressurized Furnace—Case 21, ~200 MWe Tower

Table A-2

BASIC FURNACE SIZING AND SURFACING CONSIDERATIONS
SUPERCRITICAL CO_2 CYCLE

	AFB Coal Tower	PFB Coal Barrel	PF lo Btu	PF hi Btu	Hi Temp APH HP	Hi Temp APH LP
Surfacing						
Cycle Point	1	5	6	9	1	1
Gas - Hot Fluid Flow (10^6#/hr)	1.749	1.6063	1.859	1.173		
Sectional Area (ft^2)	7x360 = 2520	4x270 = 1080	113	113	x	x
Furnace Velocity (ft/sec)	9.21	2.073	35.40	22.79		
Gas T_1/T_2 (°F)	1550/1100	1650/1600	2200/1800	2200/1150	1100/730	730/300
Working Fluid t_1/t_2 (°F)	1350/987	1350/963	1350/980	1200/855	1025/600	600/86
LMTD - ΔT) Avg. (°F)	153	448	(815)	612	100	169
Uo (Btu/hr-ft^2 °F)	44.66	48.02	(44.50)	49.93	5	5e
q"o (Btu/hr-ft^2)	6833	21511	(40717)	30556	501	845
So (10^6-ft^2)	.222	.0491	(.0173)	.0324	.389	.226
Qo (10^6- Btu/hr)	1517	1056.2	704.4	990	195	191
Surface Wt. (tons/10^6-ft^2)					2339	-
Tube size OD (in)	1¼"	1"	(2½")	2½"	2	Reg.
Weights						
Pressure Part Wt.(tons)	1467	293	(415)	664	910	397
Casing, etc. Wt. (tons)	865	440	(260)	396		
Total Wt. (tons)	2332	733	(675)	1060	910	397
Costs						
Unit cost \$ x 10^6	25.117	6.409	(9.322)	14.718	2.531	.731
Cost rate \$/ton	10,771	8744	(810)	13,885	2781	1641
Cost rate \$/ft^2	113	131	(558)	454	6.51	1.97
Cost rate \$/10^6 Btu/hr	16557	6068	(13233)	22166	12,980	3432
Pressure Part Composition						
Wt. % MO-RE 2			(20)	50		
Wt. % Hastelloy X	50	50	(80)	50		
Wt. % Inconel 601	50	50				
Wt. % TP 347H						
Wt. % TP 304H						
Wt. % T-22						
Wt. % T-2					40	
Wt. % CS					60	100

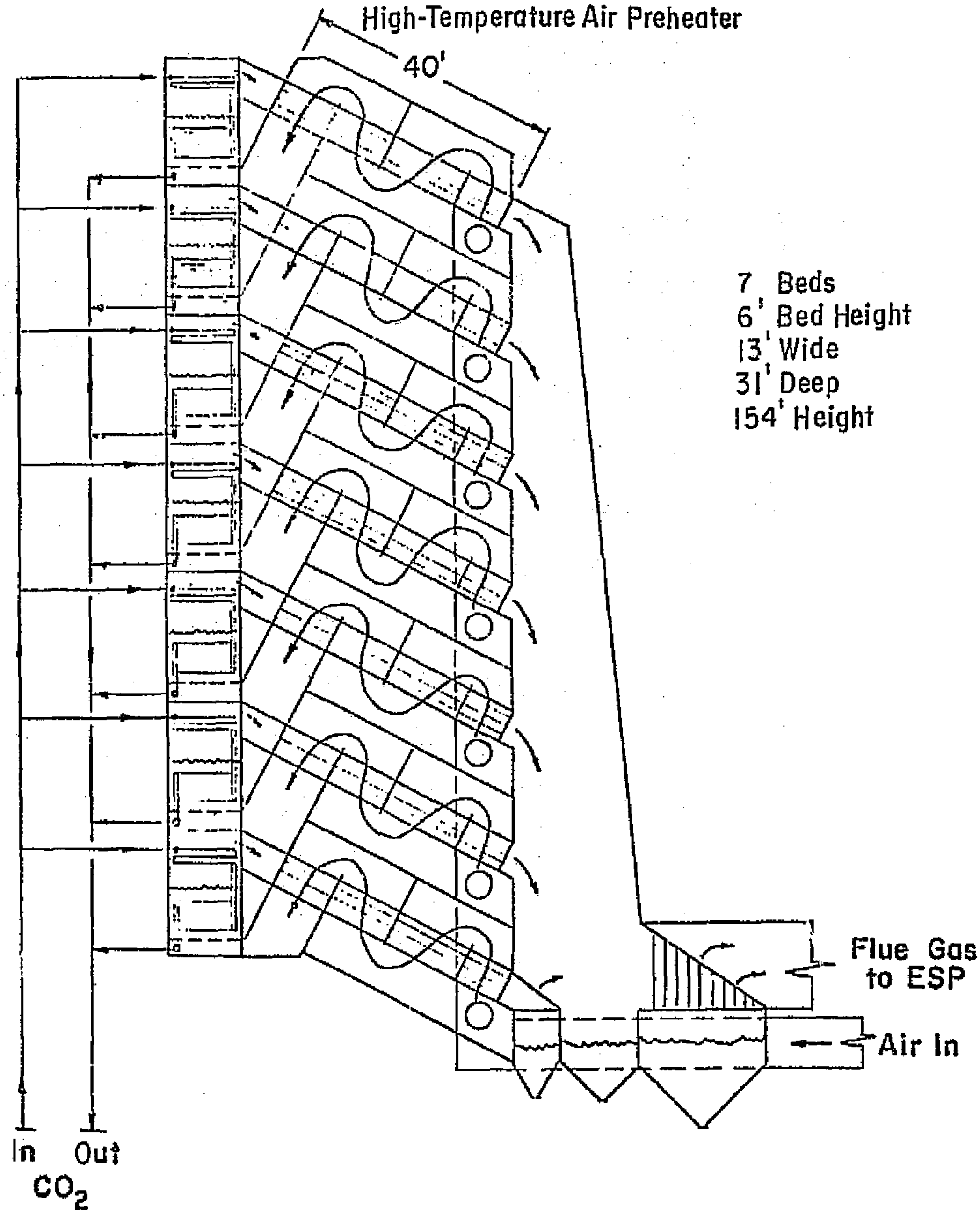


Figure A-5. Supercritical CO2 Cycle in Atmospheric Bed—Case 1

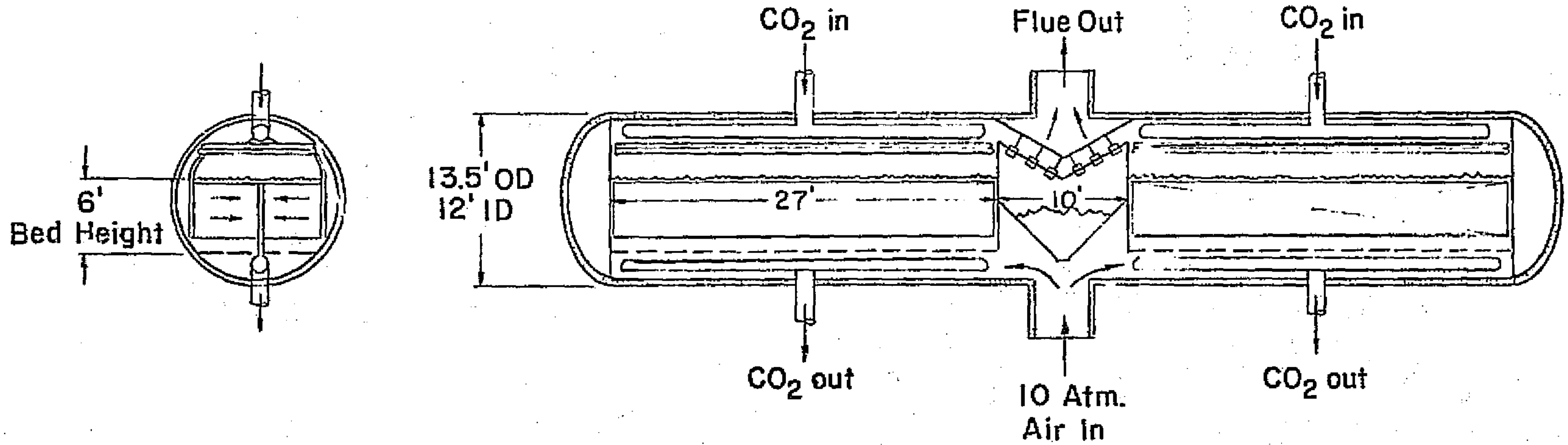


Figure A-6. Supercritical CO_2 Cycle in Pressurized Fluid Bed—Case 5
(Two per Module)

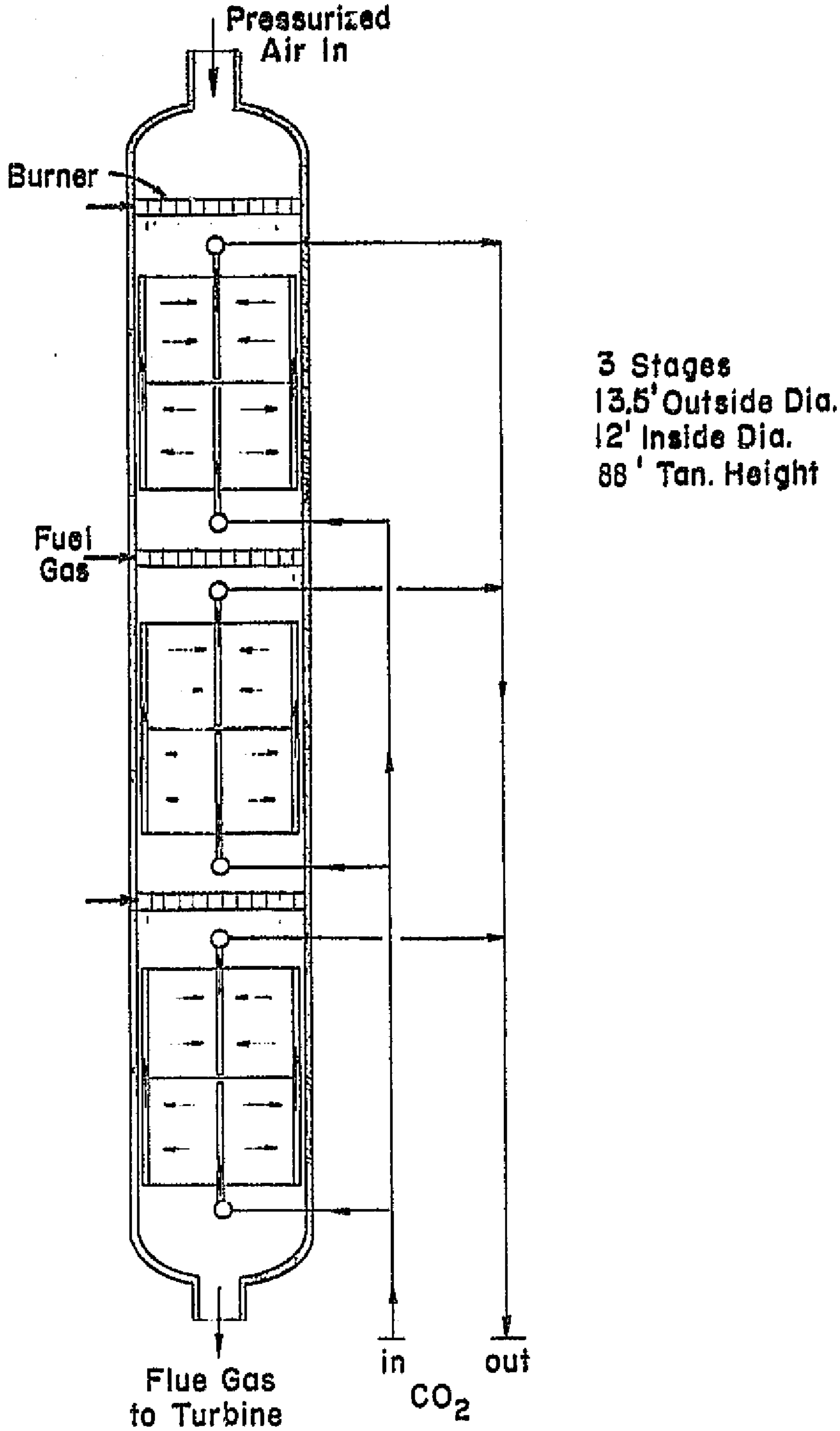


Figure A-7. Supercritical CO_2 Cycle in Pressurized Furnace—Case 6, Low-Btu Fuel

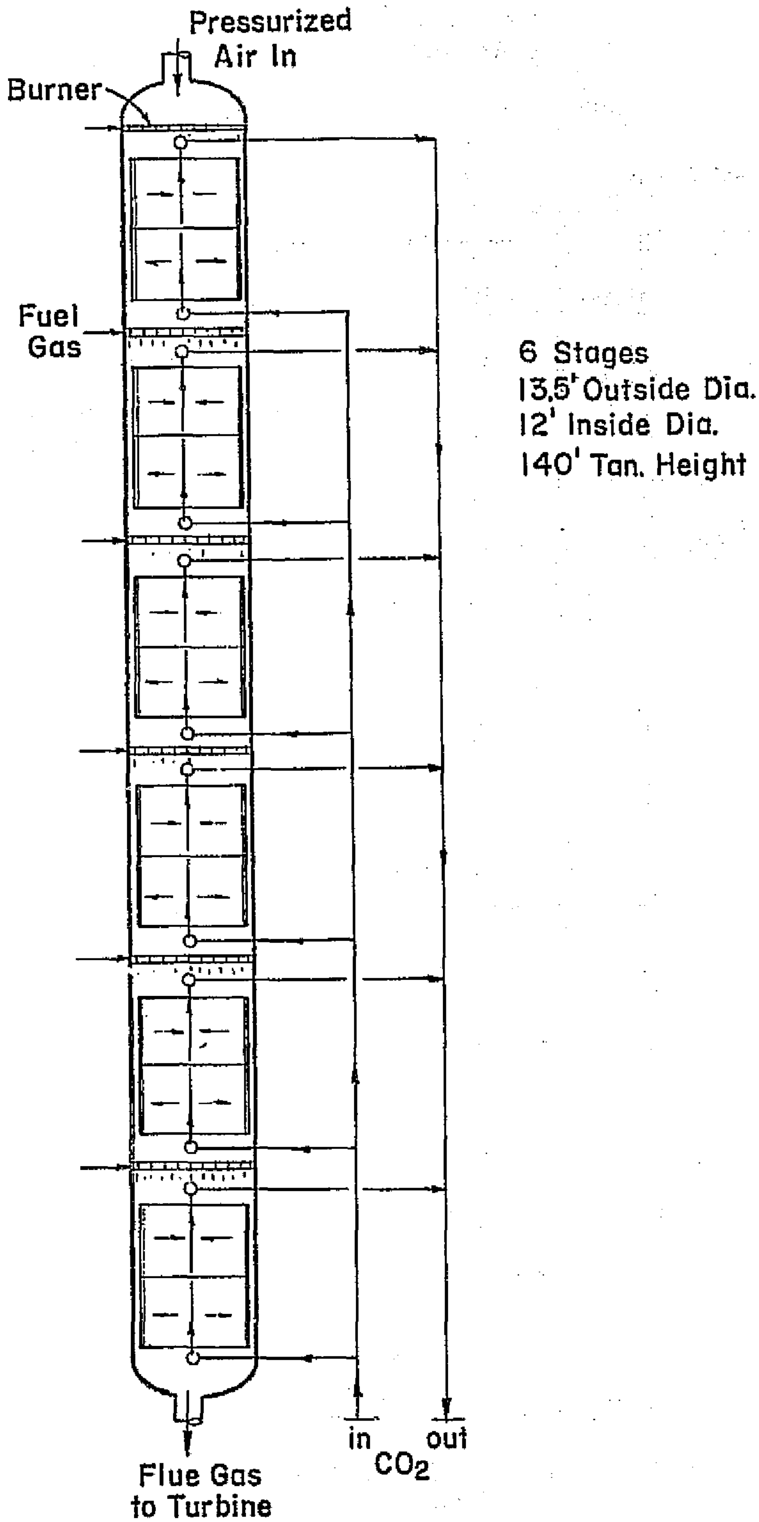


Figure A-8. Supercritical CO_2 Cycle in Pressurized Furnace—Case 9, High-Btu Fuel

Table A-3

BASIC FURNACE SIZING AND SURFACING CONSIDERATIONS
CLOSED-CYCLE HELIUM GAS TURBINE

	AFB Coal Tower	PFB Coal Barrel	PF lo Btu	PF hi Btu	Hi Temp APH HP	Hi Temp APH LP
Surfacing						
Cycle Point	1	8	4	7	1	1
Gas - Hot Fluid Flow (10^6#/hr)	1.749	1.6063	1.8590	1.173		
Sectional Area (ft^2)	7x360 = 2520	4x310 = 1240	113	113		
Furnace Velocity (ft/sec)	9.21	1.810	28.9	22.8		
Gas T_1/T_2 (°F)	1550/919	1650/1600	2200/1800	2200/1150	919/730	730/297
Working Fluid t_1/t_2 (°F)	1507/876	1507/876	1507/876	1507/876	844/602	602/86
LMTD - ΔT) Avg. (°F)	196	358	687	687	99	166
Uo (Btu/hr-ft^2 °F)	29.34	51.22	49.90	50.01	5	5e
q"o (Btu/hr-ft^2)	5751	18337	34.278	34354	495	830
So (10^6-ft^2)	.2643	.0576	.0187	.0294	.171	233
Q_o (10^6- Btu/hr)	1520	1056.2	641	1010	84.4	193.6
Surface Wt. (tons/10^6-ft^2)	3916	2951	11176	11190	2295	-
Tube size OD (in)	1¼"	1"	2½"	2½"	2	Req'd.
Weights						
Pressure Part Wt. (tons)	1242	204	251	395	471	444
Casing, etc. Wt. (tons)	920	473	263	340		
Total Wt. (tons)	2162	677	514	735	471	444
Costs						
Unit cost \$ x 10^6	20.038	5.362	5.908	9.204	1.073	.709
Cost rate \$/ton	9268	7920	11,494	12,522	2278	1597
Cost rate \$/ft^2	76	93	316	313	6.27	3.04
Cost rate \$/10^6 Btu/hr	13183	5077	9217	9113	12713	3662
Pressure Part Composition						
Wt. % MO-RE 2				20		
Wt. % Hastelloy X	40	80	100	80		
Wt. % Inconel 601						
Wt. % TP 347H						
Wt. % TP 304H	60	20				
Wt. % T-22						
Wt. % T-2					30	
Wt. % CS					70	100

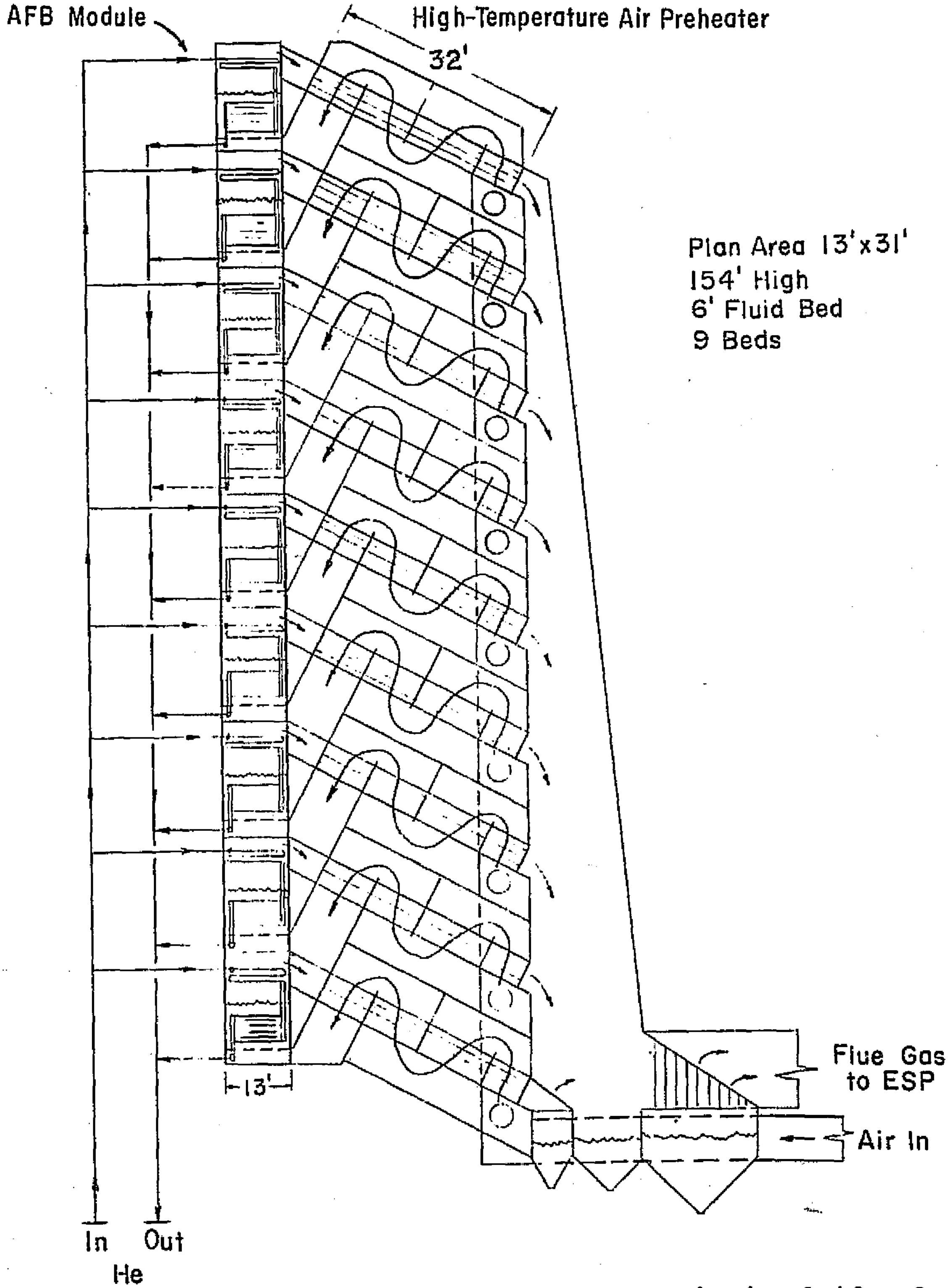


Figure A-9. Helium Gas Turbine Cycle in Atmospheric Fluid Bed—Case 1

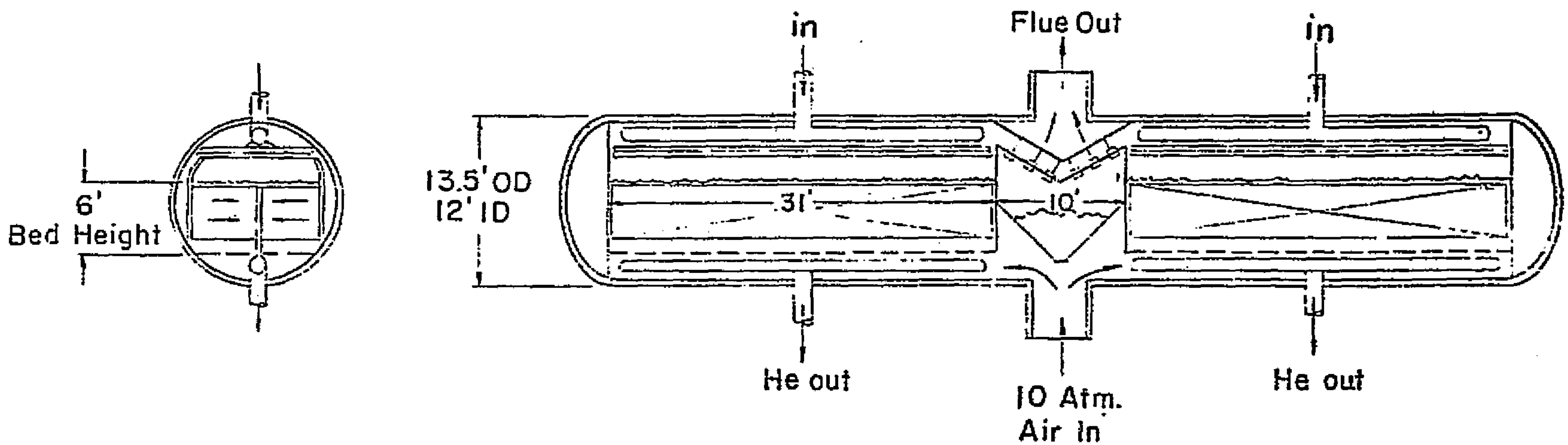


Figure A-10. Helium Gas Turbine Cycle in Pressurized Fluid Bed—Case 8
(Two per Module Required)

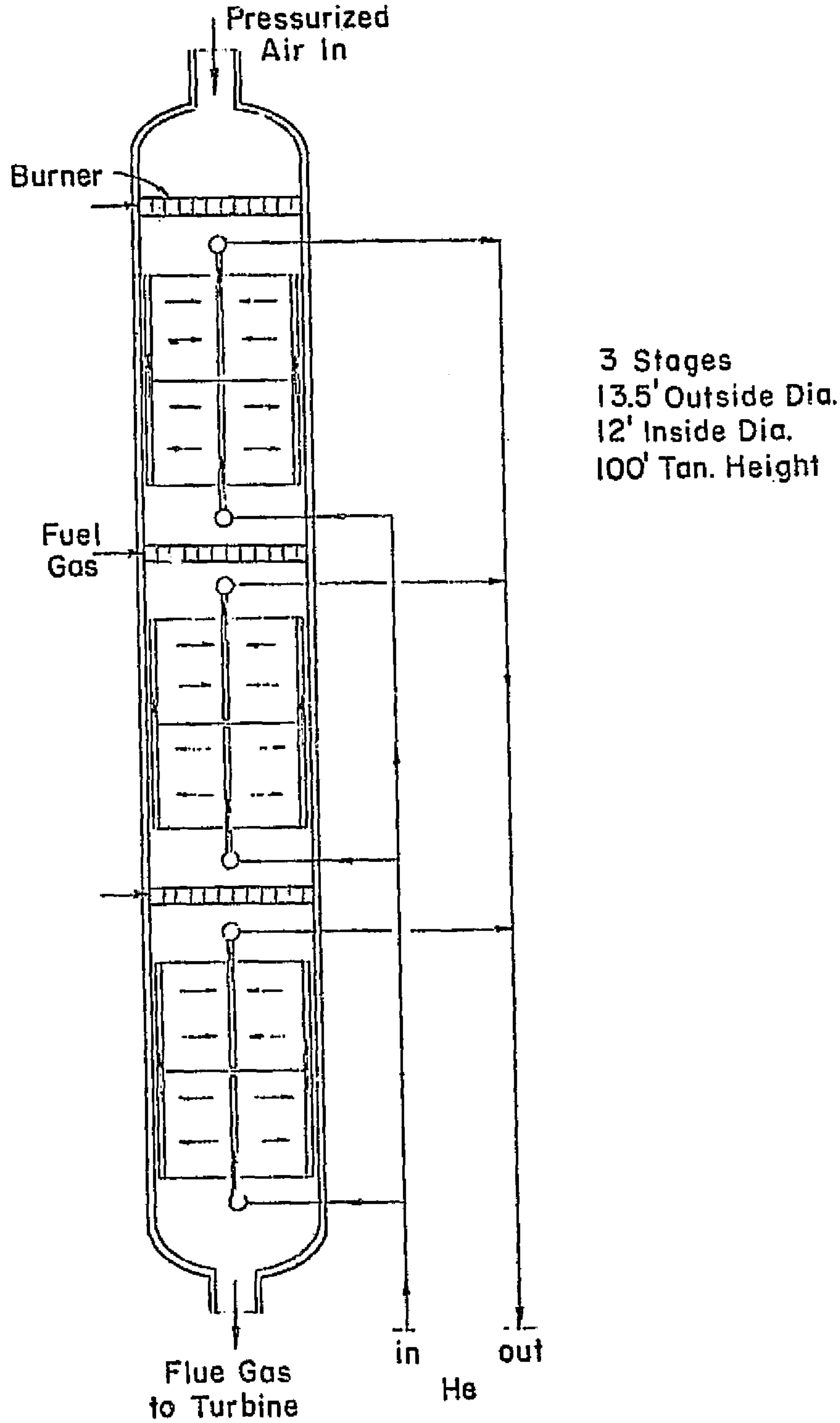


Figure A-11. Helium Gas Turbine Cycle in Pressurized Furnace—Case 4, Low-Btu Fuel

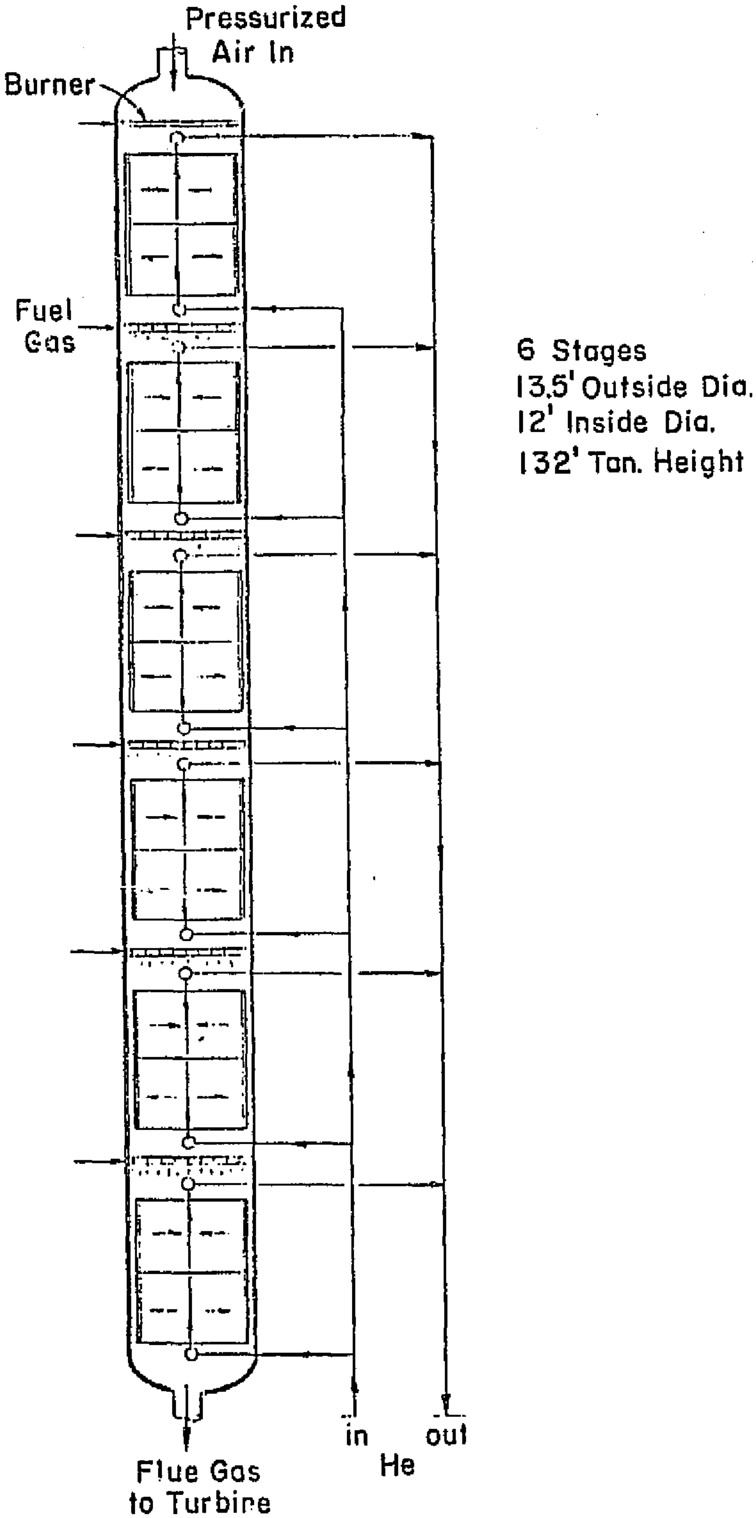


Figure A-12. Helium Gas Turbine Cycle in Pressurized Furnace—Case 7, High-Btu Fuel

Table A-4

BASIC FURNACE SIZING AND SURFACING CONSIDERATIONS
LIQUID-METAL TOPPING CYCLE

	AFB Coal Tower	PFB Coal Tower	PF lo Btu	PF hi Btu	IHX + WHB	Hi Temp APH HP	Hi Temp APH LP
Surfacing							
Cycle Point	1	9	4	7	1 Condenser	1	1
Gas - Hot Fluid Flow (10^6#/hr)	1.749	1.6063	1.859	1.173			
Sectional Area (ft^2)	15x360 = 5400	11x90.3 = 993	113	113			
Furnace Velocity (ft/sec)	4.3	2.25	28.9	22.8			
Gas T_1/T_2 (°F)	1550/1420	1650/1600	2200/1800	2200/1750	1100/1100	1420/730	730/362
Working Fluid t_1/t_2 (°F)	1400/1100	1400/1100	1400/1100	1400/1100		1345/524	524/86
LMTD - ΔT) Avg. (°F)	153	260	606	588	274	130	241
Uo (Btu/hr-ft^2 °F)	38	45.8	60	50	220	5	5e
q"o (Btu/hr-ft^2)	5820	11.895	36.355	29.378	60.275	649	1205
So (10^6-ft^2)	.256	.086	.0203	.0282	.0218	.522	.171
Qo (10^6- Btu/hr)	1490	1023	738	827	1314	339	206
Surface Wt. (tons/10^6-ft^2)	3395	3512	4435	5510			-
Tube size OD (in)	1¼"	1"	2½"	2½"	1¼"	2"	Reg.
Weights							
Pressure Part Wt.(tons)	1129	392	104	202	98	835	381
Casing, etc. Wt. (tons)	1535	561	273	404			
Total Wt. (tons)	2664	953	377	606	98	835	381
Costs							
Unit cost \$ x 10^6	27.005	9.649	2.982	6.739	.479	4.555	.660
Cost rate \$/ton	10,137	10,125	7910	11,121	4888	5455	1732
Cost rate \$/ft^2	105.49	112.20	146.9	239.97	21.97	8.73	3.86
Cost rate \$/10^6 Btu/hr	18.124	9432	4041	8149	365	13437	3204
Pressure Part Composition							
Wt. % MO-RE 2							
Wt. % Hastelloy X	100	100	100	100			
Wt. % Inconel 601							
Wt. % TP 347H							
Wt. % TP 304H						40	
Wt. % T-22					60	20	
Wt. % T-2					40	40	
Wt. % CS							100

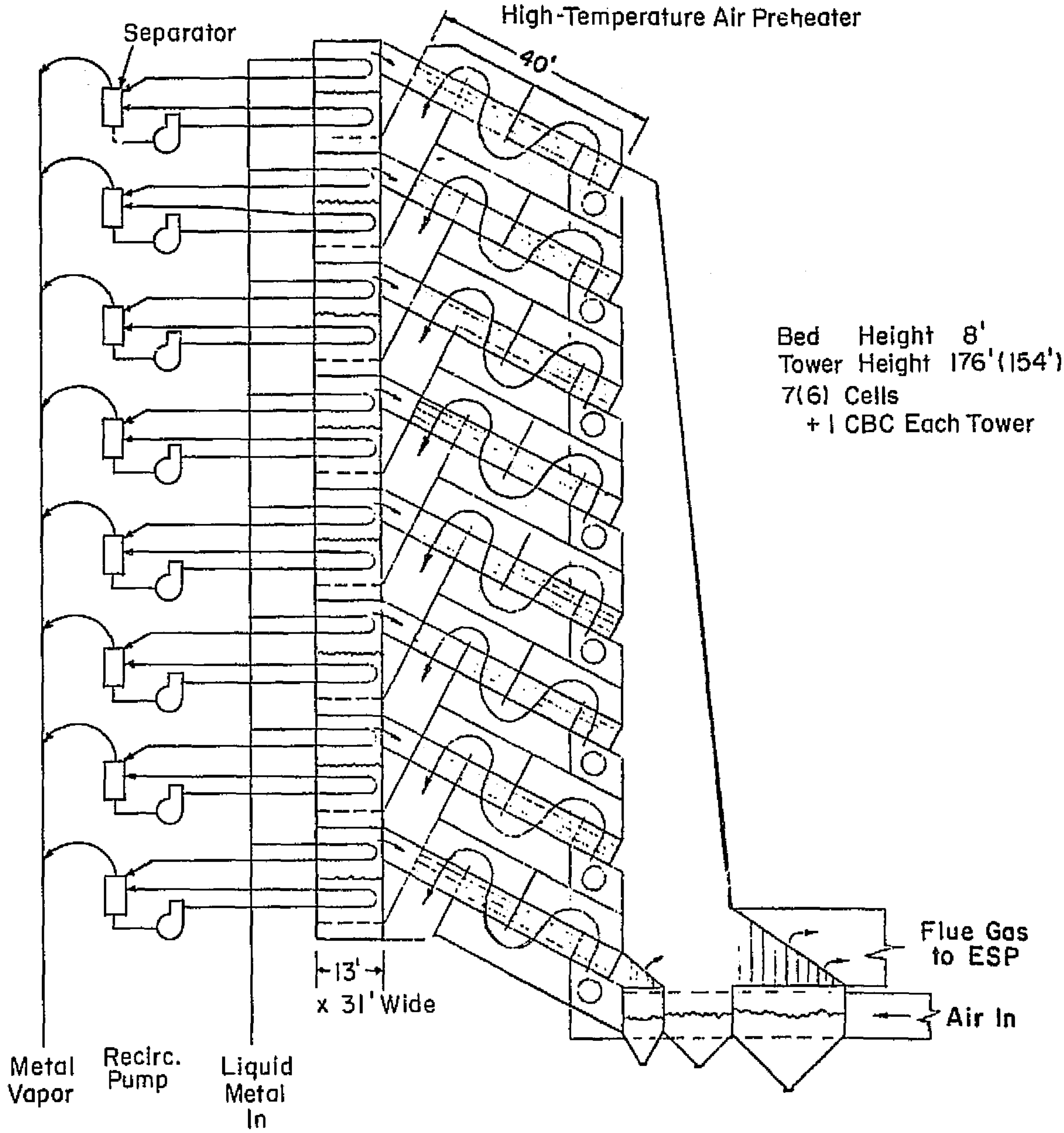


Figure A-13. Liquid-Metal Topping Cycle in Atmospheric Fluid Bed—Case 1 (Two Towers per Module)

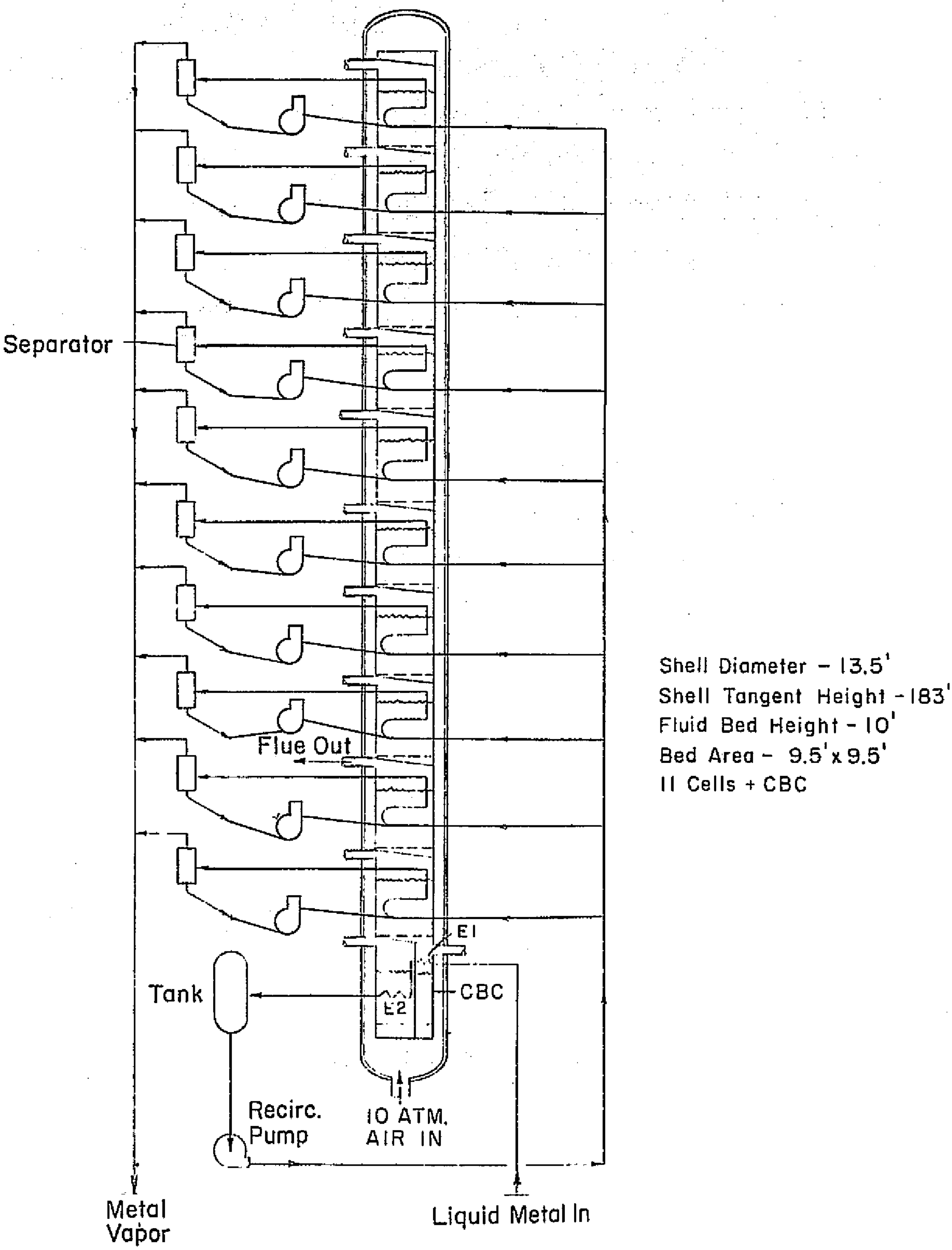
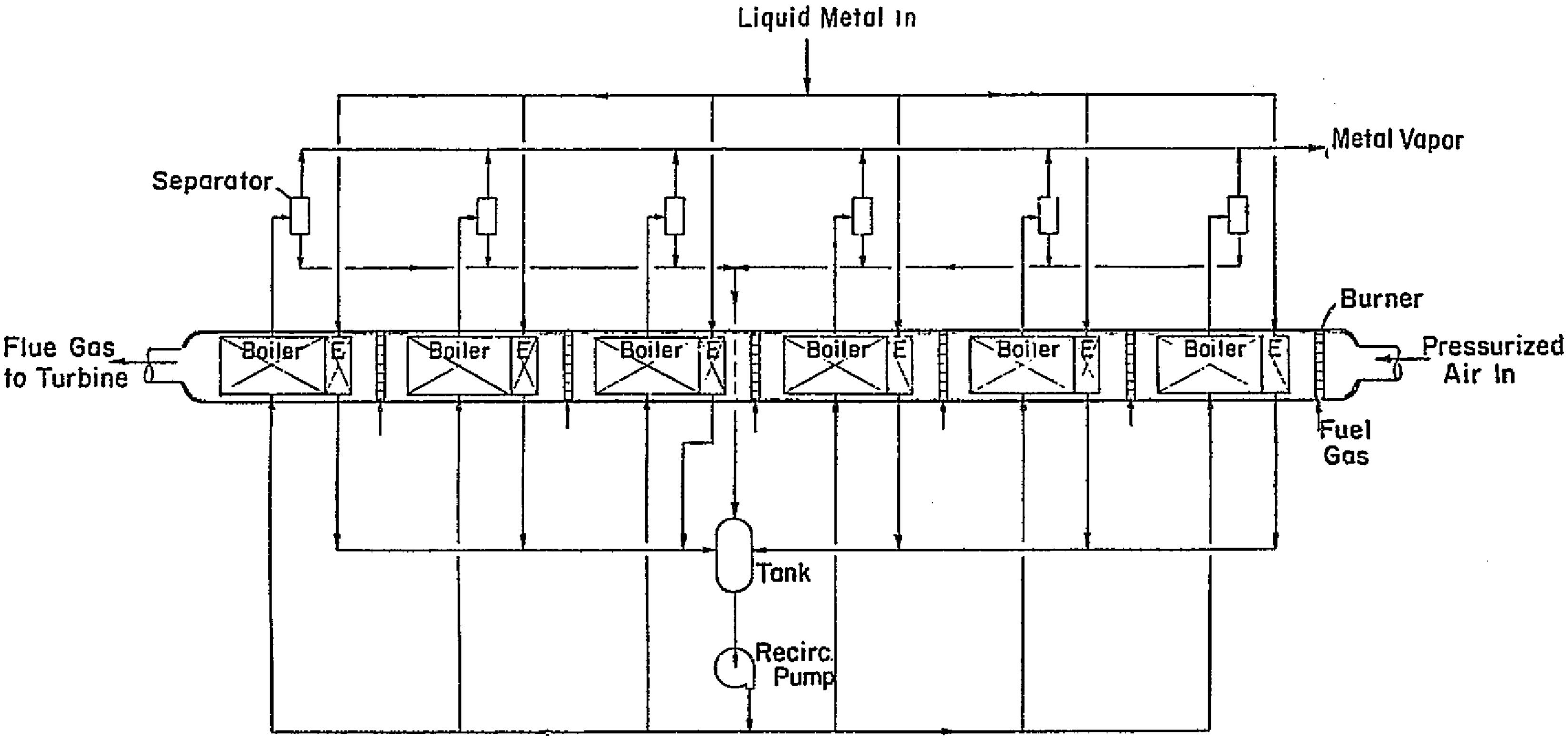


Figure A-14. Liquid-Metal Topping Cycle in Pressurized Fluid Bed—Case 9, 200 MWe Tower



Shell Diameter 13.5'
Tangent Length 161'
6 Stages

Figure A-15. Liquid-Metal Topping Cycle in Pressurized Furnace—Case 7, High-Btu Fuel

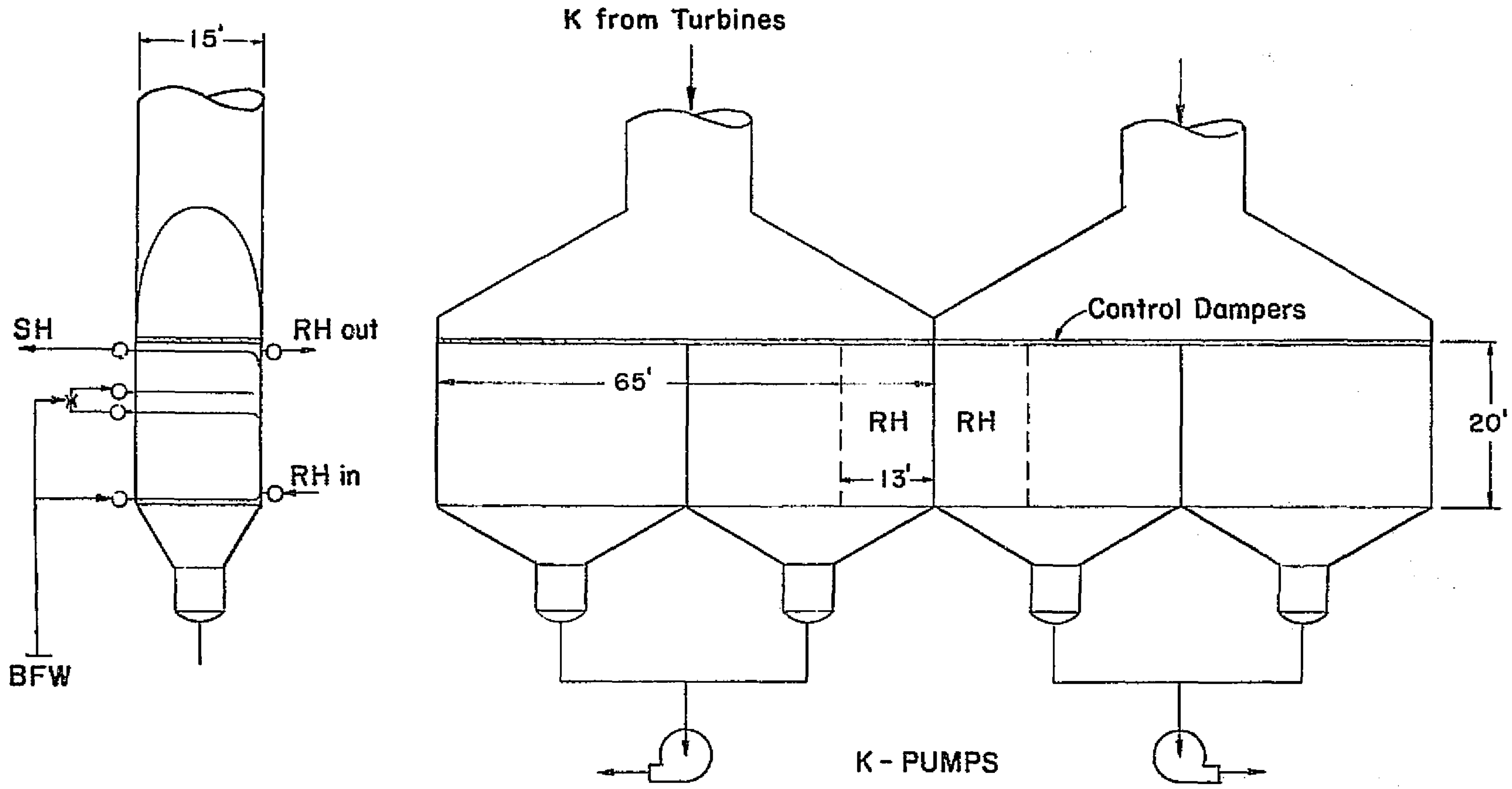


Figure A-16. Liquid-Metal Topping Cycle in Potassium Condenser—Case 1

Table A-5

BASIC FURNACE SIZING AND SURFACING CONSIDERATIONS
LIQUID-METAL MHD

	AFB Coal Tower	PFB Coal Tower	PF lo Btu	PF Hi Btu	IHX + WHB	Hi Temp APH LP
Surfacing						
Cycle Point	1	10	6	9	1	1
Gas - Hot Fluid Flow (10^6#/hr)	1.749		1.859	1.173		
Sectional Area (ft^2)	7x360 = 2520	7x90.3 = 632	113	113		
Furnace Velocity (ft/sec)	9.21	3.54	28.9	22.8		
Gas T_1/T_2 (°F)	1550/730	1650/1600	2200/1800	2200/1360	1278/197	730/252
Working Fluid t_1/t_2 (°F)	1300/610	1300/610	1300/610	1300/610	1000/117	655/86
LMTD - ΔT) Avg. (°F)	(192)	(402)	(828)	(468)	159	115
Uo (Btu/hr-ft^2 °F)	(35)	(50)	(60)	(50)	60	5e
q"o (Btu/hr-ft^2)	6716	20103	49690	23388	9540	575
So (10^6-ft^2)	.2293	.0484	.0129	.0366	.1086	.371
Q_o (10^6- Btu/hr)	1540	973	641	856	1036	213.5
Surface Wt. (tons/10^6-ft^2)	5530	2913	6822	6776	3614	-
Tube size OD (in)	1¼" + 1-3/4"	1" + 1-3/4"	2½"	2½"	1¼"	Req.
Weights						
Pressure Part Wt.(tons)	1522	169	105	297	393	-
Casing, etc. Wt. (tons)	4.43	2.93	2.45	2.46	.606	
Total Wt. (tons)	2434	544	306	653	871	452
Costs						
Unit cost \$ x 10^6	15.962	3.045	1.888	4.560	4.558	.782
Cost rate \$/ton	6558	5598	6170	6983	5233	1730
Cost rate \$/ft^2	69.61	62.91	146.36	124.60	42.25	2.11
Cost rate \$/10^6 Btu/hr	10.365	3130	2945	5327	4429	3663
Pressure Part Composition						
Wt. % MO-RE 2						
Wt. % Hastelloy X						
Wt. % Inconel 601			2	3		
Wt. % TP 347H	52	80	78	80		
Wt. % TP 304H	4	2	2	2	20	
Wt. % T-22	38	14	16	12	60	
Wt. % T-2	6	4	2	3	-	
Wt. % CS					20	100

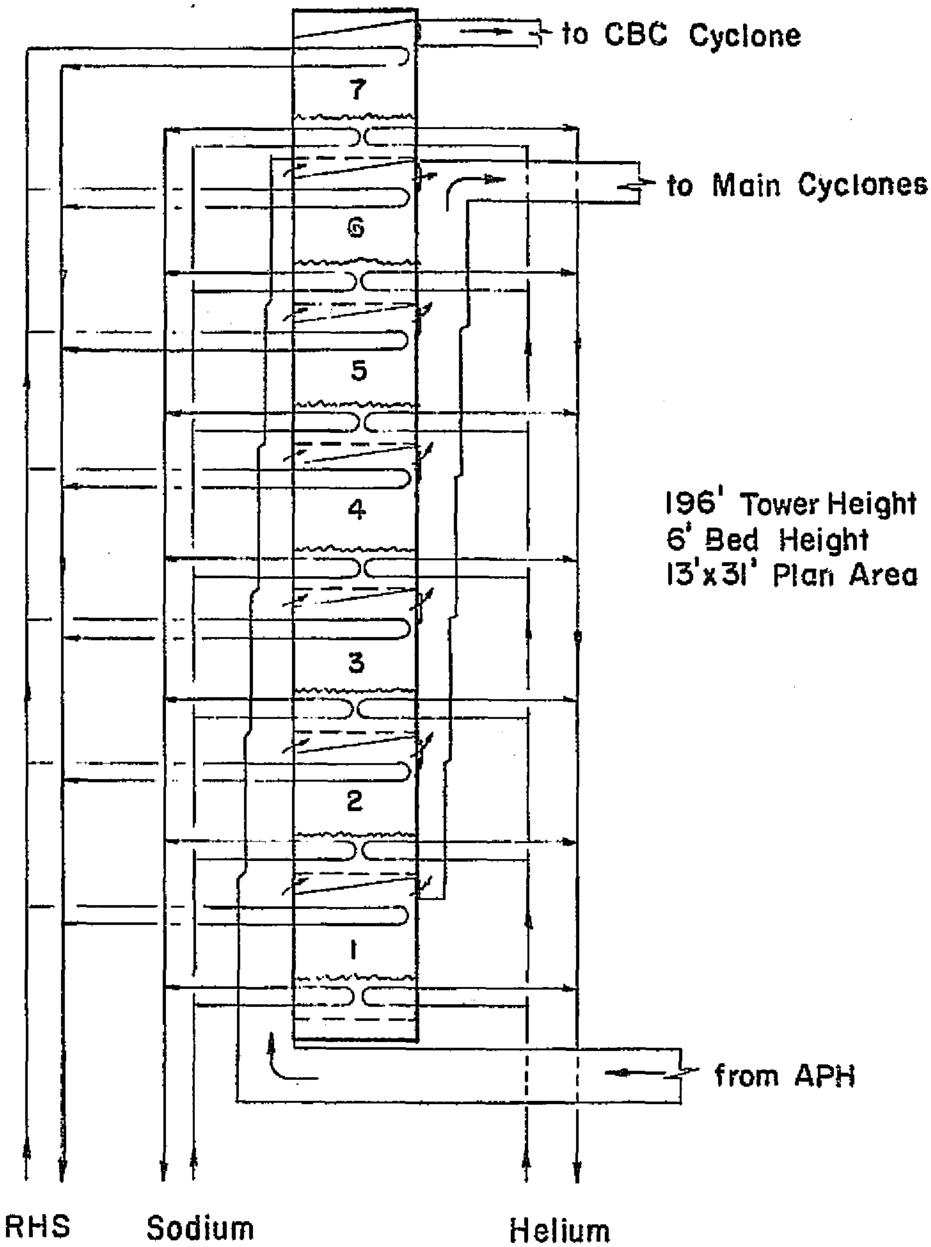


Figure A-17. Liquid-Metal MHD Cycle in Atmospheric Fluid Bed Tower—Case 1

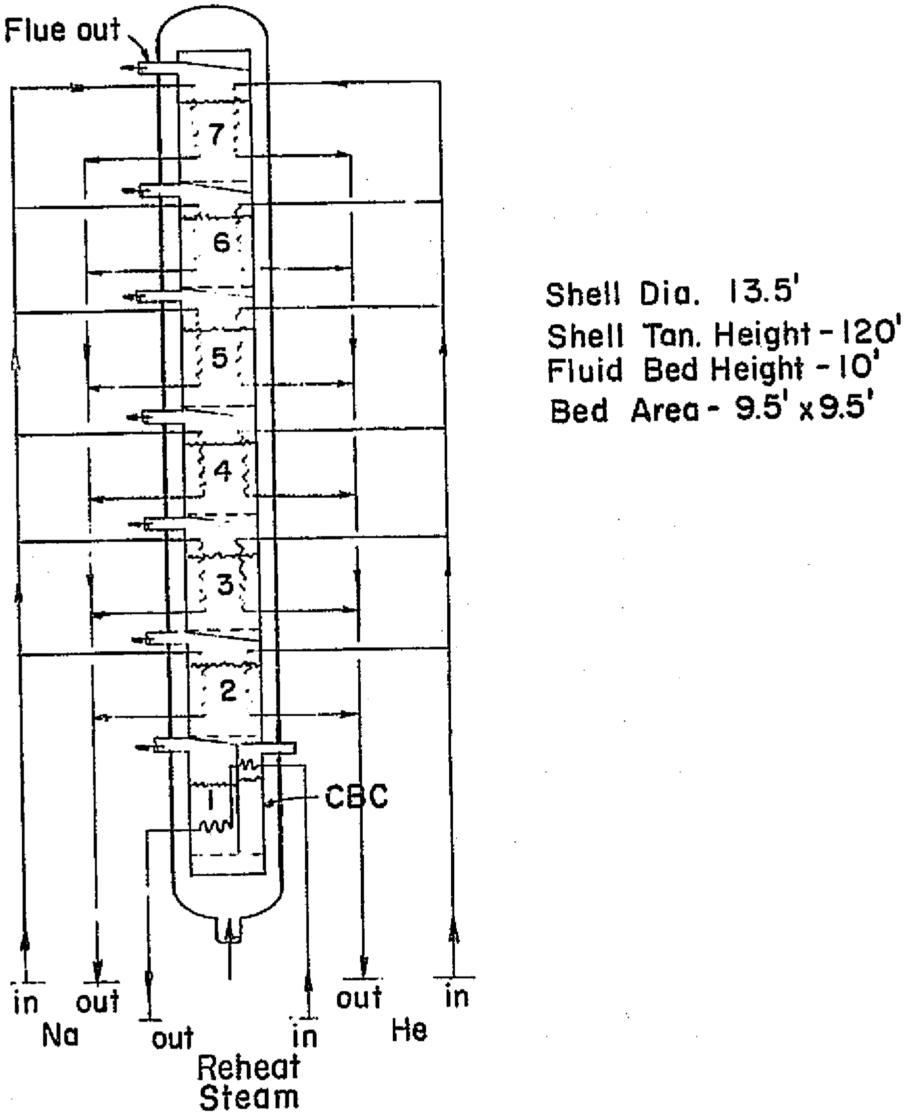


Figure A-18. Liquid-Metal MHD Cycle in Pressurized Fluid Bed—Case 10

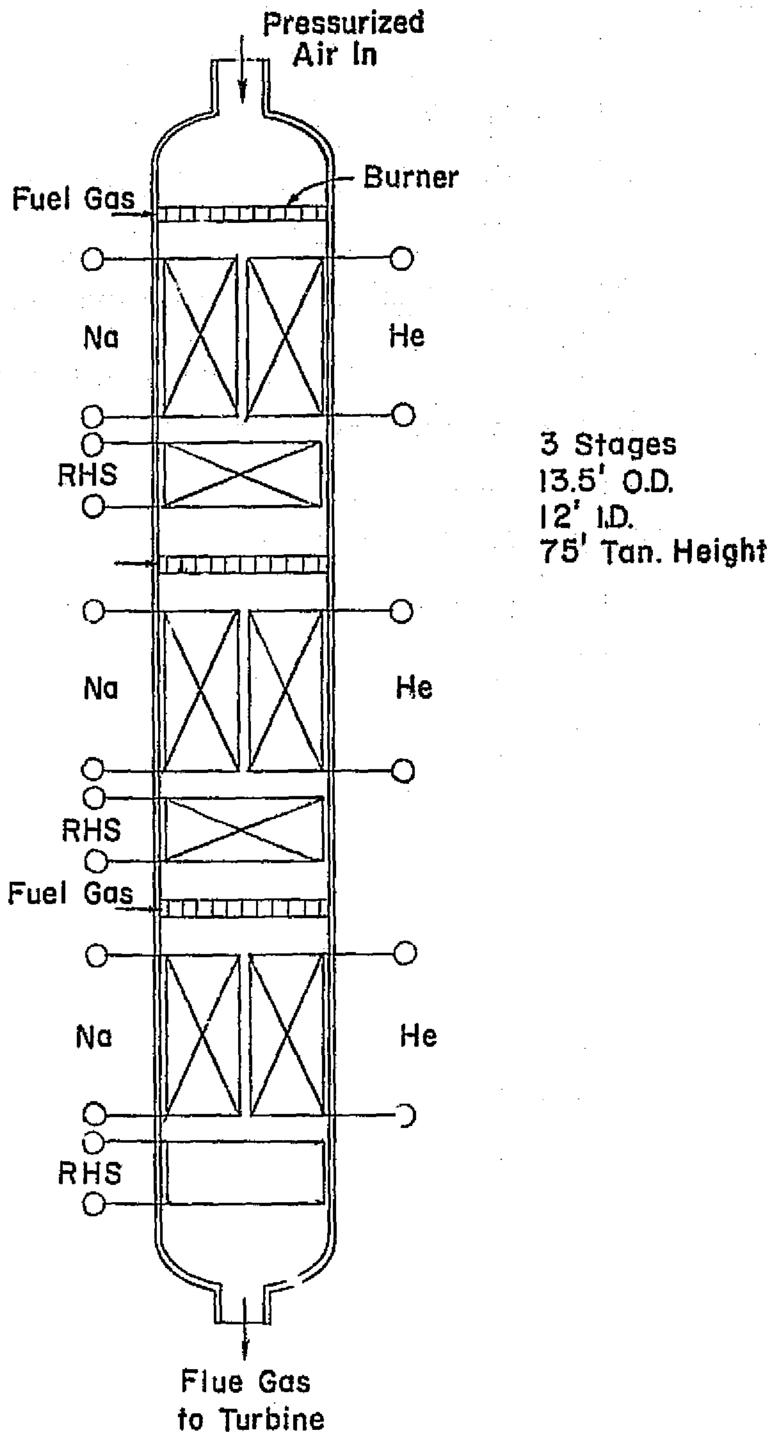


Figure A-19. Liquid-Metal MHD Cycle in Pressurized Furnace—Case 6, Low-Btu Fuel

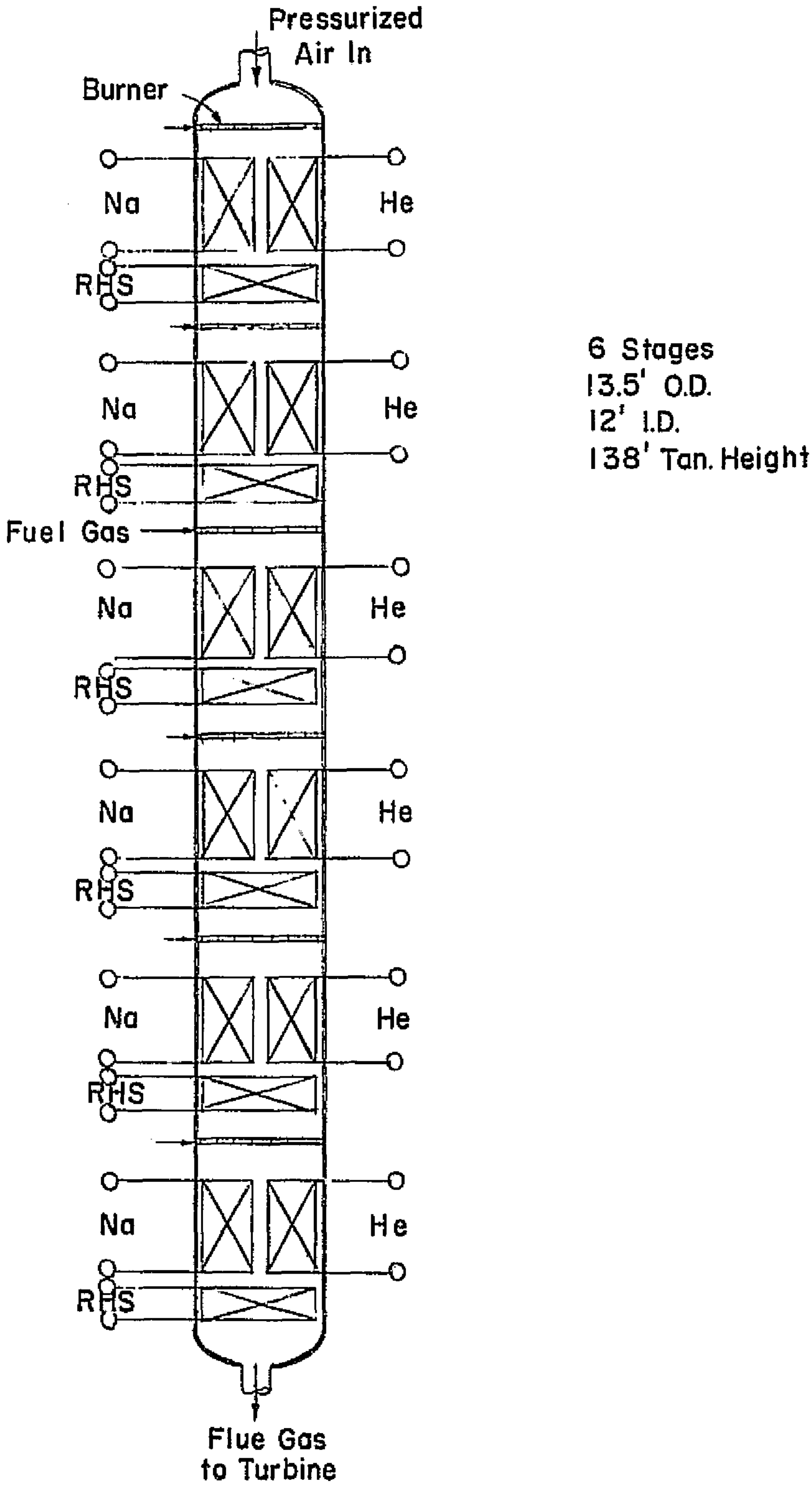


Figure A-20. Liquid-Metal MHD Cycle in Pressurized Furnace—Case 9, High-Btu Fuel

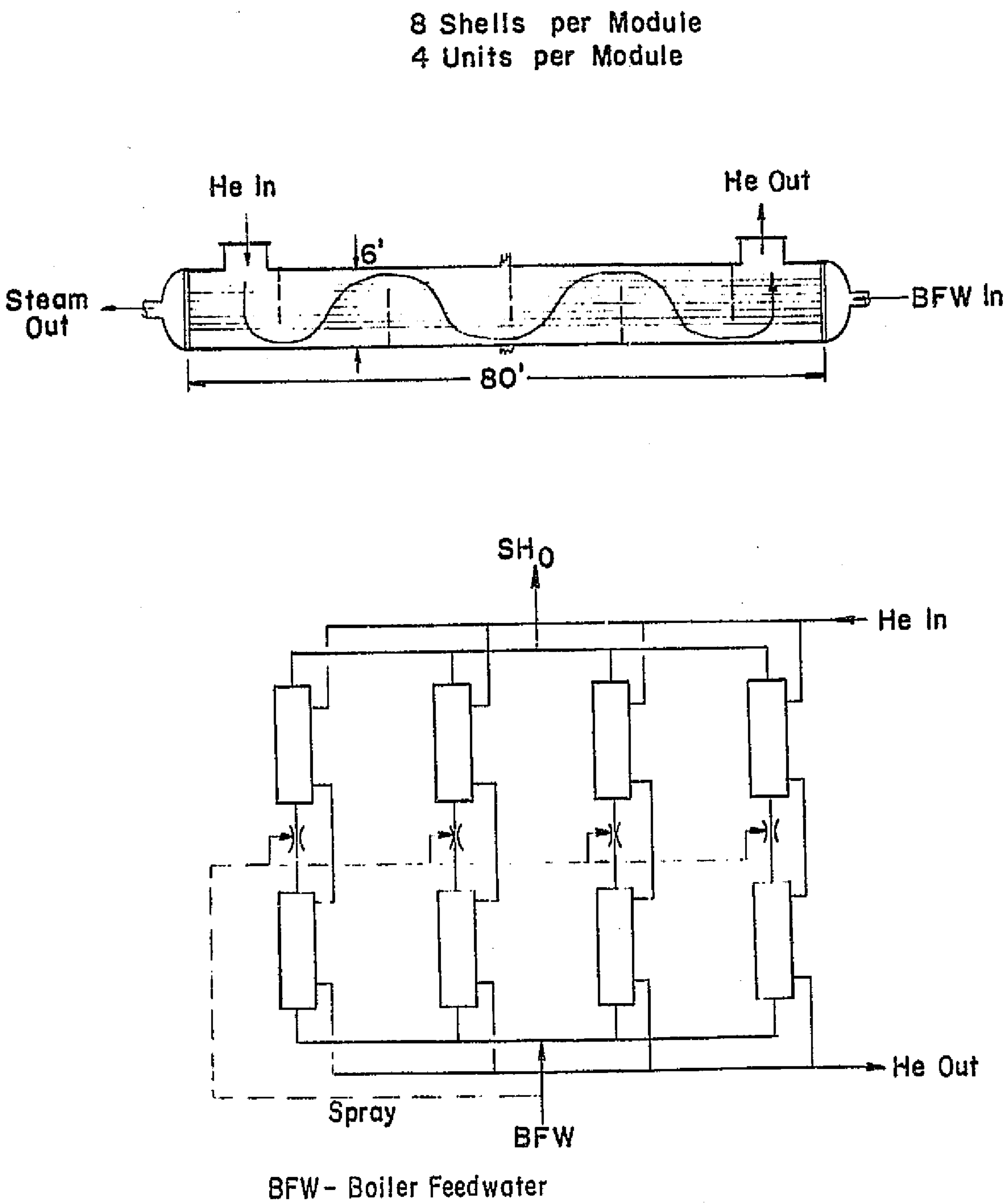


Figure A-21. Liquid-Metal MHD Cycle in Heat Recovery Boilers—Case 1

Table A-6

BASIC FURNACE SIZING AND SURFACING CONSIDERATIONS
CLOSED-CYCLE INERT GAS MHD

	MHD-CC Coal	MHD-CC SRC	IHX +	WHB	Hi Temp APH HP	
Surfacing						
Cycle Point	16	1	1	16	1	
Gas - Hot Fluid Flow (10^6#/hr)	11.407	6.621	-	11.407		
Sectional Area (ft^2)	9556	5027	1964	4600	-	
Furnace Velocity (ft/sec)	20.7	22.87	30	40		
Gas T_1/T_2 (°F)	3200	3200	1800/262	1930/540	972/344	
Working Fluid t_1/t_2 (°F)	-	-	1000/510	1000/510	897/125	
LMTD - ΔT) Avg. (°F)	-	-	238	262	122	
Uo (Btu/hr-ft^2 °F)	-	-	25	16	7.5	
q"o (Btu/hr-ft^2)	-	-	5945	4177	918	
So (10^6-ft^2)	-	-	.494	1.117	1.225	
Q_o (10^6- Btu/hr)	9906	4713	2937	4666	1124	
Surface Wt. (tons/10^6-ft^2)	-	-	3461	3461	1941	
Tube size OD (in)	-	-	1¼" + 1-3/4"	1¼" + 1-3/4"	2"	
Weights						
Pressure Part Wt.(tons)	-	-	1966	3864	2854	
Casing, etc. Wt. (tons)	3266	2215	257	102	476	
Total Wt. (tons)	3266	2215	2223	3966	2854	
Costs						
Unit cost $ x 10^6	15.247	10.341	9.692	18.182	8.358	
Cost rate $/ton	4699	4669	4360	4585	2929	
Cost rate $/$ft^2$	-	-	19.62	16.28	6.82	
Cost rate $/$10^6$ Btu/hr	1539	2194	3300	3900	7436	
Pressure Part Composition						
Wt. % MO-RE 2						
Wt. % Hastelloy X						
Wt. % Inconel 601						
Wt. % TP 347H	83%	83%				
Wt. % TP 304H	Refractory	Refractory	15	20		
Wt. % T-22			35	40	20	
Wt. % T-2			25	20	20	
Wt. % CS	17	17	25	20	60	

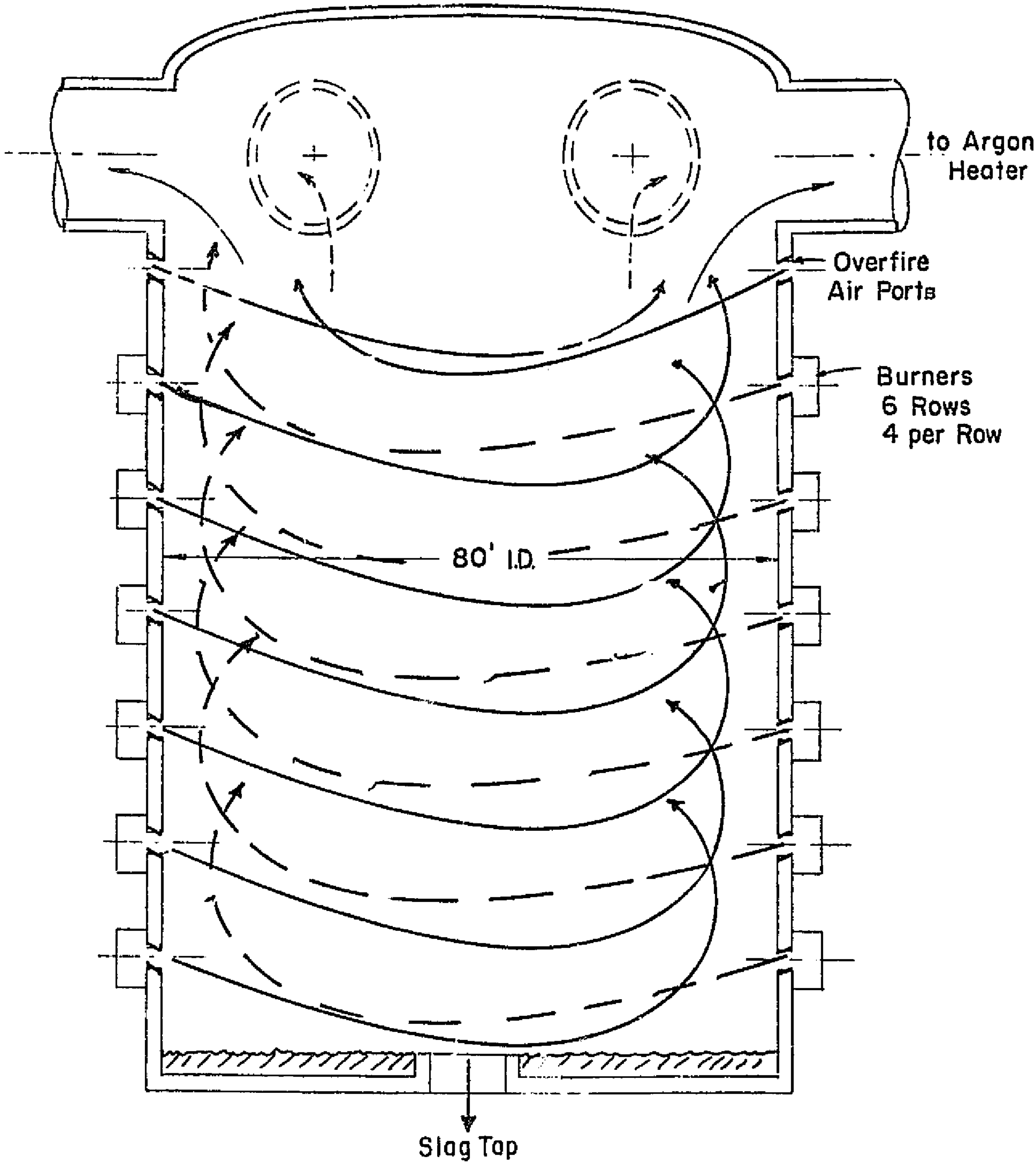


Figure A-22. Closed-Cycle Inert Gas MHD in Combustion Chamber

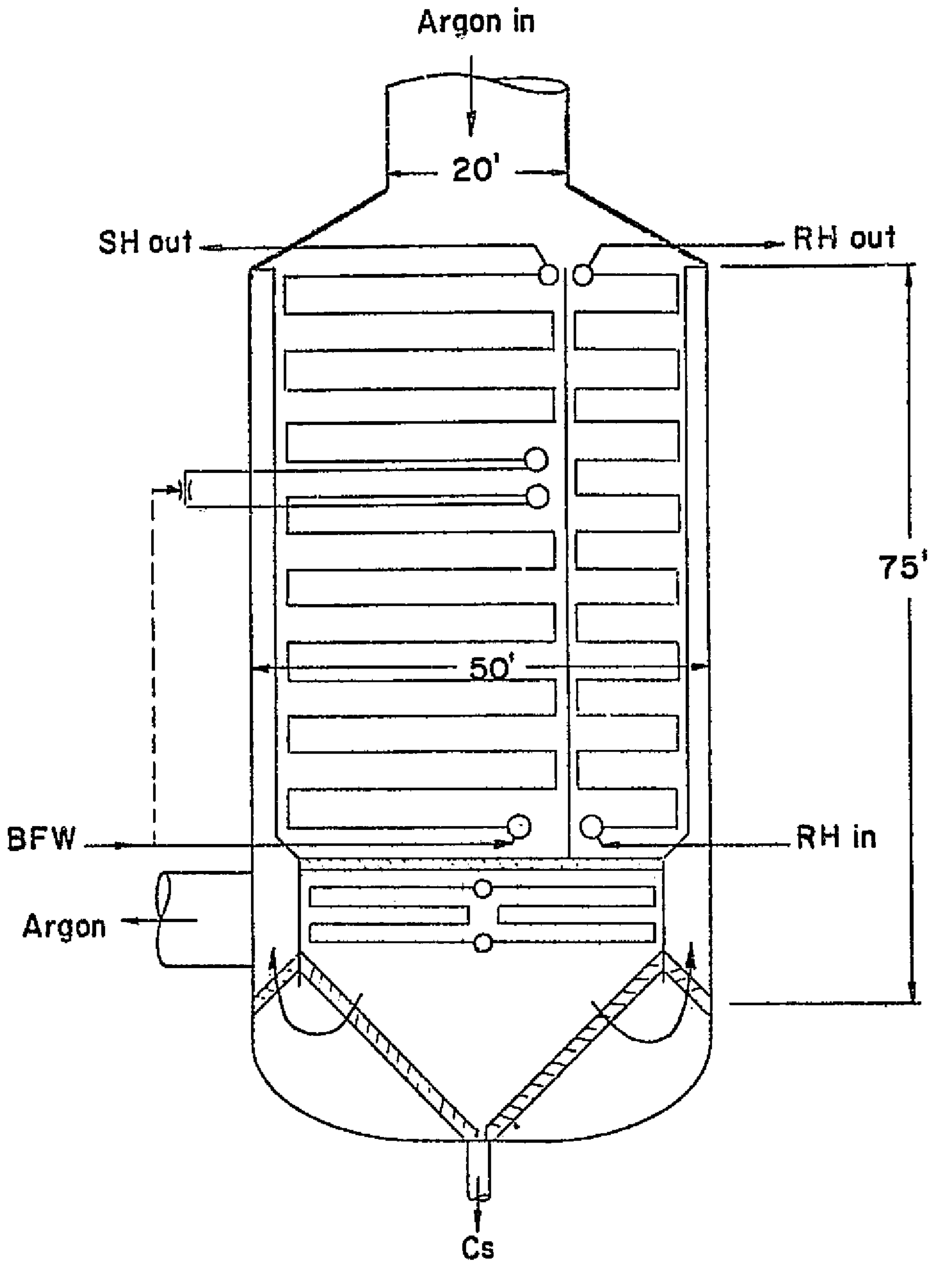


Figure A-23. Closed-Cycle Inert Gas MHD in Argon Heat Recovery Boiler

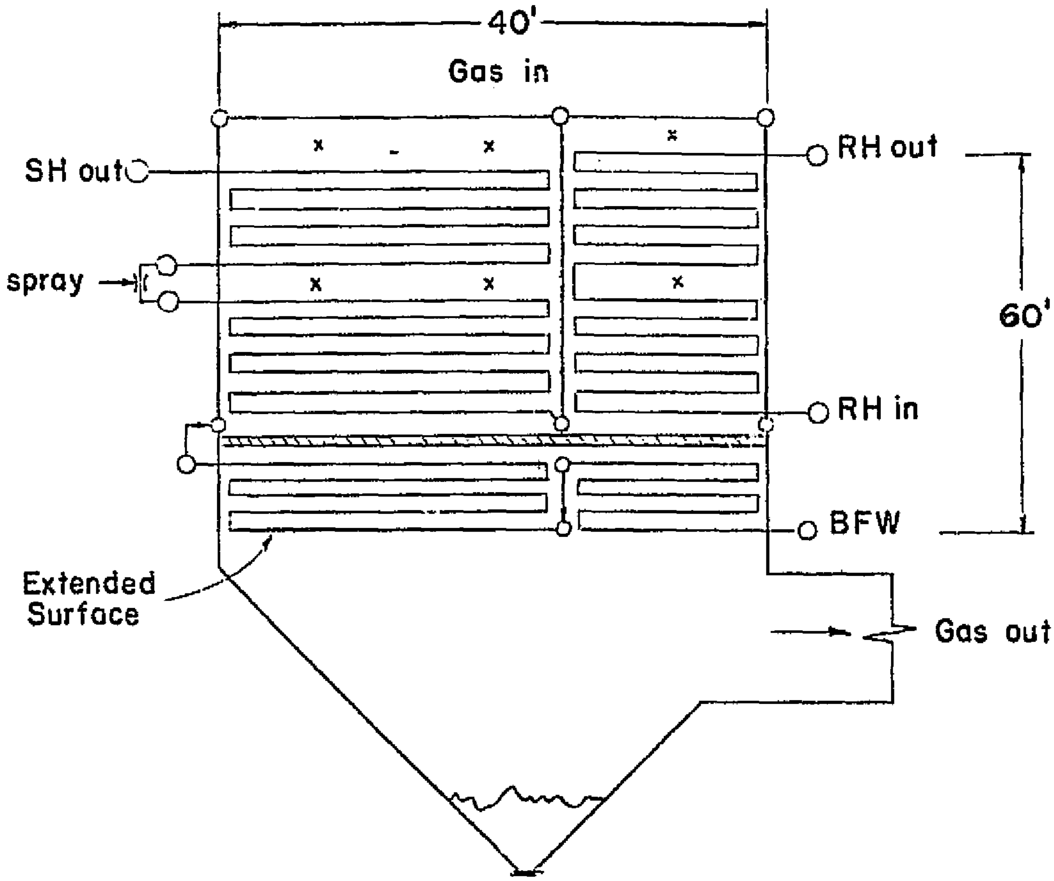


Figure A-24. Closed-Cycle Inert Gas MHD in Gas Heat Recovery Boiler

Table A-7

BASIC FURNACE SIZING AND SURFACING CONSIDERATIONS
OPEN-CYCLE MHD

	RAD. FURN.	CONV. FURN.	ECON.	LO - T APH		
Surfacing						
Cycle Point	1	1	1	1		
Gas - Hot Fluid Flow (10^6#/hr)	10.8	6.97	10.8	3.83		
Sectional Area (ft^2)	5040	3096	2616	2368		
Furnace Velocity (ft/sec)	58.2	39.9	31.8	28.4		
Gas T_1/T_2 (°F)	3650/2950	2250/825	700/300	2250/700		
Working Fluid t_1/t_2 (°F)	775/545	1000/600	417/232	1400/621		
LMTD - ΔT) Avg. (°F)	2633	373	151	325		
Uo (Btu/hr-ft^2 °F)	17	19.6	12	10		
q"o (Btu/hr-ft^2)	44764	7316	1812	3246		
So (10^6-ft^2)	.0594(flat)	.446	.565	.552		
Q_o (10^6- Btu/hr)	2659	3263	1024	1792		
Surface Wt. (tons/10^6-ft^2)	7475	10.141	1689	1919		
Tube size OD (in)	$1\frac{1}{4}$"	$1\frac{1}{4}$ - $2\frac{1}{2}$"	2" Ken Tube	2"		
Weights						
Pressure Part Wt.(tons)	466	4749	1049	1171		
Casing, etc. Wt. (tons)	23	13	48	478		
Total Wt. (tons)	489	4762	1097	1649		
Costs						
Unit cost \$ x 10^6	3.146	22.728	1.645	12.256		
Cost rate \$/ton	6434	4773	1450	7432		
Cost rate \$/ft^2	52.96	50.96	2.91	22.21		
Cost rate \$/10^6 Btu/hr	1183	6965	1606	6839		
Pressure Part Composition						
Service (Wt.)						
Fluid Temperature						
Wt. % MO-RE 2						
Wt. % Hastelloy X						
Wt. % Inconel 601						
Wt. % TP 347H						
Wt. % TP 304H	7	20		38		
Wt. % T-22	21	62		20		
Wt. % T-2	57	18		42		
Wt. % CS	15	-	100	-		

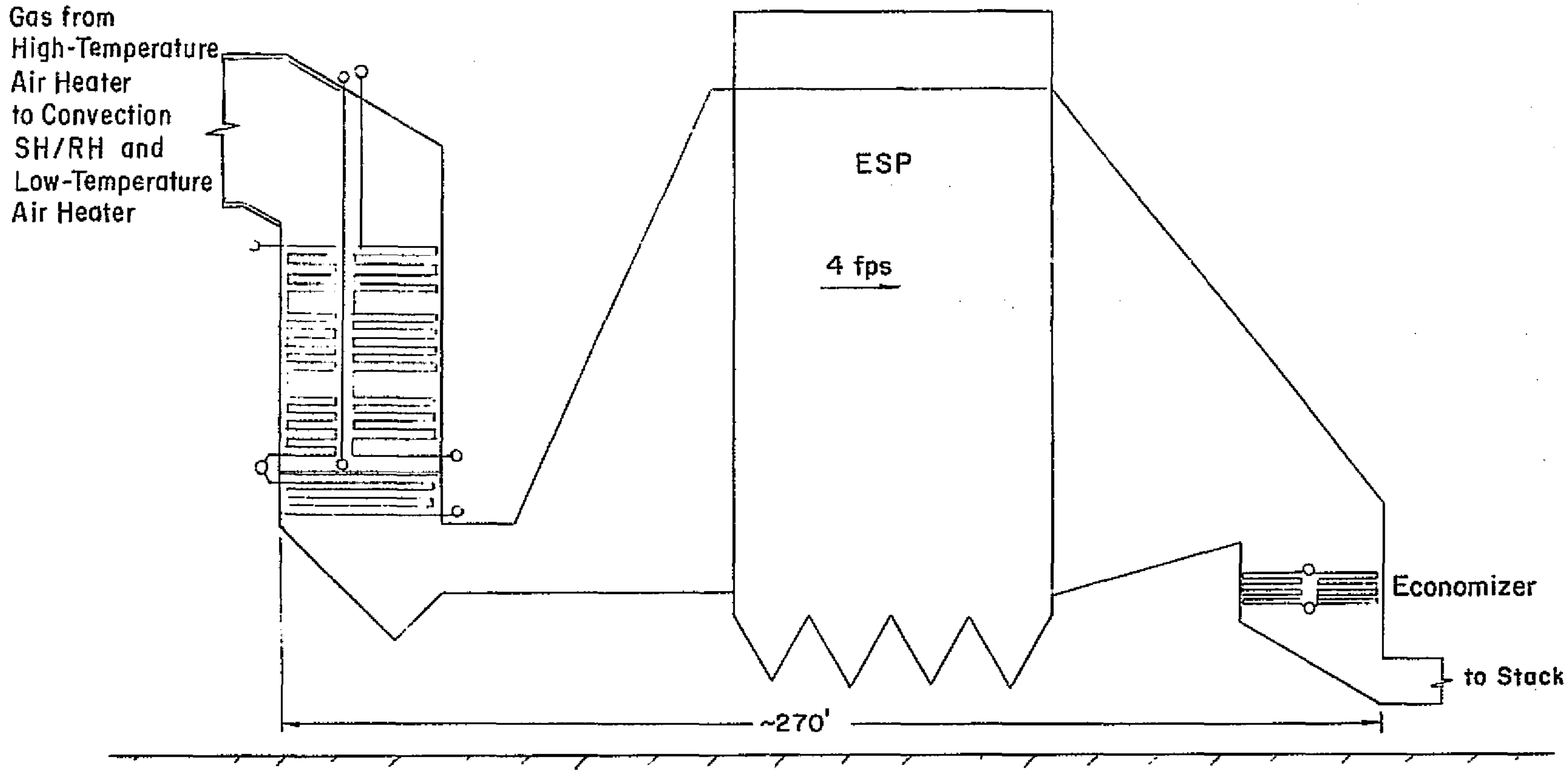


Figure A-25. Open-Cycle MHD in Convection Boiler and Economizer Arrangement

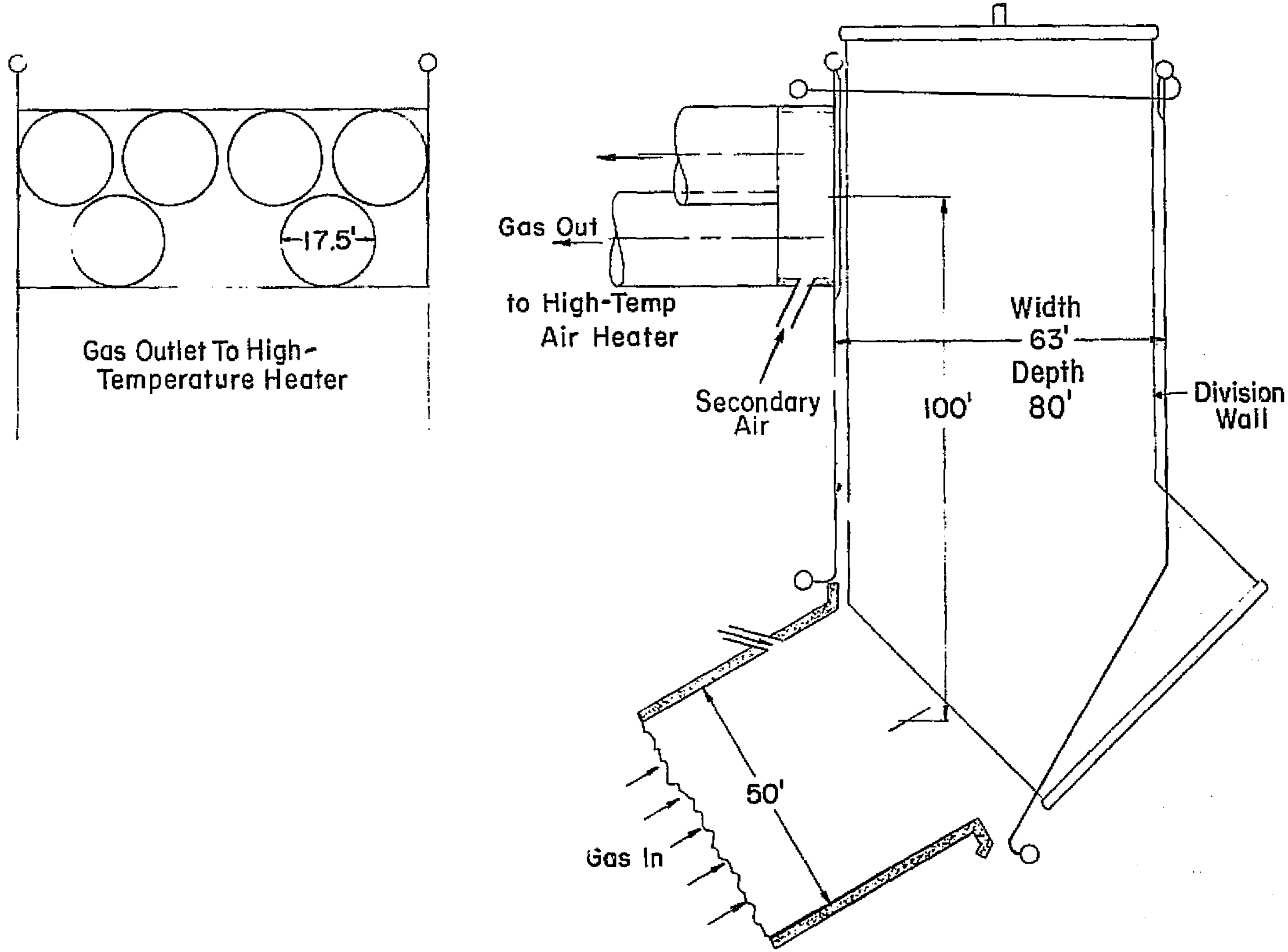


Figure A-26. Open-Cycle MHD in Radiant Furnace

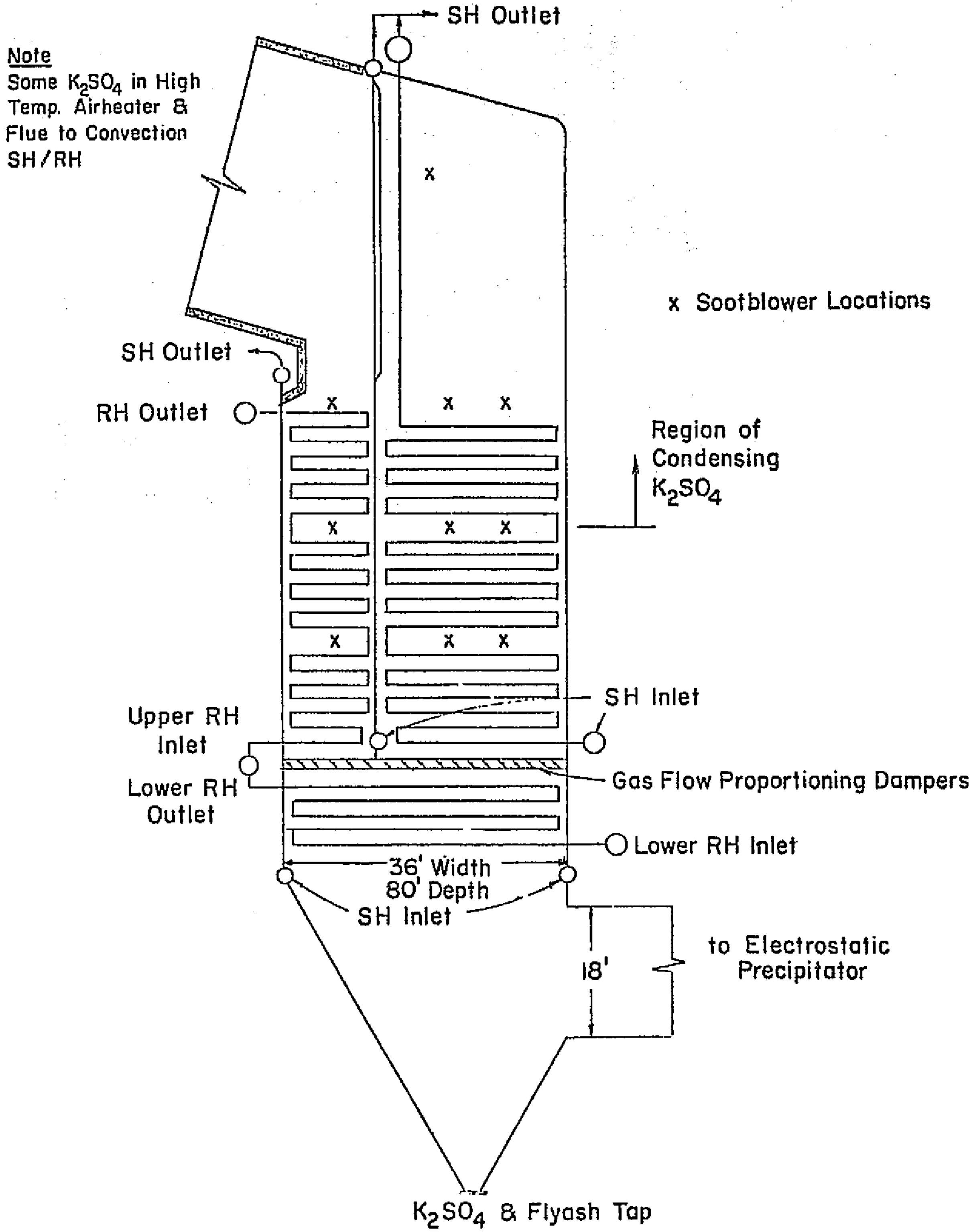
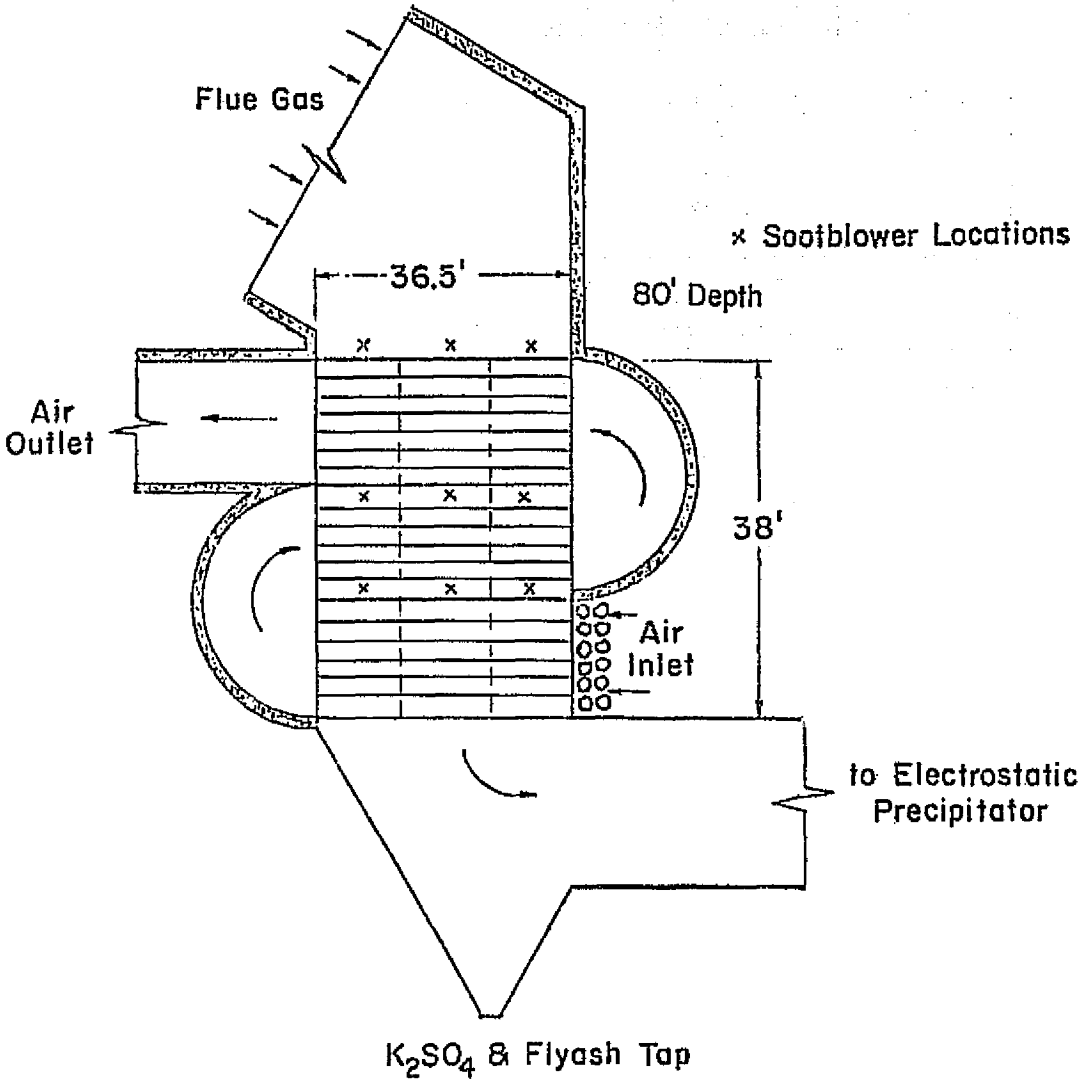


Figure A-27. Open-Cycle MHD in Convection Superheater-Reheater



<u>Note</u> Some K_2SO_4 will Condense in the High Temperature Air Heater and in the flue to the Low Temperature Air Heater.

Figure A-28. Open-Cycle MHD in Low-Temperature Air Heater

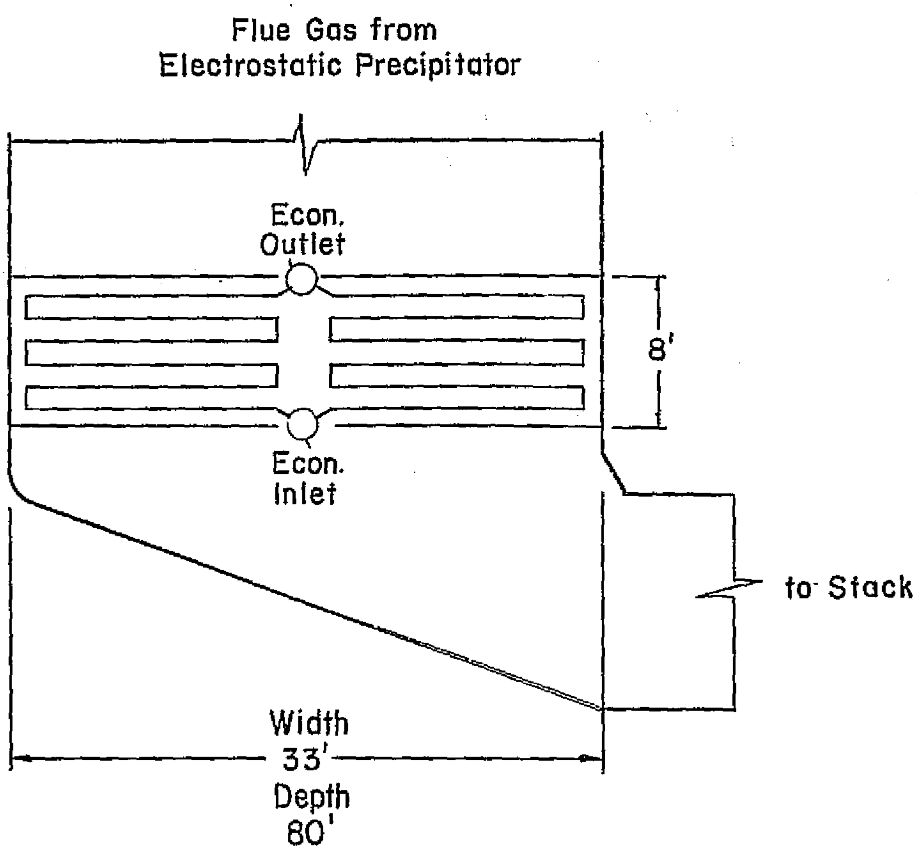


Figure A-29. Open-Cycle MHD with Economizer

Table A-8

BASIC FURNACE SIZING AND SURFACING CONSIDERATIONS
HIGH-TEMPERATURE FUEL CELLS

	CF GAS					
Surfacing						
Cycle Point	1					
Gas - Hot Fluid Flow (10^6#/hr)	9.482					
Sectional Area (ft^2)	80x77 = 6160					
Furnace Velocity (ft/sec)	27.8					
Gas T_1/T_2 (°F)	2033/540					
Working Fluid t_1/t_2 (°F)	1000/510					
LMTD - ΔT) Avg. (°F)	231					
Uo (Btu/hr-ft^2 °F)	18.7					
q"o (Btu/hr-ft^2)	4314					
So (10^6-ft^2)	.9705					
Q_o (10^6- Btu/hr)	4187					
Surface Wt. (tons/10^6-ft^2)	3294					
Tube size OD (in)	1¼" - 2¼"					
Weights						
Pressure Part Wt.(tons)	3677					
Casing, etc. Wt. (tons)	537					
Total Wt. (tons)	4214					
Costs						
Unit cost \$ x 10^6	16.230					
Cost rate \$/ton	3852					
Cost rate \$/ft^2	16.72					
Cost rate \$/10^6 Btu/hr	3876					
Pressure Part Composition						
Service (Wt.)						
Fluid Temperature						
Wt. % MO-RE 2						
Wt. % Hastelloy X						
Wt. % Inconel 601						
Wt. % TP 347H						
Wt. % TP 304H	6					
Wt. % T-22	37					
Wt. % T-2	37					
Wt. % CS	20					

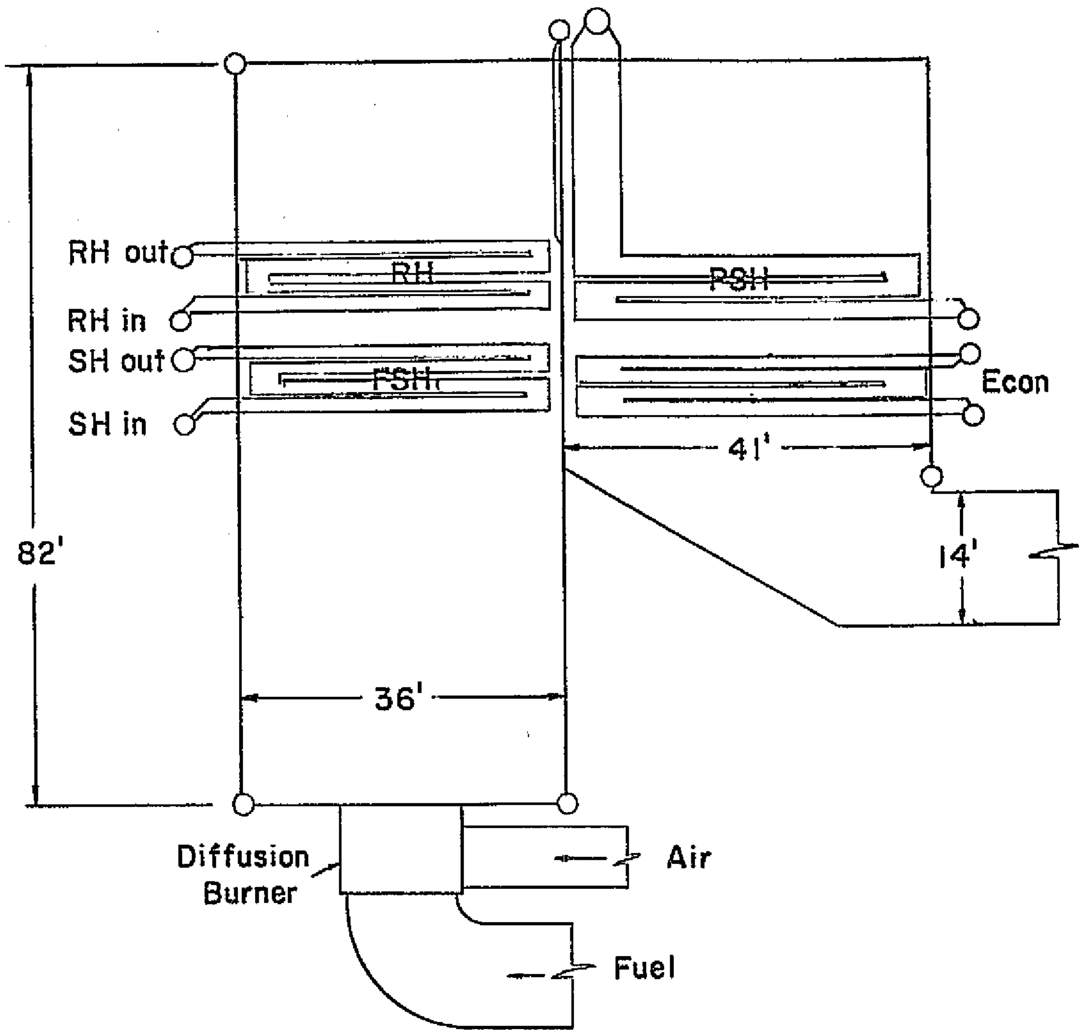


Figure A-30. High-Temperature Fuel Cell in Steam Generator

APPENDIX B

EQUIPMENT PERFORMANCE DATA FOR PRIMARY HEAT INPUT SYSTEMS

The technical data for each energy conversion system studied, are supplied on a modular basis. The final number of modules utilized for each parametric case are given in Volume II in the section describing the specific energy conversion system. The parametric point designation is also found in Volume II.

LIST OF TABLES

Table B-1 (Page 1 of 5)

ADVANCED STEAM CYCLE

Furnace Type	AFB →					
Fuel Code	I6 →					
Case Number	1	2	3	4	5	6
Output per Module, 10^6 Btu/Hr	1540	1540	1540	1540	1507	1481
Heat Exchanger Efficiency, Q_{out}/Q_{in}	0.8878	0.8878	0.8878	0.8878	0.8878	0.8878
Number of Modules	4	3	6	8	4	4
Fuel Input, 10^6 lb/Hr	0.1608				0.1573	0.1547
Combustion Air Input, 10^6 Lb/Hr	1.6003				1.5659	1.5393
Combustion Gas Flow, 10^6 Lb/Hr	1.7485				1.7109	1.6819
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0404				0.0395	0.0389
Total Solids Output 10^6 Lb/Hr	.0529				0.0529	0.0509
Percent Solids as Tap or Bottom Ash	53				53	53
SO_2 at Exit, Lb/10^6 Btu input	1.08					
NO_2 at Exit, Lb/10^6 Btu input	.27					
C O at Exit, Lb/10^6 Btu input	.21					
Solids at inlet to Final removal equip. lb/10^6 Btu	2.07					
Final particulate removal eff. reg'd. t meet 0.1 lb/10^6 Btu solids emissions	0.9517					
Fan Power Requirements, MWe	5.161				5.05	4.965
Other Auxiliary Equip. Power, MWe	3.050				2.984	2.934
Heat xchanger Width, Ft.	12	12	12	12	12	12
Heat Exch. Depth or O.D., Ft.	30	30	30	30	29.4	28.9
Heat Exchanger Height, Ft.	150	150	150	150	150	150
No. Cells or Combustion Stages/Module	7	7	7	7	7	7
Heat exchanger weight, Tons/Module	1200	1200	1200	1200	1211	1162

Table B-1 (Page 2 of 5)

ADVANCED STEAM CYCLE

Furnace Type	AFB →					
Fuel Code	I6 →					
Case Number	7	8	9	10	11	12
Output per Module, 10^6 Btu/Hr	1421	1531	1540+	1649	1606	1566
Heat Exchanger Efficiency, Q_{out}/Q_{in}	0.8878	0.8878	0.8878	0.8878	0.8878	0.8878
Number of Modules	4	4	by GE	4	4	4
Fuel Input, 10^6 lb/Hr	0.1483	0.1598	0.1608	0.1722	0.1677	0.1635
Combustion Air Input, 10^6 Lb/Hr	1.4764	1.5908	1.6003	1.7024	1.6688	1.6275
Combustion Gas Flow, 10^6 Lb/Hr	1.6132	1.7382	1.7485	1.8600	1.8233	1.7782
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0373	.0402	.0404	.0430	.0421	.0410
Total Solids Output 10^6 Lb/Hr	.0488	.0526	.0529	.0563	.0551	.0538
Percent Solids as Tap or Bottom Ash	53	53	53	53	53	53
SO_2 at Exit, Lb/10^6 Btu input	1.08	1.08	1.08			
NO_2 at Exit, Lb/10^6 Btu input	.27	.27	.27			
C O at Exit, Lb/10^6 Btu input	.21	.21	.21			
Solids at inlet to Final removal equip. lb/10^6 Btu	2.07	2.07	2.07			
Final particulate removal eff. req'd. to meet 0.1 lb/10^6 Btu solids emissions	0.9517	0.9517	0.9517			
Fan Power Requirements, MWe	4.761	5.131	5.161	5.490	5.382	5.249
Other Auxiliary Equip. Power, MWe	2.814	3.032	3.050	4.453	3.181	3.102
Heat Exchanger Width, Ft.	12	12	12	12	12	12
Heat Exch. Depth or O.D., Ft.	27.7	29.8	30+	31.9	31.3	30.5
Heat Exchanger Height, Ft.	150	150	150	150	150	150
No. Cells or Combustion Stages/Module	7	7	7	7	7	7
Heat exchanger weight, Tons/Module	1184	1293	1200+	1091	1172	1137

Table B-1 (Page 3 of 5)

ADVANCED STEAM CYCLE

Furnace Type	AFB				CF	
Fuel Code	I6		ND	MS	I6	ND
Case Number	13	14	15	16	17	18
Output per Module, 10^6 Btu/Hr	1500	1471	1540	1540	6160	6160
Heat Exchanger Efficiency, Q_{out}/Q_{in}	0.8878	0.8878	0.8216	0.8637	0.8793	0.8314
Number of Modules	4	4	4	4	1	1
Fuel Input, 10^6 Lb/Hr	0.1566	0.1536	0.2721	0.1994	0.6554	0.0754
Combustion Air Input, 10^6 Lb/Hr	1.5592	1.5289	1.7213	1.6302	6.5220	6.8024
Combustion Gas Flow, 10^6 Lb/Hr	1.7036	1.6705	1.9764	1.8144	7.1144	7.8112
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0394	.0386	.0123	.0102	--	--
Total Solids Output 10^6 Lb/Hr	.0515	.0505	.0290	.0253	0.0629	0.0667
Percent Solids as Tap or Bottom Ash	53	53	25	24	25	25
SO_2 at Exit, Lb/10^6 Btu input	1.08		0.30	0.27	7.22	2.03
NO_2 at Exit, Lb/10^6 Btu input	.27		0.26	0.26	0.70	0.70
C O at Exit, Lb/10^6 Btu input	.21		.23	.22	--	--
Solids at inlet to Final removal equip. lb/10^6 Btu	2.07		1.47	1.39	8.90	9.00
Final particulate removal eff. req'd. to meet 0.1 lb/10^6 Btu solids emissions	0.9517		0.9322	0.9282	0.9850	0.9852
Fan Power Requirements, MWe	5.029	4.93	5.950	5.464	5.300 ID 5.500FD	6.200 ID 5.700FD
Other Auxiliary Equip. Power, MWe	2.972	2.914	3.281	3.229	8.069	11.710
Heat Exchanger Width, Ft.	12	12	12	12	74.5	81.7
Heat Exch. Depth or O.D., Ft.	29.3	28.7	33.8	31.1	46	46
Heat Exchanger Height, Ft.	150	150	150	150	185	185
No. Cells or Combustion Stages/Module	7	7	7	7	32	32
Heat exchanger weight, Tons/Module	1156	1075	1350	1245	6000	6464

Table B-1 (Page 4 of 5)

ADVANCED STEAM CYCLE

Furnace Type	CF →		PF →			PFB
Fuel Code	MS	SC	LI	LN	LM	LI
Case Number	19	20	21	22	23	24
Output per Module, 10^6 Btu/Hr	6160	6160	641	616	639	972
Heat Exchanger Efficiency, Q_{out}/Q_{in}	0.8739	0.9155	0.4264	0.4385	0.4430	0.6158
Number of Modules	1	1	4	4	4	4
Fuel Input, 10^6 lb/Hr	0.7881	0.4291	0.488	0.623	0.599	0.1464
Combustion Air Input, 10^6 Lb/Hr	6.4432	6.2132	1.371	1.368	1.399	1.4568
Combustion Gas Flow, 10^6 Lb/Hr	7.1721	6.6416	1.8590	1.9910	1.9980	1.6063
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	--	--	--	--	--	0.0655
Total Solids Output 10^6 Lb/Hr	0.0591	0.0004	--	--	--	0.0624
Percent Solids as Tap or Bottom Ash	25	5	--	--	--	51
SO_2 at Exit, Lb/10^6 Btu input	1.79	0.89	--	--	--	0.72
NO_2 at Exit, Lb/10^6 Btu input	0.70	0.30	0.20	0.20	0.20	0.14
C O at Exit, Lb/10^6 Btu input	--	--	--	--	--	0.11
Solids at inlet to Final removal equip. lb/10^6 Btu	8.39	0.06	--	--	--	1.09
Final particulate removal eff. req'd. t meet 0.1 lb/10^6 Btu solids emissions	0.9841	NONE	--	--	--	0.9475
Fan Power Requirements, MWe	5.400FD 5.700ID	8.240	–	–	–	–
Other Auxiliary Equip. Power, MWe	9.257	0.599	--	--	--	2.182
Heat Exchanger Width, Ft.	75.8	69.5	--	--	--	--
Heat Exch. Depth or O.D., Ft.	46	46	13.5(12)	13.5(12)	13.5(12)	13.5 (9.2 x 9.2)
Heat Exchanger Height, Ft.	185	185	42	42	42	84
No. Cells or Combustion Stages/Module	32	32	3	3	3	5
Heat exchanger weight, Tons/Module	6039	5676	220	212	219	420

ADVANCED STEAM CYCLE

Furnace Type	PFB				CF
Fuel Code	ND	MS	I6	I6	I6
Case Number	25	26	27[1]	27*	17A
Output per Module, 10^6 Btu/Hr	902	960	1056	972	6424
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.5447	.6024	.6689	.6689	.8793
Number of Modules	4	4	4	4	1
Fuel Input, 10^6 lb/Hr	0.2402	0.1781	0.1464	0.1348	.6939
Combustion Air Input, 10^6 Lb/Hr	1.5196	1.4561	1.4568	1.3049	6.8015
Combustion Gas Flow, 10^6 Lb/Hr	1.7496	1.6247	1.6063	1.4785	7.4193
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0193	.0164	.0655	.0603	-
Total Solids Output 10^6 Lb/Hr	.0296	.0258	.0624	.0574	.0656
Percent Solids as Tap or Bottom Ash	32	31	51	51	25
SO_2 at Exit, Lb/10^6 Btu input	0.20	0.18	0.72	0.72	7.22
NO_2 at Exit, Lb/10^6 Btu input	0.13	0.13	0.14	0.14	.7
C O at Exit, Lb/10^6 Btu input	0.11	0.11	0.11	0.11	-
Solids at inlet to Final removal equip. lb/10^6 Btu	0.81	0.74	1.09	1.09	8.9
Final particulate removal eff. req'd. to meet 0.1 lb/10^6 Btu solids emissions	0.8579	0.8648	0.9475	0.9475	.9850
Fan Power Requirements, MWe	–	–	–	–	5.735FD 5.527ID
Other Auxiliary Equip. Power, MWe	2.076	1.808	2.182	2.009	8.415
Heat Exchanger Width, Ft.	--	--	--		77.7
Heat Exch. Depth or O.D., Ft.	(9.8 13.5 x 9.8)	(9.3 13.5 x 9.3)	(9.2 13.5 x 9.2)	(8.5 13.5 x 8.5)	46
Heat Exchanger Height, Ft.	84*	84*	84*	84	185
No. Cells or Combustion Stages/Module	5	5	5*	5	1
Heat exchanger weight, Tons/Module	413	377	427	423	5400
				Same Output as case 24	Rev. Conv. Steam

Table B-2 (Page 1 of 6)

SUPERCRITICAL CO_2 CYCLE

Furnace Type	AFB				PFB	PF
Fuel Code	I6		MS	ND	I6	LI
Case Number	1	2	3	4	5	6
Output per Module, 10^6 Btu/Hr	1517	1517	1520	1480	1056.2	704.4
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8744	.8744	.8520	.7944	.6685	.4687
Number of Modules	2.813	5.626	2.851	3.121	4.103	6.081
Fuel Input, 10^6 lb/Hr	.1608	.1608	.1994	.2904	.1464	.6771
Combustion Air Input, 10^6 Lb/Hr	1.6003	1.6003	1.6302	1.7103	1.4568	.8985
Combustion Gas Flow, 10^6 Lb/Hr	1.7485	1.7485	1.8144	1.9637	1.6063	1.859
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0404	.0404	.01027	.01218	.06551	--
Total Solids Output 10^6 Lb/Hr	.0529	.0529	.0253	.02883	.0624	--
Percent Solids as Tap or Bottom Ash	.0283	.0283	.00614	.00728	.03184	--
SO_2 at Exit, Lb/10^6 Btu input	1.0845	1.0845	.2683	.3048	.7231	--
NO_2 at Exit, Lb/10^6 Btu input	.2741	.2741	.2645	.2575	.1371	.2
C O at Exit, Lb/10^6 Btu input	.2142	.2142	.2178	.2292	.1071	--
Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703	2.0703	1.3919	1.4745	1.0936	--
Final particulate removal eff. reg'd. t meet 0.1 lb/10^6 Btu solids emissions	.9517	.9517	.9282*	.9322	.9475	--
Fan Power Requirements, MWe	6.82	6.82	6.89	7.43	---	42.678
Other Auxiliary Equip. Power, MWe	3.050	3.050	3.229	3.261	2.182	--
Heat Exchanger Width, Ft.	13	13	13	13	-	-
Heat Exch. Depth or O.D., Ft.	31	31	32.1	34.8	13.5 (10x27)	13.5
Heat Exchanger Height, Ft. (Note 1)	140	140	140	140	128	88
No. Cells or Combustion Stages/Module (Note 1)	7	7	7	7	4	3
Heat exchanger weight, Tons/Module	1753	1753	1845	1902	733	678

Note 1: For all AFB units - increase height by 14' and weight by 33%.
Case 1 - 4, 11 - 32

Table B-2 (Page 2 of 6)

SUPERCRITICAL CO_2 CYCLE

Furnace Type	PF				AFB	
Fuel Code	LM	LN	HI	HI	I6	
Case Number	7	8	9	10	11	12
Output per Module, 10^6 Btu/Hr	698.2	672	990	827	1517	
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.4835	.4783	.7779	.5997	.8878	
Number of Modules	6.097	6.645	4.585	4.745	2.740	2.889
Fuel Input, 10^6 lb/Hr	.7093	.7243	.06071	.06071	.1608	
Combustion Air Input, 10^6 Lb/Hr	.8429	.7972	1.1065	1.1065	1.6003	
Combustion Gas Flow, 10^6 Lb/Hr	1.998	1.991	1.1731	1.1731	1.7485	
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	--	--	--	--	.0404	
Total Solids Output 10^6 Lb/Hr	--	--	--	--	.0529	
Percent Solids as Tap or Bottom Ash	--	--	--	--	.0283	
SO_2 at Exit, Lb/10^6 Btu input	--	--	--	--	1.0845	
NO_2 at Exit, Lb/10^6 Btu input	.2	.2	.2	.2	.2741	
C O at Exit, Lb/10^6 Btu input	--	--	--	--	.2142	
Solids at inlet to Final removal equip. lb/10^6 Btu	--	--	--	--	2.0703	
Final particulate removal eff. req'd. t meet 0.1 lb/10^6 Btu solids emissions	--	--	--	--	.9517	
Fan Power Requirements, MWe	40.033	37.867	45.744	38.08	6.82	
Other Auxiliary Equip. Power, MWe	--	--	--	--	3.050	
Heat Exchanger Width, Ft.	-	-	-	-	13	
Heat Exch. Depth or O.D., Ft.	13.5				31	
Heat Exchanger Height, Ft.	87	86	140	130	140	
No. Cells or Combustion Stages/Module	3	3	6	6	7	
Heat exchanger weight, Tons/Module	571	559	1060	1041	1773	1727

Table B-2 (Page 3 of 6)

SUPERCRITICAL CO_2 CYCLE

Furnace Type	AFB					
Fuel Code	I6					
Case Number	13	14	15	16	17	18
Output per Module, 10^6 Btu/Hr	1517					
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8878					
Number of Modules	2.792	2.919	3.030	3.144	2.811	2.822
Fuel Input, 10^6 lb/Hr	.1608					
Combustion Air Input, 10^6 Lb/Hr	1.6003					
Combustion Gas Flow, 10^6 Lb/Hr	1.7485					
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0404					
Total Solids Output 10^6 Lb/Hr	.0529					
Percent Solids as Tap or Bottom Ash	.0283					
SO_2 at Exit, Lb/10^6 Btu input	1.0845					
NO_2 at Exit, Lb/10^6 Btu input	.2741					
C O at Exit, Lb/10^6 Btu input	.2142					
Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703					
Final particulate removal eff. reg'd. t meet 0.1 lb/10^6 Btu solids emissions	.9517					
Fan Power Requirements, MWe	6.82					
Other Auxiliary Equip. Power, MWe	3.050					
Heat Exchanger Width, Ft.	13					
Heat Exch. Depth or O.D., Ft.	31					
Heat Exchanger Height, Ft.	140					
No. Cells or Combustion Stages/Module	7					
Heat exchanger weight, Tons/Module	1832	1667	1588	1515	1532	1643

Table B-2 (Page 4 of 6)

SUPERCRITICAL CO_2 CYCLE

Furnace Type	AFB →					
Fuel Code	16 →					
Case Number	19	20	21	22	23	24
Output per Module, 10^6 Btu/Hr	1517					
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8876					
Number of Modules	2.813	2.800	2.734	3.112	2.813	2.813
Fuel Input, 10^6 lb/Hr	.1608					
Combustion Air Input, 10^6 Lb/Hr	1.6003					
Combustion Gas Flow, 10^6 Lb/Hr	1.7485					
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0104					
Total Solids Output 10^6 Lb/Hr	.0529					
Percent Solids as Tap or Bottom Ash	.0283					
SO_2 at Exit, Lb/10^6 Btu input	1.0845					
NO_2 at Exit, Lb/10^6 Btu input	.2741					
C O at Exit, Lb/10^6 Btu input	.2142					
Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703					
Final particulate removal eff. reg'd. t meet 0.1 lb/10^6 Btu solids emissions	.9517					
Fan Power Requirements, MWe	6.82					
Other Auxiliary Equip. Power, MWe	3.050					
Heat Exchanger Width, Ft.	13					
Heat Exch. Depth or O.D., Ft.	31					
Heat Exchanger Height, Ft.	140					
No. Cells or Combustion Stages/Module	7					
Heat exchanger weight, Tons/Module	1753	1753	1946	1839	1753	1943

Table B-2 (Page 5 of 6)

SUPERCRITICAL CO_2 CYCLE

Furnace Type	AFB					
Fuel Code	I6					
Case Number	25	26	27	28	29	30
Output per Module, 10^6 Btu/Hr	1517					
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8878					
Number of Modules	2.816	2.848	2.845	2.860	2.833	2.856
Fuel Input, 10^6 lb/Hr	.1608					
Combustion Air Input, 10^6 Lb/Hr	1.6003					
Combustion Gas Flow, 10^6 Lb/Hr	1.7485					
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0404					
Total Solids Output 10^6 Lb/Hr	.0529					
Percent Solids as Tap or Bottom Ash	.0283					
SO_2 at Exit, Lb/10^6 Btu input	1.0845					
NO_2 at Exit, Lb/10^6 Btu input	.2741					
C O at Exit, Lb/10^6 Btu input	.2142					
Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703					
Final particulate removal eff. reg'd. t meet 0.1 lb/10^6 Btu solids emissions	.9517					
Fan Power Requirements, MWe	6.82					
Other Auxiliary Equip. Power, MWe	3.050					
Heat Exchanger Width, Ft.	13					
Heat Exch. Depth or O.D., Ft.	31					
Heat Exchanger Height, Ft.	140					
No. Cells or Combustion Stages/Module	7					
Heat exchanger weight, Tons/Module	1643	1753	1679	1679	1753	1759

Table B-2 (Page 6 of 6)

SUPERCRITICAL CO_2 CYCLE

Furnace Type	AFB →				
Fuel Code	T6 →				
Case Number	31	32			
Output per Module, 10^6 Btu/Hr	1517	1517			
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8878	.8878			
Number of Modules	2.818	2.831			
Fuel Input, 10^6 lb/Hr	.1608	.1608			
Combustion Air Input, 10^6 Lb/Hr	1.6003	1.6003			
Combustion Gas Flow, 10^6 Lb/Hr	1.7485	1.7485			
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0104	.0104			
Total Solids Output 10^6 Lb/Hr	.0529	.0529			
Percent Solids as Tap or Bottom Ash	.0283	.0283			
SO_2 at Exit, Lb/10^6 Btu input	1.0845	1.0845			
NO_2 at Exit, Lb/10^6 Btu input	.2741	.2741			
C O at Exit, Lb/10^6 Btu input	.2142	.2142			
Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703	2.0703			
Final particulate removal eff. reg'd. t meet 0.1 lb/10^6 Btu solids emissions	.9517	.9517			
Fan Power Requirements, MWe	6.82	6.82			
Other Auxiliary Equip. Power, MWe	3.050	3.050			
Heat Exchanger Width, Ft.	13	13			
Heat Exch. Depth or O.D., Ft.	31	31			
Heat Exchanger Height, Ft.	140	140			
No. Cells or Combustion Stages/Module	7	7			
Heat exchanger weight, Tons/Module	1753	1759			

Table B-3 (Page 1 of 8)

CLOSED-CYCLE HELIUM GAS TURBINE

Furnace Type	AFB			PF-LO		
Fuel Code	I6	ND	MS	LI	LM	LN
Case Number	1	2	3	4	5	6
Output per Module, 10^6 Btu/Hr	1520	1561	1519	641	640	616
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8762	.8379	.8517	.4264	.4434	.4384
Number of Modules	1.840	1.792	1.841	4.364	4.374	4.541
Fuel Input, 10^6 lb/Hr	.1608	.2704	.1994	.6771	.7093	.7243
Combustion Air Input, 10^6 Lb/Hr	1.6003	1.7106	1.6302	.8985	.8429	.7472
Combustion Gas Flow, 10^6 Lb/Hr	1.7485	1.964	1.8144	1.8590	1.9980	1.9910
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0404	.01218	.0102	--	--	--
Total Solids Output 10^6 Lb/Hr	.0529	.02884	.02526	--	--	--
Percent Solids as Tap or Bottom Ash	54	25	25	--	--	--
SO_2 at Exit, Lb/10^6 Btu input	1.0845	.3048	.2683	--	--	--
NO_2 at Exit, Lb/10^6 Btu input	.2741	.2575	.2645	.2	.2	.2
C O at Exit, Lb/10^6 Btu input	.2142	.2292	.2178	--	--	--
Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703	1.4745	1.3919	--	--	--
Final particulate removal eff. req'd. to meet 0.1 lb/10^6 Btu solids emissions	.9517	.9322	.9282	--	--	--
Fan Power Requirements, MWe	6.82	7.43	6.89	0	0	0
Other Auxiliary Equip. Power, MWe	3.050	3.261	3.229	--	--	--
Heat Exchanger Width, Ft.	13	13	13	--	--	--
Heat Exch. Depth or O.D., Ft.	31	35	32	13.5(12)	13.5(12)	13.5(12)
Heat Exchanger Height, Ft.	198	198	198	100	100	96
No. Cells or Combustion Stages/Module	9	9	9	3	3	3
Heat exchanger weight, Tons/Module	2162	2161	2195	514	512	496

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CLOSED-CYCLE HELIUM GAS TURBINE

Furnace Type	PF-HI	PFB	AFB →			PF-HI
Fuel Code	HI	I6	I6 →			HI
Case Number	7	8	9	10	11	12
Output per Module, 10^6 Btu/Hr	1010	1056	1520	1520	1528	990
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.7247	.6686	.8762	.8762	.8808	.7103
Number of Modules	2.769	2.640	2.761	3.681	1.628	3.225
Fuel Input, 10^6 lb/Hr	.06138	.1464	.1608			.06138
Combustion Air Input, 10^6 Lb/Hr	1.1117	1.4568	1.6003			1.1117
Combustion Gas Flow, 10^6 Lb/Hr	1.1731	1.6063	1.7485			1.1731
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	--	.06551	.0404			--
Total Solids Output 10^6 Lb/Hr	--	.06238	.0529			--
Percent Solids as Tap or Bottom Ash	--	51	54			--
SO_2 at Exit, Lb/10^6 Btu input	--	.7231	1.0845			--
NO_2 at Exit, Lb/10^6 Btu input	.2	.1371	.2741			.2
C O at Exit, Lb/10^6 Btu input	--	.1071	.2142			--
Solids at inlet to Final removal equip. lb/10^6 Btu	--	1.0936	2.0703			--
Final particulate removal eff. req'd. t meet 0.1 lb/10^6 Btu solids emissions	--	.9475	.9517			--
Fan Power Requirements, MWe	0	--	6.820	6.820	6.820	0
Other Auxiliary Equip. Power, MWe	--	2.182	3.050	3.050	3.050	--
Heat Exchanger Width, Ft.	--	2 barrels	13	13	13	--
Heat Exch. Depth or O.D., Ft.	13.5(12)	13.5 x 2(10x31)	31	31	31	13.5(12)
Heat Exchanger Height, Ft.	132	144	198	198	154	212
No. Cells or Combustion Stages/Module	6	4	9	9	7	6
Heat exchanger weight, Tons/Module	735	677	2162	2162	1619	1129

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CLOSED-CYCLE HELIUM GAS TURBINE

Furnace Type	AFB					
Fuel Code	I6					
Case Number	13	14	15	16	17	18
Output per Module, 10^6 Btu/Hr	1520	1538	1520	1520	1520	1520
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8762	.8866	.8762	.8762	.8762	.8762
Number of Modules	1.438	.2135	1.816	1.791	1.840	1.819
Fuel Input, 10^6 lb/Hr	.1608					
Combustion Air Input, 10^6 Lb/Hr	1.6003					
Combustion Gas Flow, 10^6 Lb/Hr	1.7485					
Stone input, 10^6 Lb/Hr (Limestone or Dolomite)	.0404					
Total Solids Output 10^6 Lb/Hr	.0529					
Percent Solids as Tap or Bottom Ash	54					
SO_2 at Exit, Lb/10^6 Btu input	1.0845					
NO_2 at Exit, Lb/10^6 Btu input	.2741					
C O at Exit, Lb/10^6 Btu input	.2142					
Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703					
Final particulate removal eff. reg'd. t meet 0.1 lb/10^6 Btu solids emissions	.9517					
Fan Power Requirements, MWe	6.820					
Other Auxiliary Equip. Power, MWe	3.050					
Heat Exchanger Width, Ft.	13					
Heat Exch. Depth or O.D., Ft.	31					
Heat Exchanger Height, Ft.	286	242	264	264	198	264
No. Cells or Combustion Stages/Module	13	11	12	12	9	12
Heat exchanger weight, Tons/Module	2820	2322	2459	2443	2162	2463

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CLOSED-CYCLE HELIUM GAS TURBINE

Furnace Type	AFB					PF-HI
Fuel Code	I6					HI
Case Number	19	20	21	22	23	24
Output per Module, 10^6 Btu/Hr	1520	1540	1540	1520	1520	990
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8762	.8878	.8878	.8762	.8762	.7103
Number of Modules	1.931	2.410	2.798	1.438	1.665	3.613
Fuel Input, 10^6 lb/Hr	.1608					.06138
Combustion Air Input, 10^6 Lb/Hr	1.6003					1.1117
Combustion Gas Flow, 10^6 Lb/Hr	1.7485					1.1731
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0404					--
Total Solids Output 10^6 Lb/Hr	.0529					--
Percent Solids as Tap or Bottom Ash	54					--
SO_2 at Exit, Lb/10^6 Btu input	1.0845					--
NO_2 at Exit, Lb/10^6 Btu input	.2741					.2
C O at Exit, Lb/10^6 Btu input	.2142					--
Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703					--
Final particulate removal eff. req'd. to meet 0.1 lb/10^6 Btu solids emissions	.9517					--
Fan Power Requirements, MWe	6.820					0
Other Auxiliary Equip. Power, MWe	3.050					--
Heat Exchanger Width, Ft.	13					--
Heat Exch. Depth or O.D., Ft.	31					13.5(12)
Heat Exchanger Height, Ft.	264	220	198	286	286	196
No. Cells or Combustion Stages/Module	12	10	9	13	13	6
Heat exchanger weight, Tons/Module	2435	2134	1754	2820	2576	1035

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CLOSED-CYCLE HELIUM GAS TURBINE

Furnace Type	PF-HI	AFB				
Fuel Code	HI	I6				
Case Number	25	26	27	28	29	30
Output per Module, 10^6 Btu/Hr	990	1520	1537	1520	1520	1520
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.7103	.8762	.8860	.8762	.8762	.8762
Number of Modules	2.848	2.835	1.797	1.438	1.438	1.438
Fuel Input, 10^6 lb/Hr	.06138	.1608				
Combustion Air Input, 10^6 Lb/Hr	1.1117	1.6003				
Combustion Gas Flow, 10^6 Lb/Hr	1.1731	1.7485				
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	--	.0404				
Total Solids Output 10^6 Lb/Hr	--	.0529				
Percent Solids as Tap or Bottom Ash	--	54				
SO_2 at Exit, Lb/10^6 Btu input	--	1.0845				
NO_2 at Exit, Lb/10^6 Btu input	.2	.2741				
C O at Exit, Lb/10^6 Btu input	--	.2142				
Solids at inlet to Final removal equip. lb/10^6 Btu	--	2.0703				
Final particulate removal eff. req'd. to meet 0.1 lb/10^6 Btu solids emissions	--	.9517				
Fan Power Requirements, MWe	0	6.820				
Other Auxiliary Equip. Power, MWe	--	3.050				
Heat Exchanger Width, Ft.	--	13				
Heat Exch. Depth or O.D., Ft.	13.5(12)	31				
Heat Exchanger Height, Ft.	238	154	154	308	308	308
No. Cells or Combustion Stages/Module	6	7	7	14	14	14
Heat exchanger weight, Tons/Module	1288	1599	1629	2690	2690	2690

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CLOSED-CYCLE HELIUM GAS TURBINE

Furnace Type	PF-HI →			AFB →		
Fuel Code	HI	HI	HI	I6		
Case Number	31	32	33	34	35	36
Output per Module, 10^6 Btu/Hr	990	990	990	1520	→	
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.7103	.7103	.7103	.8762	→	
Number of Modules	2.652	3.127	3.516	1.438	→	
Fuel Input, 10^6 lb/Hr	.06138	→		.1608	→	
Combustion Air input, 10^6 Lb/Hr	1.1117	→		1.6003	→	
Combustion Gas Flow, 10^6 Lb/Hr	1.1731	→		1.7485	→	
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	--	→		.0404	→	
Total Solids Output 10^6 Lb/Hr	--	→		.0529	→	
Percent Solids as Tap or Bottom Ash	--	→		54	→	
SO_2 at Exit, Lb/10^6 Btu input	--	→		1.0845	→	
NO_2 at Exit, Lb/10^6 Btu input	.2	→		.2741	→	
C O at Exit, Lb/10^6 Btu input	--	→		.2142	→	
Solids at inlet to Final removal equip. lb/10^6 Btu	--	→		2.0703	→	
Final particulate removal eff. reg'd. to meet 0.1 lb/10^6 Btu solids emissions	--	→		.9517	→	
Fan Power Requirements, MWe	0	0	0	6.820	→	
Other Auxiliary Equip. Power, MWe	--	--	--	3.050	→	
Heat Exchanger Width, Ft.	--	--	--	13	→	
Heat Exch. Depth or O.D., Ft.	13.5(12)	13.5(12)	13.5(12)	31	→	
Heat Exchanger Height, Ft.	254	223	201	286	→	
No. Cells or Combustion Stages/Module	6	6	6	13	→	
Heat exchanger weight, Tons/Module	1385	1196	1062	2820	→	

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CLOSED-CYCLE HELIUM GAS TURBINE

Furnace Type	AFB					
Fuel Code	I6					
Case Number	37	38	39	40	41	42
Output per Module, 10^6 Btu/Hr	1520			1537	1520	
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8762			.8860	.8762	
Number of Modules	1.438			2.267	1.438	
Fuel Input, 10^6 lb/Hr	.1608					
Combustion Air Input, 10^6 Lb/Hr	1.6003					
Combustion Gas Flow, 10^6 Lb/Hr	1.7485					
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0404					
Total Solids Output 10^6 Lb/Hr	.0529					
Percent Solids as Tap or Bottom Ash	54					
SO_2 at Exit, Lb/10^6 Btu input	1.0845					
NO_2 at Exit, Lb/10^6 Btu input	.2741					
C O at Exit, Lb/10^6 Btu input	.2142					
Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703					
Final particulate removal eff. req'd. t meet 0.1 lb/10^6 Btu solids emissions	.9517					
Fan Power Requirements, MWe	6.820					
Other Auxiliary Equip. Power, MWe	3.050					
Heat Exchanger Width, Ft.	13					
Heat Exch. Depth or O.D., Ft.	31					
Heat Exchanger Height, Ft.	286			220	308	286
No. Cells or Combustion Stages/Module	13			10	14	13
Heat exchanger weight, Tons/Module	2820			2275	2690	2820

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CLOSED-CYCLE HELIUM GAS TURBINE

Furnace Type	AFB			
Fuel Code	I6			
Case Number	43	44	45	46
Output per Module, 10^6 Btu/Hr	1520	1537	1520	1520
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8762	.8860	.8762	.8762
Number of Modules	1.438	2.267	1.438	3.371
Fuel Input, 10^6 lb/Hr	.1608			
Combustion Air Input, 10^6 Lb/Hr	1.6003			
Combustion Gas Flow, 10^6 Lb/Hr	1.7485			
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0404			
Total Solids Output 10^6 Lb/Hr	.0529			
Percent Solids as Tap or Bottom Ash	54			
SO_2 at Exit, Lb/10^6 Btu input	1.0845			
NO_2 at Exit, Lb/10^6 Btu input	.2741			
C O at Exit, Lb/10^6 Btu input	.2142			
Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703			
Final particulate removal eff. req'd. t meet 0.1 lb/10^6 Btu solids emissions	.9517			
Fan Power Requirements, MWe	6.820			
Other Auxiliary Equip. Power, MWe	3.050			
Heat Exchanger Width, Ft.	13			
Heat Exch. Depth or O.D., Ft.	31			
Heat Exchanger Height, Ft.	286	220	308	176
No. Cells or Combustion Stages/Module	13	10	14	8
Heat exchanger weight, Tons/Module	2820	2275	2690	1515

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LIQUID-METAL TOPPING CYCLE

Furnace Type	AFB			PF		
Fuel Code	I6	MS	ND	LI	IM	LN
Case Number	1	2	3	4	5	6
Output per Module, 10^6 Btu/Hr (Includes Pump Work)	1507	1481	1444	746	748	723
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8588	.8206	.7605	.6468	.5750	.5601
Number of Modules	5.501	5.599	5.748	8.072	8.056	8.342
Fuel Input, 10^6 lb/Hr	.1608	.1994	.2704	.4875	.5987	.6226
Combustion Air Input, 10^6 Lb/Hr	1.6003	1.6302	1.7106	1.3175	1.3993	1.3684
Combustion Gas Flow, 10^6 Lb/Hr	1.7485	1.8144	1.9641	1.8590	1.9980	1.9910
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0404	.0102	.0122	--	--	--
Total Solids Output 10^6 Lb/Hr	.0529	.0253	.0288	--	--	--
Percent Solids as Tap or Bottom Ash	53.5	24.3	25.3	--	--	--
SO_2 at Exit, Lb/10^6 Btu input	1.0845	.2683	.3048	--	--	--
NO_2 at Exit, Lb/10^6 Btu input	.2741	.2645	.2575	.2	.2	.2
C O at Exit, Lb/10^6 Btu input	.2142	.2178	.2292	--	--	--
Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703	1.3919	1.4745	--	--	--
Final particulate removal eff. reg'd. t meet 0.1 lb/10^6 Btu solids emissions	.9517	.9281	.9322	--	--	--
Fan Power Requirements, MWe	7.880	7.958	8.585	--	--	--
Recirculation Pump Power, MWe	5.450	5.500	5.75	2.65	2.600	2.75
Other Auxiliary Equip. Power, MWe	3.050	3.229	3.261	--	--	--
Heat Exchanger Width, Ft.	13 x 31	13 x 32	13 x 35	--	--	--
Heat Exch. Depth or O.D., Ft.	--	--	--	13.5(12)	→	→
Heat Exchanger Height, Ft.	330	330	330	91	92	90
No. Cells or Combustion Stages/Module	15	15	15	3	3	3
Heat exchanger weight, Tons/Module	2664	2642	2616	377	381	373

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LIQUID-METAL TOPPING CYCLE

Furnace Type	PF	AFB	PFB	PF	AFB	
Fuel Code	HI	I6	I6	HI	I6	
Case Number	7	8	9	10	11	12
Output per Module, 10^6 Btu/Hr	836	1507	1034	830	1504	1498
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.5878	.8588	.6479	.5878	.8611	.8536
Number of Modules	7.203	8.252	4.746	5.568	5.304	5.533
Fuel Input, 10^6 lb/Hr	.0621	.1608	.1464	.0555	.1608	
Combustion Air Input, 10^6 Lb/Hr	1.1110	1.6003	1.4568	.9922	1.6003	
Combustion Gas Flow, 10^6 Lb/Hr	1.1731	1.7485	1.6063	1.0477	1.7485	
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	--	.0404	.0655	--	.0404	
Total Solids Output 10^6 Lb/Hr	--	.0529	.0624	--	.0529	
Percent Solids as Tap or Bottom Ash	--	53.5	51.0	--	53.5	
SO_2 at Exit, Lb/10^6 Btu input	--	1.0845	.7231	--	1.0845	
NO_2 at Exit, Lb/10^6 Btu input	.2	.2741	.1371	.2	.2741	
C O at Exit, Lb/10^6 Btu input	--	.2142	.1071	--	.2142	
Solids at inlet to Final removal equip. lb/10^6 Btu	--	2.0703	1.0936	--	2.0703	
Final particulate removal eff. reg'd. t meet 0.1 lb/10^6 Btu solids emissions	--	.9517	.9475	--	.9517	
Fan Power Requirements, MWe	--	7.880	--	--	11.429	7.880
Recirculation Pump Power, MWe	2.802	5.450	3.650	.804	3.200	5.450
Other Auxiliary Equip. Power, MWe	--	3.050	2.182	--	3.050	3.050
Heat Exchanger Width, Ft.	--	13 x 31	--	--	13 x 31	13 x 31
Heat Exch. Depth or O.D., Ft.	13.5(12)	—	13.5 (9.5x9.5)	13.5(12)	—	—
Heat Exchanger Height, Ft.	161	330	183	195	510	330
No. Cells or Combustion Stages/Module	6	15	11	6	15	
Heat exchanger weight, Tons/Module	606	2664	953	812	5495	3475

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LIQUID-METAL TOPPING CYCLE

Furnace Type	AFB					
Fuel Code	I6					
Case Number	13	14	15	16	17	18
Output per Module, 10^6 Btu/Hr	1507	1507	1507	1507	1491	1491
Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8588	.8588	.8588	.8588	.8588	.8588
Number of Modules	5.501					
Fuel Input, 10^6 lb/Hr	.1608					
Combustion Air Input, 10^6 Lb/Hr	1.6003					
Combustion Gas Flow, 10^6 Lb/Hr	1.7485					
Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0404					
Total Solids Output 10^6 Lb/Hr	.0529					
Percent Solids as Tap or Bottom Ash	53.5					
SO_2 at Exit, Lb/10^6 Btu input	1.0845					
NO_2 at Exit, Lb/10^6 Btu input	.2741					
C O at Exit, Lb/10^6 Btu input	.2142					
Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703					
Final particulate removal eff. req'd. t meet 0.1 lb/10^6 Btu solids emissions	.9517					
Fan Power Requirements, MWe	7.880				7.880	7.880
Recirculation Pump Power, MWe	5.450				.300	.300
Other Auxiliary Equip. Power, MWe	3.050				3.050	3.050
Heat Exchanger Width, Ft.	13 x 31					
Heat Exch. Depth or O.D., Ft.	--	--	--	--	--	--
Heat Exchanger Height, Ft.	330					
No. Cells or Combustion Stages/Module	15					
Heat exchanger weight, Tons/Module	3501				3497	3501

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CLOSED-CYCLE LIQUID-METAL MHD

	Furnace Type	AFB →					PF
	Fuel Code	I6 →			MS	ND	LI
	Case Number	1	2	3	4	5	6
1	Output per Module, 10^6 Btu/Hr	1540			1540	1540	641.0
2	Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8878			.8635	.8266	.5619
3	Number of Modules	2.766	1.383	5.532	2.766	2.766	6.646
4	Fuel Input, 10^6 lb/Hr	.1608			.1994	.2704	.4875
5	Combustion Air Input, 10^6 Lb/Hr	1.6003			1.6302	1.7106	1.3175
6	Combustion Gas Flow, 10^6 Lb/Hr	1.7485			1.8144	1.9641	1.8590
7	Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	.0404			.0102	.0122	-
8	Total Solids Output 10^6 Lb/Hr	.0529			.0253	.0288	-
9	Percent Solids as Tap or Bottom Ash	53.5			24.3	25.3	-
10	SO_2 at Exit, Lb/10^6 Btu input	1.0845			.2683	.3048	-
11	NO_2 at Exit, Lb/10^6 Btu input	.2741			.2645	.2575	.2
12	C O at Exit, Lb/10^6 Btu input	.2142			.2178	.2292	-
13	Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703			1.3919	1.4745	-
14	Final particulate removal eff. req'd. to meet 0.1 lb/10^6 Btu solids emissions	.9517			.9281	.9322	-
15	Fan Power Requirements, MWe	6.82			6.89	7.43	-
16	Other Auxiliary Equip. Power, MWe	3.050			3.229	3.261	-
17	Heat Exchanger Width, Ft.	13	13	13	13	13	-
18	Heat Exch. Depth or O.D., Ft.	31	31	31	32	35	13.5(12)
19	Heat Exchanger Height, Ft.	196	196	196	196	196	75
20	No. Cells or Combustion Stages/Module	7	7	7	7	7	3
21	Heat exchanger weight, Tons/Module	2434	2434	2434	2434	2434	306

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CLOSED-CYCLE LIQUID-METAL MHD

	Furnace Type	PF			PFB	AFB	
	Fuel Code	LM	LN	HI	I6	B6:	
	Case Number	7	8	9	10	11	12
1	Output per Module, 10^6 Btu/Hr	639.5	616.1	855.6	972.5	1540	
2	Heat Exchanger Efficiency, Q_{out}/Q_{in}	.4973	.4834	.6084	.6158	.8878	
3	Number of Modules	6.662	6.915	4.980	4.381	2.687	2.786
4	Fuel Input, 10^6 lb/Hr	.5987	.6226	.0621	.1464	.1608	
5	Combustion Air Input, 10^6 Lb/Hr	1.3993	1.3684	1.1110	1.4568	1.6003	
6	Combustion Gas Flow, 10^6 Lb/Hr	1.9980	1.9910	1.1731	1.6063	1.7485	
7	Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)	–	–	–	.0655	.404	
8	Total Solids Output 10^6 Lb/Hr	–	–	–	.0624	.0529	
9	Percent Solids as Tap or Bottom Ash	–	–	–	51	53.5	
10	SO_2 at Exit, Lb/10^6 Btu input	–	–	–	.7231	1.0845	
11	NO_2 at Exit, Lb/10^6 Btu input	.2	.2	.2	.1371	.2741	
12	C O at Exit, Lb/10^6 Btu input	–	–	–	.1071	.2142	
13	Solids at inlet to Final removal equip. lb/10^6 Btu	–	–	–	1.0936	2.0703	
14	Final particulate removal eff. req'd. t meet 0.1 lb/10^6 Btu solids emissions	–	–	–	.9475	.9517	
15	Fan Power Requirements, MWe	–	–	–	–	6.82	
16	Other Auxiliary Equip. Power, MWe	–	–	–	2.182	3.050	
17	Heat Exchanger Width, Ft.	–	–	–	–	13	
18	Heat Exch. Depth or O.D., Ft.	13.5(12)	13.5(12)	13.5(12)	13.5 x (9.5) (9.5)	31	
19	Heat Exchanger Height, Ft.	75	72	138	120	273	1092
20	No. Cells or Combustion Stages/Module	3	3	6	7	7	14
21	Heat exchanger weight, Tons/Module	305	295	653	544	3843	11,216

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CLOSED-CYCLE LIQUID-METAL MHD

	Furnace Type	AFB				
	Fuel Code	I6				
	Case Number	13	14	15	16	17
1	Output per Module, 10^6 Btu/Hr	1540				
2	Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8878				
3	Number of Modules	2.802	2.629	2.766	2.766	3.956
4	Fuel Input, 10^6 lb/Hr	1.608				
5	Combustion Air Input, 10^6 Lb/Hr	1.6003				
6	Combustion Gas Flow, 10^6 Lb/Hr					
7	Stone Input, 10^6 Lb/Hr (Limestone or / Dolomite)	1.7485 .0404				
8	Total Solids Output 10^6 Lb/Hr	.0529				
9	Percent Solids as Tap or Bottom Ash	53.5				
10	SO_2 at Exit, Lb/10^6 Btu input	1.0845				
11	NO_2 at Exit, Lb/10^6 Btu input	.2741				
12	C O at Exit, Lb/10^6 Btu input	.2142				
13	Solids at inlet to Final removal equip. lb/10^6 Btu	2.0703				
14	Final particulate removal eff. reg'd. t meet 0.1 lb/10^6 Btu solids emissions	.9517				
15	Fan Power Requirements, MWe	6.82				
16	Other Auxiliary Equip. Power, MWe	3.050				
17	Heat Exchanger Width, Ft.	13				
18	Heat Exch. Depth or O.D., Ft.	31				
19	Heat Exchanger Height, Ft.	147	140	196	196	182
20	No. Cells or Combustion Stages/Module	7	7	7	7	7
21	Heat exchanger weight, Tons/Module	2044	1940	2434	2434	2154

Table B-6 (Page 1 of 4)

CLOSED-CYCLE INERT GAS MHD (Argon + Cs Mist)

	Series - Clean Fuels						
	Furnace Type	CH					
	Fuel Code	SC			II	IM	IN
	Case Number	1	2	3	4	5	6
1	Output per Module, 10^6 Btu/Hr	4713	9426	1277	5008	4931	4929
2	Heat Exchanger Efficiency, Q_{out}/Q_{in}	.9079	.9079	.9411	.8215	.8217	.8226
3	Number of Modules	.9194	.9193	.6421	.8652	.8787	.8791
4	Fuel Input, 10^6 lb/Hr	.3310	.6620	.0865	.8869	.9114	.9462
5	Combustion Air Input, 10^6 Lb/Hr (Excess Air %)	4.753 (20)	9.506 (20)	1.242 (20)	4.450 (15)	4.308 (15)	4.227 (15)
6	Combustion Gas Flow, lb/Hr (Gas Recirculated %)	5.084 (30.23)	10.168 (30.23)	1.329 (48.23)	5.337 (25.08)	5.219 (25.00)	5.223 (24.90)
7							
8	Total Solids Output 10^6 Lb/Hr	331	662	87	-	-	-
9	Percent Solids as Tap or Bottom Ash (%)	NILL					
10	SO_2 at Exit, Lb/10^6 Btu input	.8919			-	-	-
11	NO_2 at Exit, Lb/10^6 Btu input	.3	.3	.8	.2	.2	.2
12	C O at Exit, Lb/10^6 Btu input	-	-	-	-	-	-
13	Solids at inlet to Argon Heater lb/10^6 Btu	.0638			-	-	-
14	Final particulate removal eff. req'd to meet 0.1 lb/10^6 Btu solids emissions	-	-	-	-	-	-
15	Fan Power Requirements, MWe	38.892	77.785	12.795	37.592	36.514	36.096
16	Other Auxiliary Equip. Power, MWe	.47	.93	.12	-	-	-
17							
18	Combustor Dia., Ft.	82	2-82	48	86	86	86
19	Combustor Height, Ft.	130	130	30	130	130	130
20	Burners - Combustion Stages/combustor	24	24@	4	24		
21	Combustor weight, tons/module	2215	4430	407	2336	2336	2336
22	Intermediate HX size WxDxH, Ft	50'Øx75'	2-50'Øx 75'	-	50'Ø	50'Ø	50'Ø
23	Intermediate HX Weight (tons)	(2223)	(4446)	-	(2223)	(2223)	(2223)

Table B-6 (Page 2 of 4)

CLOSED-CYCLE INERT GAS MHD (Argon + Cs Mist)

		Series - Clean Fuels					
	Furnace Type	CH →					
	Fuel Code	SC →					
	Case Number	7	8	9	10	11	12
1	Output per Module, 10^6 Btu/Hr	4713	5093	4303	4522	4673	5810
2	Heat Exchanger Efficiency, Q_{out}/Q_{in}	.9079	.9265	.8939	.9257	.9134	.8929
3	Number of Modules	.9194	.9089	.9672	.9096	.8729	.8572
4	Fuel Input, 10^6 lb/Hr	.3310	.3505	.3070	.3115	.3262	.4149
5	Combustion Air Input, 10^6 Lb/Hr	4.753	5.034	4.409	4.731	4.021	5.958
	(Excess Air %)	(20)	(20)	(20)	(26.9)hi	(3.0)	(20.0)
6	Combustion Gas Flow, 10^6 Lb/Hr	5.084	5.385	4.716	5.043	4.347	6.373
	(Gas Recirculated %)	(30.23)	(54.88)	(14.71hi)	(0.0)	(0.0)	(37.26)
7							
8	Total Solids Output 10^6 Lb/Hr	331	351	307	312	326	415
9	Percent Solids as Tap or Bottom Ash	Nill →					
10	SO_2 at Exit, Lb/10^6 Btu input	.8919	.8919	.8919	.8919	.8919	.8919
11	NO_2 at Exit, Lb/10^6 Btu input	.3	.3-	.6+	.6+	1.2+	.3
12	C O at Exit, Lb/10^6 Btu input	-	-	-	-	-	-
13	Solids at inlet to Final removal equip. lb/10^6 Btu	.0638	.0638	.0638	.0638	.0638	.0638
14	Final particulate removal eff. reg'd to meet 0.1 lb/10^6 Btu solids emissions	-	-	-	-	-	-
15	Fan Power Requirements, MWe	38.892	50.275	29.662	25.711	21.917	52.974
16	Other Auxiliary Equip. Power, MWe	.47	.49	.47	.43	.45	.57
17							
18	Combustor Dia., Ft.	82	80	82	82	82	2 x 70
19	Combustor Height, Ft.	130	130	130	130	130	150
20	Burners - Combustion Stages/combustor	24 →					14@
21	Combustor weight, tons/module	2215	1780	2492	2492	2676	4066
22	Intermediate HX size WxDxH, ft.	36'Øx130'	42'Øx133	50'Øx133	46'Øx72'	52'Øx52	2-50'Øx57
23	Intermediate HX weight tons)	(2223)	(3101)	(2727)	(1715)	(1355)	(2885)

Table B-6 (Page 3 of 4)

CLOSED-CYCLE INERT GAS MHD (Argon + Cs Mist)

		Series - Clean-Fuels			Slagging Coals		
	Furnace Type	CH	→	→	CH	→	→
	Fuel Code	SC	→	→	I6	MS	ND
	Case Number	13	14	15	16	17	18
1	Output per Module, 10^6 Btu/Hr	4459	4509	4803	9906	9706	9398
2	Heat Exchanger Efficiency, Q_{out}/Q_{in}	.9323	.9078	.9097	.8882	.8742	.8459
3	Number of Modules	.9195	.9308	.9042	.7834	.8132	.8499
4	Fuel Input, 10^6 lb/Hr	.3047	.3167	.3367	1.0338	1.2414	1.6125
5	Combustion Air Input, 10^6 Lb/Hr	4.376	4.548	4.835	10.373	10.062	9.2364
	(Excess Air %)	(20)	(20)	(20)	(22)	(20)	(10)
6	Combustion Gas Flow, 10^6 Lb/Hr	4.681	4.865	5.172	11.407	11.304	10.849
	(Gas Recirculated %)	(25.95)	(27.73)	(30.68)	(0)	(0)	(0)
7							
8	Total Solids Output 10^6 Lb/Hr	305	317	337	99200	93105	99975
9	Percent Solids as Tap or Bottom Ash	Nill	→	→	50%	→	→
10	SO_2 at Exit, Lb/10^6 Btu input	.8919			8.8988 +Fe_2O_3	8.3855	8.9986
11	NO_2 at Exit, Lb/10^6 Btu input	.3			.7	.7	.7
12	C O at Exit, Lb/10^6 Btu input	-		→	-	-	-
13	Solids at inlet to Argon Heater lb/10^6Btu	.0638			.890 +Fe_2O_3	.840	.900
14	Final particulate removal eff. req'd to meet 0.1 lb/10^6 Btu solids emissions	-			BY GE	→	→
15	Fan Power Requirements, MWe	33.568	35.550	39.490	59.389	58.065	54.023
16	Other Auxiliary Equip. Power, MWe	.42	.44	.47	8.47	10.17	13.21
17							
18	Combustor Dia., ft.	82	82	86	2-80	2-80	2-80
19	Combustor height, ft.	130	130	130	130	130	130
20	Burners - Combustion Stages/combustor	24	24	24	24@	→	→
21	Combustor weight, tons/module	2215	2215	2336	3266	3266	3266
22	Intermediate HX size WxDxH, ft.	41'Øx93	45'Øx30'	51'Øx75'	40x115x30	→	→
23	Intermediate HX weight (tons)	(1976)	(2079)	(2250)	(3051)	(3051)	(3051)

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CLOSED-CYCLE INERT GAS MHD (Argon + Cs Mist)

		Slagging Coals				
	Furnace Type	CH				
	Fuel Code	16				
	Case Number	19	20	21	22	
1	Output per Module, 10^6 Btu/Hr	10666	9906	9474	9906	.
2	Heat Exchanger Efficiency, Q_{out}/Q_{in}	.8874	.8882	.8872	.8882	*
3	Number of Modules	.7337	.7834	.7851	.7834	*
4	Fuel Input, 10^6 lb/Hr	1.1131	1.0338	.9898	1.0338	
5	Combustion Air Input, 10^6 Lb/Hr (Excess Air %)	11.169 (22)	See Case 16	9.9320 (22)	See Case 16	
6	Combustion Gas Flow, 10^6 Lb/Hr (Gas Recirculated %)	12.282 (0)	(0)	10.9220 (0)		
7						
8	Total Solids Output Lb/Hr	106,858		95,021		
9	Percent Solids as Tap or Bottom Ash	90%				
10	SO_2 at Exit, Lb/10^6 Btu input	See				
11	NO_2 at Exit, Lb/10^6 Btu input	Case				
12	C O at Exit, Lb/10^6 Btu input	16				
13	Solids at inlet to lb/10^6 Btu					
14	Final particulate removal eff. reg'd to meet 0.1 lb/10^6 Btu solids emissions	BYGE				
15	Fan Power Requirements, MWe	63.945	59.389	56.863	59.389	
16	Other Auxiliary Equip. Power, MWe	9.12	8.47	8.11	8.47	
17						
18	Combustor Dia., ft.	2-80				Rough
19	Combustor height, ft.	130				sizing
20	Burners - Combustion Stages/combustor	24@				
21	Combustor weight, tons/module	4430	3266	3266	3266	
22	Intermediate HX size WxDxH, ft.	40x135x 29	40x115x 30			
23	Intermediate HX weight (tons)	(3389)	(3051)	(2903)	(3051)	

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OPEN-CYCLE MHD

Furnace Type						
Fuel Code	I6	→	→	MS	ND	I6
Case Number	1	2	3	4	5	6
Radiant Furnace						
Duty 10^6Btu/hr	2659	1594	809	3829	3625	2768
Width, ft	80	80	65	80	→	→
Depth, ft	63	37.3	25.6	67.8	75.9	65.1
height, ft	100	→	→	→	→	→
weight, tons	488.9	330.6	203.2	619.4	600.4	499.3
Convection Furnace						
SH Duty, 10^6Btu/hr	2021	1258	680	1722	2149	1900
RH Duty, 10^6Btu/hr	1242	797	405	1343	1326	1254
Width, ft	80	80	65	80	→	→
Depth, ft	38.7	24.3	15.8	37.5	42.6	38.7
Height, ft	70	70.9	70.2	61.6	57.2	69.4
Weight, tons	4761.9	3022.8	1599.8	4057.4	4646.	4692.8
Economizer						
Duty, 10^6Btu/hr	1024	683	307	751	478	1093
Width, ft	80	80	65	80	→	→
Depth, ft	32.7	30.6	13.2	33.8	37.2	32.7
Height, ft	7.9	8.4	6.9	4.6	2.5	8.6
Weight, tons	1096.7	741.3	314.3	656 3	390.4	1202.6
Auxiliary Equipment Power, MW_e	15.12	9.787	5.037	16.841	23.883	15.124

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OPEN-CYCLE MHD

Furnace Type						
Fuel Code	16 →					
Case Number	7	8	9	10	11	12
Radiant Furnace						
Duty 10^6Btu/hr	2768	2171	5914	5460	1642	1181
Width, ft	30	→	→	→	→	→
Depth, ft	65.1	49.2	44.8	29.5	66.8	66.8
height, ft	100	→	→	→	→	82
weight, tons	499.3	410.3	958.7	1833.5	353.5	273.6
Convection Furnace						
SH Duty, 10^6Btu/hr	1900	1543	404	-	2294	2360
RH Duty, 10^6Btu/hr	1254	995	1530	853	1116	1050
Width, ft	80	→	→	→	→	→
Depth, ft	38.7	30.2	23.7	16.7	41.8	41.8
Height, ft	69.4	71.3	89.3	34.5	53.5	53.8
Weight, tons	4692.8	3717.	2309.9	544.4	4577.	4845.2
Economizer						
Duty, 10^6Btu/hr	1093	853	785	Gas/Air 768/273	922	922
Width, ft	80	→	→	→	→	→
Depth, ft	32.7	25.4	25.2	30.9	32.7	→
Height, ft	8.6	8.6	6.3	7.0	6.4	6.6
Weight, tons	1202.6	926.	675.5	870.6	892.9	913.8
Auxiliary Equipment Power, MW_e	15.124	12.290	16.498	15.124	15.124	15.124

Table B-7 (Page 3 of 5)

OPEN-CYCLE MHD

Furnace Type						
Fuel Code	16					
Case Number	13	14	15	16	17	18
Radiant Furnace						
Duty 10^6 Btu/hr	7154	2335	3358	5594	2795	2829
Width, ft	80	→	→	→	→	→
Depth, ft	57.5	62.3	63.7	64.4	63.	62.8
height, ft	100	→	→	→	→	→
weight, tons	1137.8	446.2	559.2	800.1	495.5	501.2
Convection Furnace						
SH Duty, 10^6 Btu/hr	190	2050	1930	1587	2021	2015
RH Duty, 10^6 Btu/hr	1701	1213	1333	1676	1242	1248
Width, ft	80	→	→	→	→	→
Depth, ft	34.9	38.7	→	→	→	→
Height, ft	97.7	69.7	72.8	83.9	72.6	72.7
Weight, tons	3454.9	4746.2	4847.8	4772.7	5057.5	5052.
Economizer						
Duty, 10^6 Btu/hr	939	1024	→	→	→	→
Width, ft	80	→	→	→	→	→
Depth, ft	32.7	→	→	→	→	→
Height, ft	5.9	7.9	7.7	7.2	7.9	7.9
Weight, tons	821.2	1104.1	1068.4	997.6	1096.8	1096.8
Auxiliary Equipment Power, MW_e	15.124	→	→	→	→	→

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OPEN-CYCLE MHD

Furnace Type						
Fuel Code	I6	→	→	→	→	SRC
Case Number	19	20	21	22	23	24
Radiant Furnace						
Duty, 10^6Btu/hr	2761	2522	2659	-	2659	1966
Width, ft	80	→	→	-	80	→
Depth, ft	63.7	62.5	62.8	-	63.7	58.5
height, ft	100	→	→	-	100	→
weight, tons	499.0	466.7	481	-	485.7	380.6
Convection Furnace						
SH Duty, 10^6Btu/hr	1997	2028	2014	-	2021	1858
RH Duty, 10^6Btu/hr	1266	1235	1249	-	1242	1057
Width, ft	80	→	→	-	80	→
Depth, ft	38.7	→	→	-	38.7	36.5
Height, ft	73.5	70.1	70.5	-	70.3	56.2
Weight, tons	5125.5	4760.5	4782.6	-	4771.2	4018.3
Economizer						
Duty, 10^6Btu/hr	1024	→	1058	-	1024	939
Width, ft	80	→	→	-	80	→
Depth, ft	32.7	→	→	-	32.7	28.8
Height, ft	7.8	7.9	8.2	-	7.9	6.1
Weight, tons	1089.9	1096.8	1148.2	-	1096.8	752.1
Auxiliary Equipment Power, MW_e	15.124	→	→	→	→	(1.906MW_e +45.369 mmBtu/hr

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OPEN-CYCLE MHD

Furnace Type Fuel Code Case Number	SRC → 25	26	27	28	29	30
Radiant Furnace						
Duty 10^6Btu/hr	2983	-	1966	1932	1898	2034
Width, ft	80	-	80	→	→	→
Depth, ft	55.3	-	58.5	58.5	58.6	58.4
height, ft	100	-	100	→	→	→
weight, tons	543.8	-	380.5	374.5	374.9	374.2
Convection Furnace						
SH Duty, 10^6Btu/hr	1582	2699	1854	1867	1871	1846
RH Duty, 10^6Btu/hr	1169	748	1061	1048	1044	1069
Width, ft	80	→	→	→	→	→
Depth, ft	33.7	38.7	36.5	→	→	→
Height, ft	60.0	27.4	56.3	56.1	56.0	56.4
Weight, tons	3619.6	2641.2	4014.7	4014.2	4013.	4023.6
Economizer						
Duty, 10^6Btu/hr	939	341	939	→	→	→
Width, ft	80	→	→	→	→	→
Depth, ft	28.8	→	→	→	→	→
Height, ft	7.4	1.9	6.1	→	→	→
Weight, tons	904.1	233.4	752.1	→	→	747.6
Auxiliary Equipment Power, MW_e	(1.906 MW_e + 45.369 mm Btu/hr)			→	→	→

Table B-8

HIGH-TEMPERATURE FUEL CELLS

		Furnace Type / Fuel Code	CF	CF	CF	CF		
	Case Number		1	2	3	4	5	6
1	Output per Module, 10^6 Btu/Hr		4187	4187	4146	4164		
2	Heat Exchanger Efficiency, Q_{out}/Q_{in}		0.995	0.995	0.995	0.996		
3	Number of Modules		1	1	1	1		
4	Fuel Input, 10^6 lb/Hr		3.001	3.001	2.039	2.425		
5	Combustion Air Input, 10^6 Lb/Hr		6.481	6.481	4.887	5.526		
6	Combustion Gas Flow, 10^6 Lb/Hr		9.482	9.482	6.926	7.951		
7	Stone Input, 10^6 Lb/Hr (Limestone or Dolomite)		-	-	-	-		
8	Total Solids Output 10^6 Lb/Hr		-	-	-	-		
9	Percent Solids as Tap or Bottom Ash		-	-	-	-		
10	SO_2 at Exit, Lb/10^6 Btu input		0	0	0	0		
11	NO_2 at Exit, Lb/10^6 Btu input		?	?	?	?		
12	C O at Exit, Lb/10^6 Btu input		0	0	0	0		
13	Solids at inlet to Final removal equip. lb/10^6 Btu		0	0	0	0		
14	Final particulate removal eff. req'd. to meet 0.1 lb/10^6 Btu solids emissions		-	-	-	-		
15	Fan Power Requirements, MWe		-	-	-	-		
16	Other Auxiliary Equip. Power, MWe		.052	.052	.038	.044		
17	Heat Exchanger Width, Ft.		80	80	80	80		
18	Heat Exch. Depth or O.D., Ft.		77.1	77.1	64.0	69.3		
19	Heat Exchanger Height, Ft.		82.0	82.0	69.5	75.2		
20	No. Cells or Combustion Stages/Module		1	1	1	1		
21	Heat exchanger weight, Tons/Module		4214	4214	2480	3069		

APPENDIX C

COST DATA FOR PRIMARY HEAT INPUT SYSTEMS

A tabulation of cost data for each energy conversion system studied is presented here on a modular basis. The final number of modules utilized for each parametric case is given in Volume II in the section describing the specific energy conversion system. The parametric point designation is also found in Volume II.

LIST OF TABLES

Table C-1 (Page 1 of 2)

ADVANCED STEAM CYCLE

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS				
		HEAT EXCHANGER	LOW TEMP. AIR HEATER	AUXILIARY EQUIPMENT	TOTAL PER MODULE	TOTAL COST
1	4	5.624	.731	4.719	11.074	44.296
2	3	5.680	.731	4.719	11.130	33.390
3	6	5.514	.731	4.719	10.964	65.784
4	8	5.408	.731	4.719	10.858	86.864
5	4	5.669	.715	4.630	11.014	44.056
6	4	5.624	.703	4.561	10.888	43.552
7	4	5.709	.674	4.398	10.781	43.124
8	4	6.018	.727	4.695	11.440	45.760
9	by GE					
10	4	5.481	.778	5.524	11.783	47.132
11	4	5.324	.762	4.897	10.983	43.932
12	4	5.483	.743	4.790	11.016	44.064
13	4	5.663	.712	4.612	10.987	44.048
14	4	5.087	.698	4.534	10.319	41.276
15	4	6.357	.828	5.366	12.551	50.204
16	4	5.836	.761	4.583	11.130	44.720

Table C-1 (Page 2 of 2)

ADVANCED STEAM CYCLE

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS				
		HEAT EXCHANGER	LOW TEMP. AIR HEATER	AUXILIARY EQUIPMENT	TOTAL PER MODULE	TOTAL COST
17	1	25.881	2.525	7.400	35.806	35.806
17 A	1	19.239	2.633	7.717	29.589	29.589
18	1	28.413	2.634	9.437	40.484	40.484
19	1	26.104	2.495	8.689	37.288	37.288
20	1	24.155	2.406	5.324	31.885	31.885
21	4	1.006	-	.191	1.197	4.788
22	4	.986	-	.191	1.177	4.708
23	4	1.004	-	.191	1.195	4.780
24	4	2.796	-	10.718	13.514	54.056
25	4	2.759	-	10.817	13.576	54.304
26	4	2.593	-	10.000	12.593	50.372
27[1]	4	3.076	-	10.718	15.961	
27*	4	3.047	-	9.865	14.907	

Table C-2 (Page 1 of 3)

SUPERCRITICAL CO_2 CYCLE

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS					
		HEAT EXCHANGER	HIGH TEMP. AIR HEATER	LOW TEMP. AIR HEATER	AUXILIARY EQUIPMENT	TOTAL PER MODULE	TOTAL COST
1	2.813	25.180	2.531	.731	4.943	33.385	93.912
2	5.626	25.180	2.531	.731	4.943	33.385	87.824
3	2.851	26.557	2.642	.761	4.771	34.731	99.018
4	3.121	26.944	2.670	.823	5.421	35.858	111.913
5	4.103	8.545	1.090	-	10.718	20.353	83.508
6	6.081	10.643	.426	-	.191	11.260	68.472
7	6.097	10.547	.434	-	.191	11.172	68.116
8	6.645	10.283	.442	-	.191	10.916	72.537
9	4.585	19.624	-	-	.191	19.815	90.852
10	4.745	20.840	.868	-	.191	21.899	103.911
11	2.740	25.655	2.531	.731	4.943	33.860	92.776
12	2.889	24.672	2.531	.731	4.943	32.877	94.982
13	2.792	26.791	2.531	.731	4.943	34.996	97.709
14	2.919	23.485	2.531	.731	4.943	31.690	92.503
15	3.030	21.961	2.531	.731	4.943	30.166	91.403

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SUPERCRITICAL CO_2 CYCLE

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS					
		HEAT EXCHANGER	HIGH TEMP. AIR HEATER	LOW TEMP. AIR HEATER	AUXILIARY EQUIPMENT	TOTAL PER MODULE	TOTAL COST
16	3.144	20.515	2.531	.731	4.943	28.720	90.296
17	2.811	29.507	2.531	.731	4.943	37.712	106.008
18	2.822	23.005	2.531	.731	4.943	31.210	88.075
19	2.813	25.180	2.531	.731	4.943	33.385	93.912
20	2.800	25.174	2.531	.731	4.943	33.379	93.461
21	2.734	28.917	2.531	.731	4.943	37.122	101.492
22	3.112	26.868	2.531	.731	4.943	35.073	109.147
23	2.813	25.180	2.531	.731	4.943	33.385	93.912
24	2.813	28.544	2.531	.731	4.943	36.749	103.375
25	2.816	23.017	2.531	.731	4.943	31.222	87.921
26	2.848	25.174	2.531	.731	4.943	33.379	95.063
27	2.845	23.775	2.531	.731	4.943	31.980	90.983
28	2.860	23.775	2.531	.731	4.943	31.980	91.463
29	2.833	25.174	2.531	.731	4.943	33.379	94.563
30	2.856	25.344	2.531	.731	4.943	33.549	95.816

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SUPERCRITICAL CO_2 CYCLE

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS					
		HEAT EXCHANGER	HIGH TEMP. AIR HEATER	LOW TEMP. AIR HEATER	AUXILIARY EQUIPMENT	TOTAL PER MODULE	TOTAL COST
31	2.818	25.344	2.531	.731	4.943	33.549	94.541
32	2.831	25.174	2.531	.731	4.943	33.379	94.162

Table C-3 (Page 1 of 3)

CLOSED-CYCLE GAS TURBINE

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS					
		HEAT EXCHANGER	HIGH TEMP. AIR HEATER	LOW TEMP. AIR HEATER	AUXILIARY EQUIPMENT	TOTAL PER MODULE	TOTAL COST
1	1.840	20.038	1.073	.709	4.943	26.763	49.244
2	1.792	20.053	1.427	.697	5.421	27.598	49.456
3	1.841	20.567	1.167	.719	4.771	27.224	50.119
4	4.364	5.908	-	-	.191	6.099	26.616
5	4.374	5.891	-	-	.191	6.082	26.603
6	4.541	5.720	-	-	.191	5.911	26.842
7	2.769	9.024	-	-	.191	9.215	25.515
8	2.640	5.362	.866	-	10.718	16.946	44.754
9	2.761	20.037	1.073	.709	4.943	26.764	73.895
10	3.681	20.038	1.073	.709	4.943	26.763	98.515
11	1.628	14.744	-	.908	4.943	20.595	33.529
12	3.225	13.663	-	-	.191	13.854	44.679
13	1.438	25.972	1.564	.707	4.943	33.186	47.721
14	2.135	21.380	-	.950	4.943	27.273	58.228
15	1.816	21.873	1.901	.709	4.943	29.426	53.438

Table C-3 (Page 2 of 3)

CLOSED-CYCLE GAS TURBINE

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS					
		HEAT EXCHANGER	HIGH TEMP. AIR HEATER	LOW TEMP. AIR HEATER	AUXILIARY EQUIPMENT	TOTAL PER MODULE	TOTAL COST
16	1.791	22.437	1.287	.709	4.943	29.426	52.702
17	1.840	20.038	1.073	.709	4.943	26.763	49.244
18	1.819	21.899	1.901	.709	4.943	29.452	53.573
19	1.931	22.062	1.001	.709	4.943	28.715	55.449
20	2.410	18.875	-	.783	4.943	24.601	59.288
21	2.798	15.776	-	.785	4.943	21.504	60.168
22	1.438	25.972	1.564	.707	4.943	33.186	47.721
23	1.665	23.895	1.405	.709	4.943	30.952	51.535
24	3.613	12.558	-	-	.191	12.749	46.062
25	2.848	15.532	-	-	.191	15.723	44.779
26	2.835	16.804	1.054	.709	4.943	23.510	66.651
27	1.797	14.744	-	.841	4.943	20.528	36.889
28	1.438	25.369	1.735	.709	4.943	32.756	47.103
29	1.438	25.369	1.735	.709	4.943	32.756	47.103
30	1.438	25.369	1.735	.709	4.943	32.756	47.103

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Table C-3 (Page 3 of 3)

CLOSED-CYCLE GAS TURBINE

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS					
		HEAT EXCHANGER	HIGH TEMP. AIR HEATER	LOW TEMP. AIR HEATER	AUXILIARY EQUIPMENT	TOTAL PER MODULE	TOTAL COST
31	2.652	16.676	-	-	.191	16.867	44.731
32	3.127	14.451	-	-	.191	14.642	45.786
33	3.516	12.875	-	-	.191	13.066	45.940
34	1.438	25.972	1.564	.707	4.943	33.186	47.721
35	1.438	25.972	1.564	.707	4.943	33.186	47.721
36	1.438	25.972	1.564	.707	4.943	33.186	47.721
37	1.438	25.972	1.564	.707	4.943	33.186	47.721
38	1.438	25.972	1.564	.707	4.943	33.186	47.721
39	1.438	25.972	1.564	.707	4.943	33.186	47.721
40	2.267	20.082	-	.932	4.943	25.957	58.845
41	1.438	25.368	1.735	.709	4.943	32.755	47.102
42	1.438	25.972	1.564	.707	4.943	33.186	47.721
43	1.438	25.972	1.564	.707	4.943	33.186	47.721
44	2.267	20.082	-	.932	4.943	25.957	58.845
45	1.438	25.368	1.735	.709	4.943	32.755	47.102
46	3.371	14.007	-	.783	4.943	19.733	66.520

Table C-4 (Page 1 of 2)

LIQUID-METAL TOPPING CYCLE

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS							
		HEAT EXCHANGER	HIGH TEMP. AIR HEATER	LOW TEMP. AIR HEATER	INTERMED. HEAT EXCH	AUXIL'Y EQUIP'T	IHX PUMP SEPARATOR COSTS	TOTAL PER MODULE	TOTAL COST
1	5.501	27.005	4.555	.660	.479	5.086	6.620	44.405	244.272
2	5.599	26.585	3.988	.344	.470	4.913	6.720	43.020	240.869
3	5.748	26.090	3.623	.377	.446	5.556	7.020	43.112	247.808
4	8.072	2.982	.988	-	.255	.191	3.240	7.571	61.113
5	8.056	3.010	.977	-	.255	.191	3.180	7.528	60.646
6	8.342	2.921	.955	-	.245	.191	3.360	7.590	63.316
7	7.203	6.739	-	-	.284	.191	3.381	10.595	76.316
8	8.252	27.007	3.988	.660	.470	5.086	6.620	43.831	361.693
9	4.746	9.649	-	.183	.396	10.718	4.460	25.406	120.577
10	5.568	7.772	-	-	.284	.191	.880	9.127	50.819
11	5.304	67.847	4.250	.685	.459	5.564	3.920	82.725	438.773
12	5.533	26.509	4.975	.630	.474	5.086	6.620	44.294	388.937
13	5.501	27.008	4.555	.660	.416	5.086	6.620	44.345	243.942
14	5.501	27.008	4.555	.660	.479	5.086	6.620	44.408	244.288
15	5.501	27.008	4.555	.660	.479	5.086	6.620	44.408	244.288

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LIQUID-METAL TOPPING CYCLE

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS							
		HEAT EXCHANGER	HIGH TEMP. AIR HEATER	LOW TEMP. AIR HEATER	INTERMED. HEAT EXCH.	AUXIL'Y EQUIP'T	IHX PUMP SEPARATOR COSTS	TOTAL PER MODULE	TOTAL COST
16	5.501	27.008	4.555	.660	.479	5.086	6.620	44.408	244.288
17	5.501	26.928	4.555	.660	.636	5.086	.390	38.255	210.441
18	5.501	27.008	4.555	.660	.479	5.086	.390	38.178	210.017

Table C-5

CLOSED-CYCLE LIQUID-METAL MHD

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS					
		HEAT EXCHANGER	LOW TEMP. AIR HEATER	AUXILIARY EQUIPMENT	INTERMEDIATE HEAT EXCHANGE	TOTAL PER MODULE	TOTAL COST
1	2.766	15.962	.782	4.943	4.558	26.245	72.594
2	1.383	15.962	.782	4.943	4.558	26.245	36.297
3	5.532	15.962	.782	4.943	4.558	26.245	145.187
4	2.766	15.962	.798	4.771	4.558	26.089	72.162
5	2.766	15.962	.811	5.421	4.558	26.752	73.996
6	6.646	1.888	–	.191	1.973	4.052	26.930
7	6.662	1.883	–	.191	1.967	4.046	26.954
8	6.915	1.827	–	.191	1.894	3.912	27.051
9	4.980	4.560	–	.191	2.590	7.341	36.558
10	4.381	3.045	–	10.718	2.930	16.693	73.132
11	2.687	29.323	.782	4.943	4.181	39.229	105.408
12	2.786	80.727	.782	4.943	3.767	90.219	251.350
13	2.802	14.016	.782	4.943	4.280	24.021	67.307
14	2.629	13.776	.782	4.943	4.558	24.059	63.251
15	2.766	15.962	.782	4.943	4.558	26.245	72.594
16	2.766	15.962	.782	4.943	4.558	26.245	72.594
17	3.956	11.921	.782	4.943	5.154	22.800	90.197

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CLOSED-CYCLE INERT GAS MHD

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS						
		COMBUSTOR	HIGH TEMP. AIR HEATER	LOW TEMP. AIR HEATER	AUXILIARY EQUIPMENT	INTERMEDIATE HEAT EXCHANGER	TOTAL PER MODULE	TOTAL COST
1	.9194	10.341	8.358		2.517	9.692	30.908	28.417
2	.9193	20.682	16.716		4.084	19.385	61.816	56.827
3	.6421	1.901	12.929		1.014	-	15.844	10.173
4	.8652	10.906	6.233		.191	9.696	27.026	23.383
5	.8787	10.906	6.155		.191	9.696	26.948	23.679
6	.8791	10.906	6.150		.191	9.696	26.943	23.686
7	.9194	10.341	8.358		2.517	9.692	30.908	28.417
8	.9089	8.311	7.192		2.623	13.521	31.647	28.792
9	.9672	11.634	8.003		2.517	11.890	34.044	32.927
10	.9096	11.634	5.826		2.420	7.478	27.358	24.885
11	.8729	12.493	14.088		2.500	5.908	24.989	21.813
12	.8572	18.982	15.708		2.950	12.579	50.219	43.048
13	.9195	10.341	6.600		2.380	8.615	27.936	25.687
14	.9308	10.341	6.699		2.450	9.065	28.555	26.579
15	.9042	10.906	9.362		2.550	9.810	32.628	29.502

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CLOSED-CYCLE INERT GAS MHD

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS						
		COMBUSTOR	HIGH TEMP. AIR HEATER	LOW TEMP. AIR HEATER	AUXILIARY EQUIPMENT	INTERMEDIATE HEAT EXCHANGER	TOTAL PER MODULE	TOTAL COST
16	.7834	15.247	4.083		8.195	18.182	45.707	35.807
17	.8132	15.247	4.472		9.583	18.151	47.453	38.589
18	.8499	15.247	5.062		12.005	17.800	50.114	42.592
19	.7337	20.681	4.394		8.730	21.194	54.999	40.353
20	.7834	15.247	4.083		8.195	18.182	45.707	35.807
21	.7851	15.247	3.907		7.800	17.388	44.342	34.813
22	.7834	15.247	4.083		8.195	18.182	45.707	35.807

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OPEN-CYCLE MHD

CASE #	COST IN 1974 MILLIONS OF DOLLARS						
	RADIANT FURNACE	CONVECTION FURNACE	LOW TEMP. AIR HEATER	ECONOMIZER	FLUES	AUXILIARY EQUIPMENT	TOTAL COST
1	3.146	22.728	12.256	1.645	3.241	24.162	67.178
2	2.124	14.428	8.418	1.112	2.571	15.190	43.843
3	1.302	7.626	10.596	.472	1.861	7.407	29.264
4	3.994	19.871	9.287	.985	3.294	24.666	62.097
5	3.868	22.976	9.962	.586	3.456	36.164	77.012
6	3.214	22.416	13.121	1.804	3.241	24.162	67.958
7	3.214	22.416	13.121	1.804	3.241	24.162	67.958
8	2.640	17.669	7.860	1.389	2.855	18.199	50.612
9	6.870	15.285	14.255	1.013	2.844	26.813	67.080
10	14.866	3.095	24.003	1.306	3.149	25.073	71.492
11	2.264	22.762	12.060	1.339	3.241	25.073	66.739
12	1.748	24.203	12.144	1.371	3.241	26.079	68.786
13	8.158	20.299	5.499	1.232	3.241	23.349	61.778
14	2.868	22.613	11.979	1.656	3.241	24.162	66.519
15	3.604	23.080	12.252	1.603	3.241	24.162	67.942

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OPEN-CYCLE MHD

CASE #	COST IN 1974 MILLIONS OF DOLLARS						
	RADIANT FURNACE	CONVECTION FURNACE	LOW TEMP. AIR HEATER	ECONOMIZER	FLUES	AUXILIARY EQUIPMENT	TOTAL COST
16	4.626	22.035	12.252	1.496	3.241	24.162	67.812
17	3.190	24.187	12.252	1.645	3.241	24.162	68.677
18	3.226	24.163	12.252	1.645	3.241	24.162	65.438
19	3.212	24.539	12.252	1.635	3.241	24.162	65.800
20	3.002	22.695	12.252	1.645	3.241	24.162	63.756
21	3.098	22.813	12.252	1.722	3.241	24.162	64.047
22	–	–	43.429	–	6.092	24.162	73.683
23	3.126	22.753	12.252	1.645	3.241	24.162	67.179
24	2.444	19.913	10.754	1.128	3.040	2.235	39.514
25	3.406	17.775	10.259	1.356	3.040	2.235	38.171
26	–	13.432	6.353	.350	3.040	2.235	25.410
27	2.444	19.897	10.589	1.128	3.040	2.235	39.[illegible]33
28	2.404	19.866	10.589	1.128	3.040	2.235	39.264
29	2.406	19.880	10.589	1.128	3.040	2.235	39.278
30	2.402	19.946	10.589	1.121	3.040	2.235	39.333

Table C-8

HIGH-TEMPERATURE FUEL CELLS

CASE #	MODULES	COST IN 1974 MILLIONS OF DOLLARS				
		HEAT EXCHANGER	LOW TEMP. AIR HEATER	AUXILIARY EQUIPMENT	TOTAL PER PER MODULE	TOTAL COST
1	1	16.230	2.230	.250	18.710	18.710
2	1	16.230	2.230	.250	18.710	18.710
3	1	10.610	1.629	.250	12.489	12.489
4	1	12.800	1.870	.250	14.920	14.920