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VOLUME I

ENERGY CONVERSION ALTERNATIVES STUDY
-ECAS-
WESTINGHOUSE PHASE I FINAL REPORT

Volume I – INTRODUCTION AND SUMMARY & GENERAL ASSUMPTIONS

by
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FOREWORD

A series of world and national events beginning in 1973 focused attention on the fact that this nation faces both an immediate and long-term energy crisis. Our reserves of oil and natural gas are severely limited. We have an abundant domestic reserve of coal, but face increasingly severe environmental problems from its use in power plants using conventional energy conversion techniques. There is a need to develop new energy conversion systems that are capable of using coal or coal-derived fuels in a more efficient and environmentally acceptable manner in the production of electric power.

Studies of many advanced energy conversion techniques have been performed in the past; however, new studies performed on a common basis and in the light of new national goals and current conditions are required to permit an assessment of the relative merits and potential benefits to the nation of the more promising advanced energy conversion techniques. NASA Lewis Research Center has been requested to perform the appropriate studies in the form of an Energy Conversion Alternatives Study (ECAS).

ECAS is being performed by the NASA in cooperation with the National Science Foundation, Energy Research and Development Administration and the Office of Management and Budget. The objective of the ECAS is to develop a data base that will permit decisions concerning energy research and technology to be made with a better understanding of the benefits to and impact on the nation. In addition, long range development plans and key experiments are to be defined for several advanced energy conversion systems to provide an estimate of both the development cost and probability of success of that concept development resulting in a commercially viable power plant.

To accomplish this the NASA Lewis Research Center awarded two nearly similar contracts in late December, 1974, one to General Electric Corporate Research and Development and the other to Westinghouse Electric Corporation Research Laboratories. These contracts called for a nine month program which would in Phase I perform a parametric analysis of each of either nine or ten advanced energy conversion concepts using coal or coal derived fuels; in Phase II, more detailed conceptual plant designs were to be formulated for a few selected concepts; Phase III was to be an implementation assessment for those concepts covered in Phase II.

A report entitled "Energy Conversion Alternative Study (ECAS), General Electric Phase I Final Report" will be published as report number NASA CR-134948. Another report entitled "Comparative Evaluation of ECAS Phase I Results" will be published with the number NASA TMX-71855.

ACKNOWLEDGMENTS

Sections 1 and 2 entitled "Introduction and Summary" and "General Assumptions" were centered in the Westinghouse Research Laboratories and directed by D. T. Beecher, the ECAS Program Manager.

Those making contributions were:

- M. K. Beck, Westinghouse Power Generation Systems, who projected coal costs and transportation costs.
- W. H. Comtois, Westinghouse Power Generation Systems, who prepared labor and material historical cost tables.
- G. E. Jablonka, Westinghouse Power Generation Systems, who prepared the heat rejection system description.
- W. L. Wright, Westinghouse Power Generation Systems, who reviewed wet scrubber technology and costs.
- Messrs. C. T. McCreedy, Project Manager, S. M. Scherer and A. S. Naoum, of Chas. T. Main, Inc. of Boston, who prepared balance of plant descriptions, scaling and costing relations. They also prepared overall plant arrangement, estimates of constructability, and consulted on the correctness of the capital cost summaries.

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1. INTRODUCTION AND SUMMARY

1.1 Introduction

Task I of the Westinghouse ECAS study was intended to produce a parametric analysis which would in a systematic and realistic, though preliminary, manner present the performance, economics, and natural resource requirements of nine advanced energy conversion concepts for utility applications using coal or coal-derived fuels. The nine concepts specified are given in Table 1.1.

Table 1.1 - ECAS I Concepts and Concept Responsibility

Concept	Responsible Westinghouse Division
Open-Cycle Gas Turbines	Gas Turbine Engine Division
Combined Gas-Steam Turbine Cycles	Gas Turbine Engine Division
Closed-Cycle Gas Turbines	Gas Turbine Engine Division
Metal Vapor Rankine Topping	Advanced Reactors Division
Open-Cycle MHD	Research Laboratories
Closed-Cycle MHD	Research Laboratories
Liquid-Metal MHD	Research Laboratories
Advanced Steam	Research Laboratories and Steam Turbine Division
Fuel Cell System	Research Laboratories

Commonality of assumption between the several concepts was stressed by NASA so Westinghouse decided to retain responsibility for each of the nine concepts within Westinghouse. Also shown in Table 1.1 is the Westinghouse division responsible for each concept.

Studies of the open-, combined-, and closed-cycle gas turbine systems, based on gas turbine engine technology, were undertaken by the Westinghouse Gas Turbine Engine Division at Lester, Pennsylvania. The open- and combined-cycle concepts were felt to represent extensions of existing technology that could be designed in the near future.

The study of the metal vapor Rankine topping cycles with steam bottomers was conducted by the Westinghouse Advanced Reactors Division at Madison, Pennsylvania, which has a background in the requisite liquid-metal subsystems from the Liquid Metal Fast Breeder Reactor Program.

The three MHD concepts, open-cycle, closed-cycle, and liquid-metal MHD are all topping cycles which use conventional reheat Rankine steam plants as bottomers. The MHD plants represent technologies which are not now commercial and which still have many questions to be answered before they can be the basis for commercial power plants. The responsibility for these three MHD concepts was assumed by the Westinghouse Research Laboratories in Pittsburgh, Pennsylvania.

The concept leadership for the advanced steam systems was also taken by the Westinghouse Research Laboratories because of the necessary inclusion of the gas turbine as a pressurizing unit in some plants. The Westinghouse Steam Turbine Division did, however, do the steam turbine performance and cost analysis.

The fuel cell system concept is another developing technology whose commercial applicability has yet to be demonstrated. The Westinghouse Research Laboratories, which has been active in high temperature solid electrolyte fuel cell development for many years, took responsibility for the fuel cell concept.

To assure some commonality of approach, heat rejection equipment requirements, plant cost, and estimated cost of electricity were computed centrally after receiving inputs from the concept teams.

The balance of plant equipment description and costs were prepared by Chas. T. Main, Inc. of Boston, Massachusetts, who worked closely with the concept teams and contributed greatly to the results of the study.

Combustion and gasifier technology were treated by a central group at the Westinghouse Research Laboratories which supplies subsystem information to the concept teams.

A materials effort, supported entirely by Westinghouse funds, presented information in the area of materials applicability and performance to all concept teams.

This Westinghouse ECAS Task I Report is divided into thirteen sections which are bound in twelve separate volumes. Sections 1 and 2, the Introduction and Summary and the General Assumptions, make up this, the initial volume. Section 3 presents the work on material, and Section 4 presents the work on combustion and gasification. Sections 5 to 13 each presents the work on one of the nine concepts. Sections 3 through 13 are each individually bound.

1.2 Summary

1.2.1 General Assumptions

Section 2, in addition to summarizing the NASA specified inputs of coal type, fuel cost ranges, labor rate ranges, fixed charges, and the like, discusses heat rejection apparatus requirements, balance of plant description and costing, indirect costs, and the calculation of total capital cost and the cost of electricity for each of the more than 670 plants evaluated in the ECAS Task I Study.

Heat rejected from cycles is, in general, either up a stack to atmosphere or through a coupling heat exchanger to water. The water than rejects this heat to atmosphere in a wet or dry cooling tower or to a body of water with a once-through system. Mechanical draft wet

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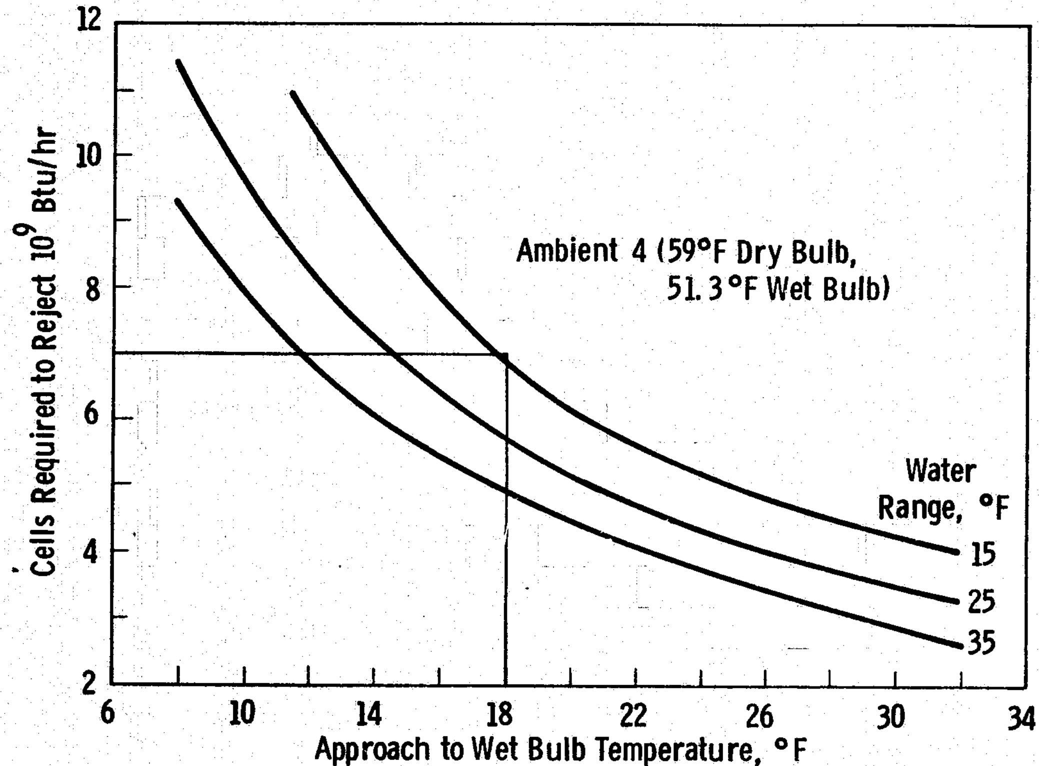

Fig. 1.1—Mechanical draft wet cooling tower capacity

cooling towers are used for most of the plants considered. The ISO ambient [288°K db, 284°K wb (59°F db, 51.4°F wb)] and a 283°K (49°F) river water temperature are specified for heat rejection. The ISO ambient is assumed adequate to provide a 6.77 kPa (2 in Hg) abs steam condensing temperature with a 2.78°K (5°F) condenser terminal temperature difference and a water range of 12.78°K (23°F).

Cooling tower performance curves were prepared for four ambients. The one for the ISO day, Figure 1.1, relates the number of cells required to reject 293 MJ/s (10^9 Btu/hr) through the circulating water to the range and approach to ambient wet bulb temperature. For example, a plant with a 10°K (18°F) approach and a 8.33°K (15°F) range would require ∿ seven cells to reject 293 MJ/s (10^9 Btu/hr) to the ISO ambient.

One or more typical or base cases were investigated for each concept. For each base case, site layouts were prepared showing site size, needed access, plant island size, fuel storage, etc., an example of which is shown in Figure 1.2.

Cost of balance of plant subsystems was generated by the A/E. General algorithms were prepared for each of the major subsystems. The relationship of raw material handling systems cost to tons/hr of material handled is shown in Figure 1.3 as a typical example. The break point in the curve [113.4 kg/s (450 tons/hr)] is the result of assuming a stacker reclaimer for the higher handling rates.

Many costs were generated by adjusting the cost experience in existing steam plants. On-site waste disposal was assumed in all cases. These base case costs were then individually adjusted for each of the more than 670 plants described.

The plant material and installation as well as the significant size parameter are reported in a code of accounts, a typical example of which required five to six pages of computer output per plant evaluated. Two copies of this detailed code of accounts for each of the 670 plants evaluated were delivered to NASA, but only the detailed accounts of the base and the preferred cases are included in this report.

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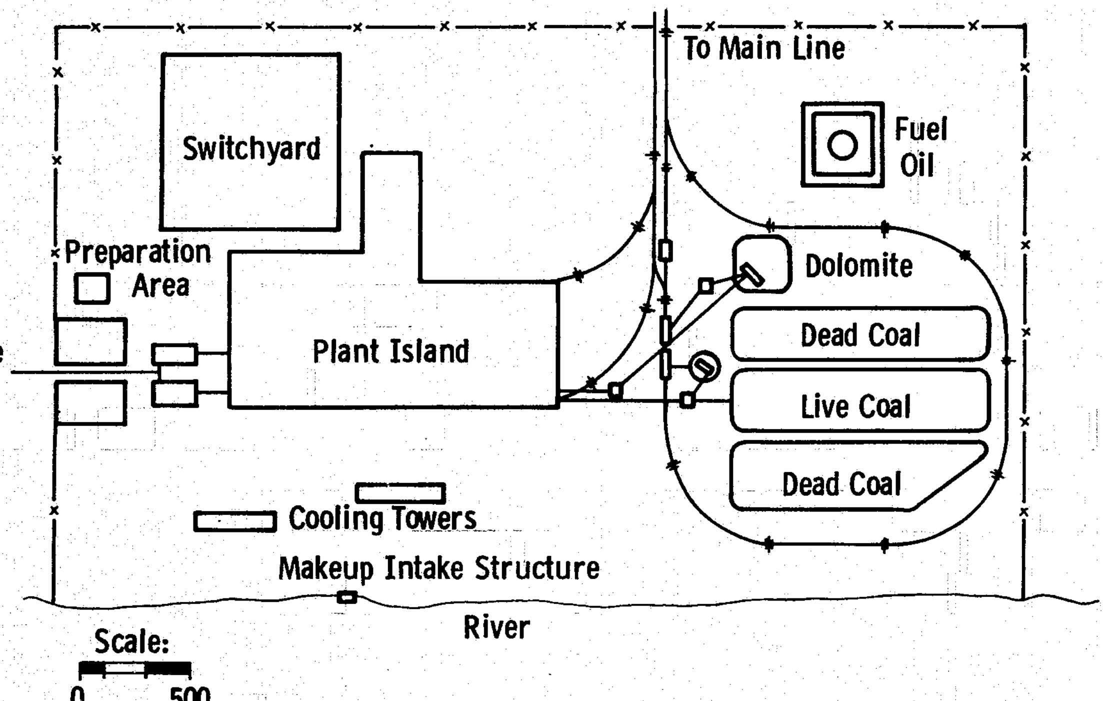

Fig. 1.2—Typical site layout

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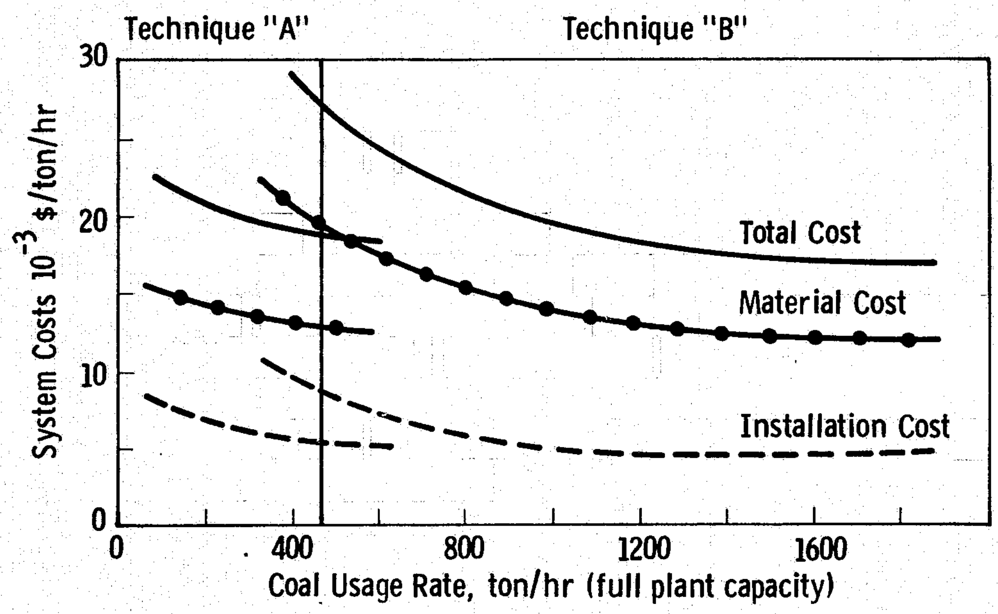

Fig. I.3—Raw material handling system costs

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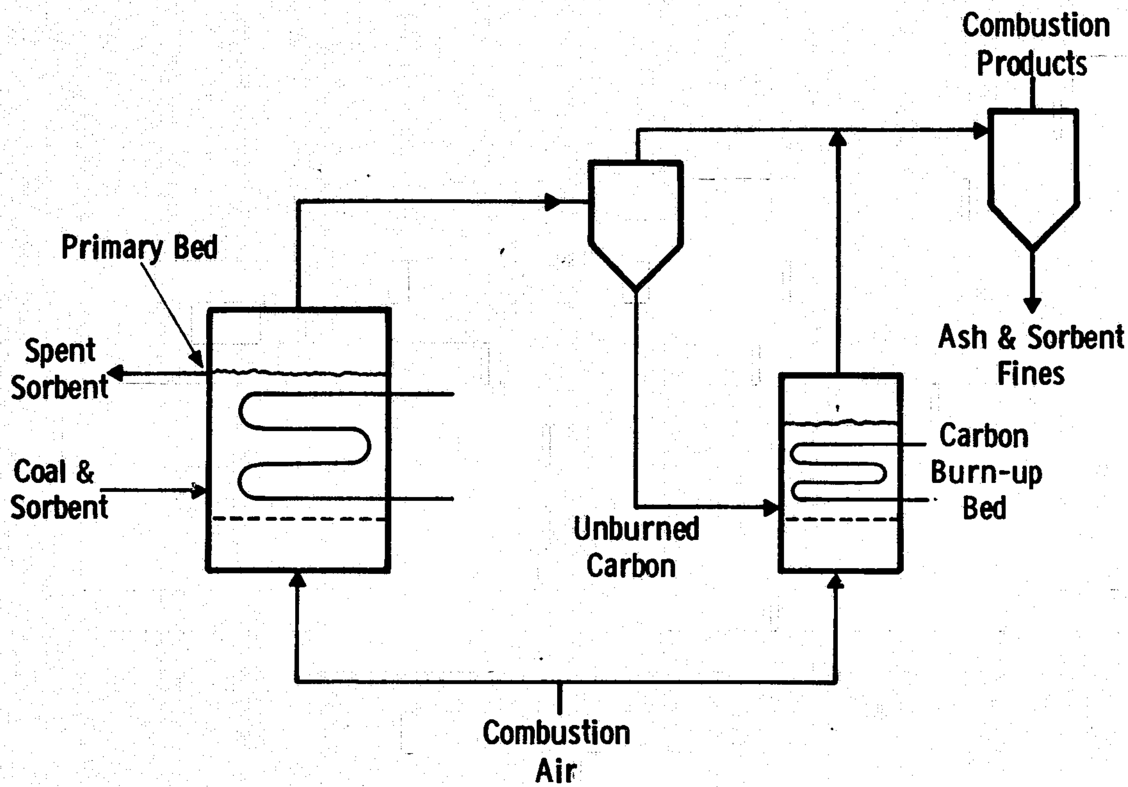

Fig. 1.4—Simplified fluidized bed combustor with heat removal

1.2.2 Materials Considerations

The material section (Section 3) presents an in-depth discussion in the area of corrosion and erosion of materials in combustion gases, and steam boiler materials. In addition, a subsection is devoted to materials problems associated with the major components of each concept.

1.2.3 Combustors, Furnaces and Low-Btu Gasifiers

Section 4 is concerned with a wide range of problems associated with fuel properties and processing, differing combustion technologies, gasification technology, and the calculation of the properties of combustion products for the several fuels.

Coal preparation is a mature technology with no recent development of any significance.

There are two notable processes under development in the field of combustion technology: fluidized bed combustion systems and staged cyclone combustors for MHD applications.

A simplified schematic of a fluidized bed combustor system with heat removal from the bed is shown in Figure 1.4. The fluidized bed combustor receives crushed coal, dolomite and combustion air which fluidizes the bed. The coal is burned at a relatively low temperature [1144 to 1255°K (1600 to 1800°F)] in the fluidized bed. At the same temperatures the dolomite gives up carbon dioxide and combines with the sulfur from the coal, removing approximately 90% of the available sulfur from the exhaust gas.

Exxon Laboratories has a 63 g/s (50 lb/hr) high pressure fluidized bed boiler miniplant under test. The inclusion of boiler tubes in the fluidized bed improves the heat transfer to the tubes outer surface and minimizes the amount of surface required. The spent sorbent or dolomite might increase the ash handling facility size requirements by as much as a factor of four.

The staged cyclone combustor being developed by the Bureau of Mines at Brushton, Pennsylvania for MHD applications fires crushed coal

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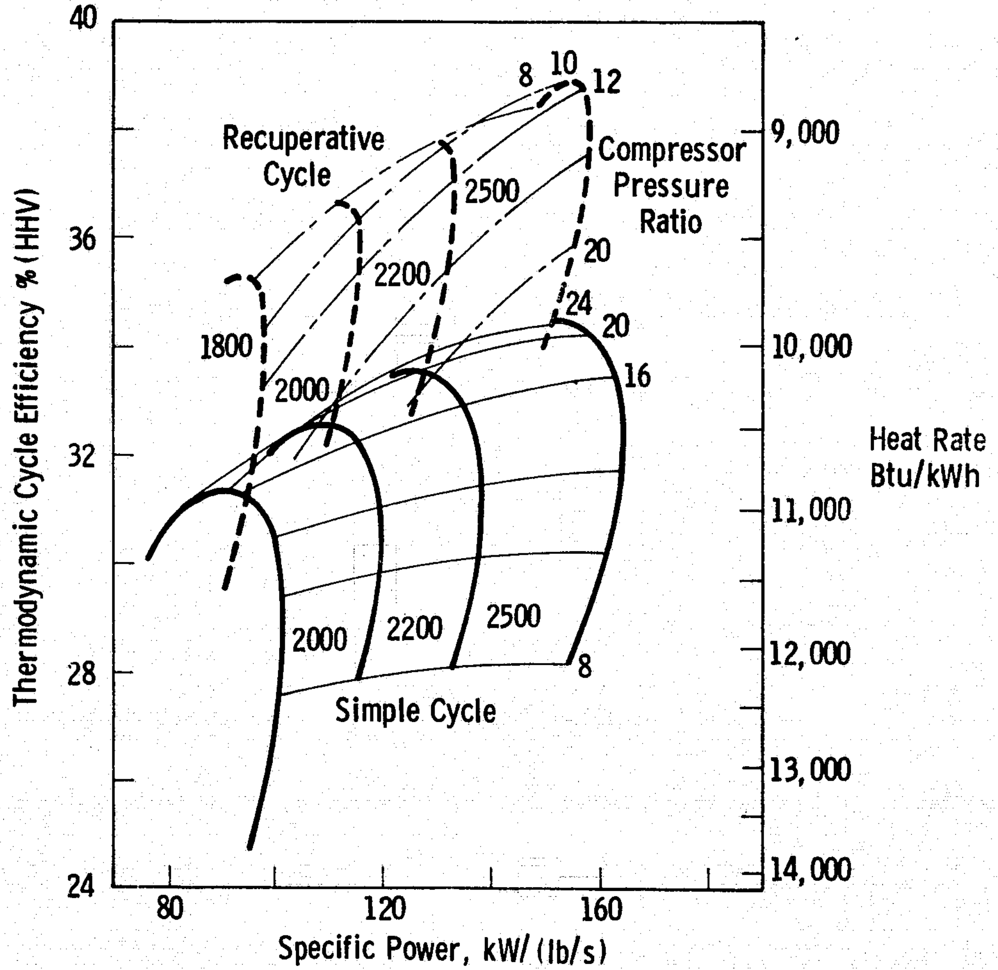

Fig. 1.5— Open cycle gas turbine efficiency

in a typical cyclone with approximately 65% stoichiometric air. Eighty percent of the ash is assumed to be tapped off in this first stage. Ninety-five percent stoichiometric air and seed are added in the second-stage combustor. The resultant products pass down the MHD duct, after which sufficient air is added to complete combustion before the heat-recover steam generator. This approach, in addition to limiting NO_x formation, also minimizes seed recovery and slagging problems.

The development of low-Btu gasification processes for utility application is being carried forward on a broad front. Westinghouse is currently operating a 151.2 g/s (1200 lb/hr) process development unit of the Westinghouse multi-stage fluidized bed gasifier at Madison, Pennsylvania. The gasifier accepts crushed coal, dolomite, and compressed air in the same manner as the fluidized bed combustor. Steam is also added to react with the char to form carbon monoxide and hydrogen. The resultant fuel gas with 90% of the sulfur removed in the bed passes through a particulate removal system and then to the designated combustion system.

1.2.4 Open Recuperated and Bottomed Gas Turbine Cycles

The first power conversion concept, open-cycle gas turbine systems, is discussed in Section 5. Included are simple or unrecuperated gas turbines, recuperated gas turbines with and without intercooling, and gas turbine topping organic vapor Rankine bottoming cycles. The parametric investigation covered gas turbine inlet temperatures from 1255 to 1644°K (1800 to 2500°F) with a base case value of 1478°K (2200°F), a modest extension of present-day state-of-the-art technology exemplified by a Westinghouse 501 gas turbine engine. Pressure ratios ranging from 6 to 24 to 1 were investigated. These gas turbines have air-cooled vanes and blades, burn clean fuels, and are fully assembled rail-shippable modules having a power output of approximately 100 MW. The generator is driven from the cold end, thereby allowing a minimum pressure loss axial arrangement of an exhaust duct, recuperator, or waste heat boiler.

Tension braze recuperators with effectiveness values of zero, the unrecuperated case, 70, 80, and 90% are considered, having a total pressure drop ratio of 3%. The efficiency for the simple and recuperated cycles is shown in Figure 1.5. It is seen that the efficiency of both

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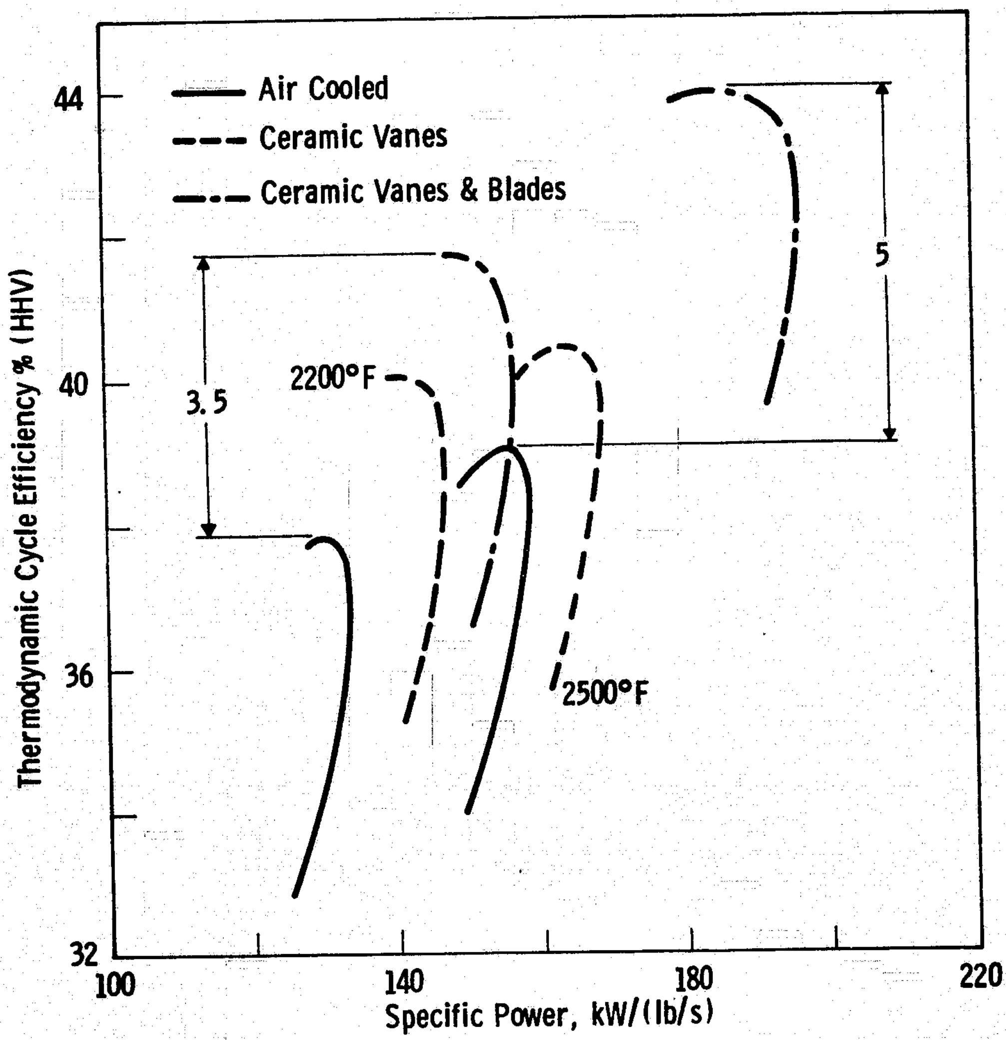

Fig. 1.6—Effect of the use of ceramic blades and vanes on the efficiency of an open-recuperated gas turbine

the simple and recuperated cycles increases with turbine inlet temperature. The optimum pressure ratio for the recuperated cycle falls around 10 to 1. For the simple cycle, higher pressure ratios produce a more efficient cycle. Higher pressure ratio machines result in higher compressor discharge temperatures and lower turbine exhaust temperatures, negating the usefulness of the recuperator.

The effect of using advanced gas turbines with ceramic vanes and blades to reduce the amount of cooling air needed is shown in Figure 1.6 for turbine inlet temperatures 1366 and 1644°K (2000 and 2500°F) An increase in cycle efficiency of 3.5 to 5 points is indicated, with the larger value accruing to the 1644°K (2500°F) cycle which uses more cooling air.

A cycle using sulfur dioxide as a bottoming fluid assumes a 1644°K (2500°F) air-cooled gas turbine with a 16 to 1 pressure ratio. The highly supercritical sulfur dioxide bottoming cycle with a nearly straight heating line is well fitted to the turbine exhaust gas cooling curve. Only a small loss in thermodynamic availability results. The sulfur dioxide throttle conditions were 17.237 MPa/811°K (2500 psi/1000°F). The sulfur dioxide vapor superheats on expansion so a desuperheating feed heater is used. This bottoming cycle has no moisture problems, therefore, and the size of the exhaust annulus is very small [27.94 cm (11 in) last row blades] compared to a steam turbine for the same duty. The 1644°K (2500°F) gas turbine used to top the sulfur dioxide cycle would have a plant efficiency of 33.5% if used in a simple cycle configuration. In a recuperated cycle it would have a plant efficiency of 37.6%, but in the sulfur dioxide combined cycle it has a plant efficiency of 47.6%.

The cost of electricity (COE) for these cycles is displayed on an overall plot of COE vs capacity factor in Figure 1.7. The light lines show the results for all concepts. The two heavy lines represent the simple cycle with a COE of 8.75 mills/MJ (31.5 mills/kWh) at a capacity factor of 65%, and the recuperated cycles with a COE of 8.19 mills/MJ (29.5 mills/kWh). The more capital intensive sulfur dioxide bottoming

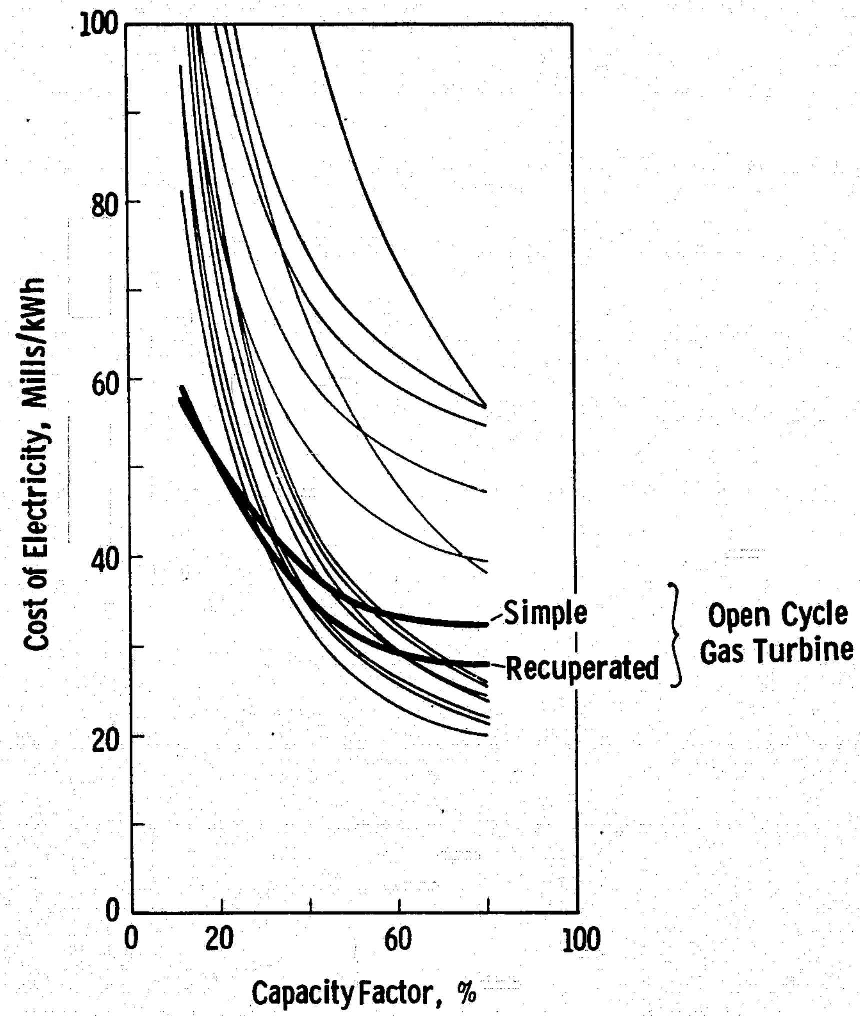

Fig. 1.7—Cost of electricity from open-cycle gas turbine plants

cycle is not shown but has a COE of 9.14 mills/MJ (32.9 mills/kWh) at a capacity factor of 65%. Both the simple and recuperated plants are most applicable to peaking and intermediate duty operation, that is to operation with a capacity factor less than 0.4. This is due to their low capital cost of 170 to 200 $/kW.

Major material problems lie in the development of ceramic elements for turbine vanes and blades [current tests have exceeded 100 cycles at 1478°K (2200°F)] and composite last-row blades.

1.2.5 Combined Gas-Steam Turbine Cycles

Section 6 treats the combined gas-steam turbine cycles. Typically, four 1478°K (2200°F), 10 to 1 gas turbines exhaust into modular heat recovery steam generators which supply a single subcritical steam turbine generator. The cycle parametric investigation is based on the use of clean distillate from coal as fuel. Specific arrangements are also evaluated which include the firing of low-Btu gas from an integrated coal gasifier. Both reheat and nonreheat steam cycles are considered. Induction of supplementary steam into the turbine cycle at one or two pressures below the throttle pressure is also considered. The inductions are into the cold reheat line and the crossover pipe between the IP and LP turbines.

The desirability for steam induction may be understood by looking at the water-steam heating-gas cooling line for a typical 16.547 MPa gauge/811°K/811°K (2400 psig/1000°F/1000°F) steam plant shown in Figure 1.8. The close correspondence of the two profiles is responsible for good efficiency of this arrangement. The broken line is the profile for a single pressure system and shows that the stack temperature would be over 478°K (400°F) for this single pressure system compared to 411°K (280°F) for a system with steam induction.

The effect of steam induction on several steam cycles is shown in Figure 1.9. Typically, steam induction can add two or three points to the cycle efficiency. The 16.547 MPa gauge/811°K/811°K (2400 psig/1000°F/1000°F) reheat steam cycle with an unfired boiler and two steam inductions after the throttle is the most efficient cycle investigated. This steam

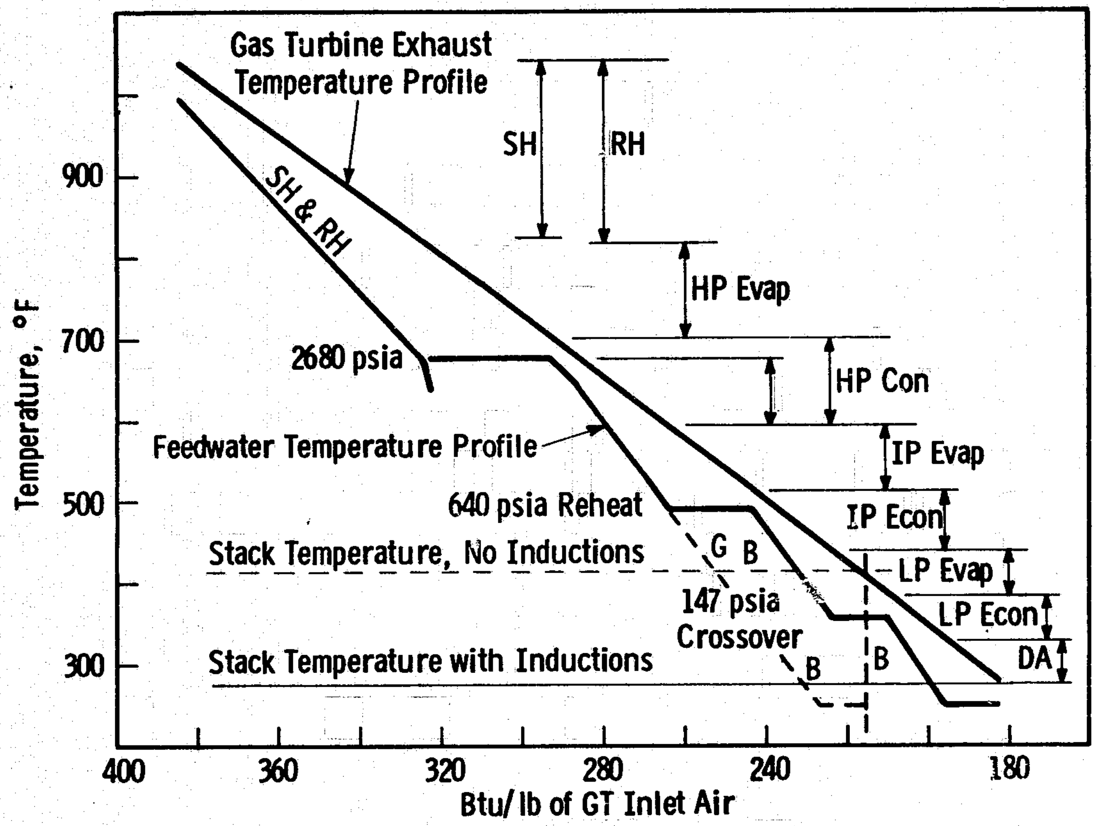

Fig. 1.8—2400 psig/1000°F/1000°F boiler temperature profile with reheat and crossover inductions

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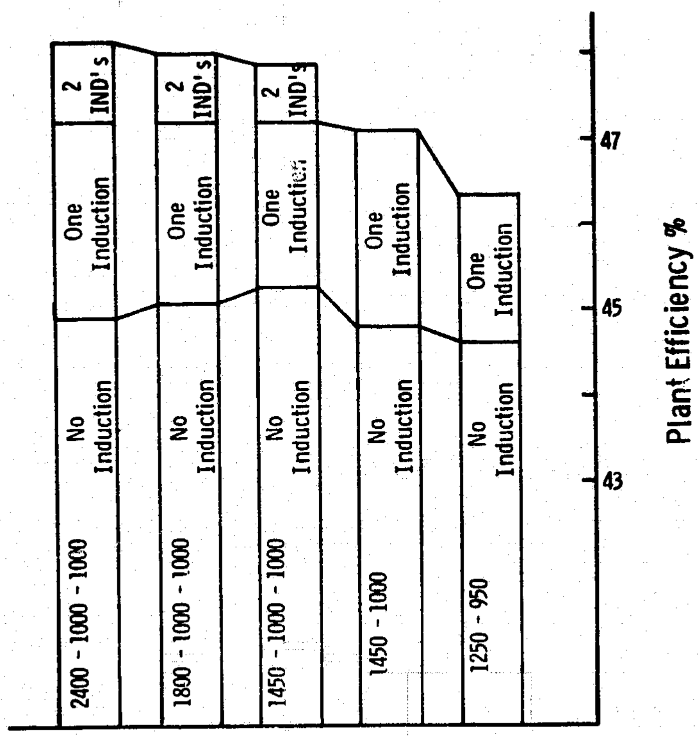

Fig. 1.9—Effect of steam induction on overall cycle efficiency for combined cycle plants in the unfired boilers (typical)

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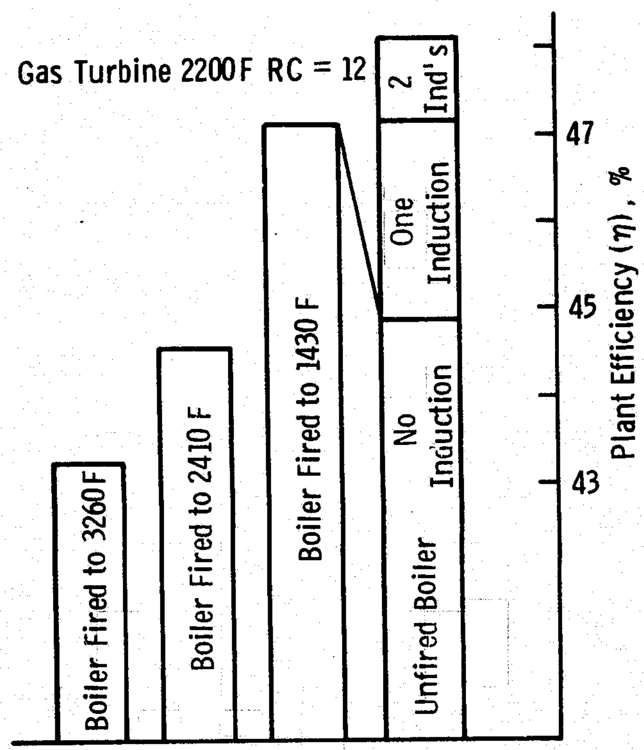

Fig. 1.10—Effect of boiler firing on combined cycle efficiency (typical)

plant with a 1478°K (2200°F) fired gas turbine and 16.7°K (30°F) approach of the exhaust gas temperature to the saturation temperature in the boilers achieves a plant efficiency of about 48% at the ISO ambient with a condenser pressure of 6.77 kPa (2 in Hg) abs. This is a 25% improvement in efficiency or 20% reduction in heat rate compared to an all-steam power plant with similar design sophistication.

Without steam induction there is little advantage in higher pressures and two inductions are required for pressures above 9.65 MPa gauge (1400 psi) to be advantageous.

The effect of post firing the boiler of a combined gas-steam turbine cycle is shown in Figure 1.10. Only steam conditions of 16.547 MPa gauge/811°K/811°K (2400 psig/1000°F/1000°F) are shown. The unfired boiler plant efficiency is included on the right for comparison. Firing the boiler increases the production of steam so induction of lower pressure steam is unnecessary. Compared to the unfired boiler cycle without induction, a little boiler firing improves efficiency about two points by balancing the heat sink in the feedwater with the heat available in the exhaust gas. Additional firing significantly worsens efficiency as shown by the three columns on the left of the figure. The unfired boiler with two inductions is more efficient than any fired boiler cycle.

Firing of the boilers adversely affects efficiency for both the nonreheat and reheat steam cycles. The capital cost, efficiency, and power cost of the nonreheat cycle decrease more with boiler firing than those of the reheated steam cycles.

Combined cycle efficiency improves significantly with increased gas turbine firing temperature. At the base firing temperature for the study of 1478°K (2200°F), the improvement is about two points per 55.6°K (100°F) in firing temperature tapering to about one point per 55.6°K (100°F) at the 1700°K (2600°F) level. A gas turbine compression ratio of 12 to 1 is close to optimum for combined cycles at all firing temperatures. Comparison of plants using integrated coal gasification and distillate from coal shows that the gasification system degrades the plant efficiency

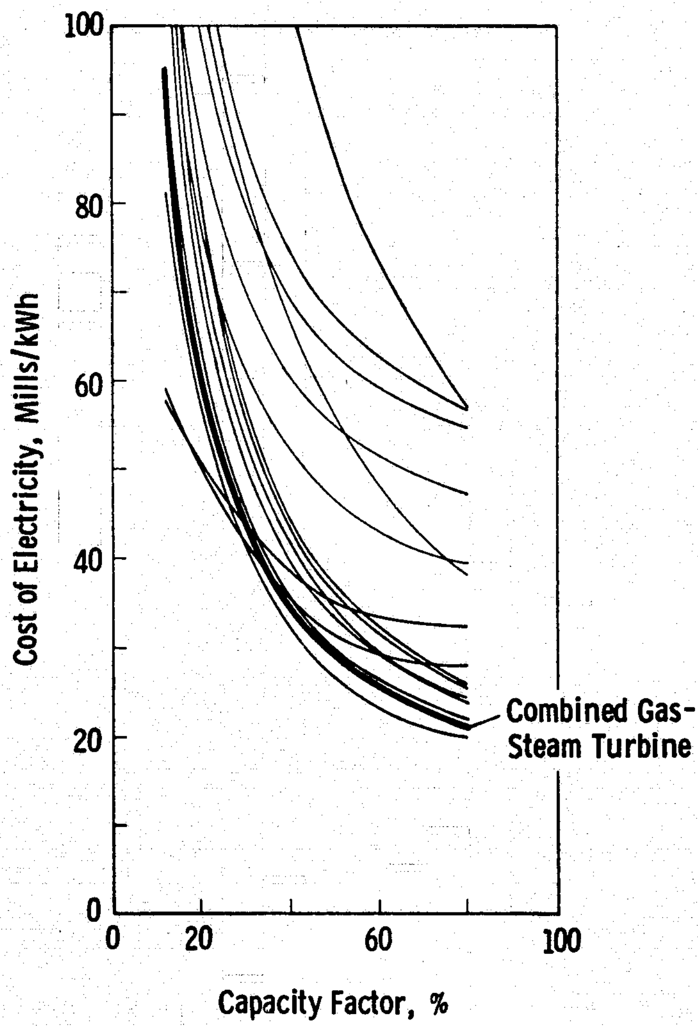

Fig. 1.11—Cost of electricity from combined gas-steam turbine plants

by about 8% (3.9 points). With a 1478°K (2200°F) gas turbine the best efficiency of a plant with gasification is about 42.3%

Figure 1.11 shows a COE of 6.74 mills (24.25 mills/kWh) for the combined cycle with gasifier. The distillate burning cycle is not shown but has a COE of 7.68 mills/MJ (27.65 mills/kWh) at the 65% capacity factor.

The coal burning plant has a higher capital cost because of the cost of the gasifier and related equipment. The lower cost of fuel for the coal burning gasifier cycle, \$0.805/GJ (\$0.85/10^6Btu) compared to \$2.46/GJ (\$2.60/10^6 Btu) for distillate, more than compensates for the additional capital cost and results in the lower cost of electricity.

Combined gas-steam turbine cycles with an integrated coal gasifier clearly offer lower cost electricity than oil from coal.

1.2.6 Closed-Cycle Gas Turbine Systems

Section 7 deals with both recuperated and combined closed-cycle gas turbine systems. The combined-cycle systems include both steam and organic vapor Rankine bottoming cycles.

Major subsystems of the recuperated closed-cycle gas turbine system are the recuperated pressurizing or pump-up gas turbine and the recuperated helium turbine which are coupled by a pressurized furnace. Pump-up gas turbine inlet temperatures of 1478, 1200, and 866°K (2200, 1700, and 1100°F) are used. The two lower temperatures are compatible with direct fluidized bed combustion of coal. Helium turbine inlet temperatures of 922, 1089, and 1255°K (1200, 1500, and 1800°F) with pressure ratios 2, 2.5, 3, and 4 to 1 are studied. The helium compressor discharge pressure is fixed at 6.89 MPa gauge (1000 psi) abs with variations of 3.45 and 10.34 MPa gauge (500 and 1500 psi) abs. Values of recuperator effectiveness of 80, 90, and 95% are assumed for both the pump-up and

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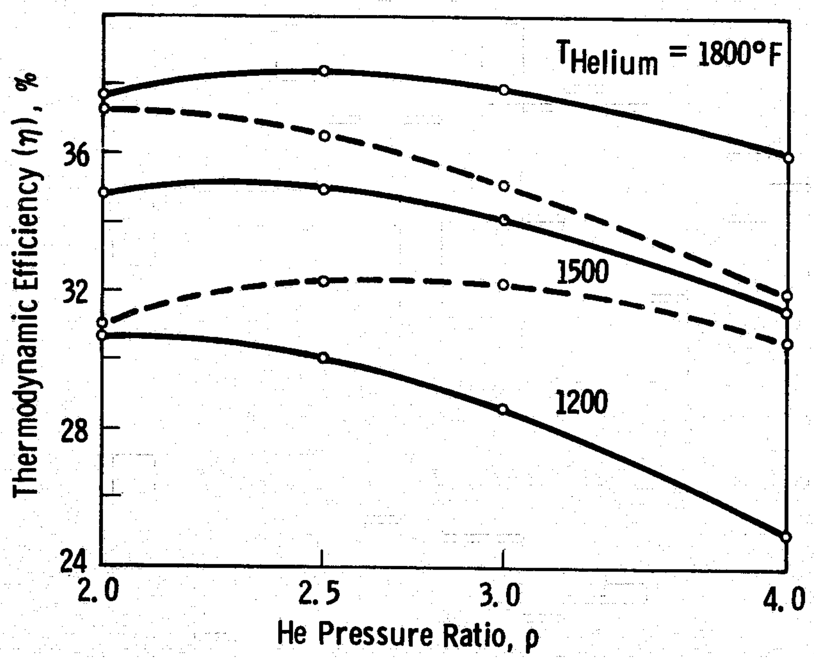

Fig. 1.12—Closed-recuperated gas turbine plant efficiency

helium turbine exhausts. Clean distillate fuel is used for the major part of the study but several cases using coal are considered. Figure 1.12 shows efficiencies of 38% are attainable at the 1255°K (1800°F) helium turbine inlet temperature with a recuperator effectiveness of 90%. These curves are based on a 1478°K (2200°F) - 10 to 1 pressurizing gas turbine. A 4.5 point increase in efficiency at the 1089°K (1500°F) helium turbine inlet temperature is observed as the recuperator effectiveness is increased from 80 to 90% at a 2.5 to 1 pressure ratio as shown by the difference in the two dashed curves. To contain 1000 psia helium at 1255°K (1800°F) is considered a major problem requiring the use of refractory alloys.

The combined closed-cycle gas turbine system using pump-up and helium gas turbine engines similar to those used in the recuperated cycle employs a reheat steam bottoming cycle. Heat from both the pump-up and helium turbine exhausts is used for heating the bottoming fluid. Pinch-point temperature differences made it necessary to terminate the helium to bottoming fluid heat transfer at about 582°K (588°F). The helium was then cooled to the required helium compressor inlet temperature in the precooler. Thermodynamic efficiencies of 43 to 45% were obtained at the 1255°K (1800°F) helium turbine temperature.

The high cost of the high temperature gas-to-gas heat exchangers results in high plant capital costs ($700/kW for a coal-burning plant and $500/kW for a plant-burning distillate). Notwithstanding this, the COE for these plants is strongly affected by fuel cost and the direct coal-burning plants shown will always have a lower COE. The composite plot, Figure 1.13, of COE vs capacity factor shows only the combined system with the steam bottomer having a COE of 8.47 mills/MJ (30.5 mills/kWh). The COE curve for the recuperated plant is similar and approximately 0.278 mill/MJ (1 mill/kWh) higher.

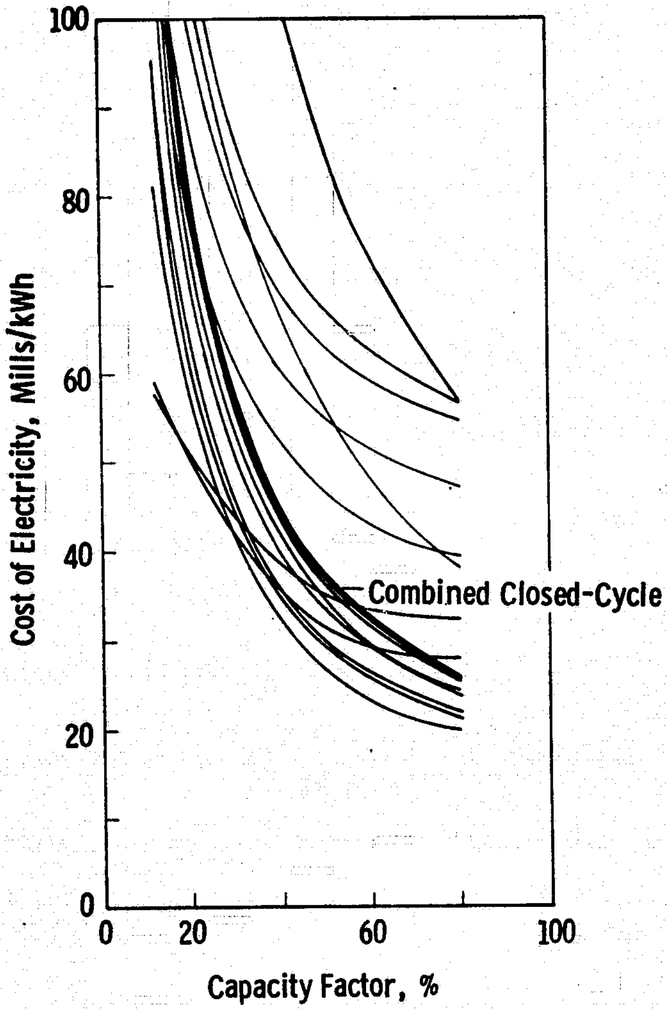

Fig. 1. 13 – Cost of electricity from a combined closed-cycle gas turbine plant

1.2.7 Metal Vapor Rankine Topping-Steam Bottoming Cycle

Section 8 discusses a metal vapor Rankine topping-steam bottoming cycle as a way to increase the mean temperature at which heat is added to the cycle and to raise the efficiency of the power plant. The majority of the study uses potassium as the working fluid with a few cesium points for comparison. The systems studied use either a pressurized fluidized bed boiler burning coal directly or a pressurized boiler burning clean fuel gas from an integrated low-Btu gasifier. Included in the cycles are a pressurizing gas turbine with its associated recuperator, and a gas economizer and feedwater heater. The base case system assumes a 1255°K (1800°F) pressurizing turbine inlet temperature and a 15 to 1 pressure ratio. The liquid-metal vapor generator is a fluidized bed boiler. The liquid-metal system uses a boiler with a 2.5 to 1 recirculation ratio, and several four-stage - 30 rps (1800 rpms) double flow - 25 MW turbine-generators which exhaust into a metal vapor condenser-steam boiler where steam is raised for a nearly conventional steam-bottoming plant.

The metal vapor enters the turbine at 1033°K (1400°F) and the condenser-steam generator at 866°K (1100°F). The steam-bottoming plant uses a 24.132 MPa (3500 psi) either single or nonreheat plant. The high pressure feedwater heating is accomplished partly by extraction steam and partially by exhaust gas feed heating. A temperature difference of 166.7°K (300°F) is assumed across the metal vapor turbine. The steam reheat and/or superheat temperature is 55.5°K (100°F) less than the metal vapor condensing temperature. These variables are not varied independently.

Calculations show the potassium-topped plant with a capitalization of $667/kW and a plant efficiency of 42.3%.

Results show the comparable cesium cycle to have an efficiency about 0.5 point higher than the potassium cycle but to have a 0.44 mill/MJ (1.6 mills/kWh) higher cost of electricity. The need for both the gasifier and pressurized furnace compared to just a pressurized fluidized bed boiler results in a 17% higher plant capitalization. The pressurized fluidized bed system is the choice for the case for further study. Also

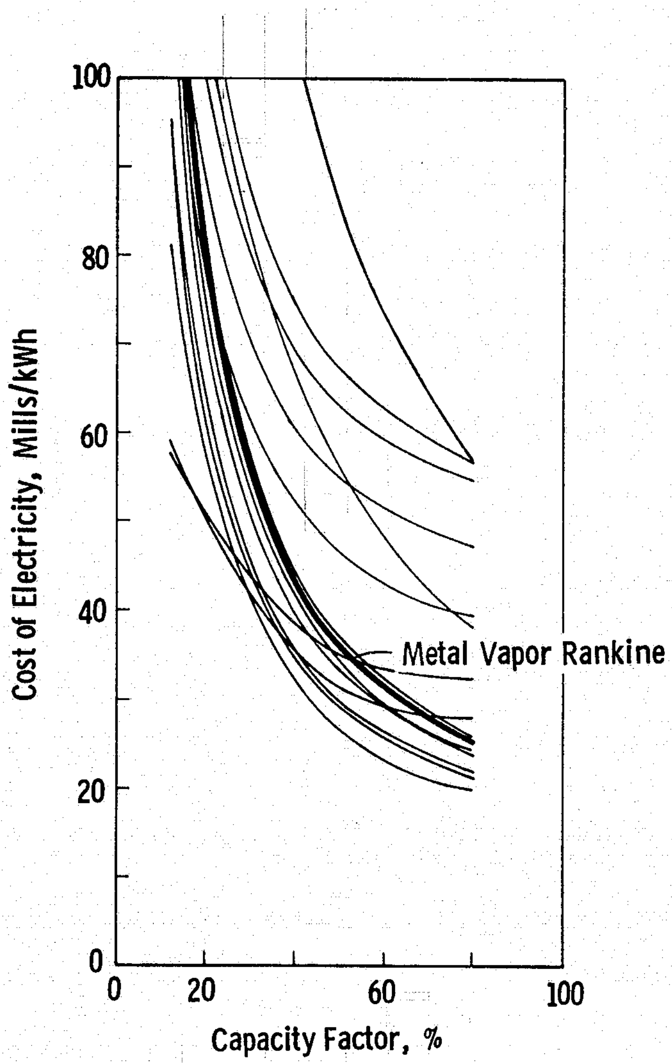

Fig. 1.14—Cost of electricity from a metal vapor Rankine topping-steam bottoming plant

indicated are a 10 to 1 - 1255°K (1800°F) pressurizing gas turbine, a 1033°K (1400°F) metal turbine inlet temperature, and a 24.132 MPa/811°K/811°K (3500 psi/1000°F/1000°F) steam-bottoming plant.

The 1200 MW plant, made up of several distinct pressurized boiler and liquid-metal turbine loops with the exception of the steam turbine which is common to all loops, can be expected to have a higher availability than a normal plant with line dependence on all major components.

The composite plot, Figure 1.14, shows a plant with a cost of electricity of 8.19 mills/MJ (29.5 mills/kWh). Extrapolation to other conditions than those calculated shows possible efficiencies of 44% with a possible capital cost of $583/kW and a COE of 6.94 mills/MJ (25 mills/kWh).

1.2.8 Open-Cycle MHD

Section 9 looks parametrically at three open-cycle MHD systems: a direct coal-burning system, a system with a separately fired air preheater and a system firing low-Btu gas from an integrated gasifier. Only the system with the lowest COE, the direct fired system, will be described here. Dried crushed coal is fed from the coal processor to the single-stage cyclone combustor. Air from a steam turbine-driven compressor is preheated to 1622°K (2460°F) as it cools the duct walls between the MHD generator and the heat recovery steam generator. It is then introduced with the coal into the combustor so that a fuel-rich mixture exists (95% stoichiometric air fuel ratio) and fired to the duct inlet condition of 2700°K (4400°F). Eighty percent of the coal ash is assumed to have been removed in the cyclone combustor. Potassium carbonate is added in the mixer as a seed to improve the plasma conductivity and to combine with the sulfur in the coal. The MHD duct inlet conditions are assumed to be 2700°K (4400°F), 0.6195 MPa (6 atm), and a Mach number of 0.75. A superconducting magnet establishes a magnetic induction of 6 T. A generator-loading coefficient of 0.82 is assumed.

The air heater is of a radiant design and the walls are assumed to be protected from corrosion due to molten slag and seed by an injected

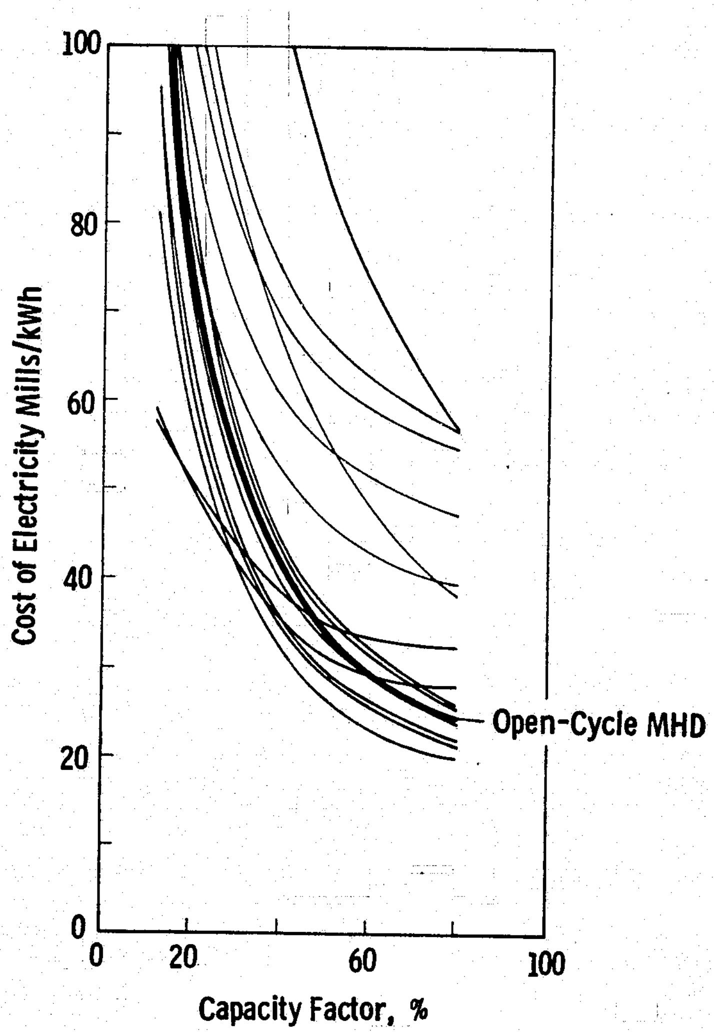

Fig. 1.15—Cost of electricity for an open-cycle MHD plant

layer of air or recycled combustion product along the wall. The exhaust products at 1650°K (2511°F) are then passed to the heat recovery steam generator, the first part of which is a ceramic-coated radiant superheater. Molten seed-ash mixture is assumed to condense on this surface and drain off for collection. The remaining seed is quenched by injected air. The dry seed-ash mixture is passed through the remainder of the steam generator and collected by the stack gas cleanup system. The seed is then recycled through a Claus plant where part of the potassium sulfate is converted back to potassium carbonate and the processed seed with appropriate makeup reinjected into the mixer. The bottoming plant is a conventional 24.132 MPa/811°K/811°K (3500 psi/1000°F/1000°F) steam plant with a modified feedwater heater string to make use of available low-grade heat from various cooling duties. The bottoming plant efficiency is approximately 42%.

This plant has an efficiency of 47% and a total capital cost of \$633/kW. An MHD duct life of 946 Ms (30 yr) is also assumed. The other two MHD system base cases have slightly better efficiencies of 48 to 48.5%, but capital costs of 823 and 811 \$/kW with a resultant higher cost of electricity. The open-cycle MHD with its high efficiency does have the potential for a future base load power system. As shown in Figure 1.15, its COE is only 7.22 mills/MJ (26 mills/kWh) and this system will become more desirable as fuel costs increase further. A final judgment on the commercial viability of the open-cycle MHD system will require better estimates for the cost of superconducting magnets, recovery heat exchangers, and the air preheaters - these items represent approximately 40% of the total direct-system cost.

1.2.9 Closed-Cycle MHD

Section 10 describes the second MHD topping system, the closed-cycle MHD topping - Rankine steam bottoming plant. The system consists of an external heating loop, the cesium seeded argon closed-cycle non-equilibrium ionization MHD loop, and the steam bottoming plant. The external heating or firing system consists of four 20.16 kg/s (80 ton/hr)

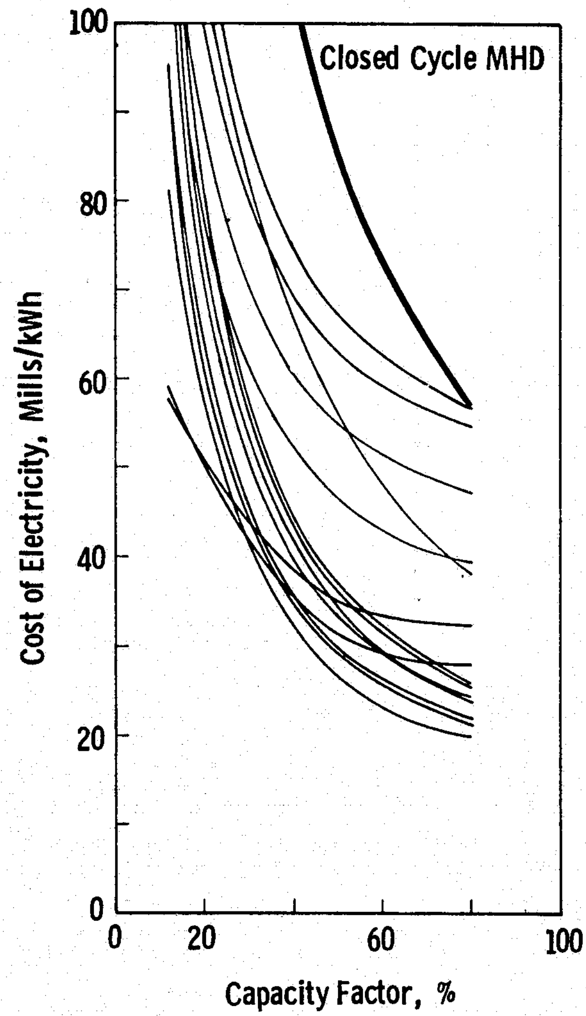

Fig. 1.16—Cost of electricity for a closed-cycle MHD plant

gasifiers. Around each gasifier are clustered an air preheater, seven argon preheaters, and 14 regenerative-type refractory stoves which heat the argon gas to the MHD duct inlet temperature of 2366°K (3800°F). The argon seeded with approximately 0.1% cesium and at a pressure of 0.942 MPa (9.3 atm) is accelerated to a duct inlet Mach number of 0.9 and expands down the duct. A magnetic field of 5 T is provided by a superconduction magnet. A generator coefficient of 0.75 is assumed.

The thermodynamic efficiency of the nonequilibrium closed-cycle MHD system is optimized to be approximately 59% after more than 900 separate case examinations.

The nominal plant output is 1000 MW of which the MHD duct provides more than 900 MW, the balance being supplied by the 24.132 MPa/811°K/811°K (3500 psi/1000°F/1000°F) steam turbine generator. Other similar bottoming steam turbines are used for the argon compressor drives. An optimistic bottoming plant efficiency of 45% was chosen rather than the readily attainable 43% for the steam turbine drives. A net overall efficiency of 46.1% is calculated for the plant. The capitalization of this plant is calculated to be $2,228/kW. A similar plant with a 1978°K (3100°F) top temperature had an overall efficiency of 42.2% and a capital cost of $1,913/kW. The COE, Figure 1.16, for this concept is, therefore, high [approximately 19.03 mills/MJ (68.5 mills/kWh) at a 65% capacity factor]. Assuming a breakthrough in design results in a 90% reduction in the cost of the external argon heating system which is 53% of the total direct cost, the COE would still be 10% mills/MJ (39.5 mills/kWh). In addition, the severe temperature duty would require major developments in the valving of the regenerative cyclic heat exchangers.

Internally-fired systems such as those fired by a HTGCR may make this system more attractive.

1.2.10 Liquid-Metal MHD Systems

Section 11 deals with liquid-metal MHD systems. Major emphasis was placed on a direct coal fire design using a bubbly two-component flow of sodium and argon in the MHD generator and a Rankine steam bottoming plant. Argon temperatures of 922 and 1089°K (1200 and 1500°F) at the

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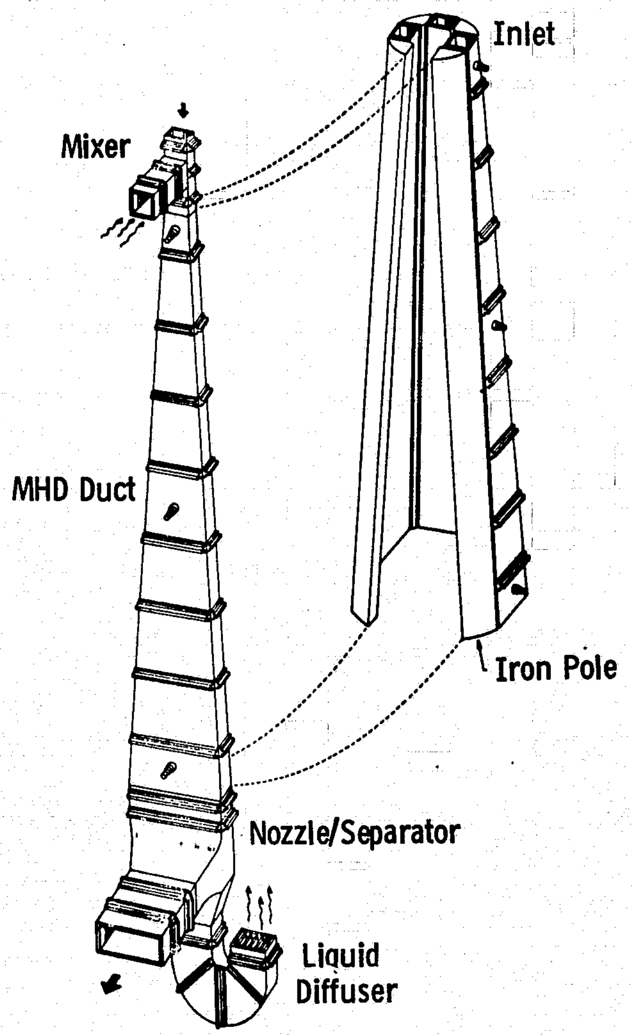


Fig. 1.17—Conceptual LM-MHD duct assembly

duct inlet were chosen. A two-component bubbly flow of sodium and argon at 8.27 MPa (1200 psi) expands to about 2.76 MPa (400 psi) in the duct system. The liquid metal and gas are then separated. The argon gives up heat to superheat, reheat, and generate steam and to preheat and the combustion air. It is then compressed and passed through the fired heater where its temperature is again increased to 1089°K (1500°F).

The MHD duct system consists of multiple ducts arranged in clusters and separated by iron magnet pole pieces. An artist's conception of one such cluster is shown in Figure 1.17. The ducts, each with an output of about 100 MW, are in parallel to the flow but connected in series electrically to provide a higher MHD voltage. Nonetheless, the inversion equipment costs are 20% of the total plant cost due to the high currents involved at low MHD output voltages.

Because of the large mass of liquid metal circulated, over 63,090 ℓ/s (1,000,000 gpm) at 5.51 MPa (800 psi) pressure difference, the liquid metal pump efficiency becomes a critical parameter.

With duct efficiencies of 80%, a pump efficiency of 90% and a 45% efficient steam-buttoming plant, the efficiency of the 1089°K (1500°F) liquid-metal cycle is only 43%. The complexity and high cost of the plant ($1,165/kW) result in a COE (Figure 1.18) greater than 12.5 mills/MJ (45 mills/kWh).

1.2.11 Advanced Steam Systems

Section 12 describes three advanced steam concepts. The first considers an atmospheric furnace burning coal directly. This is the plant common to the power industry today. The last two involve pressurized boilers. Pressurizing is accomplished by a coupled gas turbine engine in each case. The pressurized furnace requires clean fuel and is assumed to burn low-Btu gas from an integrated pressurized coal gasifier. The pressurized fluidized bed boiler accepts coal directly. Desulfurization is accomplished in the bed.

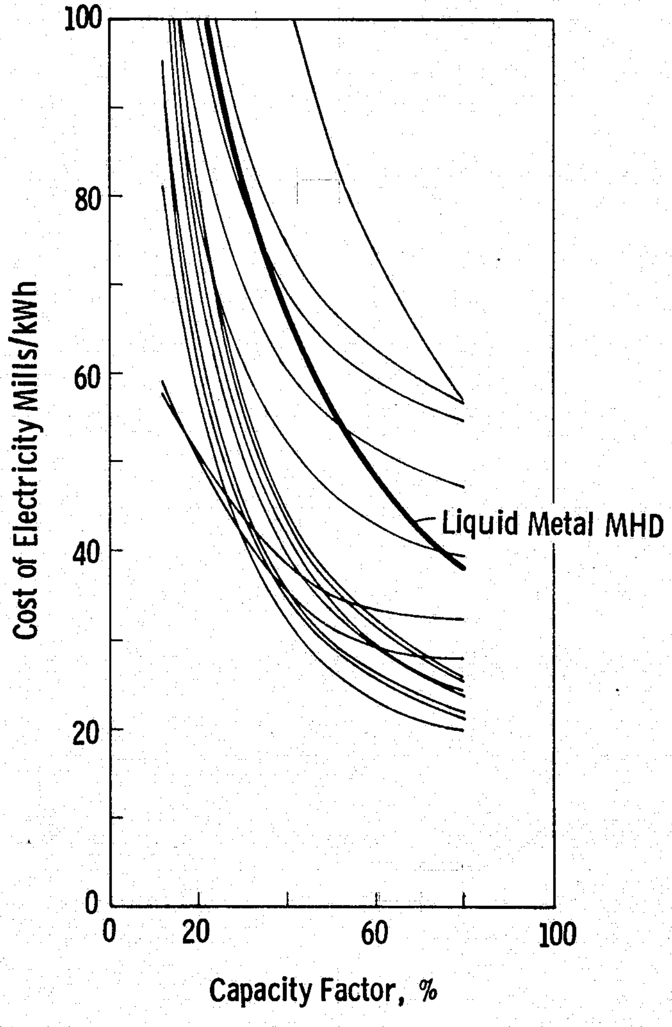


Fig. 1.18—Cost of electricity for a liquid metal MHD plant

Steam turbine throttle conditions were varied from the well established values of 16.547 MPa/811°K/811°K and 24.132 MPa/811°K/811°K (2400 psi/1000°F/1000°F and 3500 psi/1000°F/1000°F) in steps to 34.474 MPa/1033°K/1033°K (5000 psi/1400°F/1400°F) for all three systems. The results, shown in Figure 1.19, indicate that the efficiency increases by 1.5 points as the pressure increases from 16.547 to 34.474 MPa (2400 to 5000 psi) at a constant throttle and reheat temperature of 811°K (1000°F). The increase is more pronounced at higher throttle temperatures. It is also seen that the 24.132 MPa (3500 psi) plant efficiency increases by nearly four points as the throttle temperature is increased from 811 to 1033°K (1000 to 1400°F).

Although the efficiency of the plants with advanced steam turbine throttle conditions is higher, the cost of electricity, shown in Figure 1.20, is seen to be higher also, with the increased capital costs greatly offsetting the minimal fuel cost savings associated with the advanced conditions. Since high alloy materials are required in ever increasing amounts as throttle temperatures exceed 839°K (1050°F), no financial incentive was found to move toward higher throttle conditions.

The advanced steam systems with pressurized boilers require clean fuel so an integarted low-Btu gasifier was used. The pressurizing system for the furnace consists of a gas turbine with a 10 to 1 compression ratio whose turbine inlet temperature is varied from 1255 to 1644°K (1800 to 2500°F). Both extraction and gas feedwater heating are assumed to have been used. Results show a two-point increase in efficiency as the gas turbine inlet temperature increases from 1366 to 1644°K (2000 to 2500°F). Both the capital and fuel costs decrease with increasing gas turbine inlet temperature as does the resultant total cost of electricity.

The exhaust gas from the pressurized fluidized bed system after being cleaned is expanded through the gas turbine. For this reason the gas turbine inlet temperatures are fixed at temperatures slightly less than bed temperature, that is 1144 to 1255°K (1600 to 1800°F). Increasing the gas turbine inlet temperature from 1144 to 1255°K (1600 to 1800°F)

Curve 682918-A

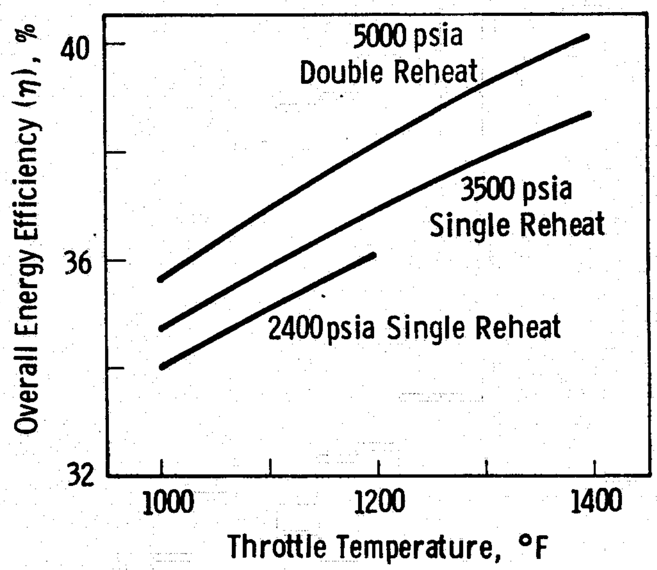


Fig. 1.19—Effect of steam turbine throttle temperature on overall efficiency

Curve 682917-A

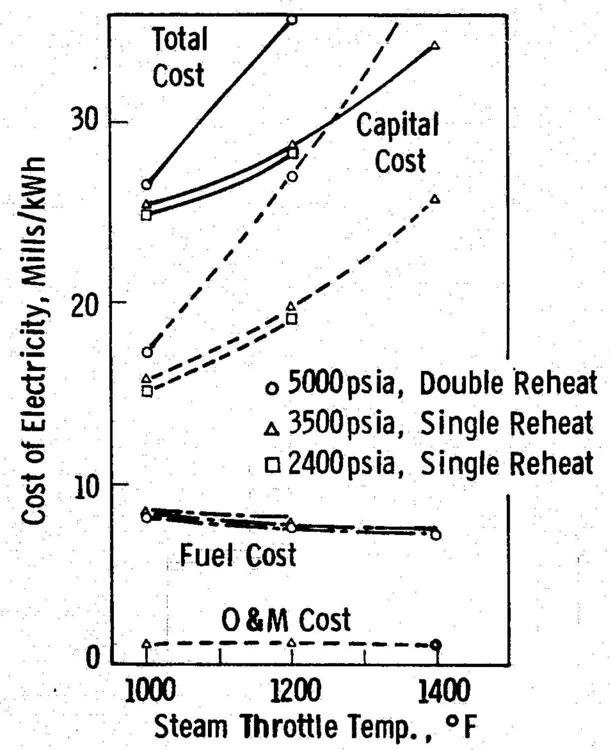


Fig. 1.20—Effect of steam turbine throttle temperature on cost of electricity

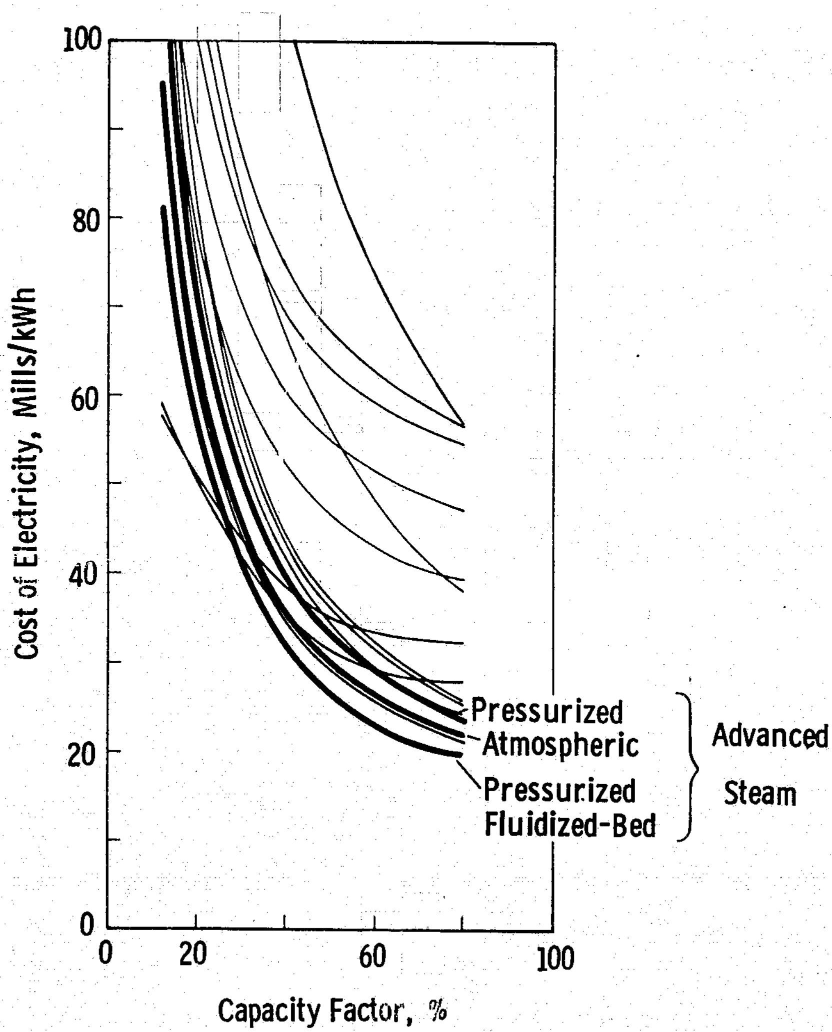


Fig. 1.21—Cost of electricity from advanced steam plants

increases the net cycle efficiency from 37.5 to 38.3%. The cost of electricity for the system is found to be between 6.11 and 6.39 mills/MJ (22 and 23 mills/kWh).

It is seen from Figure 1.21 that each of these systems has a competitive cost of electricity. The pressurized fluidized bed boiler system has the lowest COE of the three systems at 6.25 mills/MJ (22.5 mills/kWh). The COE for the base case atmospheric boiler system is 7.03 mills/MJ (25.3 mills/kWh). The pressurized furnace COE is slightly higher. It is noted that all of these systems are near-in, commercially viable power systems and that the difference in COE between the atmospheric and pressurized furnace results is within the error band in the cost estimates.

1.2.12 Fuel Cells

Section 13 covers four fuel cell systems classified by electrolyte type. The two high-temperature fuel cell systems considered are the solid electrolyte and the molten carbonate. Two low-temperature aqueous systems, the phosphoric acid and the alkaline, are also considered. The principal parameters studied are fuel cell useful life, current density, catalyst loading, voltage degradation, and electrolyte thickness. Heat recovery bottoming systems are also studied for the high-temperature systems.

The base case values used for all four fuel cell systems are a 36 Ms (10,000 hr) useful life and a 25 MW dc fuel cell system with high-Btu gas as the fuel and air as the oxidizer.

The comparison of the results for the base cases of each of the four fuel cell systems in Figure 1.18 shows plant efficiencies of 35.5 and 38.1% for the low-temperature fuel cell systems and 48.8 and 69.7% for the molten carbonate and oslid electrolyte fuel cell systems. The COE is estimated to exceed 13.89 mills/MJ (50 mills/kWh) for all systems but the solid electrolyte system.

The overall energy efficiencies (bus bar to coal) are around 25% for the low-temperature systems and 33 and 43% for the molten carbonate and solid electrolyte systems, respectively.

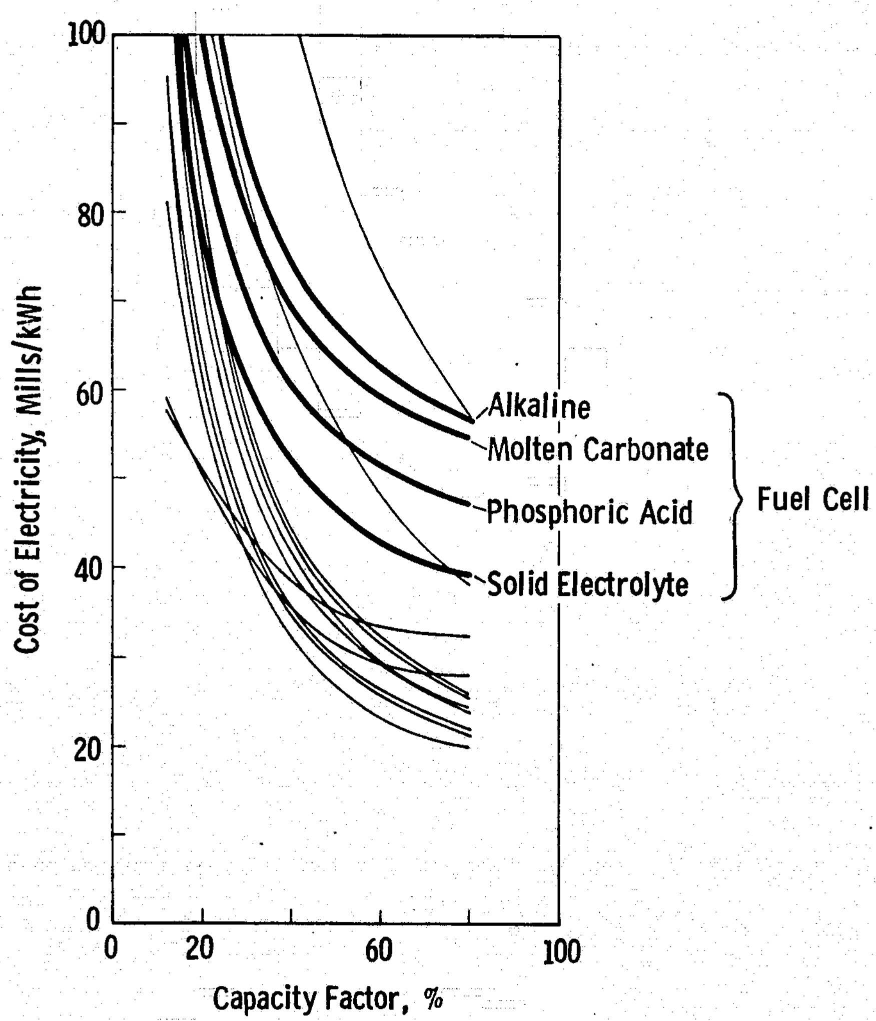


Fig. 1.22—Cost of electricity for the base case fuel cell systems

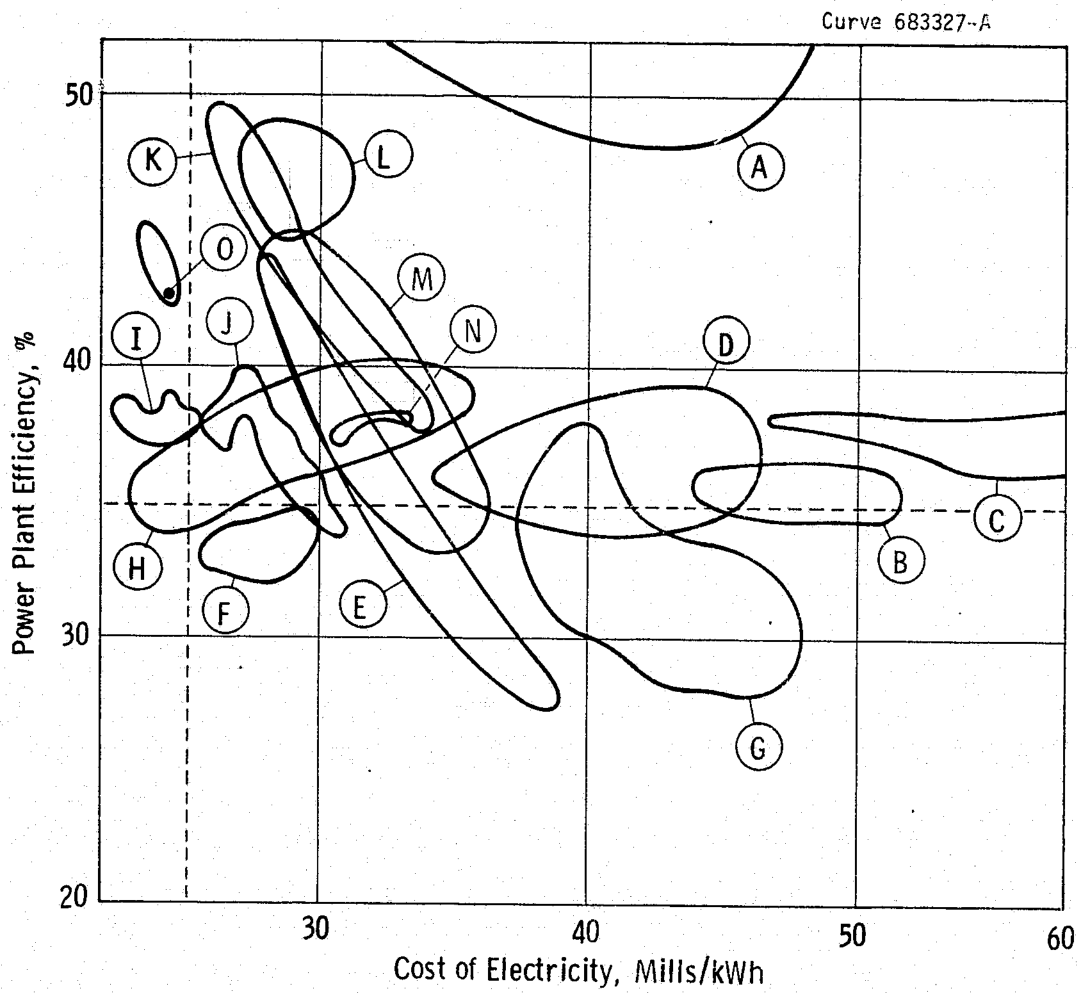
If the fuel cell useful life is increased from 36 to 360 Ms (10,000 to 100,000 hr), the cost of electricity for the acid system would decrease from 13.89 to about 11.53 mills/MJ (50 to about 41.5 mills/kWh). Like improvements are seen for the other systems. At 3/10th the base case catalyst loading, the COE would decrease from 13.89 to 12.78 mills/MJ (50 to 46 mills/kWh). Doubling the fuel cell power density reduces the COE from 13.89 to 11.72 mills/MJ (50 to 42.2 mills/kWh).

Fuel cell useful life, catalyst loading, and cell power density are the most important parameters treated in this study. For the high-temperature solid electrolyte system using medium-Btu gas as a fuel the overall efficiency is improved from 34 to 50% by the addition of a steam-bottoming plant to utilize the waste heat with a corresponding resultant decrease in COE from 14.64 to 11.67 mills/MJ (52.7 to 40.2 mills/kWh). Large high-temperature fuel cell systems should, therefore, use the available waste heat for process heat or for bottoming power generation.

Optimistic estimates of overall efficiencies and COE for the four fuel cell systems show the low-temperature systems with efficiencies of 30% and the cost of electricity in the 9.72 to 11.11 mills/MJ (35 to 40 mills/kWh) for the acid system and 11.11 to 12.5 mills/MJ (40 to 45 mills/kWh) for the alkaline system. The high-temperature systems show overall efficiencies greater than 45% with cost of electricity for the molten carbonate in the 8.35 to 9.72 mills/MJ (30 to 35 mills/kWh) range and 6.94 to 8.33 mills/MJ (25 to 30 mills/kWh) for the solid electrolyte system.

The lowest fuel cell COE found in this study was 9.72 mills/MJ (35 mills/kWh). The COE plot for each of the 25 MW fuel cell base cases is given in Figure 1.22.

A credit for locating the 25 MW plants near the load center could reduce the cited COE's by 0.55 to 5.55 mills/MJ (2 to 20 mills/kWh).



A-Fuel Cell Steam Bottoming
  1-Molten-Carbonate
  2-Solid Electrolyte
B-Fuel Cell (Phosphoric Acid)
C-Alkaline Fuel Cells
D-Liquid Metal MHD
E-Recuperated-Open-Cycle Gas Turbine
F-Closed Recuperated Gas Turbine (Coal)
G-Recuperated-Closed-Cycle Gas Turbine
H-Steam (Atmospheric Boiler)
I-Steam (Pressurized Fluidized Bed Boiler)
J-Steam (Pressurized Boiler)
K-Distillate-Burning Combined-Gas Turbine
L-Open-Cycle MHD
M-Metal Vapor Rankine Topping Cycle
N-Combined-Closed-Cycle Gas-Steam Turbine
O-Coal Burning Combined-Cycle-Gas Steam Turbine

Fig. 1.23—Advanced energy-conversion systems--Range of results

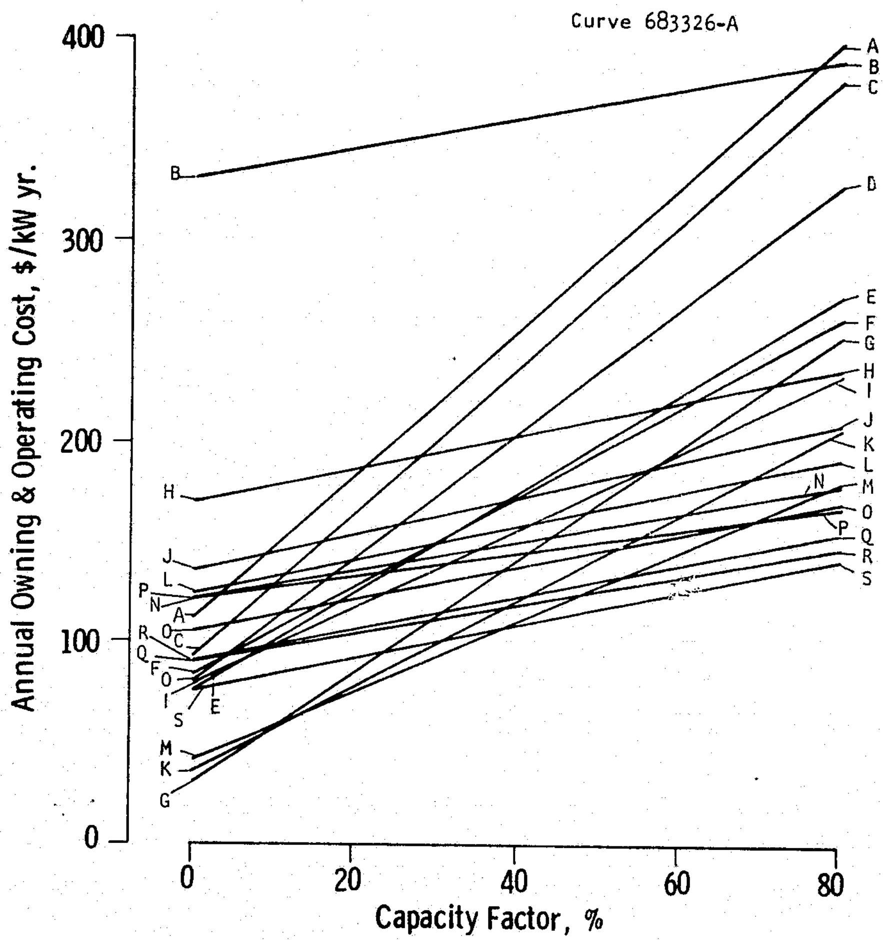
1.3 Comparison of Results

Some relative comparisons of results have been given previously in curves of COE versus capacity factors for each of the concepts. Figure 1.23 is a composite of the majority of the data, with COE plotted versus plant efficiency. The base case steam plant with a 34.7% plant efficiency, a capital cost of $499/kW, and a cost of electricity of 7.03 mills/MJ (25.3 mills/kWh) is indicated by the intersecting dashed lines on the figure which divide it into quadrants. An update of the Wash 1230 estimate of the total steam power plant capital costs agrees closely with the $499/kW capital cost calculated for the base case. Figure 1.24 gives another representation of the data, showing the annual owning and operating cost of the typical plant as a function of capacity factor. The intercepts represent the plant capital cost in $/kW times a fixed charge rate of 18%.

The advanced steam plant with the pressurized fluidized bed boiler has the lowest COE, 6.25 mills/MJ (22.5 mills/kWh), a capital cost of $419/kW, and a plant efficiency of 38.3%. The combined gas-steam turbine plant has the next lowest COE. The combined-cycle gas turbine inlet temperature assumed is only 1478°K (2200°F), yet this plant has an efficiency of 42.6%, a capital cost of $496/kW, and a COE of 6.74 mills/MJ (24.25 mills/kWh). Increasing the gas turbine firing temperature to 1589°K (2400°F) should increase the plant efficiency to 45.5%, further enhancing its position relative to the conventional steam plant. Further advances in the cost of fuel would result in a lower relative COE for advanced combined gas turbine plants.

On the basis of these very preliminary studies, Westinghouse concludes that, in fairness, these concepts should be divided into two groups: those which represent near-in technology with relative certainty of results, and those which require significant technological development to attain reasonable operating life and performance.

Power generation for peaking and intermediate duty with start-up and shut-down at least daily should fall to the simple or recuperated



A-Fuel Cell (Alkaline)
B-Closed-Cycle MHD
C-Fuel Cell (Molten-Carbonate)
D-Fuel Cell (Phosphoric Acid)
E-Fuel Cell (Solid Electrolyte)
F-Fuel Cell (Solid Electrolyte Steam Bottoming)
G-Open-Cycle Gas Turbine
H-Liquid Metal MHD
I-Combined-Closed-Cycle Gas-Steam Turbine
J-Recuperated-Closed-Cycle Gas Turbine
K-Recuperated-Open-Cycle Gas Turbine
L-Recuperated Intercooled Gas Turbine
M-Distillate-Burning Combined-Gas Turbine
N-Metal Vapor Rankine Topping Cycle
O-Steam (Pressurized Boiler)
P-Open-Cycle MHD
Q-Steam (Atmospheric Boiler)
R-Coal Burning Combined-Cycle-Gas Steam Turbine
S-Steam (Pressurized Fluidized Bed Boiler)

Fig. 1.24—Annual owning & operating costs as a function of capacity factor for the ECAS Task I concept

open-cycle gas turbine. Simple-cycle gas turbine systems previously used for this duty when convenience fuels were cheap and readily available should give way to recuperated systems with turbine inlet temperature in the 1478 to 1644°K (2200 to 2500°F) range. The development of combustion technology to minimize thermal NO_x formation and the development of air-cooled or ceramic turbines should be the subject of major efforts.

Base load duty will see the use of combined gas-steam turbine plants and steam plants with fluidized bed boilers. These plants show the greatest potential for economical power generation of any investigated. Only modest advances, if any, in steam turbine throttle conditions are forecast.

Those concepts which are not extensions of existing commercial technology, in general, need more study in areas where new equipment technology is required and especially in areas of novel equipment costing and useful life.

Those with the greatest potential, that is highest efficiency, appear to be the metal vapor Rankine topping cycles, MHD concepts, and the high-temperature fuel cell systems.

The metal vapor Rankine topping-steam bottom cycle represents a modest advance over conventional steam plants but has an expected upper efficiency limit of approximately 45%. The technology is not all state of the art but requires only specific hardware development to be so. The cycle is complicated and should high-temperature gas turbines become a reality may not show a sufficient efficiency improvement to justify its development because its complexity indicates relatively high capital cost.

Of the three MHD concepts, only the open-cycle offers the potential for moderately low-cost electricity with a cycle efficiency of 47 to 49%. First, however, necessary duct materials developments must be made and their life demonstrated. A careful review of the high-temperature air preheater is certainly necessary as well as the investigation of ash-seed corrosion of ceramic and metal parts at very high temperature.

The high-temperature fuel cell systems have yet to be demonstrated in any reasonable size. They have high potential efficiency but it remains for a prototype system to demonstrate useful life before further system studies are warranted here.

Although metal vapor Rankine topping and high-temperature fuel cells have sufficiently high efficiency to warrant interest, unless their projected capitalization can be substantially reduced, they will still not be able to compete with advanced steam or combined-cycle plants - even with assumed coal cost of \$2.37/MJ (\$2.50/10^6 Btu) approximately three times the \$0.806/MJ (\$0.85/10^6 Btu) used in this study. The COE for advanced steam and combined-cycle plants is only 10.28 mills/MJ (37 mills/kWh) compared to over 11.11 mills/MJ (40 mills/kWh) for the futuristic systems.

Westinghouse recommends priority be given to the study of near-in power conversion concepts for Phase II and III of ECAS. Primary consideration should be given to direct coal using advanced steam and combined gas-steam turbine systems for base load duty. Advanced simple or recuperated open-cycle gas turbines with their low capital cost have merit for peaking or intermediate duty (capacity factors less than 40%) even though they require convenience fuels.

More detailed design and costing information on the several types of high temperature heat exchangers and the superconducting magnet system associated with the direct fired coal burning MHD system is required before the economics of this relatively efficient concept can be properly evaluated. It is, therefore, suggested that open-cycle MHD also be studied in Phase II and III. The effort should concentrate on detailing potential designs and estimation of their cost for these components.

2. GENERAL ASSUMPTIONS

During the course of Task I of this study, information was supplied to NASA Lewis which embodied the current practice or data available in the literature to aid them in setting the design values to be used. These recommendations appeared in the monthly technical progress reports submitted to NASA, and since NASA-specified values are actually used for this study, this information will not be repeated here.

Although considerably larger vessels can be shipped by rail, at very high cost, this study arbitrarily assumes that components of 3.9 m (14 ft) diameter by 15.24 m (50 ft) long or 3.96 m (14 ft) by 5.18 m (17 ft) high in shorter lengths can be shipped. Special cars are available in very limited supply to carry 272 to 453.6 Mg (300 to 500 tons).

2.1 Service

This study is to consider all plants for base-load service, with some consideration given to the applicability of the plant to intermediate or peaking service. Base-load service is defined as that equipment historically used in any given utility system to provide 70% of the system's total electrical energy output in any one year. Historically, it has been found that 70% of the total electrical energy output has been generated by plants with capacity factors greater than about 55%; further, that the average capacity factor for base-load plants was 67.5%.

Peaking service is defined as that equipment historically used in any given utility system to provide the last 2% (98 to 100%) of that system's total electrical energy output. This energy has historically been generated by plants operating with capacity factors less than 24%.

Intermediate service is defined as that equipment historically required to produce the remaining 28% of the total electrical output.

This equipment has been found to have typical capacity factors between 24 and 55%, with an average of 45%.

In order to display properly the effect of capacity factor on the cost of electricity, each plant will be evaluated at five capacity factors: 12%, peaking; 45%, intermediate; and 50%, 65%, and 80% for base-load service. The following criteria are suggested for use in evaluating each concept design for base, intermediate, and peaking service:

- Time for cold start to full power
  - Base load - not applicable
  - Intermediate - not applicable
  - Peaking - 600 s (10 min) emergency, 1500 s (25 min) normal
- Time from hot start to full power
  - Base load - not applicable
  - Intermediate - not applicable but must stand 5% per minute load ramp
  - Peaking - 120 s (2 min) emergency, 720 s (12 min) normal
- Expected service life at defined capacity factor
  - Base load - 946.08 Ms (30 years)
  - Intermediate - 946.08 Ms (30 years)
  - Peaking - 630.72 Ms (20 years)

It is further assumed that each plant is designed to have an availability of no less than 90%.

2.2 Transmission Voltage

It is realized that a new power plant, if small, will probably be attached to an existing grid. New large plants may be used to establish a new grid working at a higher voltage level than currently used in that locale. NASA specified all output from large systems be connected to a 500 kV grid. Westinghouse may also use 69 kV for small plants

(< 50 MWe) and 230 kV for 50 to 300 MWe plants. The delivered power will be at a frequency of 60 Hz, and this study includes costs of all equipment through the transformer high-voltage bushing but does not include the distribution switchyard.

2.3 Fuels and Fuel Costs

All fuels used in this study are coal or coal-derived fuels.

2.3.1 Fuel Properties

2.3.1.1 Coal

The three coals specified for use in this study were an Illinois No. 6 bituminous (Macoupin County) to be shipped from Paducah, Kentucky; a Montana subbituminous (Rosebud Seam, Rosebud County) to be shipped from Billings, Montana; and a North Dakota lignite (Mercer County) to be shipped from Bismarck, North Dakota. The coal properties given by the Bureau of Mines (Bruceton, Pennsylvania) and specified by NASA are included here for completeness (Table 2.1). Although the NASA trace element specification did not include chlorine, 400 to 4000 ppm were assumed when considering fire-side corrosion. Coal transportation charges were found to be about 4.79 mills/Mg-km (7 mills/ton-mi) and were projected to increase to about 13.69 mills/Mg-km ($\sim$ 20 mills/ton-mi) by 1990. For this study 6.48 mills/Mg-km (10 mills/ton-mi) will be used as the transportation charge for other required raw materials (limestone and potassium or cesium ore) with a net distance of 804.7 km (500 mi) by rail from the mine to the Middletown, USA site.

2.3.1.2 High-Btu Gas

It is assumed that high-Btu gas derived from coal via a methanation of nearly all the carbon monoxide and hydrogen contained in the gaseous product of an oxygen-blown gasifier (Reference 2.1) has the properties given in Table 2.2.

The product gas from the different gasification processes would be fairly uniform in composition and in heating value per scf. The values cited above would be applicable to gas produced via a commercial Lurgi

Table 2.1 - Coal Properties

	Illinois No. 6 (Macoupin County)	Montana Subbituminous Rosebud Seam (Rosebud County)	N. Dakota Lignite (Mercer County)
Reference Material	BOM TP - 641	BOM TP - 529	BOM RI - 7158
Proximate Analysis (As Received)			
Moisture	13.0	24.3	36.7
Volatile	36.7	28.6	26.6
Fixed Carbon	40.7	39.6	30.5
Ash	9.6	7.5	6.2
Ultimate Analysis (As Received)			
Ash	9.6	7.5	6.2
Sulfur	3.9	0.8	0.7
Hydrogen	5.9	6.1	6.9
Carbon	59.6	52.2	41.1
Nitrogen	1.0	0.8	0.6
Oxygen	20.0	32.6	44.5
Higher Heating Value*	10,788	8,944	6,890
Lower Heating Value*	10,230	8,372	6,248
Average Softening Temperature °F	1979	2224	2280
Initial Deformation Temperature °F	1990-2130	2120-2410	2190-2400
Fluid Temperature °F	2090-2440	2180-2520	2330-2500
Ash Analysis			
SiO_2	46.6	22.1	17.9
Al_2O_3	19.3	15.5	9.9
Fe_2O_3	20.8	6.4	10.2
T_1O_2	0.8	1.2	0.3
P_2O_5	.24	.11	0.4
CaO	7.7	18.9	23.6
MgO	.9	6.6	6.7
Na_2O	.2	1.0	7.4
K_2O	1.7	.4	0.4
SO_3	2.4	26.2	21.8
Grindability H.G.I.			
Range	52-66	49-59	36-75
Average	55	53	50

*As received.

Table 2.1 (cont.)

	Illinois No. 6 (Macoupin County)	Montana Subbituminous Rosebud Seam (Rosebud County)	N. Dakota Lignite (Mercer County)
Free Swelling Index			
Range	1-6.5	- - -	- - -
Average	4.5	- - -	- - -
Trace Element Analysis: ppm in Coal			
Beryllium	0.6-7.6	1.0-1.1	0.1-3.9
Fluorine	50-167	60-70	60-70
Arsenic	8-45	1.2-25	- - -
Selenium	- - -	0.8	- - -
Cadmium	- - -	0.04	- - -
Mercury	0.04-0.49		0.07-0.09
Lead	8-14	3.6	5-10
Boron	13-198	84-92	78-201
Vanadium	8.7-67	14-18	5.3-29
Chromium	5-54	5-7	2.6-19
Cobalt	1.2-10	2	0.7-7
Nickel	5-37	4-6	1.5-15
Copper	3.1-25		2.8-16
Zinc	0-53	10-12	0-23
Gallium	1.5-8	3.4-3.5	1.0-13
Germanium	0.4-27	2-3	0-7
Molybdenum	0.6-8.5	8-30	0.1-3.4
Tin	0.1-5	5-15	0.2-4.3
Yttrium	1-13		1-27
Lanthanum	0.2-24		0-22
Uranium	10		50-240
Trace Element Analysis %W in Ash			
Lithium	.017-.039	.0215	.010-.022
Scandium	.007-.008	.0034	.003-.005
Manganese	.020-.062	.0456	.030-.046
Strontium	.058-.070	.2612	.061-.066
Barium	.029-.047	.3000	.265-.300
Ytterbium	.0003-.0011	.0004	.0003-.0011
Bismuth	.0001-.0002		.0001-.0002

unit, a Hygas unit (under development), or a Koppers-Totzek unit (gasifier is commercial but large-scale methanation units are not yet commercial).

Table 2.2 - High-Btu Gas Composition

CO_2	0.39% (by volume)
CO	0.08
H_2	2.49
N_2	2.81
S	--
CH_4	94.23
HHV	959.2 Btu/scf = 22,549.6 Btu/lb
LHV	864 Btu/scf = 20,311.9 Btu/lb
Mol wt	16.15
Sensible heat @ 60°F	64.15 Btu/lb
(Based on a reference temperature of 400°R)	

Yields of 0.32, 0.264, and 0.206 kg of pipeline gas per kg of coal are assumed for the three coals: bituminous, subbituminous, and lignite, respectively. These yields, which assume recovery of 67% of the coal's energy in the pipeline gas (Reference 2.2), may be optimistic. Recoveries as low as 59% may occur, depending on the use of coal within the plant to raise low-pressure steam and whether the tars and oil from the process are fired to recover heat.

2.3.1.3 Intermediate-Btu Gas

The composition and yield of the intermediate-Btu gas from a Koppers-Totzek gasifier after low-temperature desulfurization are given in Tables 2.3 to 2.5 for the three types of coals respectively. These numbers were used in this study. It should be noted that the coals were assumed to have been partially dried before entering the gasifier. Two values are given for product fuel gas to coal weight ratio, one based on the coal moisture content at the gasifier inlet and the other based on the as received coal moisture content.

Table 2.3 - INTERMEDIATE-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Koppers-Totzek/Stretford Desulf.

COAL Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature 150

Moisture Content 3

PROCESS OXYGEN

Air/Coal Ratio .790 (.709)

Temperature - °F 220

Pressure - psia 14.7

PROCESS STEAM GASIFIER

Steam/Coal Ratio .290 (.260)

Temperature - °F 400

Pressure - psia 14.7

PRODUCT FUEL GAS

Temperature - °F 100

Pressure - psia 14.7

Composition-Mole Fraction

N_2 .0043

O_2 ---

H_2 .3276

CO .5460

CO_2 .0573

H_2O .0647

H_2S ---

CH_4 ---

C_2H_4 ---

Product Fuel Gas/Coal Ratio 1.87 (1.68)

Molecular Wt 19.76

Heating Value

LHV - 265.0 Btu/scf 5089.2 Btu/lb

HHV - 281.5 Btu/scf 5405.9 Btu/lb

LHV/HHV .9414

Enthalpy (400°R Base) - 57.16 Btu/lb

Stoichiometric Fuel/Air Ratio 0.329

(Values in parenthesis are for as received coal)

Table 2.4 - INTERMEDIATE-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Koppers-Totzek/Stretford Desulf.

COAL Montana subbituminous

Lockhopper Inlet Conditions

Temperature 150

Moisture Content 20

PROCESS OXYGEN

Air/Coal Ratio .643 (.608)

Temperature - °F 220

Pressure - psia 14.7

PROCESS STEAM GASIFIER

Steam/Coal Ratio .130 (0.123)

Temperature - °F 400

Pressure - psia 14.7

PRODUCT FUEL GAS

Temperature - °F 100

Pressure - psia 14.7

Composition-Mole Fraction

N_2 .0045

O_2 ---

H_2 .3196

CO .5356

CO_2 .0757

H_2O .0647

H_2S ---

CH_4 ---

C_2H_4 ---

Product Fuel Gas/Coal Ratio 1.38 (1.31)

Molecular Wt 20.27

Heating Value

LHV - 259.4 Btu/scf 4858.3 Btu/lb

HHV - 275.5 Btu/scf 5159.5 Btu/lb

LHV/HHV .9416

Enthalpy (400°R Base) - 55.97 Btu/lb

Stoichiometric Fuel/Air Ratio 0.345

(Values in parenthesis are for as received coal)

Table 2.5 - INTERMEDIATE-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Koppers-Totzek/Stretford Desulf.

COAL North Dakota lignite

Lockhopper Inlet Conditions

Temperature 150

Moisture Content 27

PROCESS OXYGEN

Air/Coal Ratio .561 (.487)

Temperature - °F 220

Pressure - psia 14.7

PROCESS STEAM GASIFIER

Steam/Coal Ratio .110 (.095)

Temperature - °F 400

Pressure - psia 14.7

PRODUCT FUEL GAS

Temperature - °F 100

Pressure - psia 14.7

Composition-Mole Fraction

N_2	.0046
O_2	---
H_2	.3197
CO	.5268
CO_2	.0842
H_2O	.0647
H_2S	---
CH_4	---
C_2H_4	---

Product Fuel Gas/Coal Ratio 1.29 (1.12)

Molecular Wt 20.4

Heating Value

LHV -	256.6	Btu/scf	4775.6	Btu/lb
HHV -	272.7	Btu/scf	5075.1	Btu/lb
LHV/HHV	.9410			

Enthalpy (400°R Base) - 55.72 Btu/lb

Stoichiometric Fuel/Air Ratio 0.350

(Values in parenthesis are for as received coal)

Table 2.6 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS: Westinghouse Fluidized Bed/High-Temp. Desulfurization

COAL: Illinois No. 6 bituminous

Lockhopper Inlet Conditions
- Temperature: 150
- Moisture Content: 3%

SORBENT: Dolomite
- Sorbent/Coal Ratio: 0.59 (0.53)

PROCESS AIR
- Air/Coal Ratio: 2.95 (2.65)
- Temperature - °F: 350
- Pressure - psia: 250

PROCESS STEAM
- Steam/Coal Ratio: 0.462 (0.414)
- Temperature - °F: 400
- Pressure - psia: 250

PRODUCT FUEL GAS
- Temperature - °F: 1600
- Pressure - psia: 225

Composition-Mole Fraction

N_2	.4597
O_2	0
H_2	.1437
CO	.2142
CO_2	.0830
H_2O	.0681
H_2S	0
CH_4	.0313
C_2H_4	0

Product Fuel Gas/Coal Ratio: 4.31 (3.86)

Gasifier Aux. Pwr. (kW/lb/s): 14.4

Spent Sorbent Oxidizer Exhaust Products
- T_{in} - °F: 1500
- T_{out} - °F: 300
- q-Btu/lb coal: 444 (399)

Molecular Wt: 24.55

Heating Value

LHV -	136.57	Btu/scf	2111.65	Btu/lb
HHV -	146.92	Btu/scf	2271.73	Btu/lb

LHV/HHV: 0.9295

Enthalpy (400°R Base) - 539.33 Btu/lb

Stoichiometric Fuel/Air Ratio: 0.728

(Values in parenthesis are for as received coal)

2.3.1.4 Low-Btu Gas

Composition and properties of low-Btu fuel gas from three generic gasification processes were compiled for each of the three specified coals. Data on fuel gas from the Westinghouse fluidized bed gasification process currently under development are given in Tables 2.6 to 2.17. Tables 2.6 to 2.14 are for high-temperature desulfurization and Tables 2.15 to 2.17 are for low-temperature desulfurization. Tables 2.18 to 2.20 show the advantage of adding a recuperator between the gasifier exhaust and the low-temperature desulfurization process to reheat the fuel gas after desulfurization, thereby increasing its sensible heat and making its combustor design easier.

Tables 2.21 to 2.23 contain data on fuel gas from a suspension-type gasifier such as the Bituminous Coal Research process. Data are given for only the Illinois No. 6 coal and only for low-temperature desulfurization.

Tables 2.24 to 2.26 show the effect of recuperating the fuel gas from the suspension bed gasifier.

Tables 2.23 to 2.29 contain data on fuel gas from a fixed bed gasifier such as the Lurgi process. Data are given for only the Illinois coal and only for low-temperature desulfurization.

In all cases data are given for three values of process air temperature going to the gasifier: 449.82, 560.94, and 672.05°K (350, 550, and 750°F). The various ratios are given for two values of coal moisture—the lockhopper inlet or as fired value which is listed and the as received value. The ratios for the as received case are in parentheses.

2.3.1.5 Distillate from Coal

Syncrude from the H-coal process is assumed to have the compositions shown in Table 2.30.

Table 2.7 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS: Westinghouse Fluidized Bed/High-Temp. Desulfurization

COAL: Illinois No. 6 bituminous

Lockhopper Inlet Conditions
- Temperature: 150
- Moisture Content: 3%

SORBENT: Dolomite
- Sorbent/Coal Ratio: 0.59 (0.53)

PROCESS AIR
- Air/Coal Ratio: 2.86 (2.57)
- Temperature - °F: 550
- Pressure - psia: 250

PROCESS STEAM
- Steam/Coal Ratio: 0.500 (0.448)
- Temperature - °F: 400
- Pressure - psia: 250

PRODUCT FUEL GAS
- Temperature - °F: 1600
- Pressure - psia: 225

Composition-Mole Fraction

N_2	.4468
O_2	0
H_2	.1518
CO	.2147
CO_2	.0834
H_2O	.0720
H_2S	0
CH_4	.0314
C_2H_4	0

Product Fuel Gas/Coal Ratio: 4.24 (3.80)

Gasifier Aux. Pwr. (kW/lb/s): 14.4

Spent Sorbent Oxidizer Exhaust Products
- T_{in} - °F: 1500
- T_{out} - °F: 300
- q-Btu/lb coal: 444 (399)

Molecular Wt: 24.31

Heating Value

LHV -	139.04	Btu/scf	2171.24	Btu/lb
HHV -	149.81	Btu/scf	2339.43	Btu/lb

LHV/HHV: .9281

Enthalpy (400°R Base) - 545.15 Btu/lb

Stoichiometric Fuel/Air Ratio: 0.717

(Values in parenthesis are for as received coal)

Table 2.8 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Westinghouse Fluidized Bed/High-Temp. Desulfurization

COAL Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature 150

Moisture Content 3%

SORBENT Dolomite

Sorbent/Coal Ratio 0.59 (0.53)

PROCESS AIR

Air/Coal Ratio 2.77 (2.49)

Temperature - °F 750

Pressure - psia 250

PROCESS STEAM

Steam/Coal Ratio 0.535 (0.480)

Temperature - °F 400

Pressure - psia 250

PRODUCT FUEL GAS

Temperature - °F 1600

Pressure - psia 225

Composition-Mole Fraction

N_2	.4346
O_2	0
H_2	.1594
CO	.2152
CO_2	.0837
H_2O	.0757
H_2S	0
CH_4	.0315
C_2H_4	0

Product Fuel Gas/Coal Ratio 4.15 (3.72)

Gasifier Aux. Pwr. (kW/lb/s) 14.4

Spent Sorbent Oxidizer Exhaust Products

T_{in} - °F 1500

T_{out} - °F 300

q-Btu/lb coal 444 (399)

Molecular Wt 24.08

Heating Value

LHV -	141.37	Btu/scf	2228.89	Btu/lb
HHV -	152.53	Btu/scf	2404.88	Btu/lb

LHV/HHV .9268

Enthalpy (400°R Base) - 550.75 Btu/lb

Stoichiometric Fuel/Air Ratio 0.699

(Values in parenthesis are for as received coal)

Table 2.9 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Westinghouse Fluidized Bed/High-Temp. Desulfurization

COAL Montana subbituminous

Lockhopper Inlet Conditions

Temperature 150

Moisture Content 20%

SORBENT Dolomite

Sorbent/Coal Ratio 0.12 (0.11)

PROCESS AIR

Air/Coal Ratio 2.32 (2.20)

Temperature - °F 350

Pressure - psia 250

PROCESS STEAM

Steam/Coal Ratio 0.390 (0.369)

Temperature - °F 400

Pressure - psia 250

PRODUCT FUEL GAS

Temperature - °F 1600

Pressure - psia 225

Composition-Mole Fraction

N_2 .4542

O_2 0

H_2 .1517

CO .2239

CO_2 .0781

H_2O .0646

H_2S 0

CH_4 .0275

C_2H_4 0

Product Fuel Gas/Coal Ratio 3.39 (3.21)

Gasifier Aux. Pwr. (kW/lb/s) 14.4

Spent Sorbent Oxidizer Exhaust Products

T_{in} - °F 1500

T_{out} - °F 300

q-Btu/lb coal 70 (67)

Molecular Wt 24.34

Heating Value

LHV - 138.32 Btu/scf 2157.07 Btu/lb

HHV - 148.69 Btu/scf 2318.72 Btu/lb

LHV/HHV .9303

Enthalpy (400°R Base) - 540.41 Btu/lb

Stoichiometric Fuel/Air Ratio 0.727

(Values in parenthesis are for as received coal)

Table 2.10 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Westinghouse Fluidized Bed/High-Temp. Desulfurization

COAL Montana subbituminous

Lockhopper Inlet Conditions

Temperature 150

Moisture Content 20%

SORBENT Dolomite

Sorbent/Coal Ratio 0.12 (0.11)

PROCESS AIR

Air/Coal Ratio 2.25 (2.13)

Temperature - °F 550

Pressure - psia 250

PROCESS STEAM

Steam/Coal Ratio 0.421 (0.398)

Temperature - °F 400

Pressure - psia 250

PRODUCT FUEL GAS

Temperature - °F 1600

Pressure - psia 225

Composition-Mole Fraction

N_2	.4413
O_2	0
H_2	.1599
CO	.2242
CO_2	.0786
H_2O	.0685
H_2S	0
CH_4	.0275
C_2H_4	0

Product Fuel Gas/Coal Ratio 3.35 (3.17)

Gasifier Aux. Pwr. (kW/lb/s) 14.4

Spent Sorbent Oxidizer Exhaust Products

T_{in} - °F 1500

T_{out} - °F 300

q-Btu/lb coal 70 (67)

Molecular Wt 24.10

Heating Value

LHV - 140.76 Btu/scf 2217.14 Btu/lb

HHV - 151.55 Btu/scf 2387.08 Btu/lb

LHV/HHV .9288

Enthalpy (400°R Base) - 546.33 Btu/lb

Stoichiometric Fuel/Air Ratio 0.709

(Values in parenthesis are for as received coal)

Table 2.11 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Westinghouse Fluidized Bed/High-Temp. Desulfurization

COAL Montana subbituminous

Lockhopper Inlet Conditions

Temperature 150

Moisture Content 20%

SORBENT Dolomite

Sorbent/Coal Ratio 0.12 (0.11)

PROCESS AIR

Air/Coal Ratio 2.18 (2.06)

Temperature - °F 750

Pressure - psia 250

PROCESS STEAM

Steam/Coal Ratio 0.450 (0.425)

Temperature - °F 400

Pressure - psia 250

PRODUCT FUEL GAS

Temperature - °F 1600

Pressure - psia 225

Composition-Mole Fraction

N_2	.4291
O_2	0
H_2	.1677
CO	.2244
CO_2	.0791
H_2O	.0722
H_2S	0
CH_4	.0276
C_2H_4	0

Product Fuel Gas/Coal Ratio 3.31 (3.13)

Gasifier Aux. Pwr. (kW/lb/s) 14.4

Spent Sorbent Oxidizer Exhaust Products

T_{in} - °F 1500

T_{out} - °F 300

q-Btu/lb coal 70 (67)

Molecular Wt 23.87

Heating Value

LHV - 143.02 Btu/scf 2274.75 Btu/lb

HHV - 154.21 Btu/scf 2452.76 Btu/lb

LHV/HHV .9274

Enthalpy (400°R Base) - 552.09 Btu/lb

Stoichiometric Fuel/Air Ratio 0.685

(Values in parenthesis are for as received coal)

Table 2.12 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Westinghouse Fluidized Bed/High-Temp. Desulfurization

COAL North Dakota lignite

Lockhopper Inlet Conditions

Temperature 150

Moisture Content 27%

SORBENT Dolomite

Sorbent/Coal Ratio 0.11 (0.10)

PROCESS AIR

Air/Coal Ratio 1.93 (1.67)

Temperature - °F 350

Pressure - psia 250

PROCESS STEAM

Steam/Coal Ratio 0.310 (0.269)

Temperature - °F 400

Pressure - psia 250

PRODUCT FUEL GAS

Temperature - °F 1600

Pressure - psia 225

Composition-Mole Fraction

N_2	.4484
O_2	0
H_2	.1501
CO	.2293
CO_2	.0803
H_2O	.0642
H_2S	0
CH_4	.0278
C_2H_4	0

Product Fuel Gas/Coal Ratio 2.86 (2.48)

Gasifier Aux. Pwr. (kW/lb/s) 14.4

Spent Sorbent Oxidizer Exhaust Products

T_{in} - °F 1500

T_{out} - °F 300

q-Btu/lb coal 69 (60)

Molecular Wt 24.42

Heating Value

LHV -	139.98	Btu/scf	2175.62	Btu/lb
HHV -	150.31	Btu/scf	2336.11	Btu/lb

LHV/HHV .9313

Enthalpy (400°R Base) - 539.49 Btu/lb

Stoichiometric Fuel/Air Ratio 0.725

(Values in parenthesis are for as received coal)

Table 2.13 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Westinghouse Fluidized Bed/High-Temp. Desulfurization

COAL North Dakota lignite

Lockhopper Inlet Conditions

Temperature 150

Moisture Content 27%

SORBENT Dolomite

Sorbent/Coal Ratio 0.11 (0.10)

PROCESS AIR

Air/Coal Ratio 1.87 (1.62)

Temperature - °F 550

Pressure - psia 250

PROCESS STEAM

Steam/Coal Ratio 0.336 (0.291)

Temperature - °F 400

Pressure - psia 250

PRODUCT FUEL GAS

Temperature - °F 1600

Pressure - psia 225

Composition-Mole Fraction

N_2	.4357
O_2	0
H_2	.1581
CO	.2296
CO_2	.0808
H_2O	.0679
H_2S	0
CH_4	.0279
C_2H_4	0

Product Fuel Gas/Coal Ratio 2.83 (2.46)

Gasifier Aux. Pwr. (kW/lb/s) 14.4

Spent Sorbent Oxidizer Exhaust Products

T_{in} - °F 1500

T_{out} - °F 300

q-Btu/lb coal 69 (60)

Molecular Wt 24.18

Heating Value

LHV -	142.36	Btu/scf	2234.67	Btu/lb
HHV -	153.10	Btu/scf	2403.23	Btu/lb

LHV/HHV .9299

Enthalpy (400°R Base) - 545.23 Btu/lb

Stoichiometric Fuel/Air Ratio 0.704

(Values in parenthesis are for as received coal)

Table 2.14 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Westinghouse Fluidized Bed/High-Temp. Desulfurization

COAL North Dakota lignite

Lockhopper Inlet Conditions

Temperature 150

Moisture Content 27%

SORBENT Dolomite

Sorbent/Coal Ratio 0.11 (0.10)

PROCESS AIR

Air/Coal Ratio 1.81 (1.57)

Temperature - °F 750

Pressure - psia 250

PROCESS STEAM

Steam/Coal Ratio 0.361 (0.313)

Temperature - °F 400

Pressure - psia 250

PRODUCT FUEL GAS

Temperature - °F 1600

Pressure - psia 225

Composition-Mole Fraction

N_2	.4237
O_2	0
H_2	.1657
CO	.2299
CO_2	.0812
H_2O	.0715
H_2S	0
CH_4	.0279
C_2H_4	0

Product Fuel Gas/Coal Ratio 2.80 (2.42)

Gasifier Aux. Pwr. (kW/lb/s) 14.4

Spent Sorbent Oxidizer Exhaust Products

T_{in} - °F 1500

T_{out} - °F 300

q-Btu/lb coal 69 (60)

Molecular Wt 23.95

Heating Value

LHV - 144.54 Btu/scf 2290.67 Btu/lb

HHV - 155.66 Btu/scf 2466.91 Btu/lb

LHV/HHV .9286

Enthalpy (400°R Base) - 550.76 Btu/lb

Stoichiometric Fuel/Air Ratio 0.686

(Values in parenthesis are for as received coal)

Table 2.15 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Westinghouse Fluidized Bed/Low-Temp. Desulfurization

COAL Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature 150°F

Moisture Content 3%

PROCESS AIR

Air/Coal Ratio 2.73 (2.45)

Temperature - °F 350

Pressure - psia 250

PROCESS STEAM

	GASIFIER	DESULFURIZER
Steam/Coal Ratio	0.533 (0.478)	0.714 (0.640)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

Temperature - °F 230

Pressure - psia 225

Composition-Mole Fraction

N_2	.4351
O_2	0
H_2	.1738
CO	.2110
CO_2	.0538
H_2O	.0922
H_2S	0
CH_4	.0341
C_2H_4	0

Product Fuel Gas/Coal Ratio 4.06 (3.64)

Gasifier Aux. Pwr. (kW/lb/s) 5.0

Sensible Heat Recovery from Product Fuel Gas

T_{in} - °F 1600

T_{out} - °F 350

q-Btu/lb coal 1780 (1597)

Molecular Wt 23.026

Heating Value

LHV - 146.34 Btu/scf 2412.53 Btu/lb

HHV - 158.48 Btu/scf 2612.78 Btu/lb

LHV/HHV .9234

Enthalpy (400°R Base) - 90.83 Btu/lb

Stoichiometric Fuel/Air Ratio 0.642

(Values in parenthesis are for as received coal)

Table 2.16 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Westinghouse Fluidized Bed/Low-Temp. Desulfurization

COAL Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature 150°F

Moisture Content 3%

PROCESS AIR

Air/Coal Ratio 2.64 (2.37)

Temperature - °F 550

Pressure - psia 250

PROCESS STEAM	GASIFIER	DESULFURIZER
Steam/Coal Ratio	.569 (0.511)	0.707 (0.634)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

Temperature - °F 230

Pressure - psia 225

Composition-Mole Fraction

N_2	.4258
O_2	0
H_2	.1825
CO	.2117
CO_2	.0544
H_2O	.0922
H_2S	0
CH_4	.0334
C_2H_4	0

Product Fuel Gas/Coal Ratio 3.99 (3.58)

Gasifier Aux. Pwr. (kW/lb/s) 5.0

Sensible Heat Recovery from Product Fuel Gas

T_{in} - °F 1600

T_{out} - °F 350

q-Btu/lb coal 1780 (1597)

Molecular Wt 22.818

Heating Value

LHV - 148.31 Btu/scf 2467.37 Btu/lb

HHV - 160.83 Btu/scf 2675.57 Btu/lb

LHV/HHV .92218

Enthalpy (400°R Base) - 91.647 Btu/lb

Stoichiometric Fuel/Air Ratio 0.629

(Values in parenthesis are for as received coal)

Table 2.17 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Westinghouse Fluidized Bed/Low-Temp. Desulfurization

COAL Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature 150°F

Moisture Content 3%

PROCESS AIR

Air/Coal Ratio 2.54 (2.28)

Temperature - °F 750

Pressure - psia 250

PROCESS STEAM

	GASIFIER	DESULFURIZER
Steam/Coal Ratio	.603 (0.541)	0.699 (0.627)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

Temperature - °F 230

Pressure - psia 225

Composition-Mole Fraction

N_2	.4148
O_2	0
H_2	.1916
CO	.2125
CO_2	.0552
H_2O	.0922
H_2S	0
CH_4	.0337
C_2H_4	0

Product Fuel Gas/Coal Ratio 3.94 (3.53)

Gasifier Aux. Pwr. (kW/lb/s) 5.0

Sensible Heat Recovery from Product Fuel Gas

T_{in} - °F 1600

T_{out} - °F 350

q-Btu/lb coal 1780 (1597)

Molecular Wt 22.59

Heating Value

LHV -	151.33	Btu/scf	2543.02	Btu/lb
HHV -	164.34	Btu/scf	2761.52	Btu/lb

LHV/HHV .9209

Enthalpy (400°R Base) - 92.586 Btu/lb

Stoichiometric Fuel/Air Ratio 0.609

(Values in parenthesis are for as received coal)

Table 2.18 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Westinghouse Fluidized Bed/Low-Temp. Desulf./Reheat

COAL Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature 150°F

Moisture Content 3%

PROCESS AIR

Air/Coal Ratio 2.73 (2.45)

Temperature - °F 350

Pressure - psia 250

PROCESS STEAM	GASIFIER	DESULFURIZER
Steam/Coal Ratio	.533 (0.478)	0.714 (0.640)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

Temperature - °F 700

Pressure - psia 225

Composition--Mole Fraction

N_2	.4351
O_2	0
H_2	.1738
CO	.2110
CO_2	.0538
H_2O	.0922
H_2S	0
CH_4	.0341
C_2H_4	0

Product Fuel Gas/Coal Ratio 4.06 (3.64)

Gasifier Aux. Pwr. (kW/lb/s) 5.0

Sensible Heat Recovery from Product Fuel Gas

T_{in} - °F 1600

T_{out} - °F ∿ 780

q--Btu/lb coal 1198 (1075)

Molecular Wt 23.02

Heating Value

LHV -	146.34	Btu/scf	2412.53	Btu/lb
HHV -	158.48	Btu/scf	2612.78	Btu/lb

LHV/HHV .92336

Enthalpy (400°R Base) - 244.82 Btu/lb

Stoichiometric Fuel/Air Ratio 0.642

(Values in parenthesis are for as received coal)

Table 2.19 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS: Westinghouse Fluidized Bed/Low-Temp. Desulf./Reheat

COAL: Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature: 150°F

Moisture Content: 3%

PROCESS AIR

Air/Coal Ratio: 2.64 (2.37)

Temperature - °F: 550

Pressure - psia: 250

PROCESS STEAM

	GASIFIER	DESULFURIZER
Steam/Coal Ratio	.569 (0.511)	0.707 (0.634)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

Temperature - °F: 700

Pressure - psia: 225

Composition-Mole Fraction

N_2	.4250
O_2	0
H_2	.1825
CO	.2117
CO_2	.0544
H_2O	.0922
H_2S	0
CH_4	.0334
C_2H_4	0

Product Fuel Gas/Coal Ratio: 3.99 (3.58)

Gasifier Aux. Pwr. (kW/lb/s): 5.0

Sensible Heat Recovery from Product Fuel Gas

T_{in} - °F: 1600

T_{out} - °F: ∿ 780

q-Btu/lb coal: 1200 (1076)

Molecular Wt: 22.81

Heating Value

LHV - 148.31 Btu/scf, 2467.37 Btu/lb

HHV - 160.83 Btu/scf, 2675.57 Btu/lb

LHV/HHV: .9222

Enthalpy (400°R Base) - 247.01 Btu/lb

Stoichiometric Fuel/Air Ratio: 0.629

(Values in parenthesis are for as received coal)

Table 2.20 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Westinghouse Fluidized Bed/Low-Temp. Desulf./Reheat

COAL Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature 150°F

Moisture Content 3%

PROCESS AIR

Air/Coal Ratio 2.54 (2.28)

Temperature - °F 750

Pressure - psia 250

PROCESS STEAM	GASIFIER	DESULFURIZER
Steam/Coal Ratio	.603 (0.541)	.699 (0.627)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

Temperature - °F 700

Pressure - psia 225

Composition-Mole Fraction

N_2	.4148
O_2	0
H_2	.1916
CO	.2125
CO_2	.0552
H_2O	.0922
H_2S	0
CH_4	.0337
C_2H_4	0

Product Fuel Gas/Coal Ratio 3.94 (3.53)

Gasifier Aux. Pwr. (kW/lb/s) 5

Sensible Heat Recovery from Product Fuel Gas

T_{in} - °F 1600

T_{out} - °F ∿ 780

q-Btu/lb coal 1203 (1079)

Molecular Wt 22.59

Heating Value

LHV - 151.33 Btu/scf 2543.02 Btu/lb

HHV - 164.34 Btu/scf 2761.52 Btu/lb

LHV/HHV .9209

Enthalpy (400°R Base) - 249.58 Btu/lb

Stoichiometric Fuel/Air Ratio 0.609

(Values in parenthesis are for as received coal)

Table 2.21 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS: Suspension Bed/Low-Temp. Desulfurization

COAL: Illinois No. 6 bituminous

Lockhopper Inlet Conditions

- Temperature: 150°F
- Moisture Content: 3%

PROCESS AIR

- Air/Coal Ratio: 4.67 (4.19)
- Temperature - °F: 350
- Pressure - psia: 250

PROCESS STEAM

	GASIFIER	DESULFURIZER
Steam/Coal Ratio	.483 (.433)	.850 (.762)
Temperature - °F	400	281
Pressure - psia	225	50

PRODUCT FUEL GAS

- Temperature - °F: 230
- Pressure - psia: 225

Composition-Mole Fraction

N_2	.5879
O_2	0
H_2	.0951
CO	.1506
CO_2	.0605
H_2O	.0923
H_2S	0
CH_4	.0136
C_2H_4	0

Product Fuel Gas/Coal Ratio: 5.49 (4.92)

Gasifier Aux. Pwr. (kW/lb/s): 5.0

Sensible Heat Recovery from Product Fuel Gas

- T_{in} - °F: 2000
- T_{out} - °F: 350
- q-Btu/lb coal: 3135.5 (2813)

Molecular Wt: 25.42

Heating Value

- LHV - 86.74 Btu/scf, 1295.19 Btu/lb
- HHV - 92.89 Btu/scf, 1386.90 Btu/lb
- LHV/HHV: .9339

Enthalpy (400°R Base) - 82.07 Btu/lb

Stoichiometric Fuel/Air Ratio: 1.23

(Values in parenthesis are for as received coal)

Table 2.22 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Suspension Bed/Low-Temp. Desulfurization

COAL Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature 150°F

Moisture Content 3%

PROCESS AIR

Air/Coal Ratio 4.48 (4.02)

Temperature - °F 550

Pressure - psia 250

PROCESS STEAM	GASIFIER	DESULFURIZER
Steam/Coal Ratio	.483 (.433)	.827 (.742)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

Temperature - °F 230

Pressure - psia 225

Composition-Mole Fraction

N_2	.5745
O_2	0
H_2	.1033
CO	.1591
CO_2	.0572
H_2O	.0923
H_2S	0
CH_4	.0136
C_2H_4	0

Product Fuel Gas/Coal Ratio 5.34 (4.79)

Gasifier Aux. Pwr. (kW/lb/s) 5.0

Sensible Heat Recovery from Product Fuel Gas

T_{in} - °F 2000

T_{out} - °F 350

q-Btu/lb coal 3053.3 (2738)

Molecular Wt 25.16

Heating Value

LHV -	91.72	Btu/scf	1383.94	Btu/lb
HHV -	98.27	Btu/scf	1482.85	Btu/lb

LHV/HHV .9333

Enthalpy (400°R Base) - 82.85 Btu/lb

Stoichiometric Fuel/Air Ratio 1.15

(Values in parenthesis are for as received coal)

Table 2.23 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Suspension Bed/Low-Temp. Desulfurization

COAL Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature 150°F

Moisture Content 3%

PROCESS AIR

Air/Coal Ratio 4.31 (3.87)

Temperature - °F 750

Pressure - psia 250

PROCESS STEAM	GASIFIER	DESULFURIZER
Steam/Coal Ratio	.483 (.433)	.805 (.722)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

Temperature - °F 230

Pressure - psia 225

Composition-Mole Fraction

Component	Mole Fraction
N_2	.5606
O_2	0
H_2	.1116
CO	.1673
CO_2	.0542
H_2O	.0923
H_2S	0
CH_4	.0141
C_2H_4	0

Product Fuel Gas/Coal Ratio 5.20 (4.66)

Gasifier Aux. Pwr. (kW/lb/s) 5.0

Sensible Heat Recovery from Product Fuel Gas

T_{in} - °F 2000

T_{out} - °F 350

q-Btu/lb coal 2979.3 (2672)

Molecular Wt 24.89

Heating Value

LHV - 97.08 Btu/scf 1480.54 Btu/lb

HHV - 104.10 Btu/scf 1587.64 Btu/lb

LHV/HHV .9325

Enthalpy (400°R Base) - 83.68 Btu/lb

Stoichiometric Fuel/Air Ratio 1.09

(Values in parenthesis are for as received coal)

Table 2.24 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Suspension Bed/Low-Temp. Desulf./Reheat

COAL Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature 150°F

Moisture Content 3%

PROCESS AIR

Air/Coal Ratio 4.67 (4.19)

Temperature - °F 350

Pressure - psia 250

PROCESS STEAM	GASIFIER	DESULFURIZER
Steam/Coal Ratio	.483 (.433)	.850 (.762)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

Temperature - °F 700

Pressure - psia 225

Composition-Mole Fraction

N_2	.5879
O_2	0
H_2	.0951
CO	.1506
CO_2	.0605
H_2O	.0923
H_2S	0
CH_4	.0136
C_2H_4	0

Product Fuel Gas/Coal Ratio 5.49 (4.92)

Gasifier Aux. Pwr. (kW/lb/s) 5.0

Sensible Heat Recovery from Product Fuel Gas

T_{in} - °F 2000

T_{out} - °F ∿ 780

q-Btu/lb coal 2376 (2131)

Molecular Wt 25.42

Heating Value

LHV -	86.74	Btu/scf	1259.19	Btu/lb
HHV -	92.89	Btu/scf	1386.90	Btu/lb

LHV/HHV .9339

Enthalpy (400°R Base) - 220.53 Btu/lb

Stoichiometric Fuel/Air Ratio 1.23

(Values in parenthesis are for as received coal)

Table 2.25 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS: Suspension Bed/Low-Temp. Desulf./Reheat

COAL: Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature: 150°F

Moisture Content: 3%

PROCESS AIR

Air/Coal Ratio: 4.48 (4.02)

Temperature - °F: 550

Pressure - psia: 250

PROCESS STEAM	GASIFIER	DESULFURIZER
Steam/Coal Ratio	.483 (.433)	.827 (.742)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

Temperature - °F: 700

Pressure - psia: 225

Composition-Mole Fraction

N_2	.5745
O_2	0
H_2	.1033
CO	.1591
CO_2	.0572
H_2O	.0923
H_2S	0
CH_4	.0136
C_2H_4	0

Product Fuel Gas/Coal Ratio: 5.34 (4.79)

Gasifier Aux. Pwr. (kW/lb/s): 5.0

Sensible Heat Recovery from Product Fuel Gas

T_{in} - °F: 2000

T_{out} - °F: ∿ 780

q-Btu/lb coal: 2308 (2070)

Molecular Wt: 25.16

Heating Value

LHV - 91.72 Btu/scf 1383.94 Btu/lb

HHV - 98.27 Btu/scf 1482.85 Btu/lb

LHV/HHV: .9333

Enthalpy (400°R Base) - 222.55 Btu/lb

Stoichiometric Fuel/Air Ratio: 1.15

(Values in parenthesis are for as received coal)

Table 2.26 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Suspension Bed/Low-Temp. Desulf./Reheat

COAL Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature 150°F

Moisture Content 3%

PROCESS AIR

Air/Coal Ratio 4.31 (3.87)

Temperature - °F 750

Pressure - psia 250

PROCESS STEAM	GASIFIER	DESULFURIZER
Steam/Coal Ratio	.483 (.433)	.805 (.722)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

Temperature - °F 700

Pressure - psia 225

Composition-Mole Fraction

N_2	.5606
O_2	0
H_2	.1116
CO	.1673
CO_2	.0542
H_2O	.0923
H_2S	0
CH_4	.0141
C_2H_4	0

Product Fuel Gas/Coal Ratio 5.20 (4.66)

Gasifier Aux. Pwr. (kW/lb/s) 5.0

Sensible Heat Recovery from Product Fuel Gas

T_{in} - °F 2000

T_{out} - °F ∿ 780

q-Btu/lb coal 2246 (2015)

Molecular Wt 24.89

Heating Value

LHV -	97.08	Btu/scf	1480.54	Btu/lb
HHV -	104.10	Btu/scf	1587.64	Btu/lb

LHV/HHV .9325

Enthalpy (400°R Base) - 224.72 Btu/lb

Stoichiometric Fuel/Air Ratio 1.09

(Values in parenthesis are for as received coal)

Table 2.27 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS: Fixed Bed/Low-Temp. Desulfurization

COAL: Illinois No. 6 bituminous

Lockhopper Inlet Conditions

- Temperature: 150°F
- Moisture Content: 3%

PROCESS AIR

- Air/Coal Ratio: 2.06 (1.85)
- Temperature - °F: 350
- Pressure - psia: 250

PROCESS STEAM	GASIFIER	DESULFURIZER
Steam/Coal Ratio	1.12 (1.01)	.555 (.498)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

- Temperature - °F: 230
- Pressure - psia: 225

Composition-Mole Fraction

N_2	.4059
O_2	0
H_2	.2070
CO	.1655
CO_2	.0847
H_2O	.0923
H_2S	0
CH_4	.0446
C_2H_4	0

Product Fuel Gas/Coal Ratio: 3.38 (3.03)

Gasifier Aux. Pwr. (kW/lb/s): 5.0

Molecular Wt: 22.53

Heating Value

- LHV - 150.39 Btu/scf, 2533.93 Btu/lb
- HHV - 165.26 Btu/scf, 2784.37 Btu/lb
- LHV/HHV: .9101

Enthalpy (400°R Base) - 93.81 Btu/lb

Stoichiometric Fuel/Air Ratio: 0.603

(Values in parenthesis are for as received coal)

Table 2.28 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS: Fixed Bed/Low-Temp. Desulfurization

COAL: Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature: 150°F

Moisture Content: 3%

PROCESS AIR

Air/Coal Ratio: 1.99 (1.78)

Temperature - °F: 550

Pressure - psia: 250

PROCESS STEAM

	GASIFIER	DESULFURIZER
Steam/Coal Ratio	1.12 (1.01)	.548 (.492)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

Temperature - °F: 230

Pressure - psia: 225

Composition-Mole Fraction

N_2	.3960
O_2	0
H_2	.2134
CO	.1698
CO_2	.0835
H_2O	.0923
H_2S	0
CH_4	.0450
C_2H_4	0

Product Fuel Gas/Coal Ratio: 3.34 (3.00)

Gasifier Aux. Pwr. (kW/lb/s): 5.0

Molecular Wt: 22.34

Heating Value

LHV -	153.89	Btu/scf	2614.96	Btu/lb
HHV -	169.12	Btu/scf	2873.69	Btu/lb

LHV/HHV: .9100

Enthalpy (400°R Base) - 94.58 Btu/lb

Stoichiometric Fuel/Air Ratio: 0.577

(Values in parenthesis are for as received coal)

Table 2.29 - LOW-BTU FUEL GAS PROPERTIES

GASIFICATION PROCESS Fixed Bed/Low-Temp. Desulfurization

COAL Illinois No. 6 bituminous

Lockhopper Inlet Conditions

Temperature 150°F

Moisture Content 3%

PROCESS AIR

Air/Coal Ratio 1.93 (1.73)

Temperature - °F 750

Pressure - psia 250

PROCESS STEAM	GASIFIER	DESULFURIZER
Steam/Coal Ratio	1.12 (1.01)	.538 (.483)
Temperature - °F	400	281
Pressure - psia	250	50

PRODUCT FUEL GAS

Temperature - °F 230

Pressure - psia 225

Composition-Mole Fraction

N_2	.3862
O_2	0
H_2	.2198
CO	.1737
CO_2	.0826
H_2O	.0923
H_2S	0
CH_4	.0454
C_2H_4	0

Product Fuel Gas/Coal Ratio 3.28 (2.94)

Gasifier Aux. Pwr. (kW/lb/s) 5.0

Molecular Wt 22.15

Heating Value

LHV - 157.26 Btu/scf 2694.60 Btu/lb

HHV - 172.85 Btu/scf 2961.70 Btu/lb

LHV/HHV .9098

Enthalpy (400°R Base) - 95.35 Btu/lb

Stoichiometric Fuel/Air Ratio 0.564

(Values in parenthesis are for as received coal)

Table 2.30 - Syncrude from the H-Coal Process

Wt%	C	H	O	S	N	HHV, Btu/lb	LHV, Btu/lb
C_4^+ to 375°F (naphtha)	84.3	14.0	1.5	-	0.2	19,700	18,400
375 to 675°F (distillate)	88.9	10.9	-	0.1	0.1	18,700	17,700
675 to 975°F (heavy oil)	88.9	7.8	2.4	0.4	0.5	18,000	17,300
C_4^+ to 975°F	87.5	11.2	1.1	.2	.2		

The syncrude might typically be assumed to consist of 0.168 kg naphtha, 0.291 kg distillate, and 0.012 kg heavy oil per kg of bituminous coal. Similarly, 0.241 and 0.1858 kg of distillate are expected per kg of subbituminous and lignite coal, respectively.

When distillate from syncrude is burned, the overall energy efficiency calculated will be very misleading since the distillate is only a part of the syncrude product.

2.3.1.6 Methanol

Methanol can be produced from coal by employing a gasification process, followed by a 2:1/H_2:CO shift process, and then (catalytic) methanation. The higher heating value of methanol is 22.416 MJ/kg (9640 Btu/lb) and the heating value is 19.649 MJ/kg (8450 Btu/lb). Yields of 0.78, 0.62, and 0.48 kg methanol per kg of coal can be expected for the three coals, bituminous, subbituminous, and lignite, respectively.

2.3.1.7 Hydrogen

Hydrogen can be produced from coal by employing a gasifier and shift reactor. Absorption of the carbon dioxide from the gas leaves hydrogen. The higher heating of hydrogen is 142.04 MJ/kg (61,084 Btu/lb) and the low heating value is 122.15 MJ/kg (52,532 Btu/lb).

Yields of 0.098, 0.077, and 0.060 kg of hydrogen per kg of coal can be expected for the three coals, bituminous, subbituminous, and lignite, respectively.

2.3.2 Fuel Costs

NASA has specified the fuel costs in $\$/10^6$ Btu, as given in Table 2.31. In each case these are assumed to be delivered costs. Coal, distillate, and methanol are assumed to be delivered in bulk by rail to the site and stored in sufficient quantity to maintain operation for 60 days. High- and intermediate-Btu gas and hydrogen are assumed to be delivered from a pipeline and require no on-site storage. For each fuel listed the second number is the base cost used in most calculations.

Table 2.31 - Fuel Costs, \$/1.0549 GJ ($\$/10^6$ Btu)

Illinois No. 6 Bituminous	0.50	0.85	1.50	2.50	(1.2)(0.85)
Montana Subbituminous	0.30	0.85	1.50	2.50	(1.2)(0.85)
North Dakota Lignite	0.25	0.85	1.50	2.50	(1.2)(0.85)
High-Btu Gas	1.50	2.60	4.00	(0.8)(2.60)	(1.2)(2.60)
Intermediate-Btu Gas	1.20	2.00	3.10	4.00	(1.2)(2.00)
Distillate from Coal	1.50	2.60	4.00	(0.8)(2.60)	(1.2)(2.60)
Methanol	1.60	2.70	4.20	(0.8)(2.70)	(1.2)(2.70)
Hydrogen	1.45	2.50	3.80	(0.8)(2.50)	(1.2)(2.50)

Low-Btu gas is assumed to be generated on site in a close-coupled gasifier. The cost of that fuel is, therefore, a function of the size of the plant and the gasification system used. This will be discussed further in Section 4.

2.4 Heat Rejection Systems

Dry cooling towers are similar in operation to automobile radiators. Water leaving the condenser is pumped through finned-tube heat exchangers where atmospheric air absorbs and carries away heat without directly contacting the circulating water.

Heat transmission in a wet cooling tower is a combination of sensible heat transfer between hot water droplets and ambient air, and evaporative heat transfer from the water droplets. This process achieves cooling by pumping the warm circulating water to a distribution system in the tower and allowing it to splash down in cascade fashion through numerous layers of fill. The wet cooling towers may have either natural or mechanical draft. Since evaporation takes place in wet towers, such adverse environmental effects as fogging, drift, and icing may occur. The advantages of mechanical- over natural-draft wet towers include relatively lower capital cost, a lesser effect of ambient air humidity on tower performance, and lesser scenic impact; the disadvantages include high operating and maintenance costs, a tendency to recirculate during high wind conditions, possible fan noise problems, and more probable adverse local environmental effects of fogging, icing, and/or salt deposition.

The wet-dry cooling tower is a new engineering design specifically aimed at environmental control. The tower offers the combined performance characteristics of both wet cooling and dry cooling. A unique feature of the wet-dry cooling tower is its capability of plume abatement by utilizing a finned-tube heat exchanger to superheat the exhaust plume.

Once-through cooling might include cooling ponds, lakes, or spray ponds, but for the purpose of this study only natural water (a river) is considered, with an option of putting in a mixing canal if the usual water temperature rise in the condenser (the range) is too large from the environmental standpoint to allow the water to be returned directly to the river.

Table 2.32 - Cost Estimates of Different Cooling Systems[1]
(Cost Basics - 1973 $)

System	Once-Through	Mech. Draft Cooling Tower	Natural Draft Cooling Tower	Spray Canal	Cooling Pond
Cooling Range, °F	44	30	31	23	19
Approach,[2] °F	--	14	17	21	15.7
Capital Cost, $\$x10^{-6}$	55.90	44.37	55.53	41.46	61.93
($/kWe)	(23.3)	(18.5)	(23.1)	(17.3)	(25.8)
Evaluated Cost,[3] $\$x10^{-6}$	67.97	73.16	88.06	69.28	88.07
($/kWe)	(28.3)	(30.5)	(36.7)	(28.9)	(36.7)

[1]For two 1,200 MWe nuclear units (Reference 2.3).

[2]Design dry bulb (db) 90°F, wet bulb (wb) 75°F.

[3]Capital cost and capitalized operating costs.

The relative cost of typical heat rejection systems of these types is shown in Table 2.32 taken from Reference 2.3. This table indicates that once-through cooling is the cheapest and provides the best plant efficiency. It also shows that the net cost of the natural-draft tower system is higher than the equivalent mechanical-draft tower system. For this reason this study will deal with mechanical-draft wet towers only, while recognizing that some sites may require the natural-draft towers to minimize plume problems. Selected points will be evaluated using river water (once-through) and/or mechanical-draft cooling towers. This discussion is limited to power conversion systems rejecting heat through a surface condenser to circulated water from a tower or once-through water from a river.

The total cooling system capital cost information developed includes estimates for wet cooling towers; dry cooling towers; condensers; and the circulating water piping and equipment for each heat rejection system. The heat rejection system cost and performance data presented are based on a 750 MWe power plant which rejects 1000 MWt (3.413×10^9 Btu/hr), based on an assumed overall plant efficiency of 42.9%. The data are presented, however, on a per billion Btu/hr heat rejected basis to enable investigators of the various energy conversion systems to develop their own total cooling system capital cost and total cooling system auxiliary power requirements by multiplying the per billion Btu/hr values provided by their particular heat rejection loads. Although there were no specific investigations of economies of scale for plants either larger or smaller than the 750 MWe base plant, it is felt that the data presented are representative and, therefore, can be applied to power plants over the range of sizes considered in this study.

The cooling system auxiliary power requirement information provided includes estimates for the cooling tower fan and circulating water pump power requirements for each heat rejection system.

This study has specified the International Organization for Standardization (ISO) ambient 288.33°K (59°F) dry bulb (db) temperature

Table 2.33 - Heat Rejection

	Mechanical-Draft Wet Tower		Mechanical-Draft Dry Tower		Once-through Cooling	
Ambient	ISO	5% day	ISO	5% day	ISO	5% day
Sink Temperature, °F	51.4	77	59	93	49	75
		wet bulb		dry bulb		river water
Approach to Sink °F	22.0	15	29.6	30	0	0
Water Temperature, °F (condenser inlet)	73.4	92	88.6	123	49	75
Water Range, °F	23	23.5	28.75	29	35.22	21
Water Temperature, °F (condenser exit)	96.4	115.5	117.35	152	84.22	96
Condenser Terminal Temperature difference, °F	5	5.06	6.25	5.08	7.5	5.14
Steam Condensing temperature, °F	101.4	120.56	123.6	157.08	91.72	101.14
Condenser Pressure in Hg abs	2	3.5	3.8	9.0	1.5	2.0

with 60% relative humidity corresponding to ∿ 283.94°K (51.4°F) wet bulb (wb). Also given as the design ambient until the halfway point of the study was the 5% day [307.05°K db, 298.16°K wb (93°F db, 77°F wb)]. Many parts of this study were made for the 5% day heat rejection conditions. Not all points were recalculated for the ISO ambients, but sufficient calculations were done at both to show a valid comparison. Table 2.33 shows the condenser terminal temperature differences and circulating water range actually used in the study for ISO day. The choice has not been optimized but is presented as realistic.

2.4.1 Wet Cooling Tower Performance and Cost

Historically, wet cooling towers have either been built entirely from wood or else used wooden fill. A more durable and relatively fireproof construction having a lower maintenance cost has been assumed here. This tower basin is poured concrete; and the sides, dividing wall, louvers and distribution deck are assembled from precast reinforced concrete slabs made off site. Where multiple cells are required, many cells are aligned using common walls. Each cell measures approximately 13.11 m wide, 25.91 m deep, and 19.81 m high (43 ft, 85 ft, and 65 ft). The tower's two faces 12.19 x 12.19 m (40 x 40 ft) receive ambient air at velocities between 1.83 and 3.05 m/s (6 and 10 ft/s). The air is passed horizontally through 4.87 to 5.49 m (16 to 18 ft) of fill, through a drift eliminator, and thence upward toward the fan. The fill, perforated polyvinylchloride slating, is carried by a network of plastic-coated wire. The drift eliminators are also fabricated from plastic and turn the flow upward. The fan, near the neck of the fan stack, is 9.75 m (32 ft) in diameter and driven by a 149.2 shaft kW (200 hp) motor through a gear reducer, both of which are located atop the dividing wall just below the fan. The fan utilizes eight steel reinforced polyurethane airfoil blades on a 2.74 m (9 ft) diameter hub. A fiberglass stack is used to reduce the leaving velocity of the air and thereby decrease the fan power required. These dimensions are summarized in Table 2.34.

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Table 2.34 - Wet Tower Dimensions

Cell Length	43 ft
Water Travel	40 ft
Air Travel	18 ft
Wet Sides per Cell	2
Fans per Cell	1
Fan Diameter	32 ft
Hub Diameter	9 ft
Fan Stack Diameter	36 ft
Fan Stack Height	20 ft
Fan Tip Speed	200 ft/s
Fan Speed	119.37 rpm
Number of Fan Blades per Fan	8
Fan Total Efficiency	0.80
Gear Efficiency	0.95
Motor Efficiency	0.92
Fan Shaft Horsepower	200 hp

A picture of a single cell is shown in Figure 2.1. This cell actually was a wet-dry tower, as is evident from the presence of the four heat exchangers on either face, but the construction is similar.

2.4.1.1 Wet Tower Performance

A Westinghouse computational program was used to calculate the performance of the mechanical-draft wet cooling system described above. The program utilizes the appropriate heat and mass transfer relationship, together with empirical data for a specific packing, to arrive at a relationship between the air velocity and the circulating water range, temperature, and approach. Calculations were performed for the four ambients shown in Table 2.35.

Table 2.35 - Ambients for Wet Tower Performance Calculations

Ambient	Dry Bulb Temperature, °F	Wet Bulb Temperature, °F	Relative Humidity, %
1	96	80	50
2 (5% day)	93	77	50
3	84	70	50
4 (ISO)	59	51.37	60

The tower performance [number of cells or towers required to reject 293.021 MWt (10^9 Btu/hr)] vs water approach to wet bulb temperature for water ranges of 8.33, 13.89, and 19.44°K (15, 25, and 35°F) are displayed graphically in Figures 2.2 to 2.5 for the four ambients, respectively. It is noted that 4.33 cells are required to reject 293.021 MWt (10^9 Btu/hr) to the 5% ambient with a 8.33°K (15°F) approach and a 13.89°K (25°F) range. At the ISO condition this number of cells would require an approach of 13.44°K (24.2°F) to reject this amount of heat with the same range. The assumed approach and back pressure given in Table 2.33 are not self-consistent (different numbers of cells are required for the two ambients) and, in fact, a system designed for the 95°F ambient could not reach the 6.7537 kPa (2 in Hg) absolute back pressure desired on the ISO day. All wet tower systems associated with a 6.7537 kPa (2 in Hg abs) condenser pressure are, therefore, designed for the ISO day. Those associated with 11.819 kPa (3-1/2 in Hg abs) pressures were sized for the 5% day.

2.4.1.2 Wet Tower Capital Costs

An estimated installed capital cost for a wet mechanical-draft concrete cooling tower is $230,000 per cell. The cost breakdown for the tower is given in Table 2.36.

Table 2.36 - Breakdown of Wet Tower Total Installed Cost

	Cost per Cell
Tower Materials	$120,000
Tower Installation	50,000
Basin Materials	12,500
Basin Installation	12,500
Tower Electrical Equipment	21,000
Tower Electrical Installation	14,000
	$230,000

The tower materials and installation cost ($170,000 per cell = $120,000 + 50,000) reflects the competitive price used for Westinghouse cells in mid-1974.

The basin materials and installation cost ($25,000 per cell = $12,500 + $12,500) is based upon an assumed basin materials and installation costing rate of $\$70.29/m^2$ ($6.53/ft^2$) of basin area (cooling tower plan area).

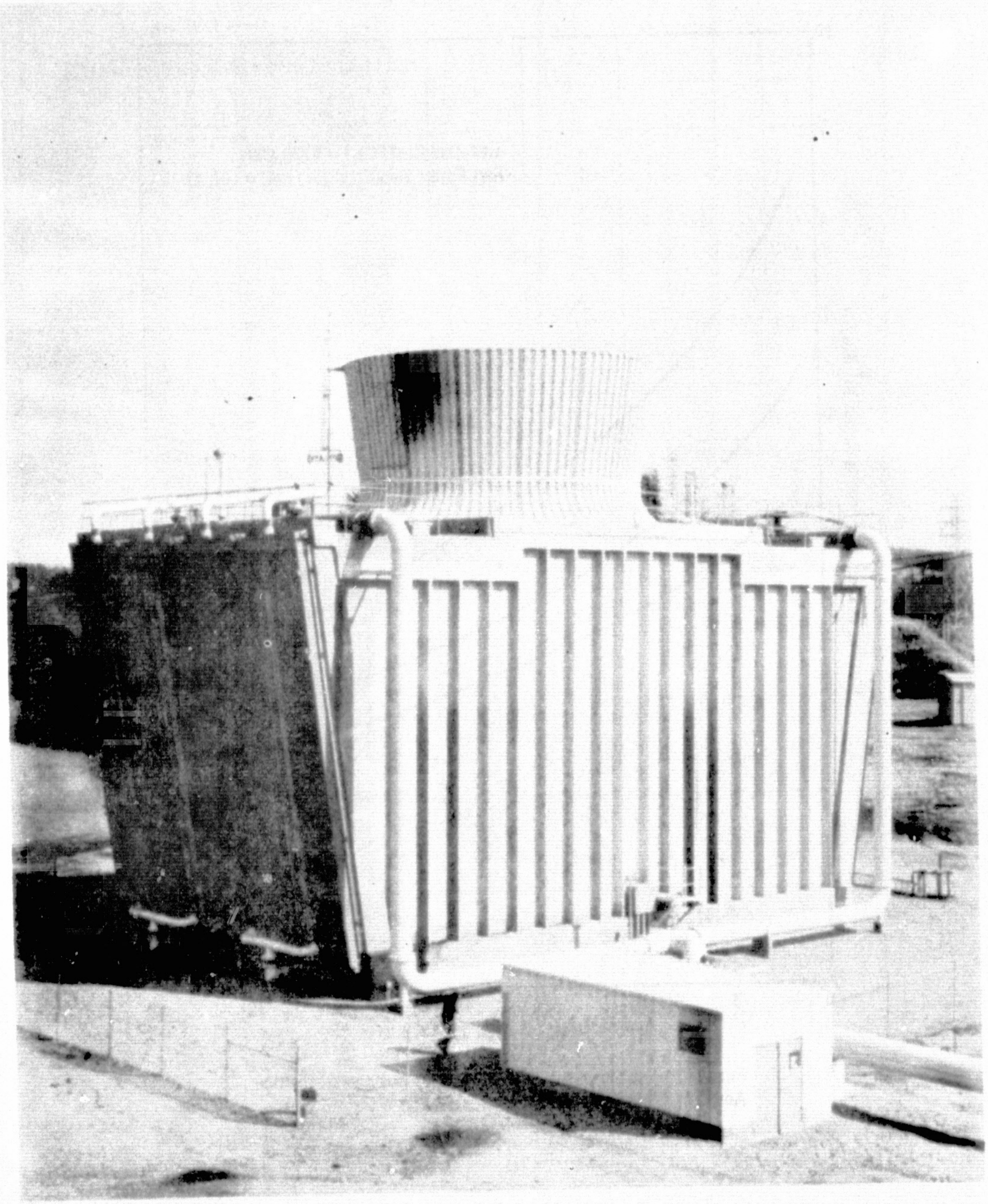


Fig. 2.1—Mechanical-draft wet-dry cooling tower (concrete construction)

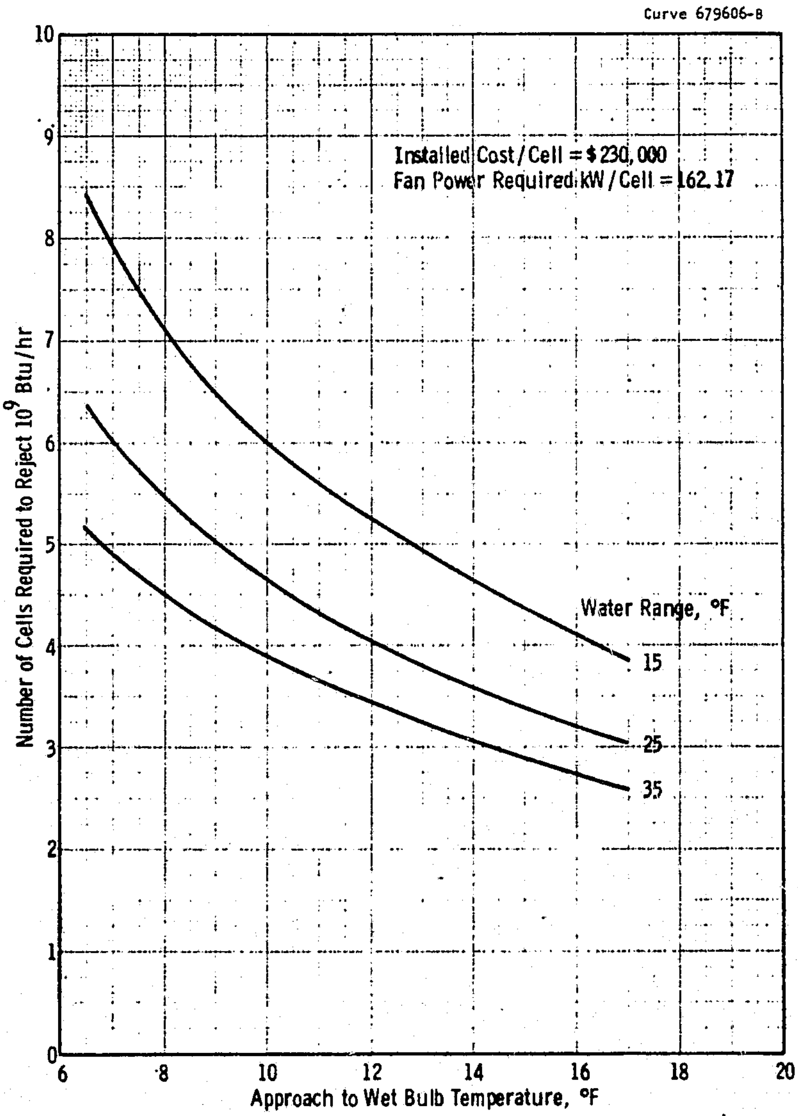


Fig. 2.2—Mechanical-draft wet cooling tower performance
Ambient 1 (96°F dry bulb, 80°F wet bulb)

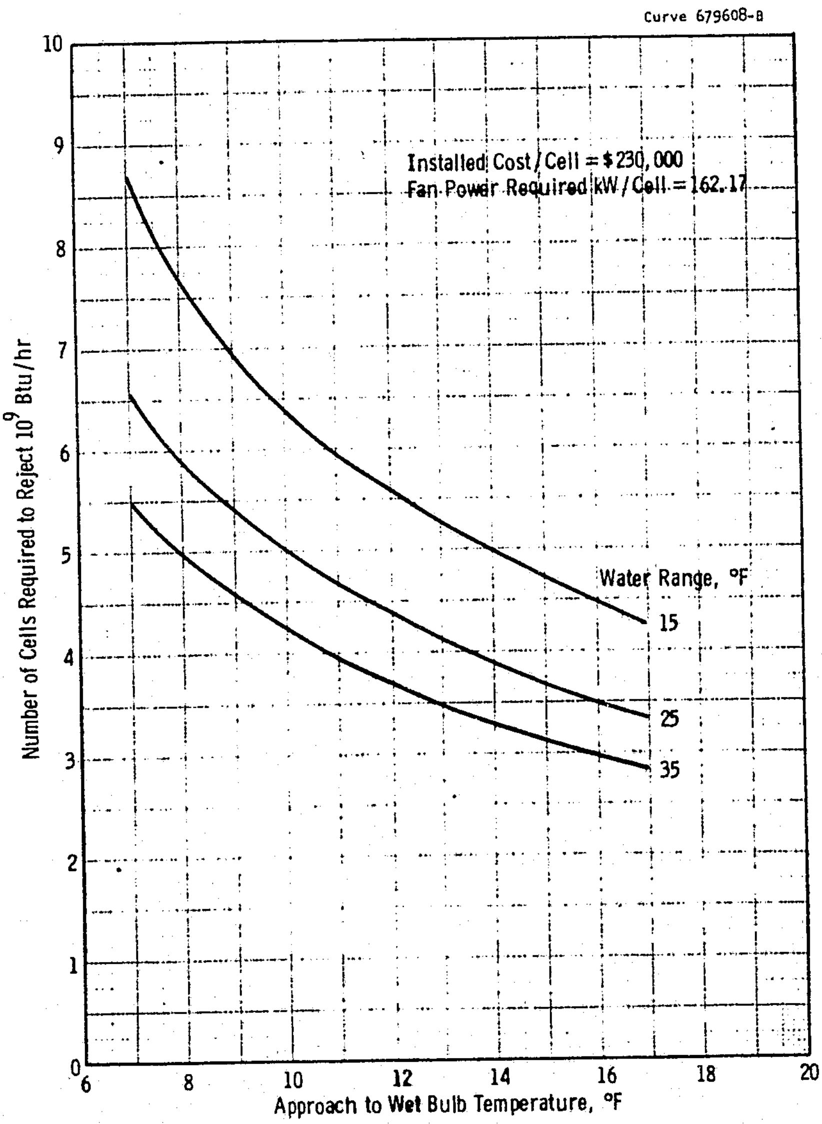


Fig. 2.3—Mechanical-draft wet cooling tower performance. Ambient 2 (93°F dry bulb, 77°F wet bulb)

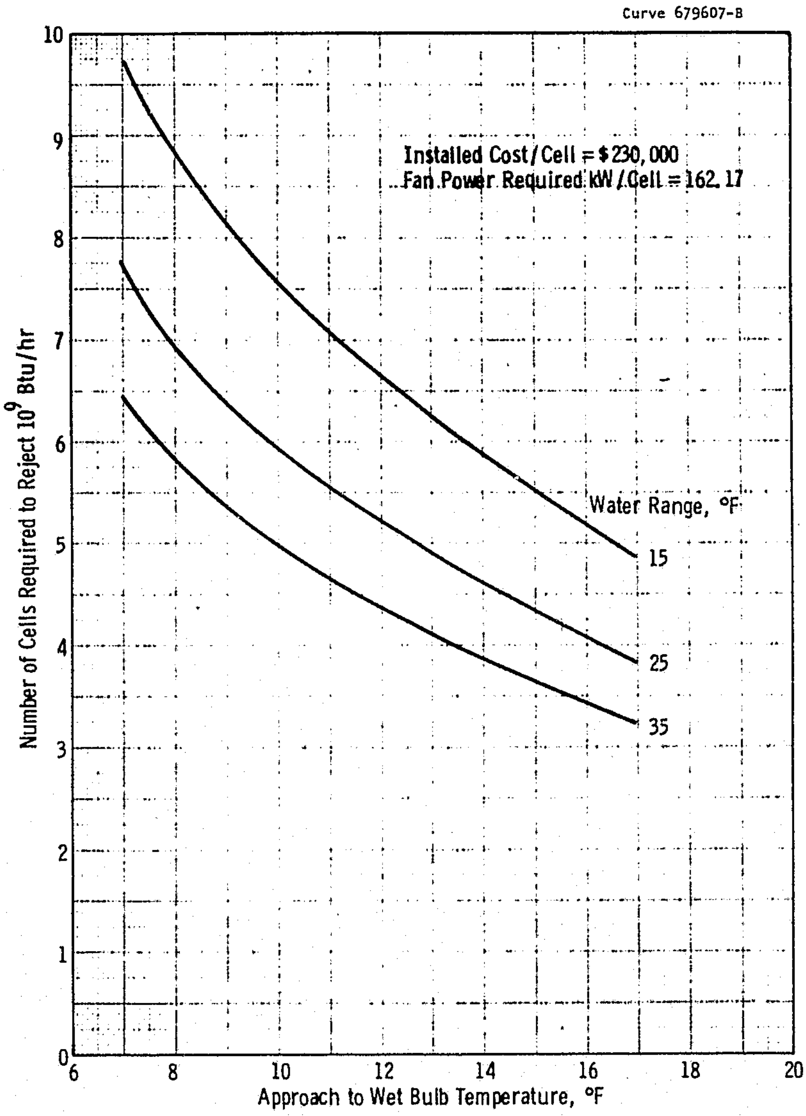


Fig. 2.4—Mechanical-draft wet cooling tower performance
Ambient 3 (84°F dry bulb, 70°F wet bulb)

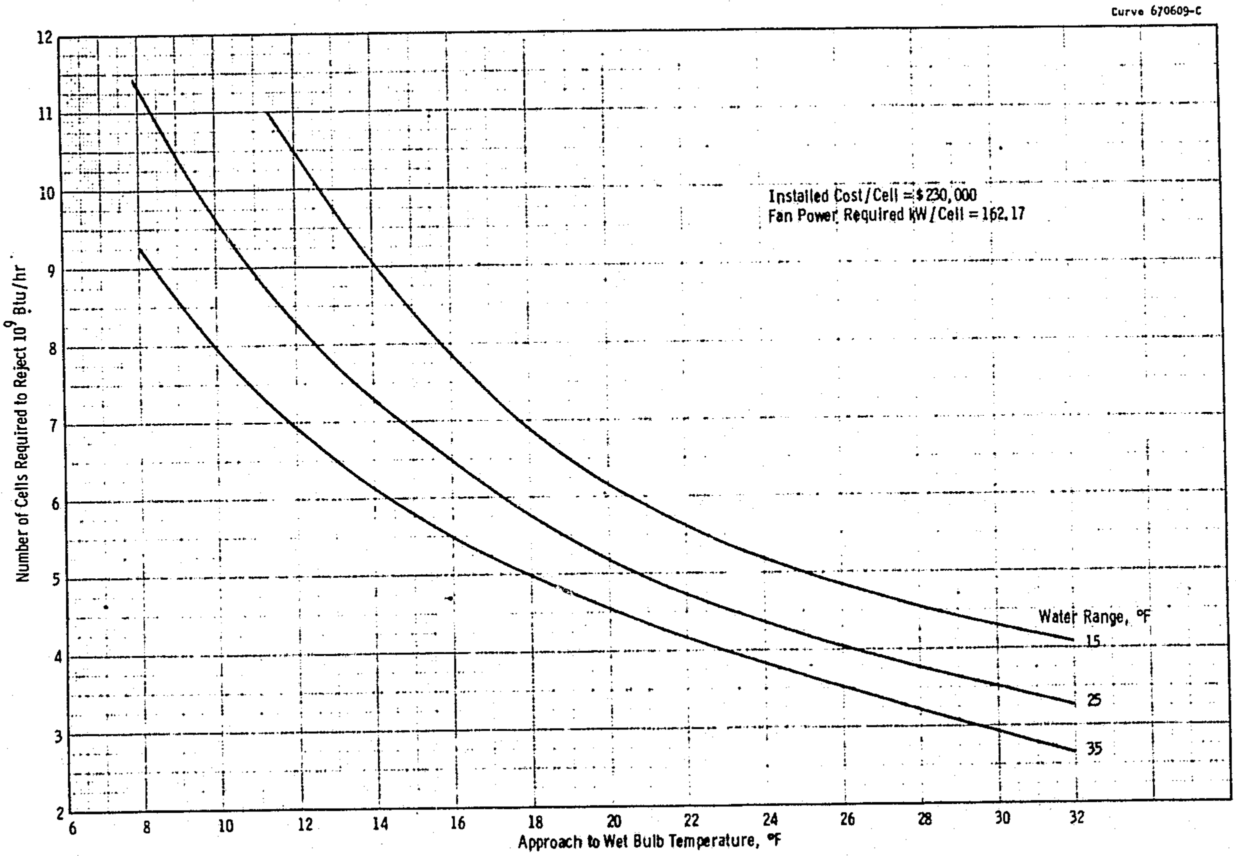


Fig. 2.5—Mechanical-draft wet cooling tower performance
Ambient 4 (59°F dry bulb, 51.3°F wet bulb)

Table 2.37 - Dry Tower Dimensions

Cell Length	43 ft
Dry Length	40 ft
Dry Sides per Cell	2
Tube Length	40 ft
Tube Rows per Pass	2
Number of Passes	4
Tube od	1 in
Tube id	0.93 in
Fin od	2.25 in
Fin Thickness	0.013 in.
Fins per Inch	11
Fin Thermal Conductivity	100 Btu/hr-ft-°F
Tube Thermal Conductivity	65 Btu/hr-ft-°F
Tube Spacing	
Transverse pitch	2.5 in
Longitudinal pitch	2.17 in
Fans per Cell	1
Fan Diameter	32 ft
Hub Diameter	9 ft
Fan Stack Diameter	36 ft
Fan Stack Height	20 ft
Fan Tip Speed	200 ft
Fan Speed	119.37 rpm
Number of Fan Blades per Fan	8
Fan Total Efficiency	0.8
Gear Efficiency	0.95
Motor Efficiency	0.92
Fan Shaft Horsepower	350 hp

The tower electrical equipment cost of $21,000 per cell includes the following equipment associated with the cooling tower:

Fan motor fusible disconnect

Gear box immersion heater

Gear box pressure switch

Circuit breaker (DHP metalclad, 1200 A, 50 DHP 240)

Transformer for water pump power supply (4.16 kV, 18.1 MVA)

Transformer for cooling tower fan motor (440 V, 25 MVA)

Lights and outlets

Quick disconnect fan motor terminals

Fan motor cable.

The tower electrical installation cost was assumed to be 67% of the tower electrical equipment cost.

From the above it is apparent that $76,500 of the $230,000 is estimated to be field installation labor. In addition, it can be expected that factory labor will account for about $40,000 of the remaining $153,500 material costs, thereby making factory and field labor account for about $116,500 of the $230,000 total installed wet cooling tower cell capital cost.

2.4.2 Dry Cooling Tower Performance and Cost

A concrete mechanical-draft dry cooling tower similar to that described in Section 2.4.1 is assumed with the following differences: the depth was 12.19 m (40 ft) since no wet packing was present; four 3.048 m (10 ft) wide by 12.19 m (40 ft) long, 8-row, 4-pass spirally wrapped fin-tube heat exchangers were placed on each face; the fan motor power was increased to 261.1 kW shaft (350 hp) to compensate for the increased pressure drop. The finned tubes were 25.4 mm (1 in.) diameter and of admiralty metal. The fins were made from aluminum and were tapered with an average thickness of 0.3302 mm (0.013 in.). Pertinent data are given in Table 2.37.

Curve 680920-A

McElroy overlapped "L" foot spiral wrapped finned tubes
1.0 in. OD 90-10 Cu-Ni tubes on 2.5 in. equilateral pitch
11.0 Al fpi of 2.25 in. OD

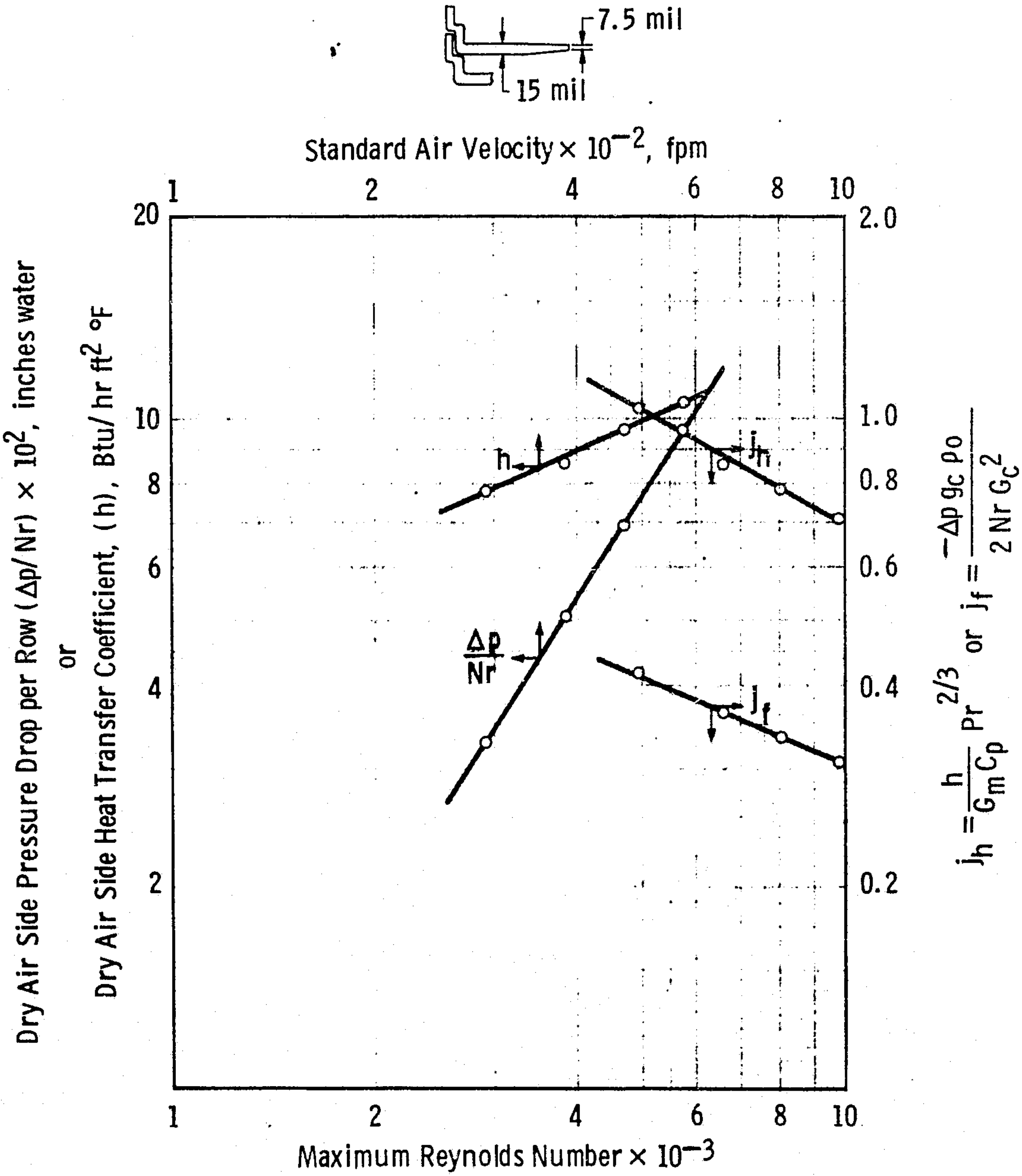


Fig. 2.6—Spiral wrapped fin-tube performance

2.4.2.1 Dry Tower Performance

The overlapped "L" foot fins made at the Westinghouse Specialty Metals Division on a McElroy finning machine were evaluated at the Westinghouse Research Laboratories in the heat exchanger test loop. Figure 2.6 is a copy of some of the test data obtained. The nomenclature used is that of Kays and London, Reference 2.4. Both the air-side heat transfer coefficient and pressure drop/row and the nondimensional j and f factors are plotted as a function of maximum Reynold's number and face velocity. These data were used to predict the performance of the dry cooling tower given in Figure 2.7, which shows the number of cells required to reject 293.02 MWt (10^9 Btu/hr) as a function of the approach of the circulating water temperature to the dry bulb temperature for ranges of 11.1, 16.66, 22.22, and 27.77°K (20, 30, 40, and 50°F). Although these data were calculated for the 5% day ambient, they will apply with only small error to other ambients.

2.4.2.2 Dry Tower Capital Costs

The estimated installed capital cost for concrete mechanical-draft dry cooling towers was $350,000 per cell. A breakdown of these costs is given in Table 2.38.

Table 2.38 - Breakdown of Dry Tower Total Installed Capital Cost

	Quantity per Cell	Unit Cost (installed)	Total Cost per cell (installed)
Heat Exchangers (2 row - 4 pass)	8	$27,000	$216,000
Motor and Gear	1	15,000	15,000
Fan	1	11,000	11,000
Structure (installed)	1	15,000	15,000
Piping & Valves	1	10,000	10,000
Electricals (installed)	1	50,000	50,000
Fan Stack, Shipping, Stairs, Hardware, Etc.	1	33,000	33,000
			$350,000

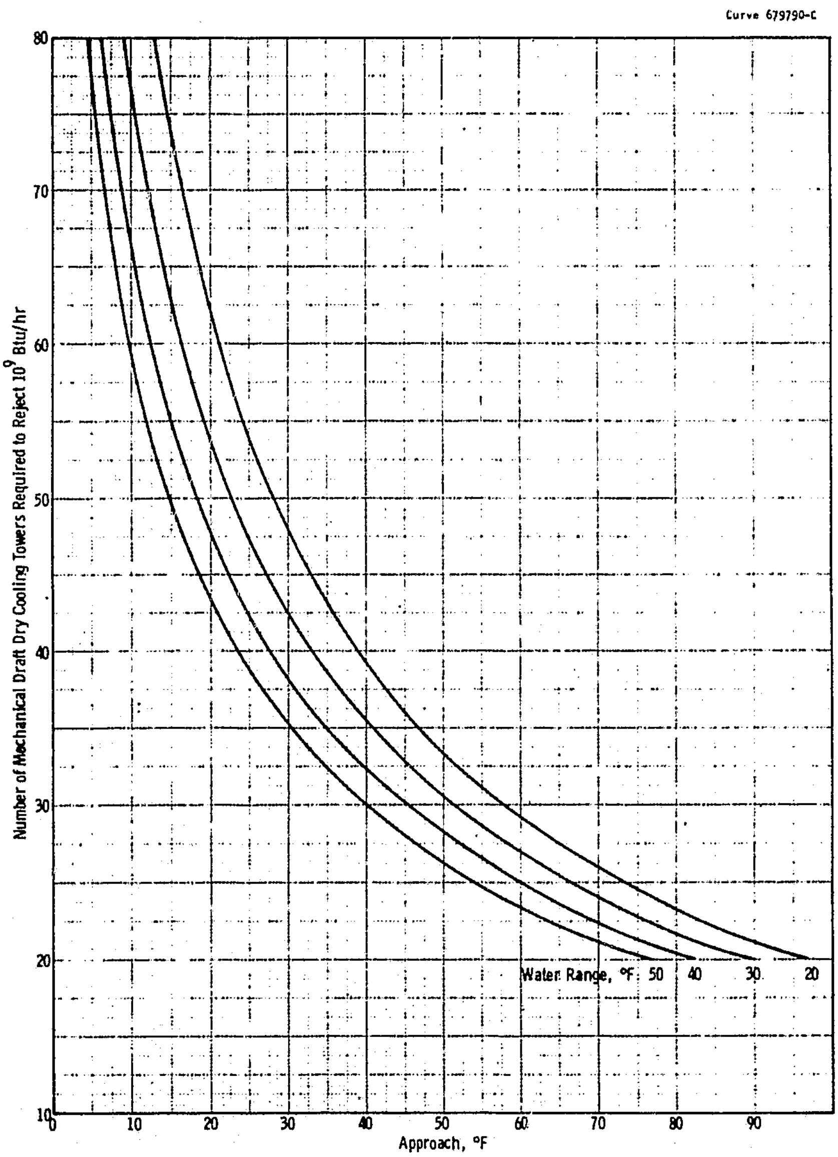


Fig. 2.7 —Dry cooling tower performance

Of the $350,000 installed capital cost per dry cell, about 55% ($192,500 per dry cell) is attributable to material cost, about 25% (87,500 per dry cell) to field labor, and about 20% ($70,00 per dry cell) to factory labor. The above material, field labor, and factory labor cost estimates are weighed average values which were determined on the basis of reasonable assumptions for material, field labor, and factory labor cost breakdowns for each of the major dry cooling tower cost components.

2.4.3 Once-through Cooling Systems

The once-through cooling system is intended to include the necessary intake structure (including the trash rack), pumps, piping, and discharge structures. Where by-pass cooling channels are used, water is assumed to be raised by pumps [3.05 m (~10 ft)] where a part is sent to the condenser and the remainder used to reduce the temperature of the condenser discharge water before it is returned to the river proper. A typical sketch is shown in Figure 2.8.

2.4.4 Condenser Performance & Cost

The radial flow type of surface condenser was developed by Westinghouse to insure adequate steam distribution to all parts of the condensing surface with a minimum of pressure drop. The basic principle is the same for all Westinghouse condensers whether they are of the small, round shell design or of the very large, rectangular shell design.

In the large radial flow condenser, the tube banks are usually rectangular in cross section with the corners well rounded to promote the smooth flow of steam. An air-offtake core is located near the center of each tube bank so that the flow of steam is radially inward from the steam space surrounding the banks toward the central core. The tube arrangement is such that as the flow nears the center of the bundle, the free flow area becomes progressively smaller. As the weight flow is reduced through condensation, the velocity remains substantially constant.

Dwg. 6352A40

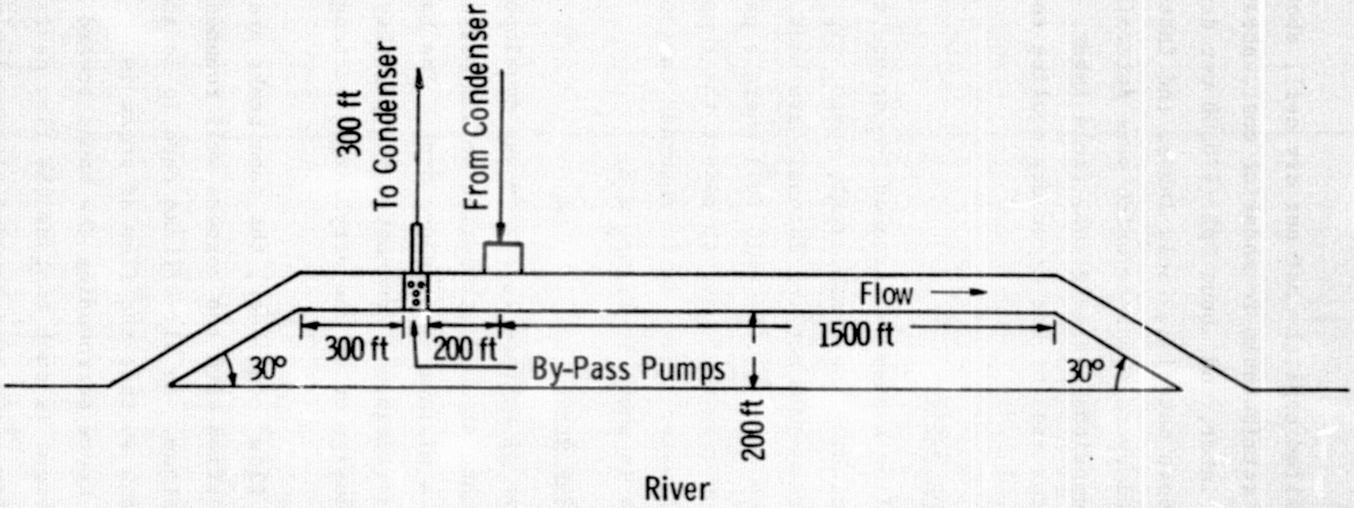


Fig. 2.8—Assumed arrangement of once-through cooling system with by-pass canal

Dwg. 6352A40

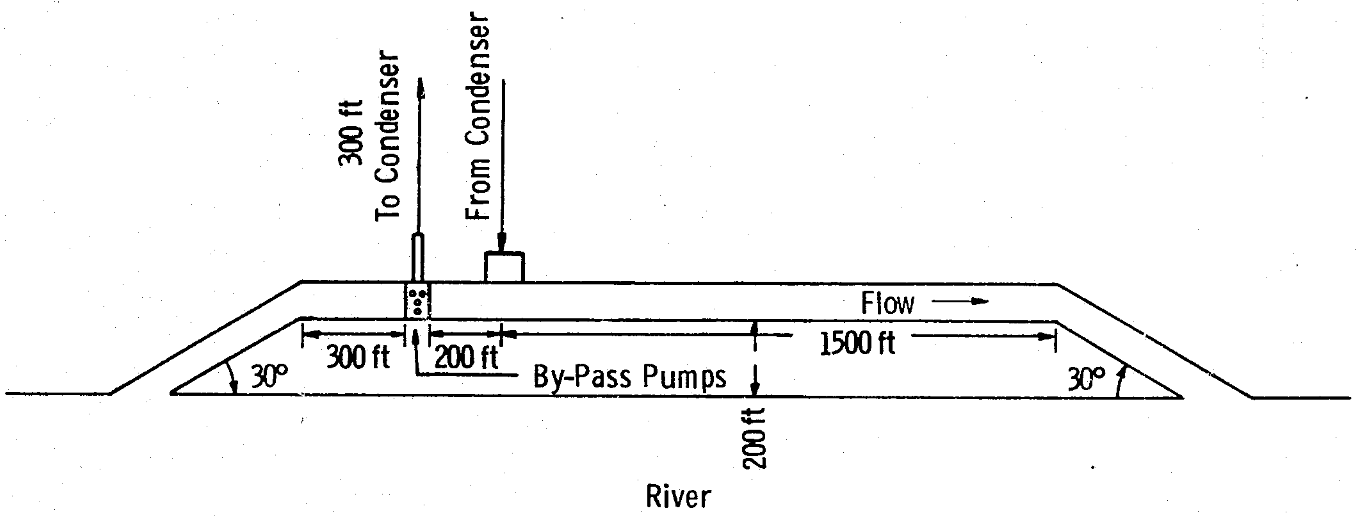


Fig. 2.8—Assumed arrangement of once-through cooling system with by-pass canal

The banks are arranged so that the pressure loss for travel from any point on the circumference of a bank to the central core is approximately the same as from any other point on the circumference to the core. Equal venting of all portions of the tube banks is thus assured.

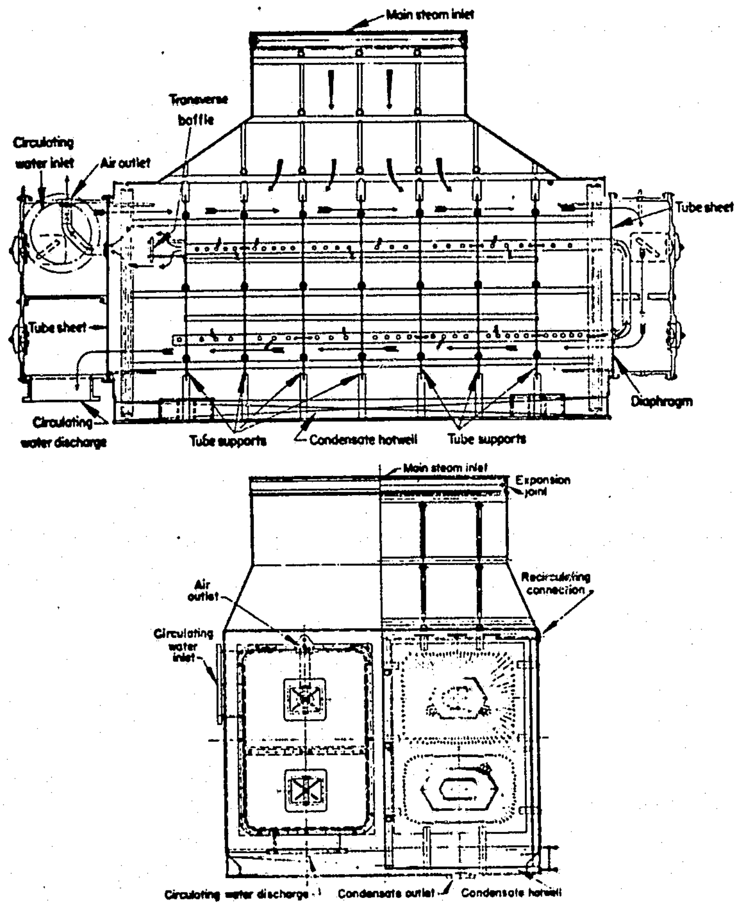
The mixture of air and steam reaching the core enters the air-offtake pipe through orifices which are carefully sized to control the venting along the length. Air and gases flow through the vent pipe to the cold end, where a baffle arranged transversely to the flow causes the vent mixture to cross several rows of the coldest tubes before it enters the offtake pipe leading to the air-removal apparatus. The final pass over these coldest tubes reduces the amount of steam in the vented gases to a minimum value, thus making the best use of the capacity of the air-removal equipment.

A very important function of the surface condenser is to ensure that the leaving condensate is free of dissolved gases, particularly oxygen, that cause corrosion to other parts of the cycle. The quantity of gas that a liquid will hold in solution depends on the nearness of the liquid temperature to its saturation temperature. If there is no subcooling, there can be no gas. In a radial flow type of condenser, the condensate formed in the tube bank must fall through the lower portion of the bank in counterflow with the entering steam. It then falls through a generously sized reheating belt where the temperature is brought as close as is possible to the saturation temperature. Gas-free condensate is thus assured.

Condensers are generally arranged for either one or two passes of the circulating water. More passes can be used but in practice seldom are. Generally speaking, a single-pass condenser is used where circulating water is plentiful and the fixed pumping head is not high. Single-pass condensers usually require less tube surface area but more circulating water than a two-pass condenser. The choice of the number of passes is based on economic considerations and plant layout.

TWO-PASS RECTANGULAR SURFACE CONDENSER

CORE AIR-OFFTAKE



FEATURES

1. Steam admitted to tube banks at top, sides and bottom—large entrance area—low steam velocity.
2. Air removed from center of tube banks—short steam flow paths—low pressure loss.
3. Longitudinal control of vent flow allows variations in steam side pressure loss with changing circulating water temperature.
4. Condensate dripping through steam below tube bank is reheated and de-aerated.
5. Open design minimizes concentration of corrosive non-condensibles by dilution with falling condensate.
6. Deep water boxes—low approach velocity to tube end—long tube life.
7. Shell end diaphragms—absorb differential expansion between tubes and shell.
8. Rectangular design effectively utilizes space in turbine foundation.

Fig. 2.9—Typical two-pass condenser with core air-offtake

In a two-pass condenser, the tubes of each pass are usually arranged into a separate tube bank with central core, as described above (see Figure 2.9). The first pass of circulating water is through the top bank. This bank, because of the colder water, is capable of condensing more steam than the second pass. A ratio of 60 to 40% is a reasonable approximation of the division of heat load between passes. Because of this difference in duty, the first pass tube bank is larger than the second, although it has the same number of tubes (because of a more compact tube layout). Pressure drops for steam flowing into both banks are thus more nearly equalized.

Figure 2.9 shows sections of a typical two-pass condenser of the core air-offtake design. Note the air-offtake jumper pipe in the reverse waterbox connecting the air-offtake sections of the first and second passes. The inlet box contains the pipe leading from the first-pass core to the air-removal equipment. Placing these pipes in the waterboxes allows the vented gases to be removed from the center of the banks without leaving any untubed lanes to the perimeters of the tube banks.

In some of the small condensers variations in the venting scheme are used. All, however, use a central core and employ the radial flow principle. Figure 2.10 depicts a typical one-pass condenser showing central core arrangement and core air-offtake design.

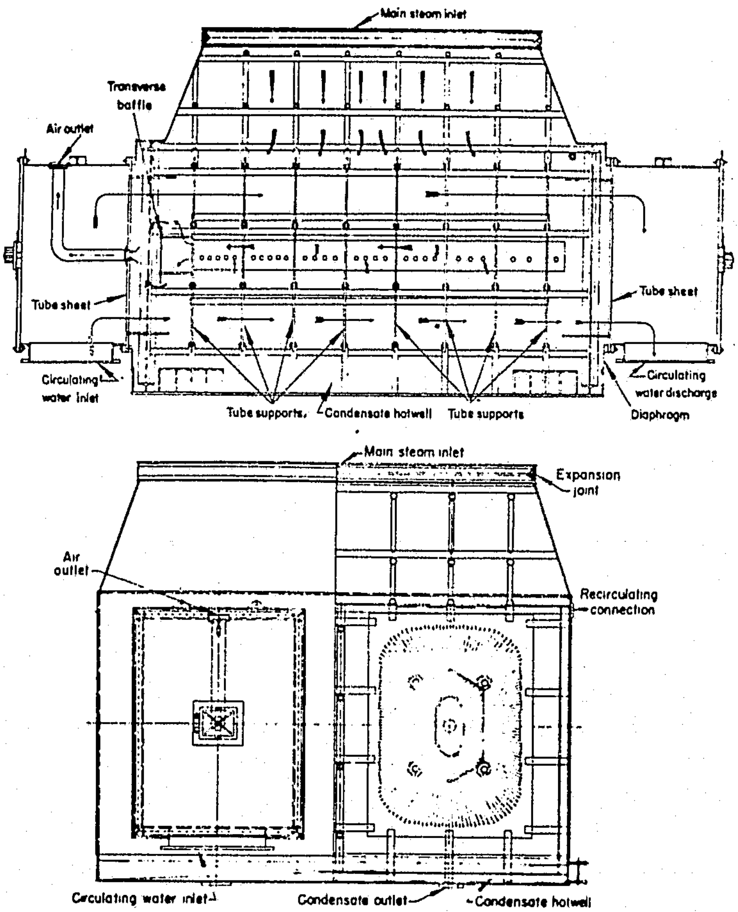
The condenser shell must withstand atmospheric pressure which, because of the size of the units, becomes a very considerable force. The flat shell plates of a rectangular condenser are well braced by support pipes which, in turn, transmit the load to the tube support plates.

Support for the tube sheets, where the pressure may be rather high, is provided by the tubes which are rolled into the tube sheets at each end.

Differential expansion between shell and tubes is provided for by the use of a steel diaphragm plate located between the shell and

ONE-PASS RECTANGULAR SURFACE CONDENSER

CORE AIR-OFFTAKE



FEATURES

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7. Shell end diaphragms—absorb differential expansion between tubes and shell.
8. Rectangular design effectively utilizes space in turbine foundation.

Fig. 2.10—Typical one-pass condenser with core air-offtake

shell-end flange. This plate flexes as the tubes expand or contract with respect to the shell because of the differential temperature.

Figure 2.11 shows a typical waterbox-to-shell joint, as well as the diaphragm construction. Gaskets are provided between waterbox and tube sheet as well as between tube sheet and shell flange. Note that every fifth stud, in this typical view, provides for holding the tube sheet to the shell joint when the waterbox is removed.

If the condenser is rigidly supported, provisions are made for relative movement between condenser and turbine by means of a rubber or stainless steel expansion joint at the steam inlet. If the condenser is rigidly connected to the turbine, the condenser is supported on springs that are adjusted in such a manner that the loads imposed on the turbine casing due to condenser weight, circulating water pipe reactions, and so forth, are within the allowable loads specified by the turbine manufacturer.

2.4.4.1 Condenser Performance

The performance of a condenser is dependent on several factors. They are:

- Condenser heat load
- Circulating water flow
- Circulating water temperature
- Cleanliness of inside tube surfaces
- Thermal conductivity of tube material
- Total surface area of tubes.

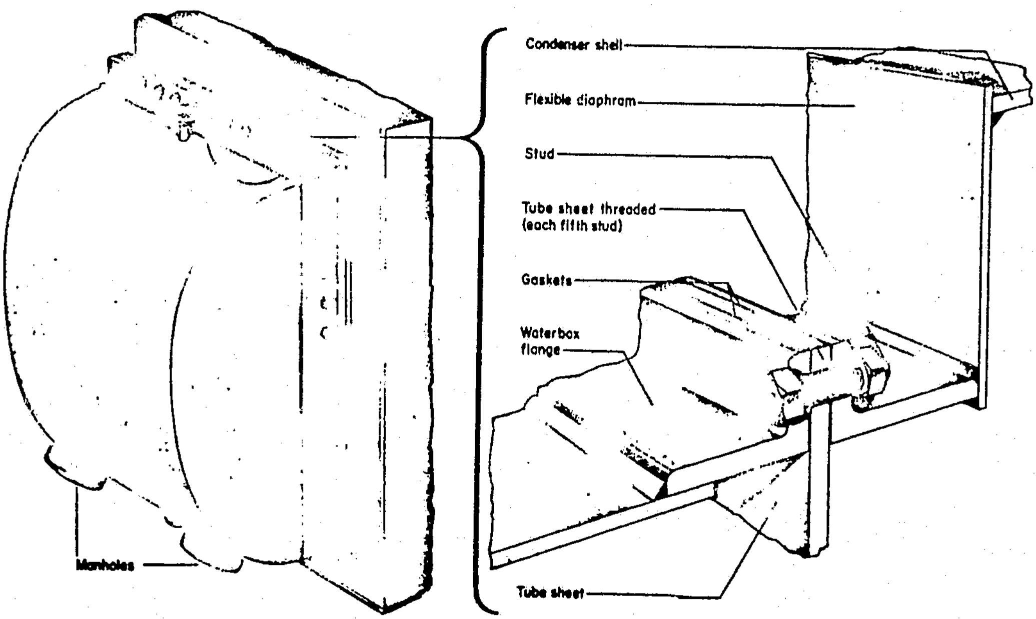
Performance may be expressed very simply by the following three equations:

$$Q = UA\theta = UA \frac{T_o - T_i}{\ln \frac{(T_s - T_i)}{(T_s - T_o)}} \qquad (2.1)$$

$$Q = W_c \; C_p \; (T_o - T_i) \qquad (2.2)$$

$$Q = W_s \Delta h \qquad (2.3)$$

WATERBOX—TUBE SHEET ATTACHMENT
AND
FLEXIBLE STEEL SHELL DIAPHRAGMS



FEATURES

1. Flexible steel diaphragms permit differential expansion between tubes and shell.
2. Diaphragm eliminates expansion joint from shell.
3. When water box is removed, tube sheet-to-shell joint is maintained by every fifth stud which is threaded into the tube sheet.
4. Gaskets of cotton duck, impregnated with red lead and linseed oil, assure a permanently tight joint between shell and tube sheet.
5. Diaphragm provides complete support for tube sheet and waterbox, eliminating need for external waterbox supports.

Fig. 2.11—Waterbox-tube sheet attachment and flexible steel shell diaphragms

where Q = Heat transferred by condenser, Btu/hr

U = Overall heat transfer coefficient, Btu/hr-ft^2-°F

θ = Log mean temperature difference, °F

W_c = Circulating water flow, lb/hr

C_p = Specific heat of water which may be taken as one over the range of condenser operation, Btu/lb-°F

T_s = Condensing temperature (saturation), °F

T_o = Circulating water outlet temperature, °F

W_s = Turbine exhaust steam flow, lb/hr

Δh = Enthalpy difference between exhaust steam and condensate, Btu/lb.

A curve of U vs water velocity is presented on Figure 2.12. Note that there are three curves, one for 15.875 and 19.05 mm (5/8 and 3/4 in) tubes, one for 22.22 and 25.4 mm (7/8 and 1 in) tubes, and one for 28.58 and 31.75 mm (1-1/8 and 1-1/4 in) tubes. The relation between U and condenser tube water velocity is given by Equation 2.4.

$$U = J\sqrt{V} \qquad (2.4)$$

where J = 259 for 15.8 mm (5/8 in) od tubes
263 for 25.4 mm (1 in) od tubes
267 for 31.75 mm (1.25 in) od tubes

V = Tube water velocity in ft/s.

This figure gives a basic, uncorrected "U" which must be modified by three correction factors. These are C_t for circulating water inlet temperature, C_m for tube material and gauge factor, and C_c for tube cleanliness. The factor C_t is presented on Figure 2.13; the factor C_m is presented on Table 2.39. The factor C_c is dependent on the fouling that is allowed to accumulate on the inside surface of the tubes. A design value of C_c was taken as 0.85. For particularly good water, a

Curve 680365-A

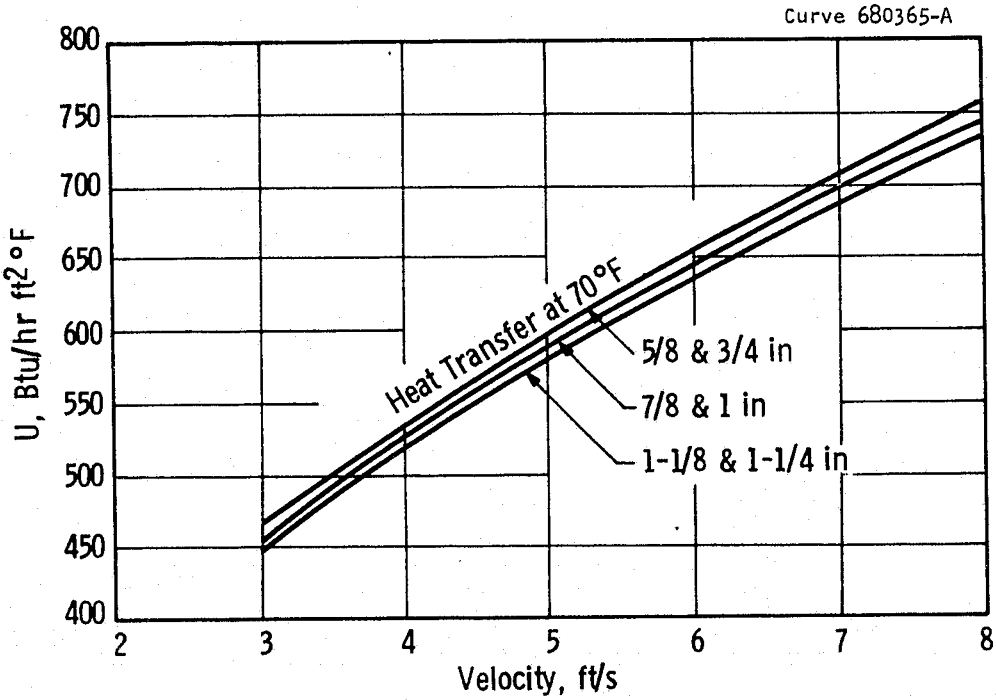


Fig. 2.12—Condenser heat transfer coefficient

design value of 0.9 may be used; for bad water, a design C_c as low as 0.7 is sometimes used. In checking the performance of a condenser, the C_c value is an unknown and must be estimated or determined.

This study assumed single-pass condensers, with 25.4 mm (1 in) od 1.245 mm (18 gauge, 0.049 in) wall admiralty metal tubes. The length of the tubes was calculated to give sufficient area after the number of tubes had been determined, based on a water velocity of 2.134 m/s (7 ft/s). (The number of condenser shells was assumed equal to the number of double-flow, low-pressure turbine ends.)

Figure 2.14 indicates the condenser surface area required to reject 293.02 MWt (10^9 Btu/hr) as a function of condenser terminal temperature difference for circulating water ranges of 8.33, 11.11, 13.89, 16.67, 19.44, 22.22, and 27.78°K (15, 20, 25, 30, 35, 40, and 50°F). This figure assumed a cleanliness factor of 0.875, a gauge factor of 1, a tube water velocity of 2.134 m/s (7 ft/s), and a circulating water inlet temperature of 310.94°K (100°F). A table on the face of the figure gives a multiplier to be used for other inlet water temperatures. This table is presented for the convenience of the reader and was not used in the study. This temperature correction chart applies only to Figures 2.14 and 2.15 and not to the earlier figures.

2.4.4.2 Condenser Costs

Condenser shell pricing is a function of the number of tubes per shell, and the diameter and length and material of the tubes. The Westinghouse Price List No. 1312 data have been reduced to equation form in Equation 2.5.

$$
\begin{aligned}
\text{Condenser shell cost } \$ &= 800\ (97.3 + 7.69\ \text{TN}) \text{ for } 3/4 \text{ in. od tubes} \\
&= 800\ (102.2667 + 8.8033\ \text{TN}) \text{ for } 7/8 \text{ in. od tubes} \\
&= 800\ (102.5 + 10.835\ \text{TN}) \text{ for } 1 \text{ in. od tubes} \\
&= 800\ (102.25 + 12.975\ \text{TN}) \text{ for } 1\text{-}1/8 \text{ in. od tubes} \\
&= 800\ (102.15 + 15.225\ \text{TN}) \text{ for } 1\text{-}1/4 \text{ in. od tubes}
\end{aligned}
\tag{2.5}
$$

where TN is the number of tubes/shell divided by 1000.

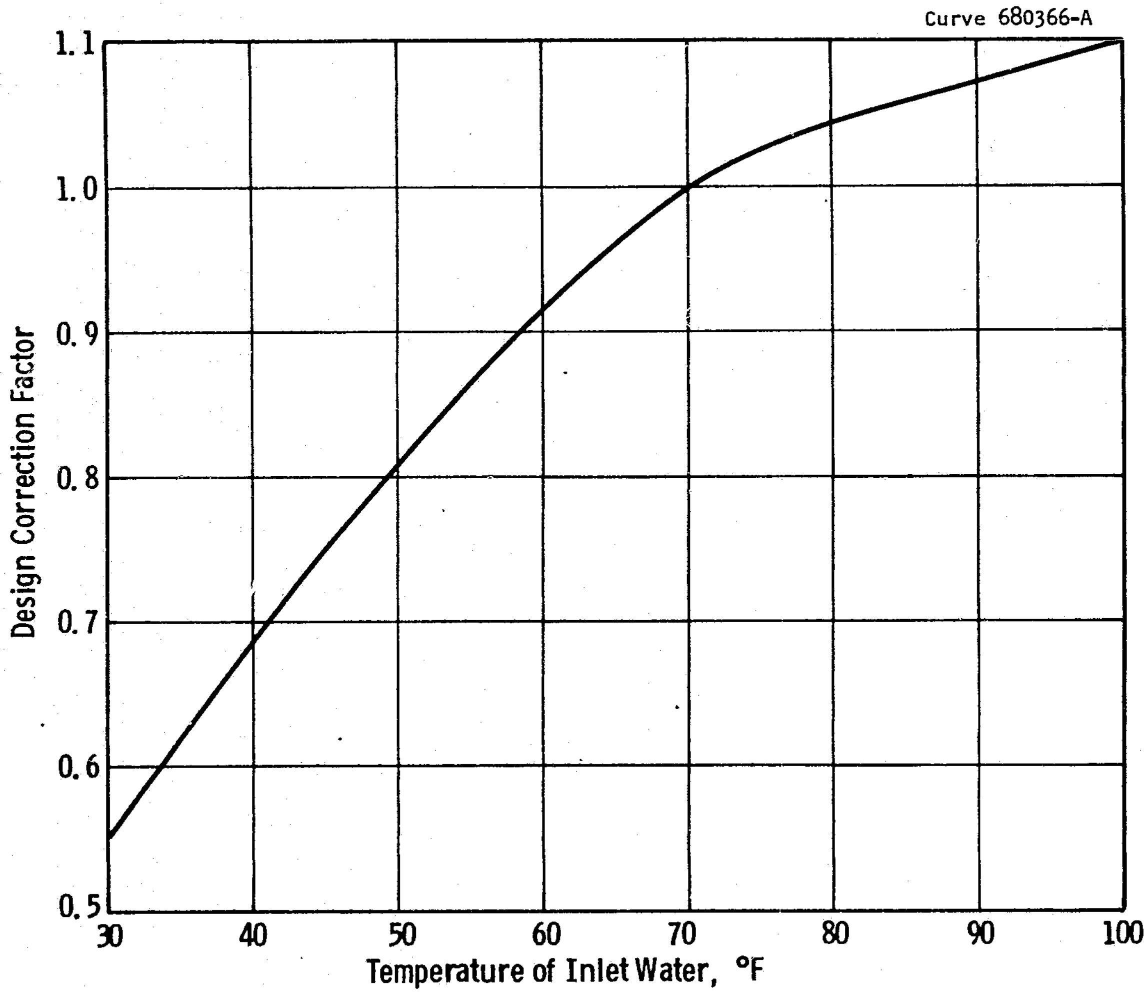


Fig. 2.13— C_t-Heat transfer coefficient temperature correction factor

Table 2.39 - C_m—Tube Material and Gauge Factor

Tube Materials	Tube Wall Gauge - BWG						
	24	22	20	18	16	14	12
Admiralty Metal	1.06	1.04	1.02	1.00	0.96	0.92	0.87
Arsenical Copper	1.06	1.04	1.02	1.00	0.96	0.92	0.87
Aluminum	1.06	1.04	1.02	1.00	0.96	0.92	0.87
Aluminum Brass	1.03	1.02	1.00	0.97	0.94	0.90	0.84
Aluminum Bronze	1.03	1.02	1.00	0.97	0.94	0.90	0.84
Muntz Metal	1.03	1.02	1.00	0.97	0.94	0.90	0.84
90-10 Cu-Ni	0.99	0.97	0.94	0.90	0.85	0.80	0.74
70-30 Cu-Ni	0.93	0.90	0.87	0.82	0.77	0.71	0.64
Cold-Rolled Low-Carbon Steel	1.00	0.98	0.95	0.91	0.86	0.80	0.74
Stainless Steels Type 410/430	0.88	0.85	0.82	0.76	0.70	0.65	0.59
Type 304/316	0.83	0.79	0.75	0.69	0.63	0.56	0.49
Type 329	0.78	0.76	0.74	0.69	0.65	0.60	0.54
Titanium (tentative)	0.85	0.81	0.77	0.71	----	----	----

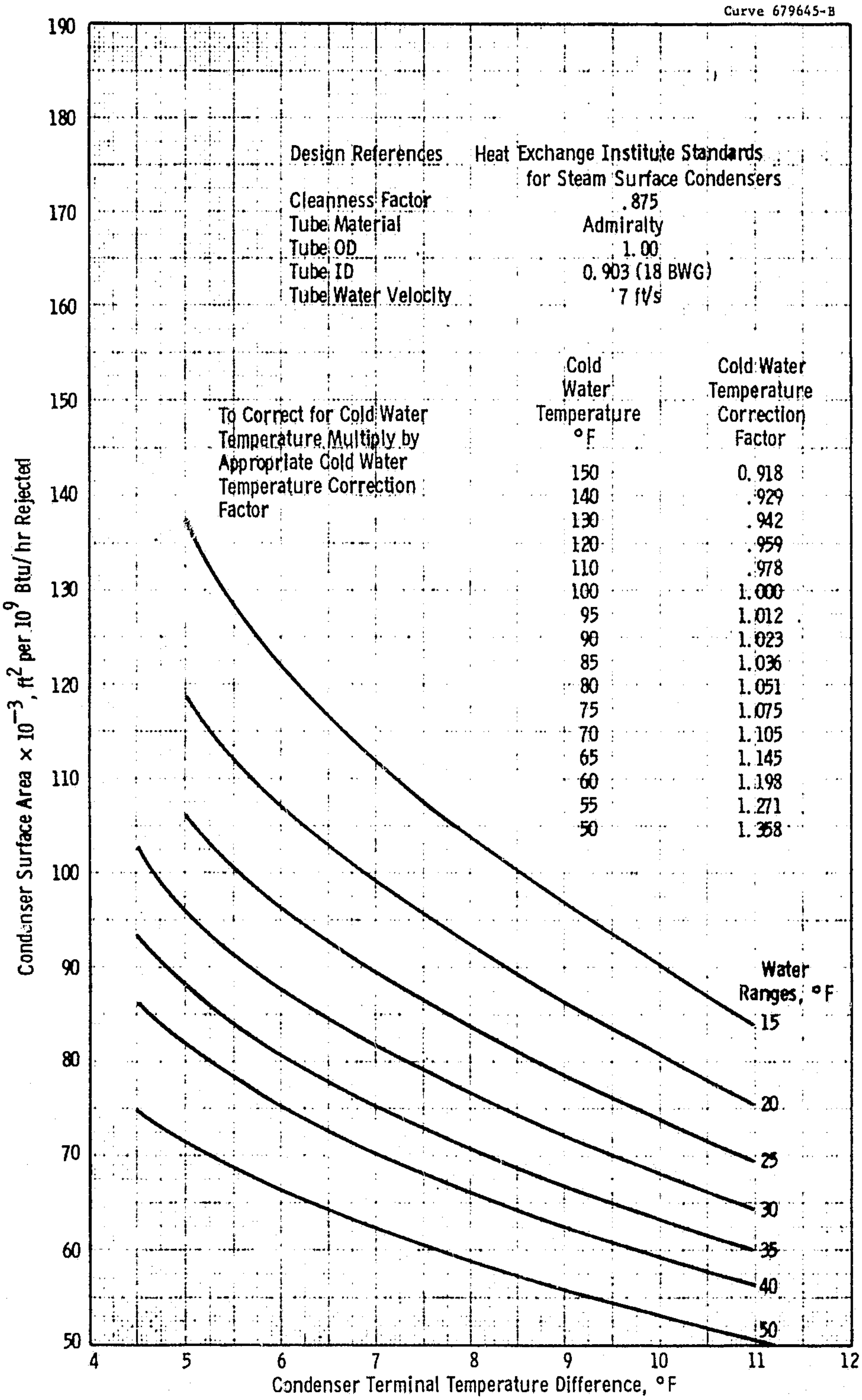


Fig. 2.14— Single-pressure condenser area requirements

This cost is then corrected for the length of the tubes using the shell length correction multiplier given in Equation 2.6.

$$\text{shell length multiplier} = 0.5725 + 0.015375\ L \qquad (2.6)$$

where L is the tube length in ft.

The admiralty metal tube is assumed to cost \$2.93/kg (\$1.33/lb); the price of rolling in is included in the shell cost. Other tube material would be priced as follows:

Type 304 SS	\$5.165/kg	(\$2.343/lb)
90-10 Cu-Ni	\$3.479/kg	(\$1.578/lb)
70-30 Cu-Ni	\$4.268/kg	(\$1.963/lb)

The condenser installation cost is assumed to be \$7.53/m^2 (\$0.70/ft^2) of tube surface.

Figure 2.15 indicates the required condenser installed capital cost per 293.02 MWt^3 (10^9 Btu/hr) heat rejected as a function of condenser terminal temperature difference and range for condenser designs represented in Figure 2.14.

The total condenser installed capital costs indicated in Figure 2.15 were determined by summing the condenser shell cost, the condenser tube cost, and the condenser installation cost.

The specific condenser shells which were priced for Figure 2.15 were two-shell, sinlge-pressure designs which had 28.57 mm (1-1/8 in) thick Muntz-metal tube sheets, 206.87 kPa (30 psi) semicylindrical carbon steel waterboxes, and 228.6 mm (9 in) deep hot wells.

Condenser tube costs of \$19.38 m^2 (\$1.80/ft^2) surface area were used for Figure 2.15. This costing rate estimate was obtained from C. T. Main, Inc. It was based on an average tube costing rate determined by averaging data obtained from several recent condenser bids. Subsequent information obtained from the Westinghouse Heat Transfer Division

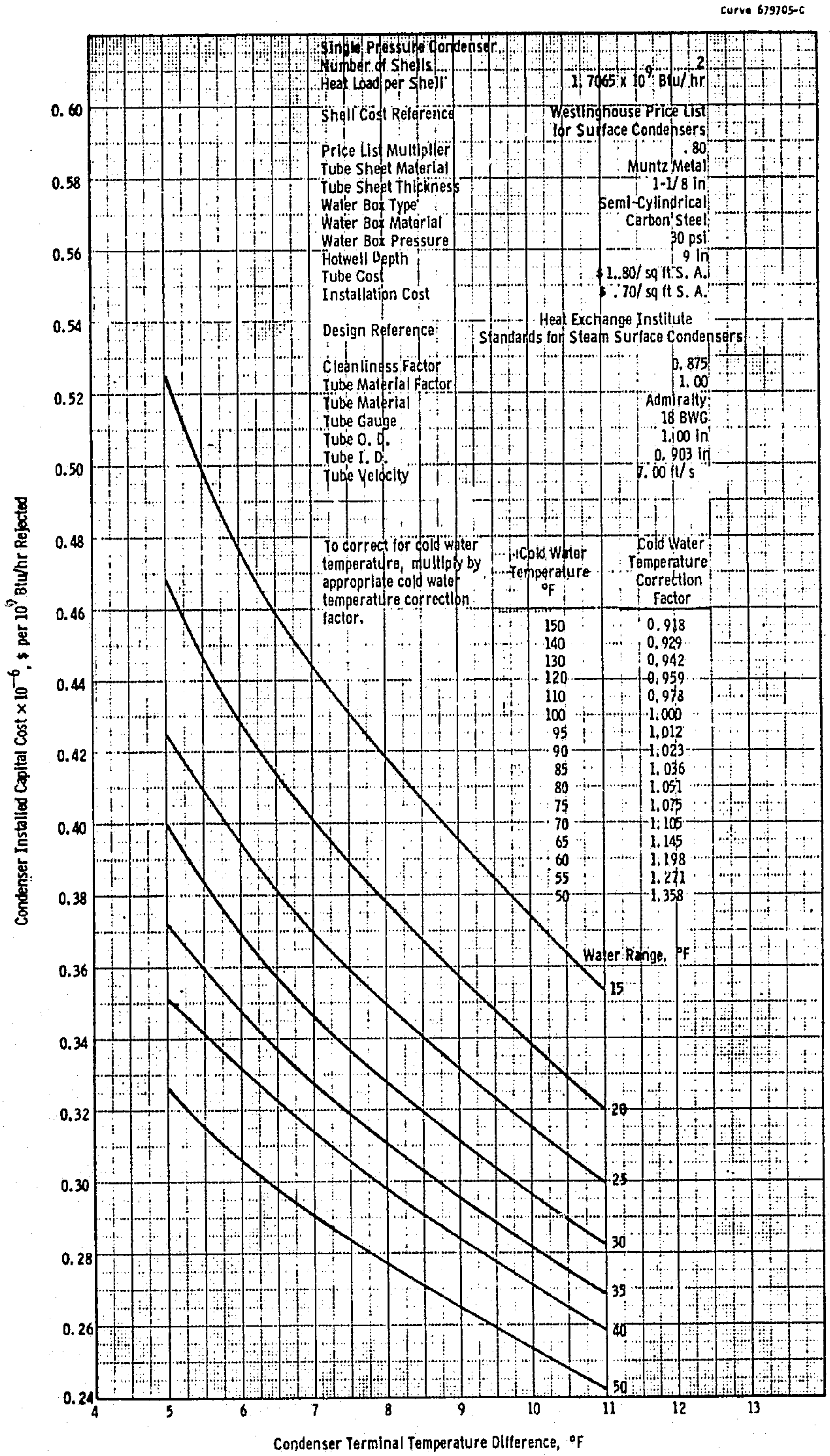


Fig. 2.15—Total condenser installed capital cost

Purchasing Department, however, indicates that for the specific material under consideration [25.4 mm od (1 in. od) 18 BWG admiralty], the costing rates indicated previously would be more appropriate. The use of the previous values would, of course, result in higher condenser installed capital costs than those indicated in Figure 2.15.

2.4.5 Circulating Water Pumps and Piping

The circulating water system is assumed to include all piping, the circulating water pumps and drive motor, the pump house, and the makeup and blowdown and water treatment equipment, including the inlet, discharge, screens, trash racks, and so on.

2.4.5.1 Piping Layouts

The assumed piping layouts for the wet, dry, and once-through systems are shown in Figures 2.16, 2.17, and 2.8 respectively. These would, in practice, vary with site and plant size. For this study these numbers have been assumed fixed and a function of circulating water range (temperature change in the condenser). The once-through system shown in Figure 2.8 includes the use of a mixing canal in which the condenser effluent is mixed with sufficient river water that the canal discharge is within 2.78°K (5°F) of the river water temperature. The once-through cooling systems used in this study did not use the mixing canal concept.

2.4.5.2 Piping and Circulating Water Equipment Capital Cost

The cost of the piping was determined by using an installed cost of \$516.67/m dia-m length (\$4.00/in. dia-ft length), of which about 37.5% was material and the remainder labor for both main and circulating water piping and branch piping, including valving. All piping was sized on the basis of a circulating water velocity of 3.048 m/s (10 ft/s).

The main circulating water pumps were estimated at \$19,812 per m^3/s (\$1.25 per gpm) circulating water flow, assuming the water was not brackish. This installed cost is assumed to be half labor and half material.

The circulating water pump motor was costed at \$48.25 per rated shaft kW (\$36 per hp) with an assumed pump efficiency of 85%. These

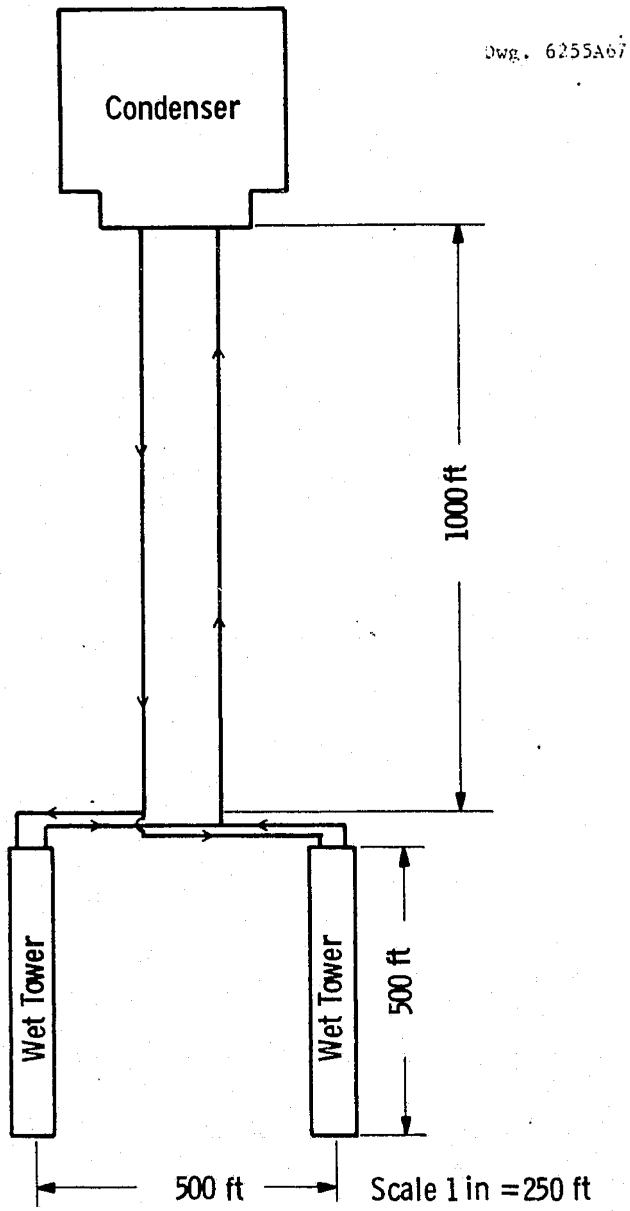


Fig. 2.16—Assumed wet cooling tower arrangement

costs are assumed to be half labor and half material. The pump house cost was assumed to be \$19,020 per m^3/s (\$1.20 per gpm) circulating water flow of which 54.17% was material and the remainder installation.

Wet cooling tower makeup and blowdown equipment including inlet, discharge, screens, trash racks, etc., were assumed to cost \$19,813 per m^3/s (\$1.25 per gpm) circulating water flow, 36% of which was material.

The assumed piping layout for all wet tower systems, as shown in Figure 2.16, has 609.6 m (2000 ft) of main circulating water piping and 371.8 m (1220 ft) of branch piping handling half the total flow. The circulating water pump is assumed to provide a 298.89 kPa (100 ft of water) head. The total cost of the circulating water system for the wet cooling tower system (the sum of the costs cited above) is given by Equation 2.7.

$$\text{Total Capital Cost} = \left(\frac{\text{Dollars}}{10^9 \text{ Btu/hr}}\right) = \frac{1{,}776{,}447}{R^{1/2}} + \frac{9{,}541{,}997}{R}$$

$$\text{Material Cost} = \left(\frac{\text{Dollars}}{10^9 \text{ Btu/hr}}\right) = \frac{667{,}167}{R^{1/2}} + \frac{4{,}520{,}898}{R} \qquad (2.7)$$

$$\text{Installation Cost} = \left(\frac{\text{Dollars}}{10^9 \text{ Btu/hr}}\right) = \frac{1{,}110{,}279}{R^{1/2}} + \frac{4{,}021{,}098}{R}$$

where R is the range in °F.

Figure 2.18 shows the cost of the wet tower circulating system required to reject 293.02 MWt (10^9 Btu/hr) as a function of range.

The assumed piping layout for the circulating water system associated with a dry cooling tower and shown in Figure 2.17 consists of 609.6 m (2000 ft) of main circulating water piping, 323.09 m (1060 ft) of branch piping carrying half flow, and 668.12 m (2192 ft) of branch piping carrying quarter flow. A pump head of 224.17 kPa (75 ft of water) was assumed for the dry tower circulating water system. The total cost

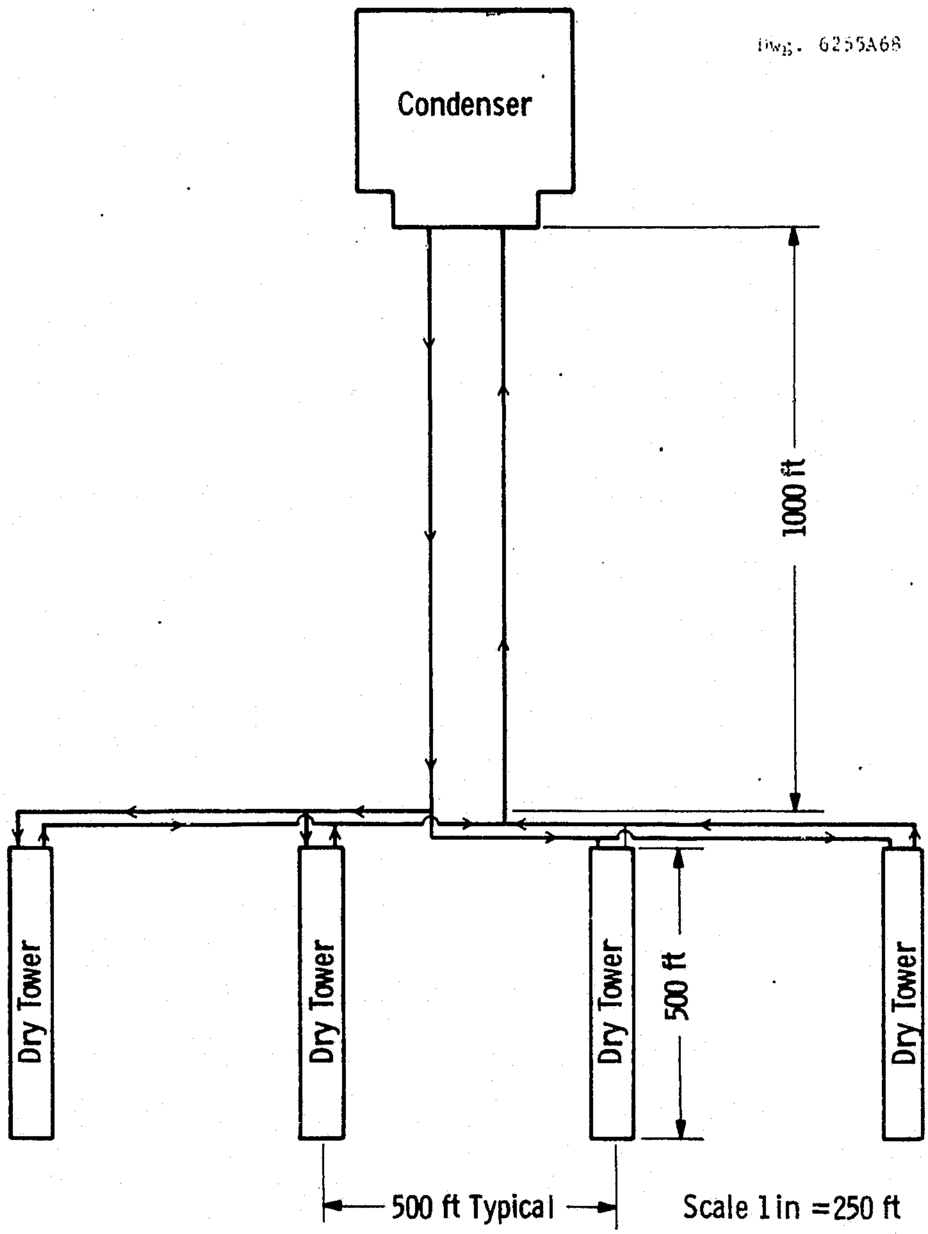


Fig. 2.17—Assumed dry cooling tower arrangement

of the circulating water system for the dry cooling tower system is given by Equation 2.8.

$$\text{Total Capital Cost} = \left(\frac{\text{Dollars}}{10^9 \text{ Btu/hr}}\right) = \frac{2{,}386{,}360}{R^{1/2}} + \frac{6{,}506{,}238}{R}$$

$$\text{Material Cost} = \left(\frac{\text{Dollars}}{10^9 \text{ Btu/hr}}\right) = \frac{894{,}887}{R^{1/2}} + \frac{3{,}353{,}158}{R} \qquad (2.8)$$

$$\text{Installation Cost} = \left(\frac{\text{Dollars}}{10^9 \text{ Btu/hr}}\right) = \frac{1{,}491{,}478}{R^{1/2}} + \frac{3{,}153{,}078}{R}$$

where R is the range in °F.

The dry system piping and circulating water equipment installed capital cost in Equation 2.8 includes cost for the dry system piping, circulating water pump, the circulating water pump motor, and the pump house.

The total dry cooling tower system piping and circulating water equipment installed capital cost per 293.02 MWt (10^9 Btu/hr) heat rejected is plotted as a function of range in Figure 2.19.

The piping layout for the once-through cooling system assumed 182.88 m (600 ft) of main circulating water piping which handles the entire condenser water flow. A pumping heat of 149.44 kPa (50 ft of water) was assumed. The installed cost of the intake and discharge facilities was taken as \$76.076 per m^3/s (\$4.80 per gpm) circulating water, which was regarded as being 36.4% material and the balance installation. The assumed cost of a conventional once-through piping and circulating water system is given by Equation 2.9.

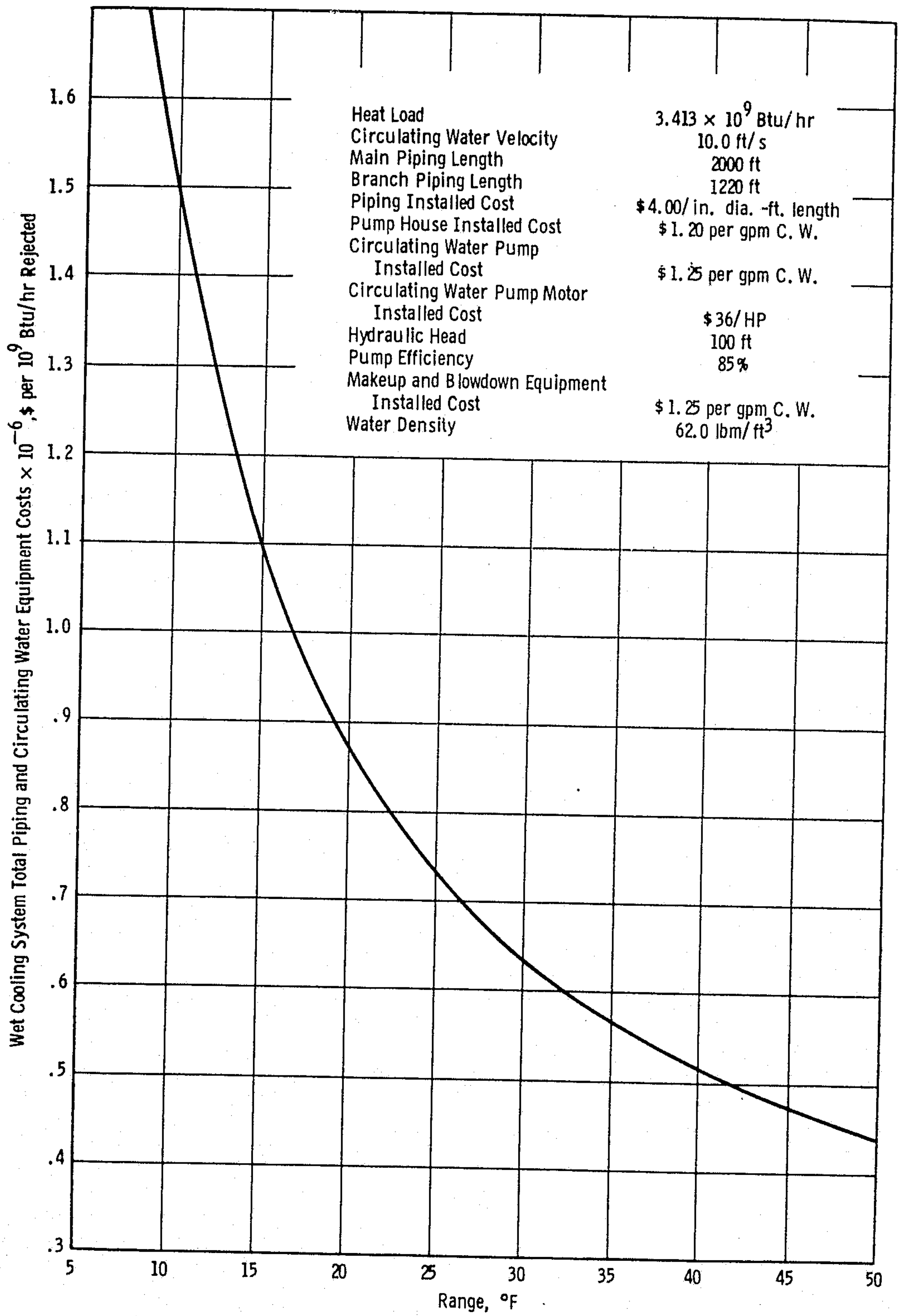


Fig.2.18—Cost of wet tower circulating water system (pump, piping and associated equipment)

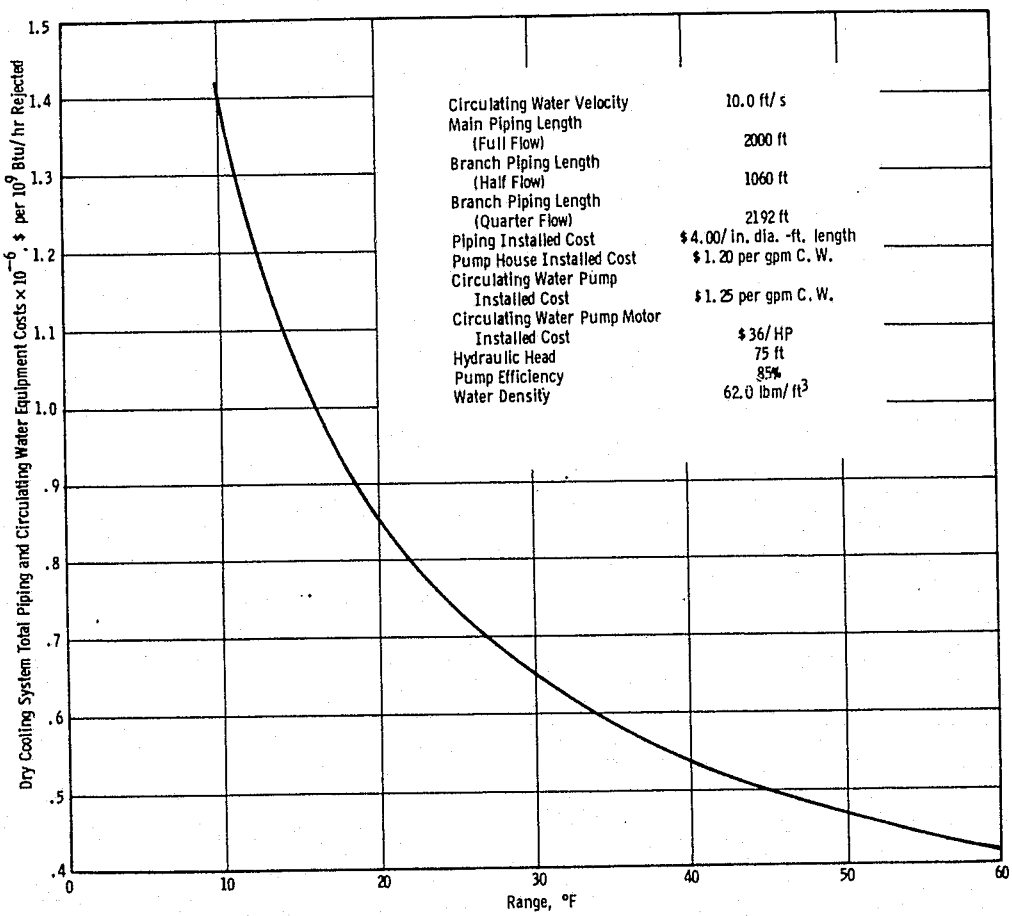


Fig. 2.19 – Cost of dry cooling tower circulating water system (pump, piping and associated equipment)

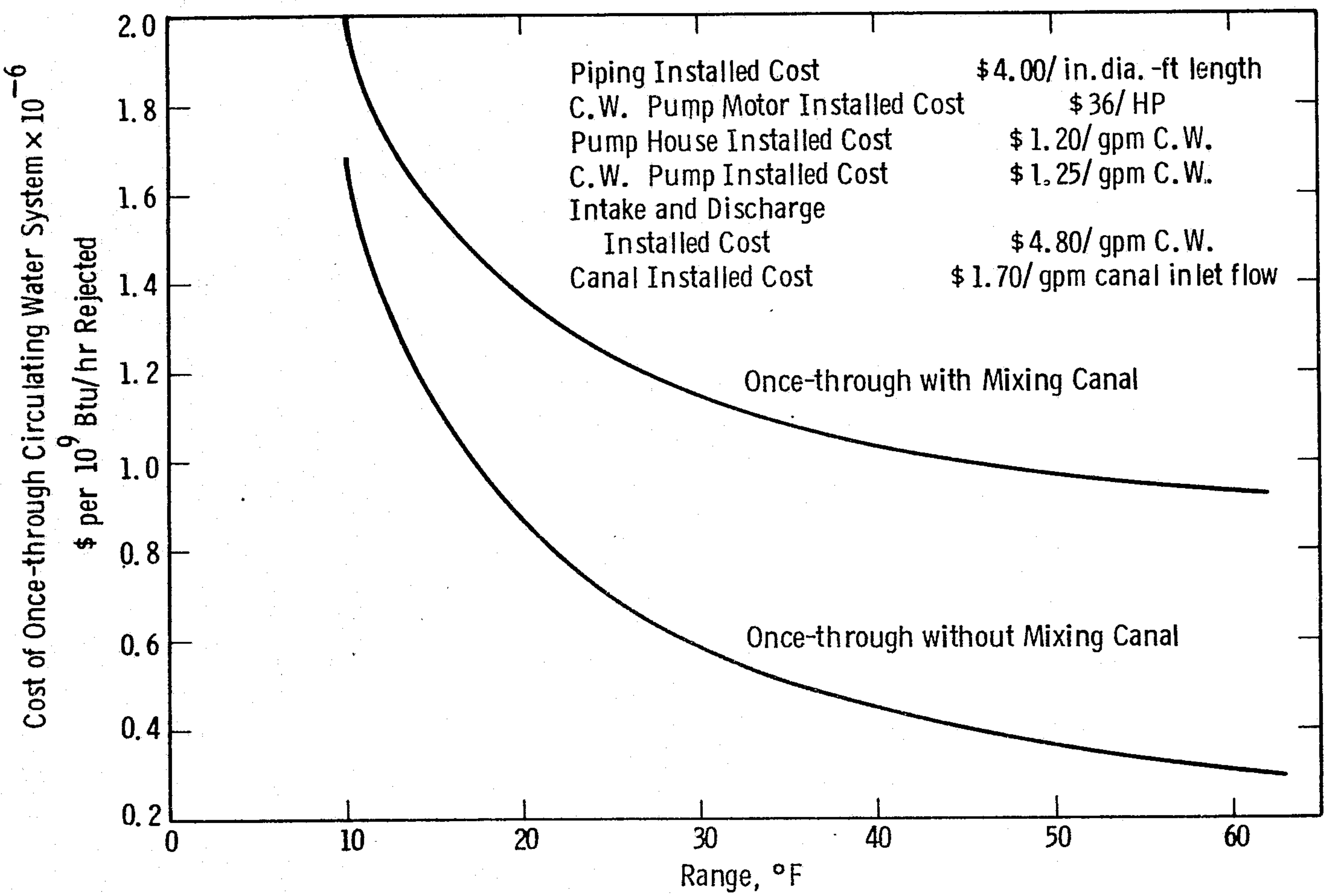


Fig. 2.20— Cost of once-through circulating water system (pump, piping and associated equipment)

$$\text{Total Capital Cost} = \left(\frac{\text{Dollars}}{10^9 \text{ Btu/hr}}\right) = \frac{372{,}333}{R^{1/2}} + \frac{15{,}575{,}316}{R}$$

$$\text{Material Cost} = \left(\frac{\text{Dollars}}{10^9 \text{ Btu/hr}}\right) = \frac{139{,}624}{R^{1/2}} + \frac{6{,}587{,}178}{R} \qquad (2.9)$$

$$\text{Installation Cost} = \left(\frac{\text{Dollars}}{10^9 \text{ Btu/hr}}\right) = \frac{232{,}708}{R^{1/2}} + \frac{8{,}988{,}138}{R}$$

where R is the range in °F.

The total cost of the conventional once-through cooling circulating water systems described in Equation 2.8 is shown graphically in Figure 2.20 and includes installed cost of the intake piping, pump, pump house, and discharge required to reject 293.02 MWt (10^9 Btu/hr). The assumed cost of a 2.78°K (5°F) mixing canal, with associated equipment, was \$2,694 per m^3/s (\$1.70 per gpm) canal flow, 47% material, and the remainder installation. Equation 2.10 shows the cost of this alternative circulating water system. It is displayed graphically as the upper curve in Figure 2.20.

$$\text{Total Capital Cost} = \left(\frac{\text{Dollars}}{10^9 \text{ Btu/hr}}\right) = 680{,}000 + \frac{372{,}333}{R^{1/2}} + \frac{12{,}175{,}316}{R}$$

$$\text{Material Cost} = \left(\frac{\text{Dollars}}{10^9 \text{ Btu/hr}}\right) = 320{,}000 + \frac{139{,}624}{R^{1/2}} + \frac{4{,}987{,}178}{R} \qquad (2.10)$$

$$\text{Installation Cost} = \left(\frac{\text{Dollars}}{10^9 \text{ Btu/hr}}\right) = 360{,}000 + \frac{232{,}708}{R^{1/2}} + \frac{7{,}188{,}138}{R}$$

where R is the range in °F.

The by-pass canal assumed a pump at the inlet with a total head of 2.99 kPa (10 ft of water).

Curve 679791-C

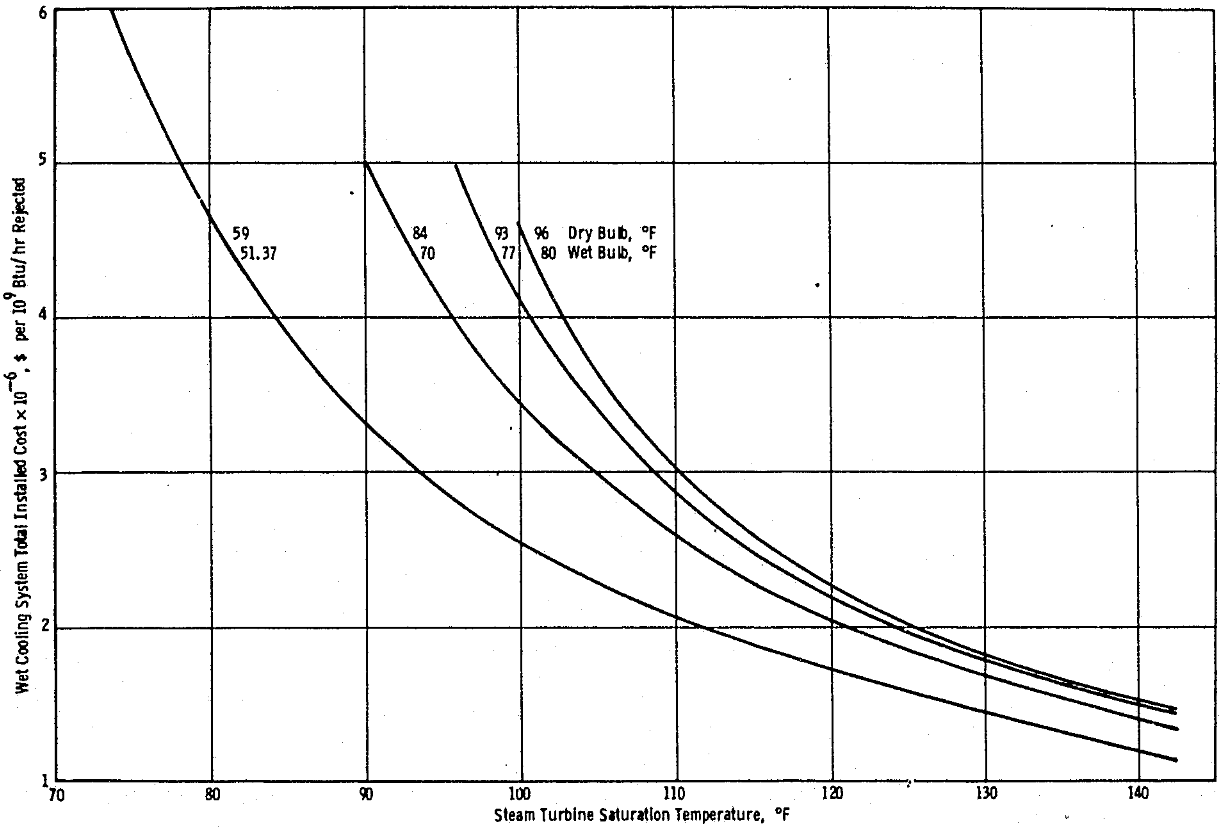


Fig. 2.21 — Wet cooling system total installed cost

2.4.6 Total Cooling System Installed Capital Cost

2.4.6.1 Total Installed Capital Cost for a Wet Cooling Tower System

In order to assist the power plant designer in estimating the effect of turbine back pressure (condensing temperature) on the total heat rejection system installed cost, seven-point curves (Figure 2.21) were generated for each of the four previously specified ambients. Each of these seven points was based on a different range, approach, and terminal temperature difference. The choices may not have been optimum but are felt to be indicative of the cost of a well-designed system. The costs include the wet cooling tower, circulating water, makeup and blowdown system, and the condenser.

2.4.6.2 Total Installed Capital Cost for a Dry Cooling Tower System

Figure 2.22 is a useful curve for the approximation of typical costs of a dry cooling tower heat rejection system, including the tower, circulating water system, and condenser, as a function of the difference in temperature between the condensing temperature and the ambient dry bulb. The curve is based on 10 data points with different nonoptimized values of range and approach.

2.4.7 Operation and Maintenance Costs

2.4.7.1 Cooling Tower Fan Power Requirements

The wet cooling tower fan power requirements may be obtained by multiplying the required number of wet cooling tower cells obtained from Figures 2.2, 2.3, 2.4, or 2.5 by 162 kW/cell. This value is based on an assumed fan motor efficiency of 92%.

The dry cooling tower fan power requirements may be obtained by multiplying the required number of dry cooling tower cells obtained from Figure 2.7 by 283.80 kW/cell. This value is also based on an assumed fan motor efficiency of 92%.

Curve 679792-C

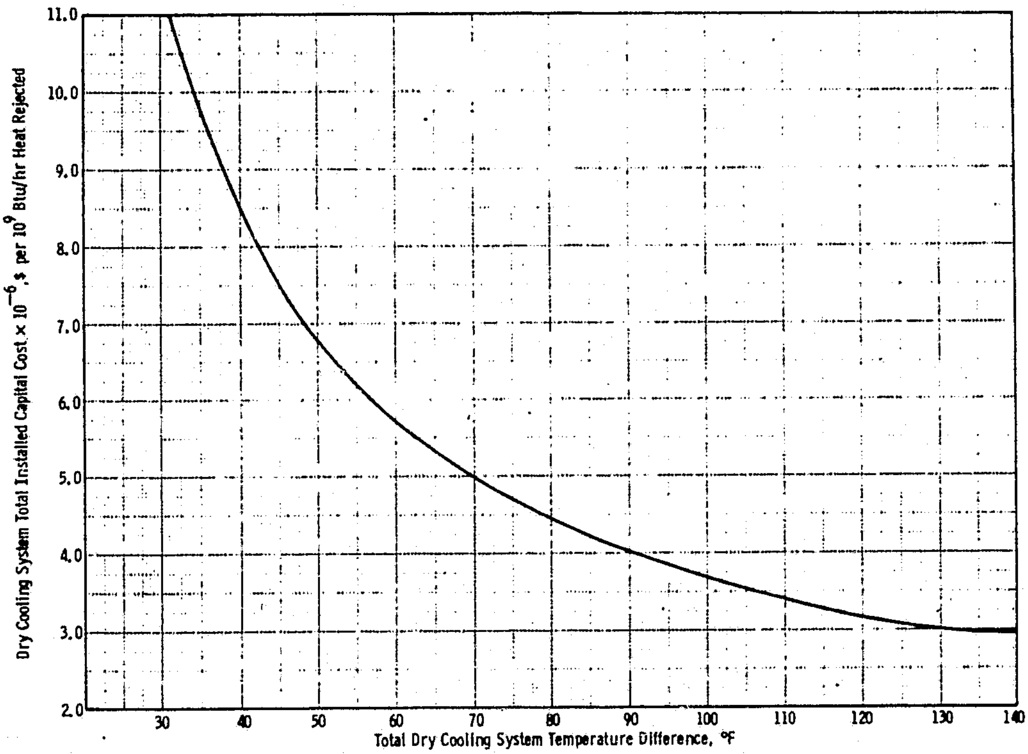


Fig. 2.22—Dry cooling tower system total installed capital cost

2.4.7.2 Circulating Water Pumping Power Requirements

Assuming a pump motor efficiency of 95% and a pump efficiency of 85%, the circulating water pumping power requirements are plotted as a function of range and hydraulic head in Figure 2.23.

Alternatively, making the same assumptions, the pumping power may be determined from Equation 2.11:

$$\text{Pumping Power, } \frac{\text{kW}}{10^9 \text{ Btu/hr}} = \left(\frac{\text{Head, ft}}{\text{Range, °F}}\right) 466.4 \tag{2.11}$$

For wet, dry, and conventional once-through systems, hydraulic heads of 30.48, 22.86, and 15.24 m (100, 75, and 50 ft) respectively, were used.

For once-through cooling systems with a by-pass canal, the total water pumping power requirements for the circulating water and the canal water may be determined from Equation 2.12.

$$\text{Pumping Power, } \left(\frac{\text{kW}}{10^9 \text{ Btu/hr}}\right) = 466.4 \left(2 - \frac{40}{R}\right) \tag{2.12}$$

This formula is based upon the following assumptions:

Range	°F
Pump motor efficiency	95%
Pump efficiency	85%
Canal hydraulic head	10 ft
Condenser flow hydraulic head	50 ft

2.4.7.3 Makeup Water Costs for Wet Cooling Tower Systems

Wet cooling tower makeup water costs ($/hr) can be determined from Equation 2.13.

$$\text{Makeup Water Cost, \$/hr} = \frac{(\text{Heat Load}) \times (\text{Ratio Wet}) \times \left(\frac{CC}{CC-1}\right) \times (\text{Water Cost})}{(\text{Latent Heat}) \times (8.33)} \tag{2.13}$$

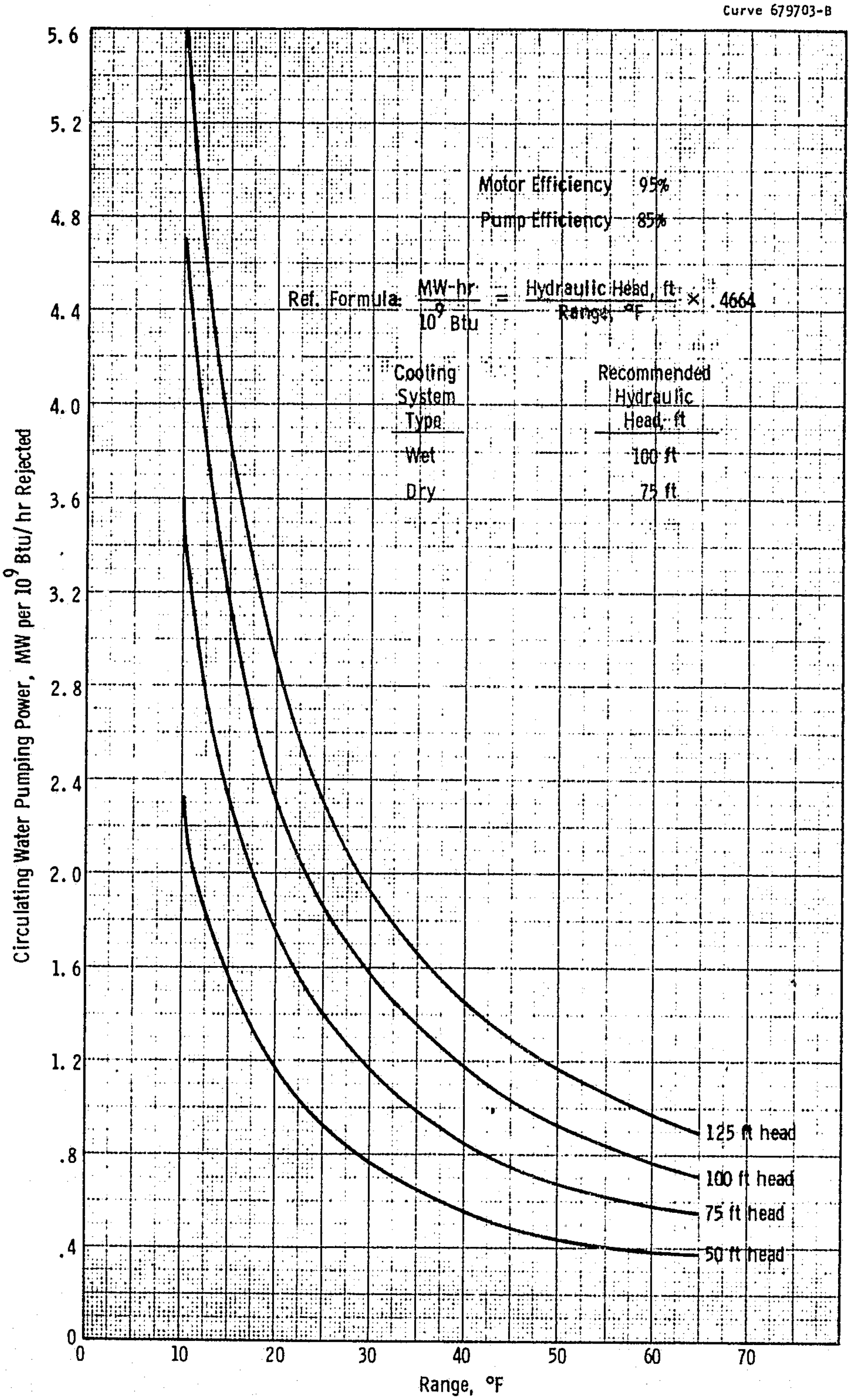


Fig. 2.23–Circulating water pump power requirement

making the following assumptions:

Ratio Wet (the portion of the total heat rejected from the wet cooling tower due to evaporative heat transfer) = .85 (i.e., 15% of the heat rejected is due to heating of incoming air)

CC (cycles of concentrations of salts in circulating water) = 2

Water Cost = $\$0.0211/m^3$ [$.00008/gal (i.e., 8 cents per thousand gallons)]

Latent Heat = Btu/lb

Heat Load = total heat rejected, Btu/hr.

2.4.7.4 Annual Maintenance Costs

The annual maintenance costs for all of the cooling systems considered in this study were approximated as 1.5% of the total cooling system installed capital costs.

2.5 Power Plant Labor Rates

The Middletown Site is 25 miles from a medium-sized city. The population center is large enough to assure a reasonable labor supply. A project agreement will be established. Travel and subsistence payments will not be required, and other labor practices will be reasonable. Productivity is assumed to be average. The many labor classification (traces) specified by NASA have been combined into two groupings—civil and electromechanical. Average wages for each have been specified by NASA. Further, it has been decided to use an average wage for the two groupings for the purpose of the Task I study. The values specified for civil labor were 2.222, 2.722, and 5.55 $/ks (8.00, 9.80, and 20.00 $/hr); for electromechanical labor 2.5, 3.153, and 6.389 $/ks (9.00, 11.35, and 23.00 $/hr), the second mentioned value in each case being the recommended base value. The averages actually used were

2.361, 2.944, and 5.972 \$/ks (8.50, 10.60, and 21.50 \$/hr) for the Task I study. In addition to these average values, labor rates of 1.667 and 4.167 \$/ks (6.00 and 15.00 \$/hr) were added. The installation charges for each point were modified by the ratio of one of these rates to the base rate of 2.944 \$/ks (10.60 \$/hr) to show properly the effect of construction labor rates on power plant costs.

Although not actually used in arriving at installation costs, Equation 2.14 has been suggested by the A/E to translate installation dollars into labor hours.

$$\text{Total Labor Hours} = \frac{\text{Total Direct Installation Cost}}{11.88} \qquad (2.14)$$

Equation 2.14 was developed assuming that an engineer-constructor would perform all work except certain specialty work such as piling, earthwork, chimney, cooling tower construction, waterfront construction, and like items. Engineer-constructor indirect costs are included in separate accounts (Subsection 2.6.2.1), but similar costs for the specialty-subcontracted portions are included in the direct cost estimates, since these services are usually purchased on a firm basis and subcontractor indirects are buried in the price.

Equation 2.14 is a product of 80% engineer-constructor labor at a labor rate of 10.60 + 20% at 10.60 x 1.6 = 11.88. The 1.6 multiplier is for subcontractor indirects included in direct installation cost accounts. These numbers are typical of several conventional plants whose costs were reviewed.

2.6 Power Plant Capitalization and Cost of Electricity

In order to minimize omissions and institute a uniform method of reporting results, the estimated costs associated with the purchase and installation of equipment, as well as site purchase and development, have been divided into 21 accounts. No attempt has been made to follow the Federal Power Commission (FPC) accounting system. Instead, the accounts were set up to apply somewhat generally to any power plant concepts. Some accounts may have no entries for one concept and yet prove to be very important to another. The accounts themselves have been broken into subaccounts. Different subaccounts were used for each concept as they seemed appropriate. In general, the tendency has been to present a far more detailed breakdown than the accuracy of this study may warrant. Each person responsible for inputing numbers to these accounts has been encouraged to identify a subaccount for any fraction of the total that he has identified in his cost estimating procedures. Since many of the components involved are not now commercially available, this will better enable the reader to assess the reasonableness of the results.

The 21 accounts selected to order the direct costs are presented in Table 2.40. Of these accounts, 1 through 7, 14, and 16 through 21 were treated in a general manner as balance of plant by the A/E, Chas. T. Main, Inc. In particular instances additional subaccounts will be added to some of these accounts for particular concepts as required.

2.6.1 Balance of Plant Engineering and Cost Assumptions

Generalized algorithms were developed to represent the cost of materials and the cost of installations associated with a power plant. These assumptions or the basis for these algorithms are presented in the following sections. They are based on historical data for existing steam plants, as modified to fit the site choosen and the particular power conversion process. Each base case was approached individually, with considerable attention given to the reasonableness of the balance of plant assumptions. The remainder of the parametric points were treated superficially, with allowance made for deviations from the base cases estimates.

All of the power systems studied are assumed to be commercial systems; that is, no R&D or unusual engineering or construction risks are included in estimates. Further, environmental requirements are for 1974 standards and will not change.

Table 2.40 - Direct Cost Accounts

Account No., K	Account Name
1	Site Development
2	Excavation and Piling
3	Plant Island Concrete
4	Heat Rejection System
5	Structural Features
6	Buildings
7	Fuel Handling
8	Fuel Processing
9	Firing System
10	Vapor Generators (fired)
11	Energy Converter
12	Coupling Heat Exchanger (unfired boilers, etc.)
13	Heat Recovery Heat Exchangers (recuperators)
14	Water Treatment (Demin and Polish)
15	Power Conditioning
16	Auxiliary Mech. Equipment
17	Pipe and Fittings
18	Auxiliary Electrical Equipment
19	Control and Metering
20	Process Waste Systems
21	Stack-Gas Cleaning

2.6.1.1 Site Development (Account 1)

The Middletown site described in WASH 1230 modified for a 1000 MW coal-fired fossil plant (UEC-AEC-720630) is being used for all large base-load plants studied. In these cases significant land and natural resource use and waste-handling or heat dispersion requirements result in the Middletown site being the most economical site.

In some intermediate and peaking plant cases, and especially for small fuel cell plants, the Middletown site may not be the best choice. Because these plants are small, clean, and attractive, a site in a semi-developed area closer to the user should be selected. Although land cost is higher for such industrial or commercial sites, transmission, site development, and other related costs may make using the Middletown site uneconomical.

Accordingly, two alternate types of sites will be used: an industrial site located on the outskirts of a large city near a commercial or industrial area, adjacent to highway and/or rail facilities, and a central city site (commercial property) for use with a fuel cell power system which would be directly connected to a commercial complex or group of users and might be factored into a total energy system to better utilize the low-temperature waste heat available. Higher land costs are more than offset by the benefit of being connected directly to the user.

The selection of a site has been made on the basis of judgement. Since transmission line costs are not included, specific cost comparisons have not been made. However, sites other than Middletown have not been selected unless they were clearly the better choice.

The following subaccounts are included in the Site Development Account.

Land Costs (Subaccount 1.1). Land costs include all costs associated with land acquisition for the plant except the access railroad and waste disposal area, including fees, condemnations, removals, and relocations, etc. Costs of \$24,7105/km^2 (\$1000/acre) were used for the

Middletown site, $2,471,050/km^2 ($100,000/acre) for an industrial site, and $24,710,538/km^2 ($100,000/acre) for a central city commercial site. The size of the site is a function of the plant layout (plot plan) for each base line case. These will be presented in the discussion of each conversion concept.

Clearing and Grading Costs (Subaccounts 1.2 and 1.3). For the Middletown site, contour intervals resulting in an average cut and fill of 0.762 m (2.5 ft) over the entire site were assumed. This resulted in general cut and fill (yard grading) of 379,851 m^3/km^2(2000 yd^3/acre). The flood plain is lightly forrested, resulting in light clearing over $\frac{1}{3}$ of the site. This results in costs for clearing of $148,263 x $\frac{1}{3}$ ($600/acre x $\frac{1}{3}$) = $49,421/km^2 ($200/acre). Grading costs were assumed to be $377,851/m^3/km^2 = $1.96/m^3 (2,000 yd^3/acre x 1.50 yd^3) = $741,316/km^2 ($3,000/acre).

The above figure is 100% installation cost, and although there would be smaller quantities per acre in the industrial and commercial sites, higher unit costs would offset any savings. Therefore, the same figure was used for all sites.

Access Railroad (Subaccount 1.4). Both the Middletown and industrial sites are assumed to require rail transport, both for the delivery of equipment and for fuel and other raw materials such as limestone. The Middletown site is assumed to require 8.046 km (5 mi) of access track from main line to the site. The industrial site is assumed to be located adjacent to an existing line and requires no access track. The access track cost of $71,458/km ($115,000/mi) for material includes $12,427/km ($20,000/mi) for a 45.72 m (150 ft) wide right-of-way, the cost of new and relayer rail (averages), concrete ties, and ballast. An installation charge of $68,351/km ($110,000/mi) includes $40,389/km ($65,000/mi) for grading [an average 3.05 m (10 ft) cut and fill is assumed] together with such final grading as may be required and the actual placement of roadbed material, ties, and track.

Loop Track (Subaccount 1.5). Unit trains delivering coal and other raw materials are assumed to require a loop track, including a

1,609 km (1 mi) long passing track parallel to the access track on the same right-of-way. The rail system layout is shown on each base plot plan. The minimum size of loop track is governed by the minimum radius for curves [168 m (550 ft)] and the needed straight sections [152 m (500 ft)] minimum at the unloading and thaw facilities. A typical loop track layout showing minimum track requirements is shown in Figure 2.24. This layout assumes 70-car unit trains [average car length, 16.7 m (55 ft)]; engine length, 30.5 m (100 ft); caboose length, 15.2 m (50 ft). All unit trains include provision for spare cars and engines.

In Figure 2.24, the loop length is 4X + 1.450 km (4X + 4756 ft). A 70-car train with a single engine would be easily accommodated by this loop. The number X is equal to the (total number of cars in the unit train - 70) times 4.27 m (14 ft) for unit trains longer than 70 cars.

The cost of all materials is assumed to be $74,665/km ($120,000/mi) on the same basis as the access track but recognizing the increased number of switches, etc. Installation cost is assumed to be $43,496/km ($70,000/mi) for the loop track.

Siding Track (Subaccount 1.6). Where plant size is such that fuel shipment might be smaller than that required for a loop system, or at industrial sites, siding or ladder track is used instead of loop track. A typical system might include two parallel sidings. Deliveries are assumed to be made in 20- to 40-car trains, and each spur is twice as long as required to handle half the cars in a train with an unloading station at the center of the siding. In this way the full cars can be pushed to one end of this siding (away from the main line) with a car at the unloading station. As the cars are emptied, they end up between the unloading station and the main line.

Materials costs of $77,671/km ($125,000/mi) and installation costs of $49,710/km ($80,000/mi) were assumed for these sidings.

Other Site Costs (Subaccount 1.7). Other site costs, including the access road (final paving of existing road), on-site roads, wells, parking, surfacing, water supply, storm sewers, holding ponds, sewage

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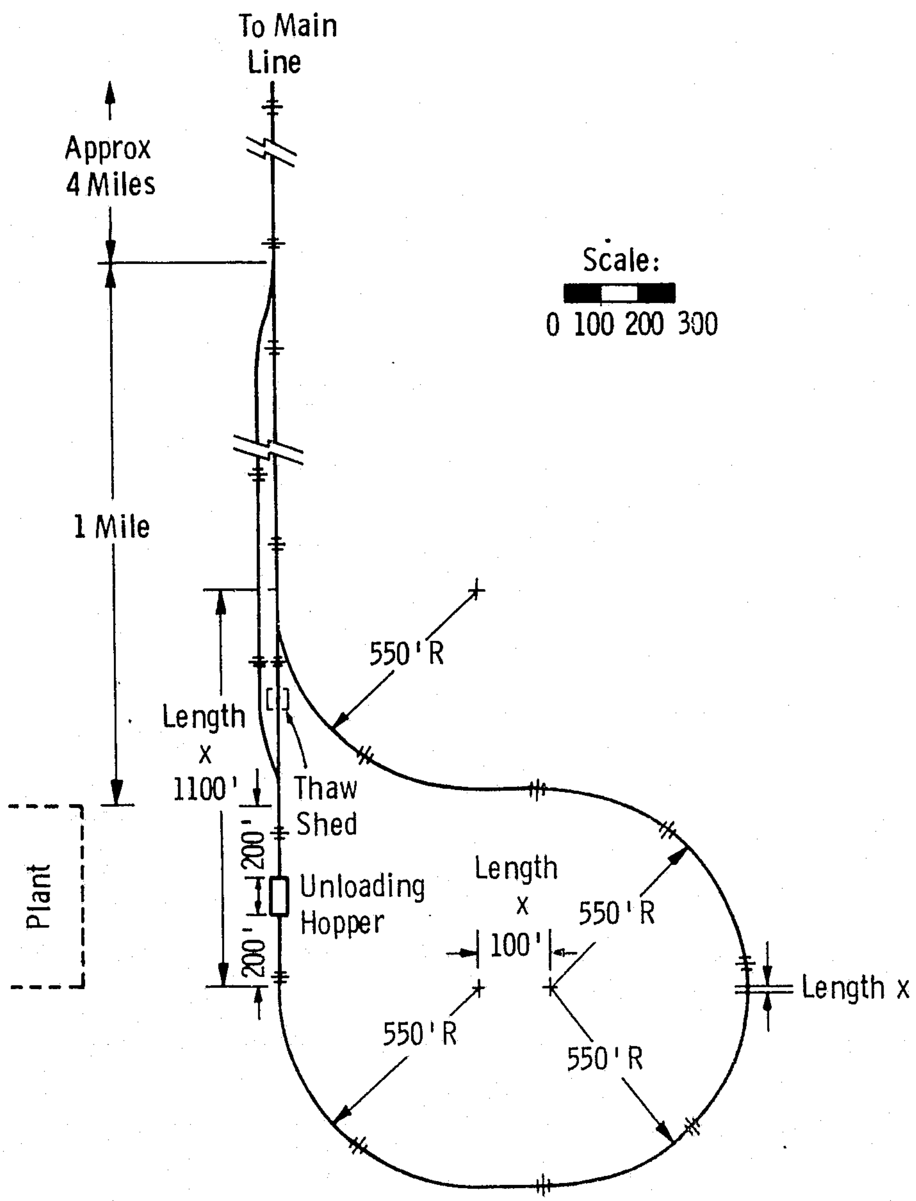

Fig. 2.24 —Minimum track requirements based on 70-car unit trains

treatment, yard fire protection, fencing, landscaping, etc. are carried as a single cost item. Figure 2.25 was developed by studying these costs (in mid-1974 \$) for several recent coal-fired power plants in sizes ranging from 400 to 1600 MWe. The assumptions are considered valid for industrial and commercial sites, again because higher unit costs for certain items are offset by lower quantities. The curve in Figure 2.25 has been approximated by a fifth-order polynomial in Equation 2.15:

$$\text{Other site costs, \$} = [4.9904722 - 0.57614414 \left(\frac{A}{100}\right) + 0.11711559 \left(\frac{A}{100}\right)^2 - 0.01420278 \left(\frac{A}{100}\right)^3 + 0.00084507 \left(\frac{A}{100}\right)^4 - 1.9139018 \times 10^{-5} \left(\frac{A}{100}\right)^5] (A) (10^3) \qquad (2.15)$$

where A = site total area in acres.

This cost is assumed to be 50% material and 50% installation.

2.6.1.2 Excavation and Piling (Account 2)

Common Excavation (Subaccount 2.1). Excavation for all plant island features is lumped together and carried as a single item. Except in certain special cases where deep foundations are required, excavation costs are not significant and therefore do not deserve any greater analysis. Excavation costs for items not included in the plant excavation volume are included with the item.

The excavation quantity is a function of foundation type and spacing, trenching for piping, and electrical ductwork and other considerations, which result in a range of 1.147 to 3.06 m^3 (1-1/2 to 4 yd^3) of excavation for each 0.765 m^3 (1 yd^3) of concrete. The smaller plants usually have an excavation range near the lower value, but higher unit costs (for the lower quantity) cancel out most of the difference. Accordingly, a quantity of 2.274 m^3 (3 yd^3) excavation for each 0.765 m^3 (1 yd^3) of concrete was used in all estimates, except in special cases which are carried as a separate item.

Curve 679884-A

Costs include roads, surfacing, landscaping, fencing, drainage including sedimintation and treatment, sewage treatment, yard fire protection.

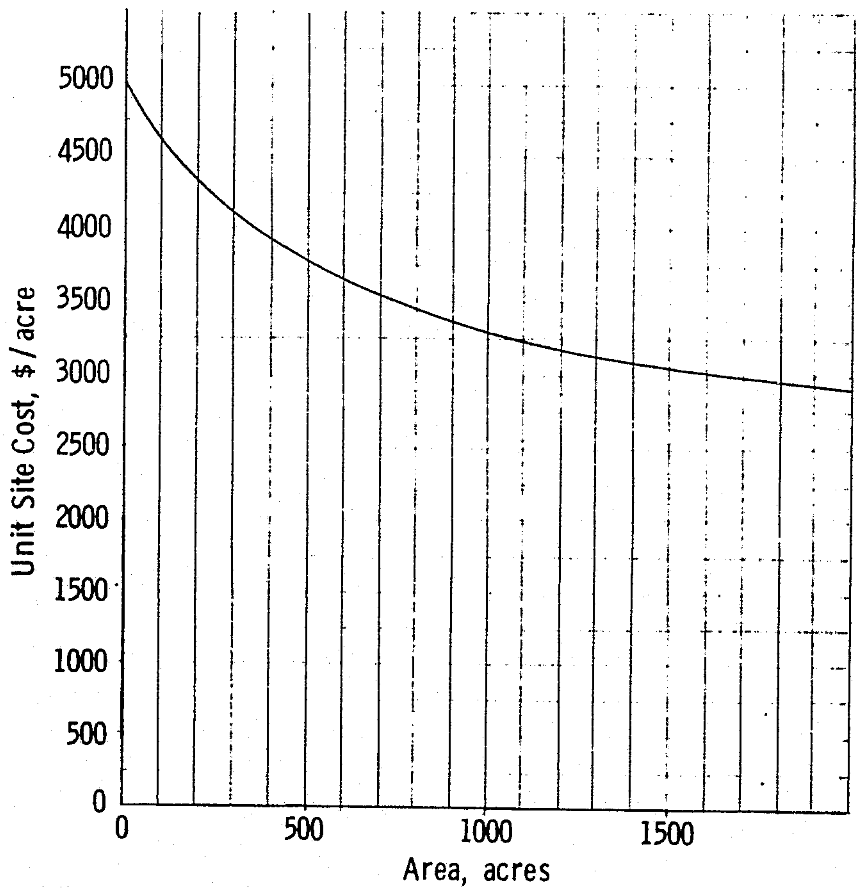

Fig. 2.25—Other site costs, 50% material 50% installation

General excavation costs include those of cut, stockpiling, backfill, spoil disposal, trench shoring, and casual dewatering. An installation cost of $3.92/m^3 ($3.00/yd^3) was used except for special cases or outlying structure excavations where costs for cofferdams, ground water control, and so on are considered.

Piling (Subaccount 2.2). The allowable soil-bearing pressure of 41.3685 MPa (6,000 lb/ft^2) assumed for the Middletown site is adequate for some smaller power systems (gas turbines and fuel cells) and for certain outlying structures such as coal hoppers and conveyor foundations, intake structures, and mechanical-draft cooling towers. These structures are constructed with spread footings and monolithic foundations. Major plant island foundations require piling. A concrete-filled shell pile was used for all pile-supported structures. The average length is approximately 13.716 m (45 ft).

Three recent pile-supported, coal-fired stations were studied to establish a relationship of from 2.13 to 2.74 m (7 to 9 ft) of piling per 0.765 m^3 (1 yd^3) of concrete when converted to 13.796 m (45 ft) lengths. Accordingly, a ratio of 3.189 km/m^3 (8 ft/yd^3) was used for all pile-supported structures. A unit cost of $21.33 m ($6.50 ft) material and $27.89 m ($8.50 ft) installation, based on these same studies, was used for all estimates.

2.6.1.3 Plant Island Concrete (Account 3)

All concrete in the plant island is included in this account. Concrete for outlying structures is included with the specific account (i.e., circulating water system, material handling, exit gas cleaning, etc.). Virtually all the concrete in power plants is used for foundations and floor slabs. In general, 20.685 MPa (3000 psi) reinforced concrete is specified. The design of foundations is a function of dead and live load, rotating mass, seismic and wind load, etc. Rules of thumb were used in conceptual estimates based on the total load and type of structure carried by the foundation.

The Task 1 estimates are based on the following dead load to foundation concrete rules of thumb, adjusted as required for overturning in high structures and similar considerations.

Table 2.41 - Dead Load to Concrete (Rules of Thumb)

Item	Yd^3 Foundation Concrete to Tons Dead Load Ratio
Power Plant Structure (including all supported equipment and component weights)	.50 yd^3/ton
Major Rotating Equipment (supported at or near ground level on foundation not a part of the power plant structure)	2.0 yd^3/ton
Major Rotating Equipment (supported on pedestal above ground level)	5.0 yd^3/ton

These relationships guided the determination of total plant island foundation concrete. Special foundations and shielding structures were studied on an individual basis. Ground floor paving was included, based on covering 50% of the plant island with 0.203 m (8 in) thick slabs.

For general foundations and paving (quantities determined on a yd^3/ton basis) an average unit price of \$196.19/m^3 (\$91.56/m^3 material, \$104.63/m^3 installation) [\$150.00/yd^3 (\$70.00/yd material, \$80.00/yd installation)] was used. For special foundations or shielding structures an appropriate unit price was used and noted. The minor amounts of concrete included in suspended floors and for architectural purposes are included in buildings costs.

2.6.1.4 Heat Rejection (Account 4)

Cooling Towers (Subaccount 4.1). Cooling towers may be either wet or dry, as specified. If wet, they will have been designed for the

ISO ambient of 288.16°K (59°F) db, 283.94°K (51.4°F) wb with an assigned range of 12.78°K (23°F). These towers, described in detail in Section 2.4. are modular concrete structures with plastic fill and an electric motor-driven fan rated at 149.2 kW (200 hp) shaft power. Although adjacent modules use a common wall, it is assumed that sufficient modules are present so that this end effect can be neglected. The cells are each assumed to have a material cost of $153,500 and an installation cost of $76,500.

If dry towers are used, a similar modular concrete tower construction is used with eight 3.048 m wide by 12.192 m high (10 ft by 40 ft) spirally wrapped fin-tube heat exchangers instead of the wet packing. These are described in Subsection 2.4.2. A range of 15.97°K (28.75°F) was used for the dry tower systems. They have 261.1 kW (350 hp) shaft power electric fan motor drives. Each module is assumed to cost $262,500 for materials and $87,500 for installation.

The cost of once-through cooling systems, where used, has been included in the next subsection—Circulating Water Systems.

Circulating Water Systems (Subaccount 4.2). The circulating water system associated with a wet tower system is described in Section 2.4.1. It assumes that the circulating water pump develops a head of 0.29889 MPa (100 ft) of water. The cost of materials and installation for this system are direct functions of the heat load and the range of the cooling water selected, as given in Equations 2.16 and 2.17 and explained in Section 2.4.

$$\text{Cost of Material, \$} = \left(\frac{\text{Heat Rejected}}{10^9}\right)\left(\frac{666167}{\sqrt{\text{RANGE}}} + \frac{4520898}{\text{RANGE}}\right) \qquad (2.16)$$

$$\text{Cost of Installation, \$} = \left(\frac{\text{Heat Rejected}}{10^9}\right)\left(\frac{1110279}{\sqrt{\text{RANGE}}} + \frac{5021098}{\text{RANGE}}\right) \qquad (2.17)$$

The heat rejected is given in Btu/hr and the range in °F in these equations.

The circulation system for the dry towers system is described in Subsection 2.4.2.2. It assumes a pump head of 0.22417 MPa (75 ft) Equations 2.18 and 2.19 using English units give the cost of material and installation for the dry tower circulating water system.

$$\text{Cost of Material, \$} = \left(\frac{\text{Heat Rejected}}{10^9}\right)\left(\frac{894887}{\sqrt{\text{RANGE}}} + \frac{3353158}{\text{RANGE}}\right) \qquad (2.18)$$

$$\text{Cost of Installation, \$} = \left(\frac{\text{Heat Rejected}}{10^9}\right)\left(\frac{1491478}{\sqrt{\text{RANGE}}} + \frac{3153078}{\text{RANGE}}\right) \qquad (2.19)$$

The once-through cooling system is described in Subsection 2.4.3 The circulation system assumes a 0.14945 MPa (50 ft) head,and,if a mixing canal is used,the canal pump assumes a 0.02988 MPa (10 ft) head. The total cost of the canal, inlet system,and circulating water system is given by Equations 2.20 and 2.21 if a mixing canal is not used; if the mixing canal is used, Equation 2.20 is to be modified by the addition of Equation 2.22 and Equation 2.21 is to be modified by the addition of Equation 2.23 .

$$\text{Cost of Material, \$} = A(1,2,4) = \left(\frac{\text{Heat Rejected}}{10^9}\right)\left(\frac{139624}{\sqrt{\text{RANGE}}} + \frac{6587178}{\text{RANGE}}\right) \qquad (2.20)$$

$$\text{Cost of Installation, \$} = A(2,2,4) = \left(\frac{\text{Heat Rejected}}{10^9}\right)\left(\frac{232708}{\sqrt{\text{RANGE}}} + \frac{8988138}{\text{RANGE}}\right) \qquad (2.21)$$

$$\text{Cost of Material (mixing canal adder), \$} = 320000 - \frac{1600000}{\text{RANGE}} \qquad (2.22)$$

$$\text{Cost of Installation (mixing canal adder, \$} = 360000 - \frac{1800000}{\text{RANGE}} \qquad (2.23)$$

Condenser (Surface) (Subaccount 4.3). The surface condenser, assuming standard materials with 25.4 mm (1 in) od, 22.94 mm (0.903 in) id admiralty tubes, a cleanliness factor 0.85, and tube water velocities of 2.134 m/s (7 ft/s), has been designed as described in Subsection 2.4.4 and the shell prices taken directly from Westinghouse list PL 1312 for surface condensers. The number of shells was taken as 1 for 1 or 2 low-pressure (LP) ends, 2 for 4 LP ends, and 3 for 6 LP ends. Admiralty tubes were assumed to cost \$2.336/m (\$0.712/ft). An installation charge for the tubes and the shell of \$7.53/m^2 of surface (\$0.70/ft^2) of surface) was also assumed.

2.6.1.5 Structural Features (Account 5)

Station Structural Steel (Subaccount 5.1). This account includes all of the structural and miscellaneous steel which is associated with a multipurpose station structure (station building) and which is usually designed and purchased by the A/E. It does not cover any single-purpose structural supports (boiler, precipitator, gasifier, coal conveyor, etc.). These are included as part of that equipment's cost. The ratio of weight of support steel to weight of supported materials and equipment is similar for several types of structures found in a major power plant. Typically, this ratio is 0.5:1, or for a deadweight of 1.0 kg, 0.5 kg of structural steel is required. This ratio (0.5) has been used as a guide when an existing design was not available. It has been modified as required by the particular structural shape, height, and complicity. Specific structural layouts were not made, but the general concept was considered. Such structural steel is assumed to cost \$716.50/Mg (\$650/ton), plus an installation cost of \$192.90/Mg (\$175/ton). These costs include miscellaneous steel, stairs, and walkways, as well as major support members.

Silos and Bunkers (Subaccount 5.2). The amount of coal storage or holding facilities when used have been assumed to be a direct function of the firing rate in Mg/s (tons/hr). This includes all concrete, silo support steel, the silo, tripper, and the tripper enclosure. These silo subsystems are assumed to cost \$0.55114 per Mg/s firing rate (\$1800 per

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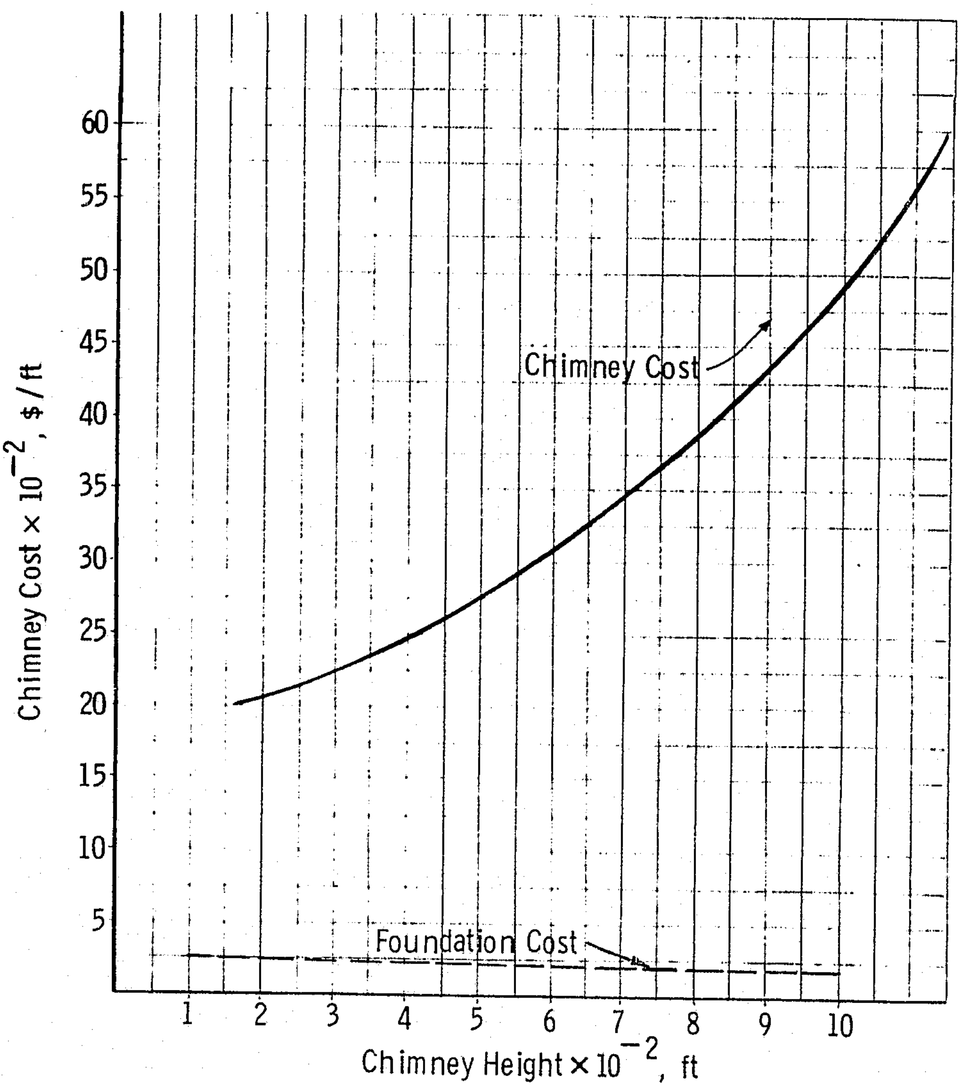

Fig. 2.26—Costs for concrete chimneys w/liners, 40% material 60% installation

ton/hr firing rate) with an installation cost of \$0.22964 per Mg/s firing rate (\$750 per ton/hr firing rate).

Coal silos for live storage were assumed to be between 7.315 m (24 ft) and 12.192 m (40 ft) in diameter with a maximum lower hooper angle of 55° and contain uncompacted coal with a density of 800.94 kg/m^3 (50 lb/ft^3). Sufficient silos are installed to provide coal for 28.8 ks (8 hr) firing at full capacity.

Chimney (Subaccount 5.3). Chimneys imply concrete structures of 121.9 m (400$^+$ ft) height or higher. Smaller stacks or lesser structures are included with the cost of the coupled equipment and not treated in this subaccount. Chimney costs as a function of height are given in Figure 2.26, and the total chimney cost is the sum of the costs of the foundation and the chimney proper. This cost is assumed to be 40% material and 60% installation. The total chimney cost has been represented by a polynomial in height in Equation 2.24, where the height, H, is in feet.

$$\text{Chimney Cost} = 100^* H\,[23.470705 - 0.00314217\,H + 3.1113099 \times 10^{-5}\,H^2] \qquad (2.24)$$

Special Structural Features (Subaccount 5.4). This subaccount includes such specialty items as permanent station cranes, elevators, and other similar equipment. They are assigned a lump sum price for each station as a whole.

2.5.1.6 Buildings (Account 6)

Station Buildings (Subaccount 6.1). All of the enclosure architectural features and building services included in the station building or buildings are included in this account. The structural steel and foundations accounts 03 and 05 for these multipurpose buildings are not included in this account since they also have the function of supporting power-generating equipment. Items such as siding, roofing, doors and windows, miscellaneous masonry construction, plumbing and drainage, heating, ventilating and air conditioning are included in this account. Base line cases include the definition and sizing of the station building. Unless otherwise noted, turbine rooms are assumed to be enclosed; boilers, combustors, gasifiers, exhaust gas treatment systems are not.

Once the enclosure was defined, its volume in cubic meters was calculated and costs based on studies of recent coal-fired power plants were assigned on a value per cubic meter. For large turbine room enclosures a total cost of $11.30/m^3 ($0.32/ft^3) was used. The cost was assumed to be equally divided between material and installation. The cost of enclosures requiring special heating and ventilating equipment, unusual environments, etc. was adjusted, using the $11.30/m^3 ($0.32/ft^3) value as a guide.

Administration Building (Subaccount 6.2). An administration section, either as a separate building or adjacent to the station building, consistent with the size and staff of each power plant, has been included as one of the features of the plant. This building was assumed to have a metal frame and an insulated metal enclosure with good quality interior finish, lighting, and services. These structures were assumed to cost $172.22/m^2 ($16/ft^2) for material and $150.70/m^2 ($14/ft^2) for installation. These costs include foundations, all structural and enclosure steel, building services, etc. required for a complete and functional building.

Warehouse and Shop (Subaccount 6.3). Warehouse, garage, and shop areas were also assumed to be metal frame buildings with little or no interior finish and with lighting and building services consistent with their intended use. The buildings are assumed to cost $129.17/m^2 ($12/ft^2) for material and $86.11/m^2 ($8/ft^2) for installation. As before, these costs include foundations, structural and enclosure steel, and the necessary services and facilities for its intended use.

2.6.1.7 Fuel Handling and Storage (Account 7)

Coal-Handling System (Subaccount 7.1). Due to the wide range of coal requirements for the various concepts, it was decided to divide the coal-handling and storage into two distinct designs, denoted as techniques A and B. The division was based on current industry practices reflecting both economics and design requirements of medium- and large-size central stations. Where a particular concept was unable to use either of the techniques as described, or required a modification to a technique, exceptions were noted in the particular concept description. Rotary car dumpers were included in technique B due to the large quantities o_ coal to be handled. Coal silos were sized to provide 28.8 ks (8 hr) live capacity. Costs of silos, when not included in fuel processing subsystems, were included in Subaccount 5.2.

Technique A was developed for smaller central stations on the Middletown site. The design incorporated a single coal pile containing both active and dead storage, a stockpile conveyor, telescopic chute to stack out the pile, and an underground reclaim hopper and collecting belt. This design has, in general, been applied to plants with firing rates of less than 0.1134 Mg/s (450 tons/hr). A rough layout showing the major components employed in technique A is shown in Figure 2.27.

This break point was suggested on the basis of the following assumptions:

- A single-unit coal train arriving daily
- Seventy loaded cars per unit train
- Car capacity 90.72 Mg (100 tons)
- Average train speed 0.011175 km/s (25 mph)
- Distance from origin of the unit train to the plant 1207 km (750 mi)
- Coal unloading time, 14.4 ks (4 hr)
- Coal loading time, 14.4 ks (4 hr)
- Down time, 14.4 ks (4 hr)

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Major Equipment

Coal Unloading:

1. Thaw Shed
2. Unloading Hopper
3. Tower, Stock Pile Conveyor, & Telescopic Chute

Coal Reclaim:

4. Reclaim Hopper & Collecting Belt
5. Conveyor & Portal Tunnel
6. Transfer Tower & Crusher to Additional Transfer Towers & Conveyors as Required

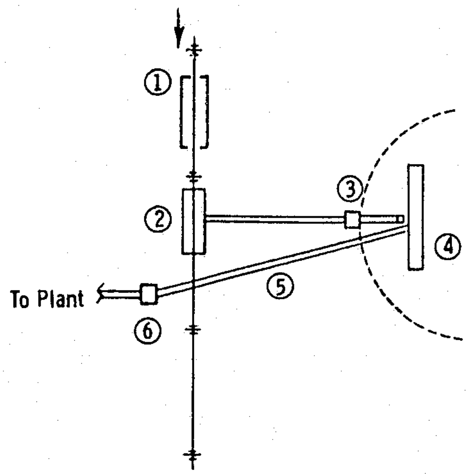

Fig.2.27—Coal-handling and storage technique A*
* Plants with an annual coal usage rate less than 450 tons of coal per hr

- 259.2 ks (72 hr) per round trip
- Plant capacity factor 0.65
- Maximum firing rate .11307 Mg/s (448.7 tons/hr).

Technique B was developed for large-size central stations on the Middletown site. The design incorporated separate active and dead storage piles. The active storage piles are stacked out and reclaimed with a traveling stacker/reclaimer. The dead storage can be reclaimed in emergency situations by underground reclaim hoppers and collecting belts. For rapid coal unloading a lowering well and reclaim hopper were included. A rough layout showing the major components employed in technique B is shown as Figure 2.28.

Both technique A and B require auxiliaries including bulldozers, instrumentation, controls, and electrical equipment for a fully automated system; fire protection equipment; dust suppression equipment; and dust removal equipment at all transfer points.

The coal pile description used in technique A is shown in Figure 2.29 and embodies the following assumptions:

- Sixty-day storage capacity
- Maximum angle of repose = 35°
- Maximum coal pile height = 12.19 m (40 ft); mean height = 9.144 m (30 ft)
- Unit weight of uncompacted coal = 720.8 to 881.0 kg/m^3 (45 to 55 lb/ft^3). Use 800.94 kg/m^3 (50 lb/ft^3)
- Compacted unit weight of coal = 1041.2 to 1153.3 kg/m^3 (65 to 72 lb/ft^3). Use 1121.3 kg/m^3 (70 lb/ft^3).

The volume of a 60-day coal storage, assuming half the storage pile is compacted [a mean density of 961.12 kg/m^3 (60 lb/ft^3)], is equal to (52,910)(firing rate, Mg/hr)[(48,000)(firing rate, tons/hr)]. The

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Major Equipment

1. Unloading Hopper & Rotary Car Dumper
2. Unloading & Reclaim Transfer Tower & Crusher
3. Conveyor (Unloading & Reclaim)
4. Rail for Stacker/Reclaimer
5. Stacker/Reclaimer
6. Dead Storage Reclaim Conveyor & Portal Tunnel
7. Dead Storage Reclaim Hopper (s)
8. Lowering Well & Reclaim Hopper for Rapid Unloading
9. Additional Conveyors & Transfer Towers to Plant as Required
10. Thaw Shed

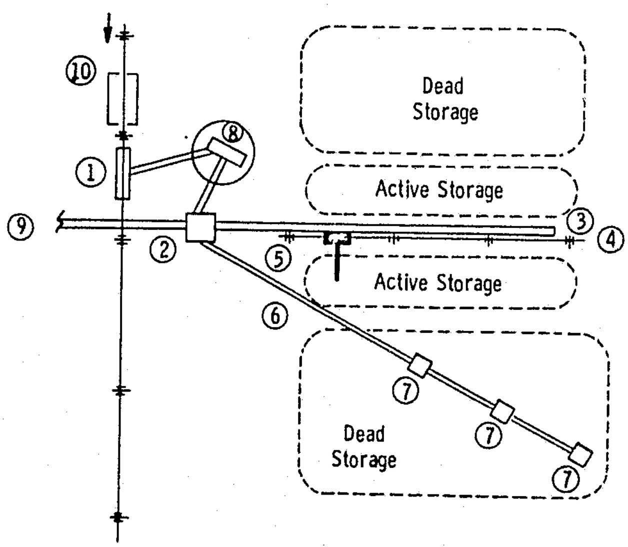

Fig.2.28—Coal-handling and storage technique B*
*Plants with an annual coal usage rate greater than 450 tons per hr

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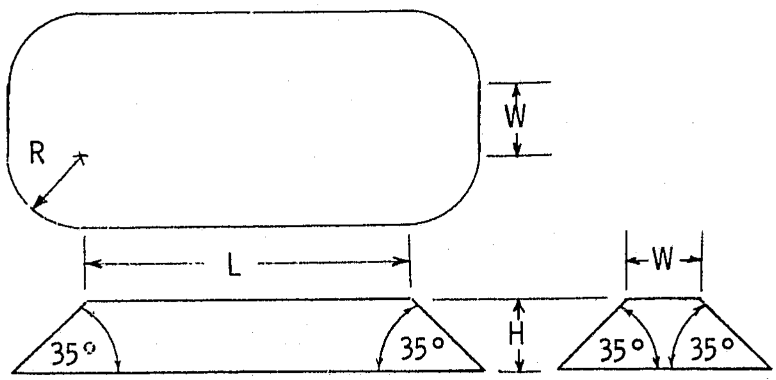

Fig.2.29—Optimum configuration

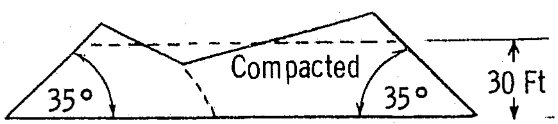

Fig2.30—Actual configuration

Notes:

H = 30 Ft

∡ = 35°

∴ R = 43 Ft

Total Length, L + 86 Ft

Total Width, W + 86 Ft

Coal-handling and storage technique A*

* Plants with an annual coal usage rate less than 450 tons of coal per hr

coal pile volume is calculated using Equation 2.25 and the variable defined in Figures 2.29 and 2.30 .

$$\text{Coal Pile Volume} = LWH + (W + L)\ R \cdot H + \frac{\pi}{3} R^2 H \qquad (2.25)$$

If it is further assumed that the depth is 9.144 m (30 ft) and that L = W, then Equation 2.25 can be rewritten as follows.

$$\text{Coal Pile Volume} = 30\ L^2 + 2570.7\ L + 57669 \qquad (2.26)$$

The rectangular area to be reserved for the coal pile would be (L + 2H/tan 35°) (W + 2H/tan 35°). Again assuming L = W and H = 9.144 m (30 ft), the area in acres required would be given by Equation 2.27.

$$\text{Coal Pile Area, acres} = \frac{(L + 85.688)^2}{43560} \qquad (2.27)$$

where L is in feet.

When coal-handling and storage technique B was used the dead storage was assumed to be that shown in Figure 2.29 with the following assumptions:

- 5.184 Ms (60-day) storage capacity
- Maximum angle of repose = 35°
- Maximum height of coal pile = 12.19 m (40 ft), mean height = 9.144 m (30 ft)
- Coal totally compacted with a density of 1121.3 kg/m^3 (70 lb/ft^3)
- L = 3W.

The coal pile volume and area are given by Equations 2.28 and 2.29, respectively, in English units of ft^3 and acres.

$$\text{Coal Pile Volume, ft}^3 = 90\ W^2 + 5141.33\ W + 57668.5 \qquad (2.28)$$

$$\text{Coal Pile Area, acres} = \frac{(W + 85.688)(3\ W + 85.688)}{43560} \qquad (2.29)$$

where W is in feet.

The relationship between dead storage volume and area is shown in Figure 2.31 which has an auxiliary scale showing coal-firing rate as an abscissa as well as coal pile volume.

For dead storage multiple piles may be used in some instances. For large systems [0.1134 Mg/s (450 ton/hr)] using a stacker-claimer two dead and two active storage piles are assumed.

Active storage coal piles are considered to be uncompacted and to have the configuration shown in Figure 2.32 . The total storage in the two piles is based on the following assumptions:

- Seven-day storage at 100% capacity factor
- Maximum angle of repose = 35°
- Maximum height 12.19 m (40 ft)
- Density of uncompacted coal 800.94 kg/m^3 (50 lb/ft^3).

For plants with a coal usage rate greater than 0.4596 Mg/s (1800 tons/hr) multiple stacker/reclaimers were used.

The active coal pile area is given by Equations 2.30 and 2.31 .

Curve 680013-B

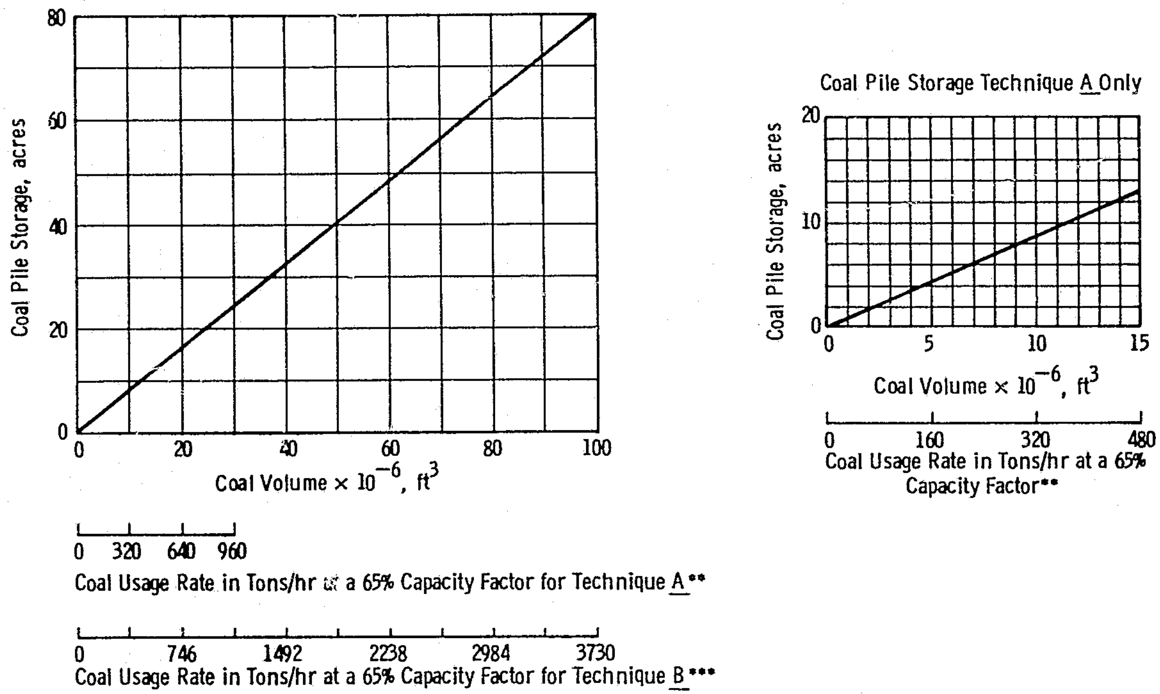

Fig.2.31— Coal pile storage

* This graph applicable for technique A storage and technique B dead coal storage only

** Rate = (30.9×10^{-6}) (ft^3)

*** Rate = (36.0×10^{-6}) (ft^3)

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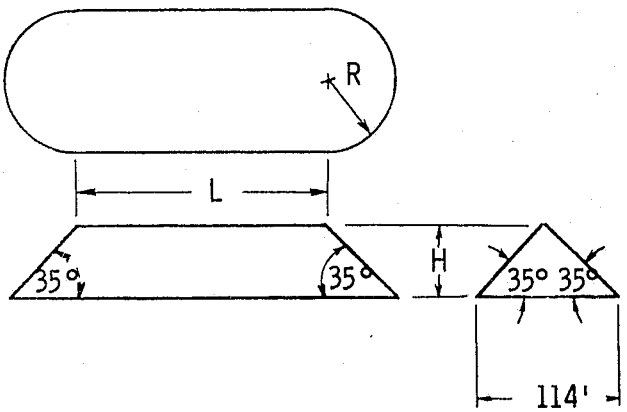
Active Storage (each pile)



Notes:

$H = 40$ Ft

$\measuredangle = 35°$

$\therefore$ $R = 57$ Ft

Total Length, L + 114 Ft

Total Width, W = 114 Ft

Fig.2.32—Coal-handling and storage technique B*

* Plants with annual coal usage rate greater than 450 tons of coal per hr

Curve 679886-A

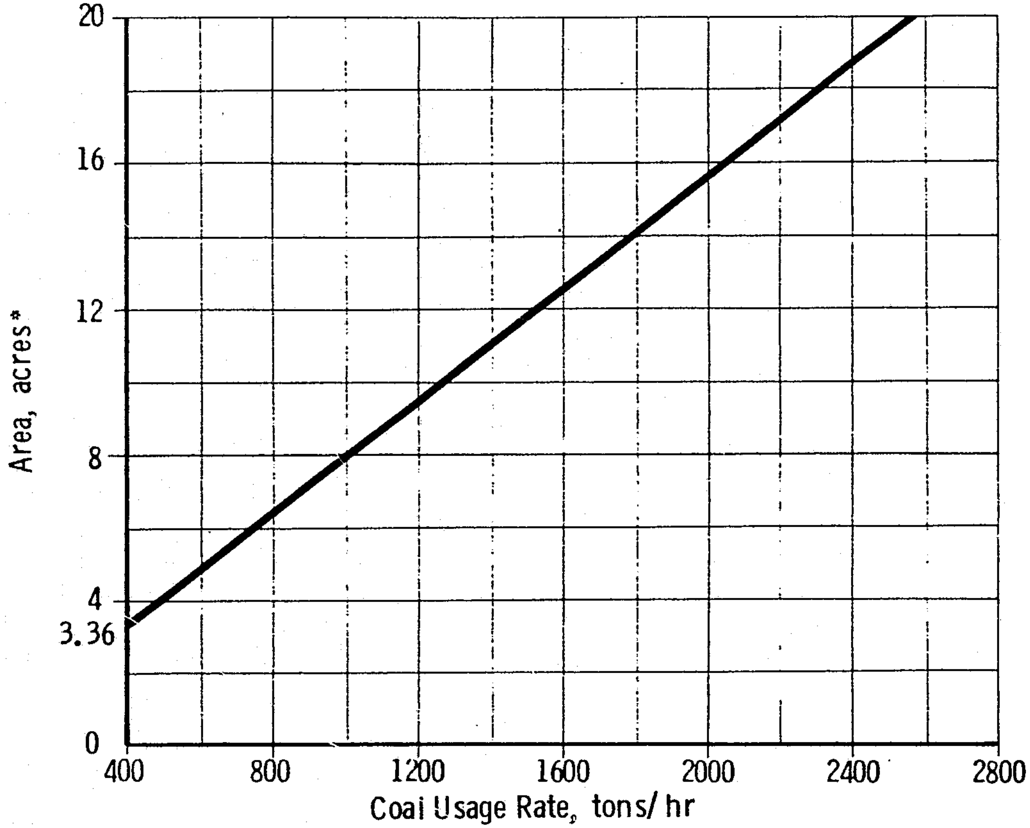


Fig.2.33— Active coal pile storage coal usage vs area*
* Area in acres = total area of two active coal piles

The volume (in ft^3) of each coal pile is 3360 times the firing rate (in tons/hr).

$$\text{Active Coal Storage Volume per pile, } ft^3 = (2285)\ L + 136696 \quad (2.30)$$

where L is in feet.

$$\text{Rectangular Area of each Active Coal Pile, acres} = \frac{114.25\ L + 13053.5}{43560} \quad (2.31)$$

where FR is the firing rate in tons/hr.

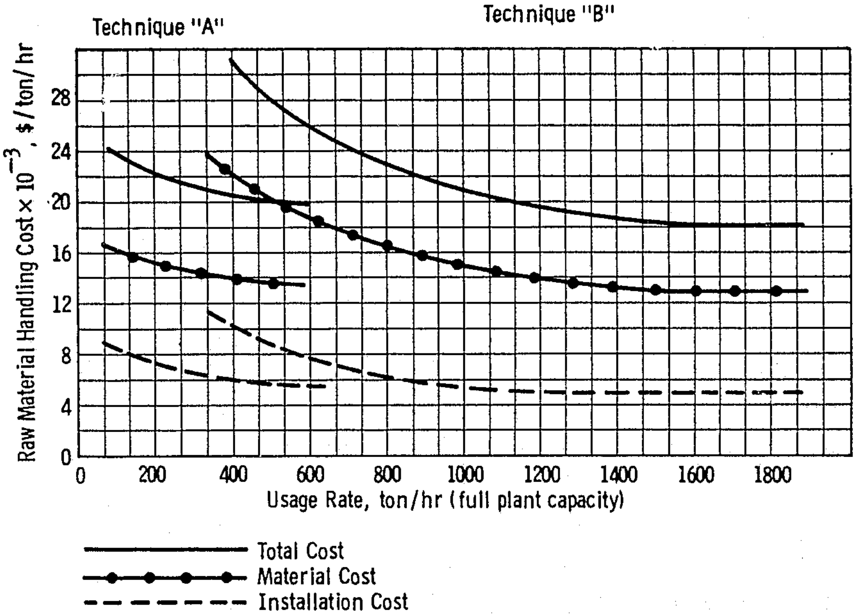
Figure 2.33 shows a curve of total required area as a function of firing rate.

The cost of the material and installation of a coal-handling system is shown in Figure 2.34 for both techniques A and B. These curves have been approximated by closed form relationship Equations 2.32 and 2.33 .

$$\begin{aligned} \text{Cost of Coal-Handling System Material} &= (12967)(FR) \text{ for } FR\ \ 1400 \\ &= \left[(22300)\left(\frac{400}{FR}\right)^{0.43276}\right](FR) \text{ for } 450 < FR < 1400 \\ &= \left[(16700)\left(\frac{67}{FR}\right)^{0.098698}\right](FR) \text{ for } FR < 450 \end{aligned} \quad (2.32)$$

$$\begin{aligned} \text{Cost of Coal-Handling System Installation} &= (4840)\ FR \text{ for } FR > 1200 \\ &= \left[(10000)\left(\frac{400}{FR}\right)^{0.66024}\right](FR) \text{ for } 450 \le FR \le 1200 \end{aligned} \quad (2.33)$$

Curve 679912-B



The above curves represent the average unit price of material handling systems handling coal, limestone and dolomite from point of delivery by rail to silo storage at or near combustor or gasifier.

Fig.2.34– Raw material handling system costs

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Major Equipment

Unloading:
1. Thaw Shed
2. Unloading Hopper
3. Transfer Tower & Conveyors
4. Telescopic Chute (Coal)
5. Telescopic Chute - (Limestone or Dolomite)

Reclaim:

6. Reclaim Hopper & Collecting Belt (Coal)
7. Reclaim Hopper & Collecting Belt (Limestone or Dolomite)
8. Conveyor & Portal Tunnel
9. Transfer Tower & Crusher (Coal)
10. Trans. Tower & Crusher (Dolomite or Limestone)
11. Additional Transfer Towers & Conveyors as Required

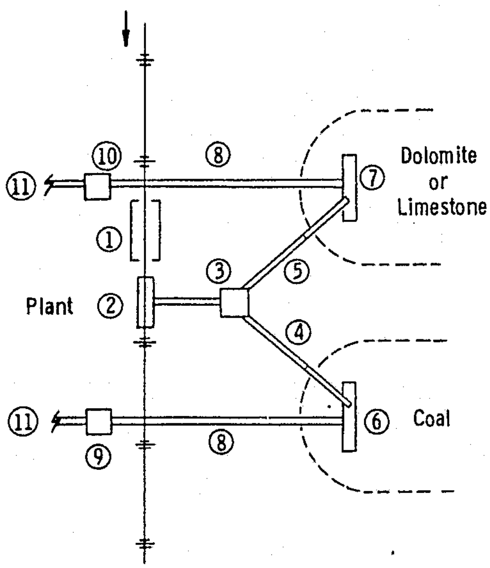

Fig.2.35—Solid material-handling and storage technique C*
* Plants with usage rate of coal less then 450 tons per hr

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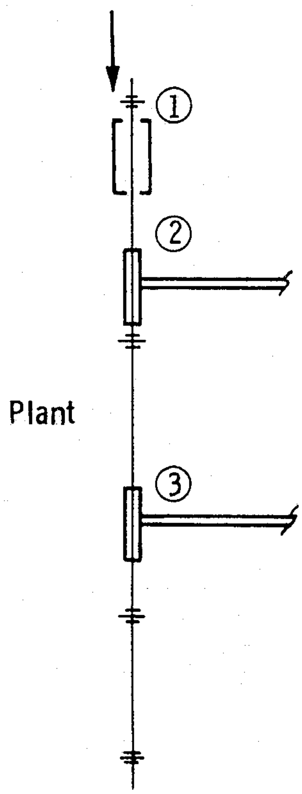
Major Equipment

Unloading:
1. Thaw Shed
2. Unloading Hopper or Rotary Car Dumper (Dolomite or Limestone)
3. Unloading Hopper or Rotary Car Dumper (Coal)



Limestone or Dolomite

Usage less than 450 tons per hr
use technique A

Coal

Usage greater than 450 tons per hr
use technique B

Fig.2.36– Solid material-handling and storage technique D*
* Plants with usage rate of coal greater then 450 tons per hr

Cost of Coal-Handling System Installation

$$= [(9000)(\frac{67}{FR})^{0.22693}] (FR) \text{ for FR} < 450$$

FR is the firing rate in tons/hr.

Costs include engineering, manufacturing, and erection of foundations, structures, all mechanical equipment, and the electrical controls, small motors, etc., normally furnished with manufactured material-handling systems.

Dolomite-Handling and Storage (Subaccount 7.2). The following techniques were developed for the handling and storage of the variety of solid materials, other than coal, required in the various concepts. For the purpose of developing assumptions and algorithms, only the major solid materials, limestone and dolomite, were considered. In concepts where solid materials other than limestone and dolomite are required the techniques developed in this section were utilized, with the major change being the material unit weight. Due to the similarity between handling and storing coal and limestone, or dolomite, the comments, assumptions, and algorithms for coal handling and storage are generally applicable to this section. Where assumptions and algorithms in this section are not completely developed, refer to the coal-handling and storage section for additional details.

Due to the wide range of coal and solid material requirements for the various concepts it was determined to divide the handling and storage into two distinct designs, denoted as techniques C and D. Technique C was for smaller central stations located at the Middletown site, requiring both coal and limestone or dolomite with a firing rate less than 0.11339 Mg/s (450 tons/hr) (see Figure 2.35). Technique D was for large-size central stations located at the Middletown site which had a firing rate greater than 0.11339 Mg/s (450 tons/hr) (see Figure 2.36). Both techniques C and D include a thaw shed for unloading materials arriving in a frozen state; instrumentation, controls and electrical equipment for a fully automated system; bulldozers; fire protection equipment; dust suppression equipment; and dust removal equipment.

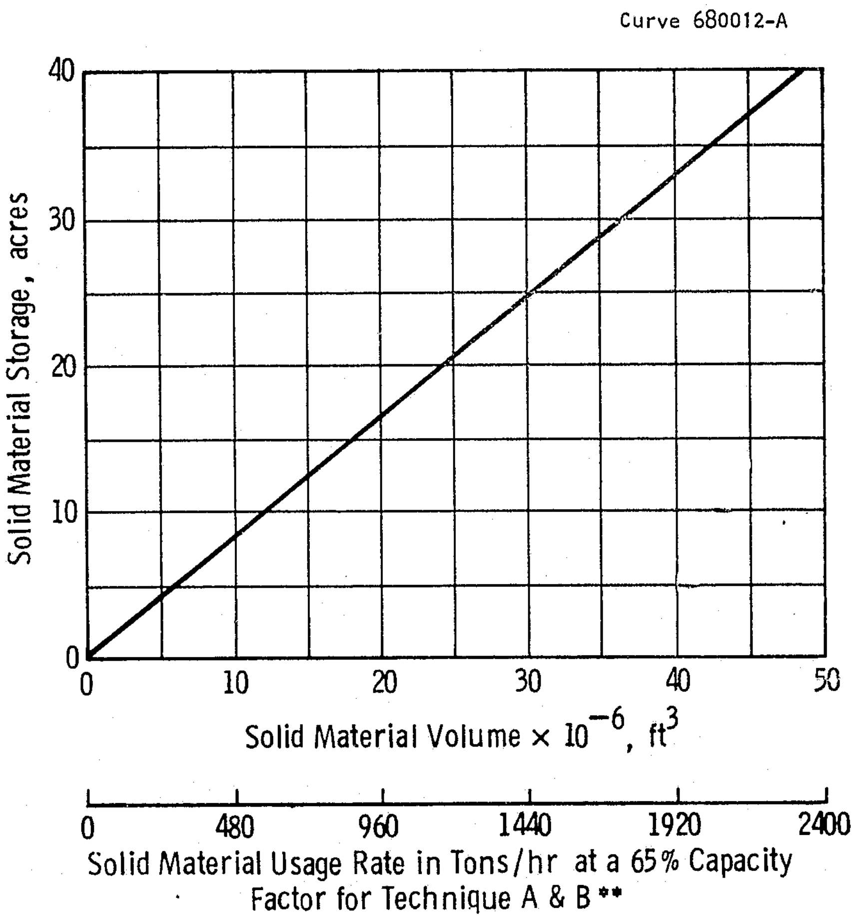


Fig. 2.37— Solid material storage*

*This graph applicable for technique C storage and technique D dead storage only

** Rate = (48.0×10^{-6}) (ft^3)

Because of their similarity to the coal-handling system, the cost of the dolomite- or limestone-handling systems are computed in the same manner. A factor (less than unity) is then applied to allow for duplication of components.

The area required for dolomite or limestone storage was computed in a similar manner with the following exceptions:

- Density of dolomite = 1441.7 to 1601.9 kg/m^3 (90 to 100 lb/ft^3). Use 1521.8 kg/m^3 (95 lb/ft^3), density of limestone = 1361.6 to 1441.7 kg/m^3 (85 to 90 lb/ft^3). Use 1361.6 kg/m^3 (85 lb/ft^3). Average for the two materials = 1441.7 kg/m^3 (90 lb/ft^3).
- 5.184 Ms (60 day) storage volume, ft^3 = (32,000) (capacity factor)(tons per hr used)
- Seven-day active storage volume, ft^3 = (3733.3) (tons per hr used) a capacity factor of 1 is assumed.

The length in feet of each active storage pile and total area in acres for active storage are given by Equations 2.34 and 2.35.

$$L = 0.8169 \text{ (tons per hr used)} - 59.82 \qquad (2.34)$$

$$\text{Total active storage, acres} = 2\left(\frac{114.25\ L + 130535}{43560}\right) \qquad (2.35)$$

Figure 2.37 shows the area required in acres for storage as a function of either volume of dolomite used/hr or use rate in tons/hr.

Fuel Oil Storage and Handling System (Subaccount 7.3). For the various concepts the coal distillate or fuel oil handling and storage requirements were divided into three distinct designs, denoted as ignition, stand-by and primary. Stand-by refers to a single tank, earthen retention dike, and the associated auxiliaries for unloading, transfer, fire protection, and drainage system oil collection equipment for use in concepts where fuel oil was used both for stand-by, start-up, auxiliary, or emergency fuel. Primary refers to a minimum of two or more tanks, earthen retention dikes, and auxiliaries as described for the stand-by system for use in concepts where fuel oil was used as the primary fuel. Ignition refers to a single tank, earthen dike, and auxiliaries as described for the stand-by system for use in concepts where fuel oil was used for ignition and warm-up only.

The following criteria were assumed for fuel oil storage and handling:

- 0.6048 Ms (7-day) storage for stand-by systems
- 2.59 Ms (30-day) storage for primary systems
- 20% of [0.432 Ms (5-day)] storage for ignition systems [capacity for approximately 0.864 Ms (5 complete start-ups)]
- An individual basin for each tank
- Earthen material dikes with a maximum height of 1.83 m (6 ft), a 0.61 m (2 ft) freeboard, and a minimum sideslope of 3 to 1
- API standard construction storage tanks on compacted sand foundations
- Dike walls of a minimum of 1/2-tank diameter from storage tanks
- Average railroad tank car capacity 75.7 m^3/car (20,000 gal/car)
- Average railroad tank car length 16.76 m/car (55 ft/car)
- Separate unloading facility on existing track for stand-by and ignition systems
- Separate unloading facilities and ladder tracks for primary systems.

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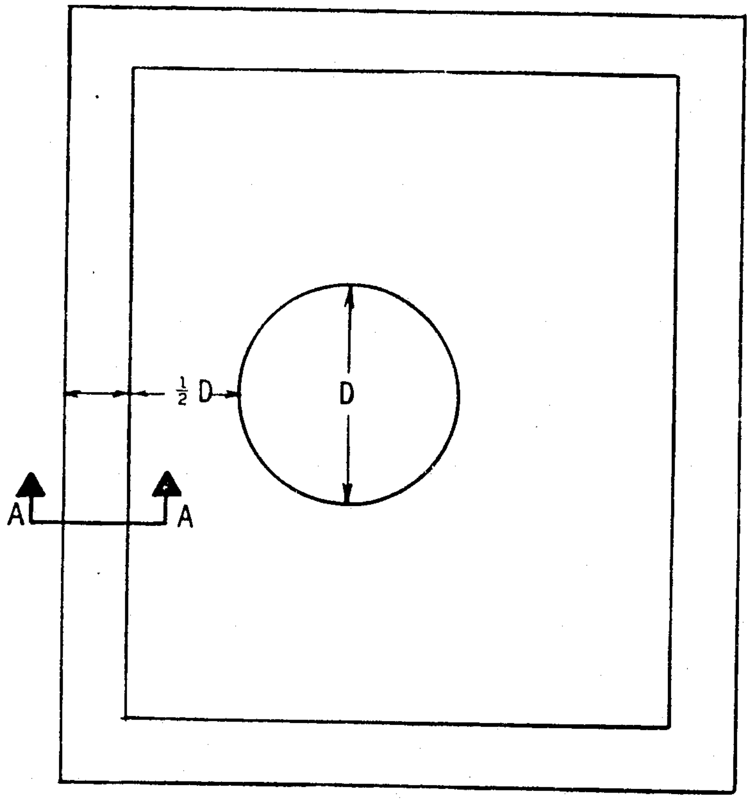
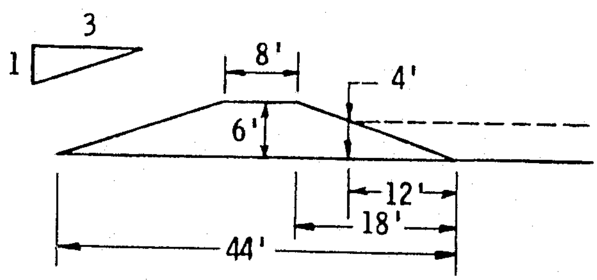
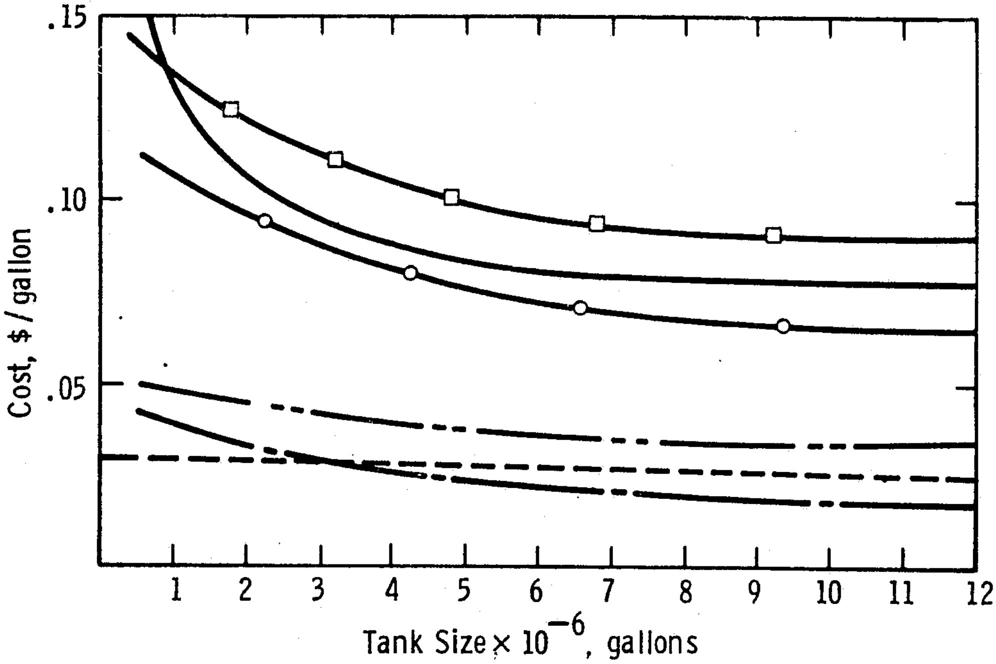




Section "A-A"

Fig. 2.38—Fuel oil storage

Curve 680011-A



Total Material Unit Cost —□—
Total Installation Unit Cost —○—

Basis:

1. Oil Tank Cost Curve ————
   a. Material Cost = .65
   b. Installation Cost = .35
2. Fire Protection System – – – – –
   a. Material Cost = .60
   b. Installation Cost = .40
3. Retention Dike, Misc Earthwork Fndn, Ect ——-——-——
   a. Material Cost = .10 Middletown; 35 Other Sites
   b. Install Cost = .90 Middletown; 65 Other Sites
4. Fuel Oil Unloading & Transfer System ——-·-——-·-——
   a. Material Cost = .60
   b. Install Cost = .40

Fig. 2.39– Distillate/fuel oil storage and handling cost

Figure 2.38 shows an example of a typical storage tank and dike arrangement.

The fuel oil storage requirement per day is given by Equation 2.36 where (sp gr) is the specific gravity of the oil and is assumed to be 0.825 for coal distillate and 0.865 for No. 2 fuel oil for this study.

$$\text{Storage, } \frac{\text{gal}}{\text{day}} = (\text{No. of Units})(\text{FR})\left(\frac{10350.7}{\text{sp gr}}\right) \qquad (2.36)$$

where FR is the firing rate in lb/s.

A square dike arrangement like that shown in Figure 2.34, with a 3.658 m (12 ft) road around it, would require an area given by the relation in Equation 2.37:

$$L = \frac{152 + \sqrt{152 - 4\left[76 - \left(\frac{\text{Tank Volume}}{4}\right)\right]}}{2} \qquad (2.37)$$

$$\text{Area, acres/tank} = \frac{(L + 24)(L + 12)}{43560}$$

The assumed cost of coal distillate fuel oil handling systems is shown in Figure 2.39. Costs are expressed in \$/gpm total storage for the complete system. Separate curves are shown for tank costs, fuel unloading and transfer equipment, fire protection system and retention dike, and miscellaneous civil structures; but the estimates carry only the total cost amount. The installation cost has been approximated by Equation 2.38 and the material cost by Equation 2.39:

$$\text{Installation Cost} = 0.106 \left[\frac{10^6}{\text{Vol}}\right]^{0.19923} (\text{Vol}) \qquad (2.38)$$

$$\text{Material Cost} = 0.134 \left[\frac{10^6}{\text{Vol}}\right]^{0.1890} (\text{Vol}) \qquad \text{Vol} < 8 \times 10^6 \text{ gal}$$

$$= (0.0904)(\text{Vol}) \qquad \text{Vol} > 8 \times 10^6 \text{ gal} \tag{2.39}$$

where Vol = tank volume in gallons.

2.6.1.8 Water Treatment Equipment (Account 14)

Demineralizers (Subaccounts 14.1). Water treatment was treated as a special case because of the wide variables in makeup water requirements for gasifiers, PFB boilers, carbonizers, etc. Raw water was assumed to come from wells. Pretreatment is assumed to include filtration through sand bed filters followed by postfiltration using activated carbon.

The makeup demineralizer system consists of dual or multiple trains of sufficient capacity to supply 100% makeup with one train out of service. Trains would consist of anion and cation exchangers and fixed bed polishing. The quantity of water to be demineralized was calculated using Equation 2.40 :

$$\text{Flow to demineralization, gpm} = \left[\left(\frac{8000 \text{ lb}}{\text{MW - hr}}\right)(\text{steam plant power, MW})(0.01) + (\text{FR})(\text{sc ratio})(200)\right] \frac{1}{500} \tag{2.40}$$

where FR is the gasifier firing rate, tons/hr

sc ratio is the steam-coal ratio required by the gasifier.

Pretreatment and demineralization costs are expressed in \$/gpm makeup requires. Cost of chemical storage, distribution, and water analysis equipment is included. The makeup quantity is based on 1% for steam generators plus other special requirements for advanced cycle components. The demineralizer is assumed to represent a cost of \$11,010 per m^3/s (\$2,500/gpm) for units with a capacity less than 12,620 m^3/s (2000/gpm) and \$8,810/per m^3/s (\$2,000/gpm) for larger plants. Installation costs are assumed to be \$3,080 per m^3/s (\$700/gpm) for the smaller plant and \$2,470 per m^3/s (\$560/gpm) for the larger ones.

Condensate Polishing (Subaccount 14.2). One hundred percent condensate polishing is provided for high-pressure [24.1316 MPa (3500 psi) or greater] once-through steam generators and any other special cycles which are noted. The material and installation charge for the polishing system were taken as $1.25/kWe and $0.30/kWe respectively.

2.6.1.9 Auxiliary Mechanical Equipment (Account 16)

The auxiliary mechanical equipment normally associated with a commercial steam, gas turbine, or combined-cycle plant is found in portions of the cycles studied, particularly in bottoming plants. In addition, similar equipment may be required for advanced cycle portions of various plants.

In all base line case studies auxiliary mechanical equipment requirements were evaluated to determine what equipment was included in the advanced cycle portion vs the equipment included in topping or bottoming plants which was similar to large coal-fired stations currently in operation.

Equipment that could be so identified as part of, or similar to a commercial steam or gas turbine cycle was priced on the basis of $/kW. An estimate of the relationship to equipment requirements and current costs for similar components in a 750 MW, 16.5517 MPa/810.94°K/810.94°K (2400 psig/1000°F/1000°F) coal-fired steam plant was made first. Prices determined on that basis were further adjusted for cycle or sub-cycle pressure and temperature, and then for size, on the basis of the curve shown in Figure 2.40, which compares the cost of components in a 750 MW plant with similar components in larger and smaller cycles. This curve has been approximated by the polynomial in Equation 2.41.

$$\text{Size Multiplier} = 1.468916 - 9.910901 \times 10^{-4}\ (\text{Power}) + 4.9203 \times 10^{-7}\ (\text{Power})^2 \tag{2.41}$$

where power is the station power in MW.

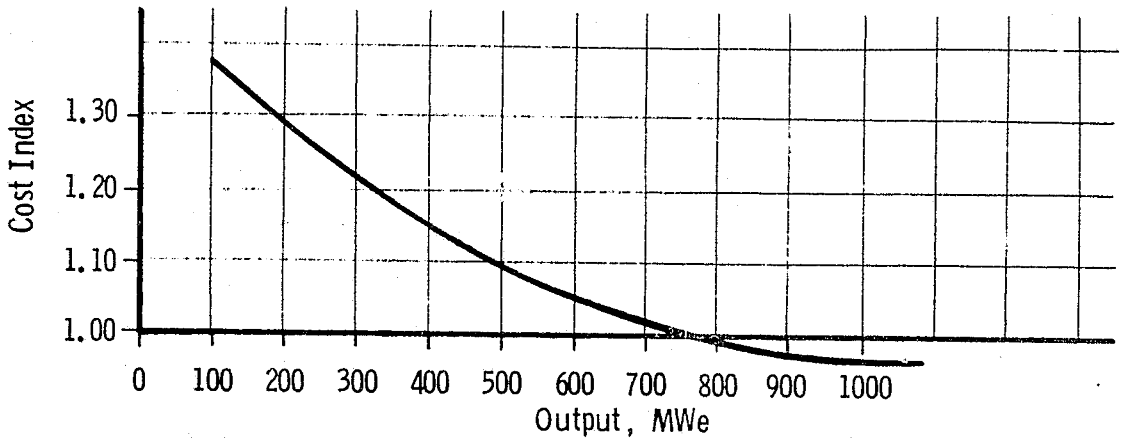


Fig. 2.40—Plant or component cycle, output - cost index

In this study we have modified the power generated by steam powered equipment by applying the size multiplier before calculating the auxiliary mechanical equipment costs.

The unit costs assumed for material and installation costs are given in Table 2.42.

Table 2.42 - Auxiliary Mechanical Equipment Costs

	Multiplier	Material	Installation
1. Boiler Feed Pump and Drive			
a. 2400 psia steam	(kWe)*	1.43	0.08
b. 3500 psia steam	(kWe)*	1.67	0.10
c. 5000 psia steam	(kWe)*	2.00	0.12
2. Other Pumps	(Factor)(kWe)*	0.88	0.12
3. Misc. Service Systems	Factor (kWe)*	1.17	0.73
4. Auxiliary Boiler	lb/hr	4.00	0.80

(kWe)* = (kWe)(size multiplier)

Costs for auxiliary mechanical equipment were determined on a kWe basis. In this study, in order to arrive at an equivalent kWe value for other pumps and miscellaneous equipment, the steam bottoming portion has been compared with a coal-fired station of like size. The other portions of the cycle are then compared to a coal-fired steam plant, and a simple calculation is then made to equate the weighted value to the total MWe for the plant.

For example, an unfired bottoming plant may be assumed to require only 80% of the auxiliaries required for a coal-fired station, the MHD portion may require only 40% and the gas turbine portion 40%. If a given plant has an MHD power output of 1000 MWe, a steam output of 600 MWe and gas turbine output of 400 MWe, the calculation is made as follows:

$$\text{Factor} = \frac{1000(.4) + 600(.8) + 400(.4)}{2000} = \frac{1040}{2000} = .52$$

Boiler Feed Pump and Drive (Subaccount 16.1). The boiler feed pump and drive costs have been assumed to be related to the steam cycle throttle pressure, and different costs have been suggested in Table 2.42 for throttle pressures of 16.5517, 24.1379, and 34.4827 MPa (2400, 3500, and 5000 psig).

Electrically driven feed pumps were assumed where the steam bottoming portion was 300 MWe or less ($.55/kW material cost, $.04/kW installation cost).

Other Pumps (Subaccount 16.2). This includes all pumps (condensate, heater drain, sump and service, etc.) exclusive of the boiler feed and circulating water pumps.

Estimated costs are a function of the kWe equivalent times the size multiplier.

Miscellaneous Service Systems (Subaccount 16.3). Miscellaneous service systems include cooling and water; compressed air; gas storage and distribution; condensate storage; and other miscellaneous service systems required for both the steam bottoming and the advanced cycle equipment.

Estimated costs are a function of the kWe equivalent times the size multiplier.

Auxiliary Boilers (Subaccount 16.4). Auxiliary boilers were included in some systems for start-up and auxiliary service. These boilers were priced on the basis of their output in lb/hr of steam. The size of the auxiliary boiler was calculated on the basis of the individual concept requirements of each base line case.

2.6.1.10 Piping Systems (Account 17)

Conventional Piping (Subaccount 17.1). The total weight of piping in the steam and water cycles relating to steam generators and the weight of other auxiliary piping was estimated on the basis of comparisons made with coal-fired stations in the 500-to-800 MWe range. The total cost of piping systems, including valves, trim piping and fittings, and insulation was studied to determine the average unit price of a system, based on the weight of fabricated piping.

The material cost of alloy piping (main steam, hot reheat, high-pressure extraction) operating at over 644.3°K (700°F) is greater than for other piping, but these systems have fewer valves, less trim, etc., and these factors result in an average unit price for all systems of $3308.90/Mg ($3,000/ton) material cost.

Installation cost varies more widely because of size, handling, and welding procedure considerations but for Task I it was considered valid to use $1984.13/Mg ($1,800/ton) for installation cost.

Other specialized piping may be included in additional subaccounts or as a subdivision of a major component cost in another account.

The quantity of piping in the steam bottoming plant plus conventional piping related to advanced cycle components is determined by using an equivalent value technique similar to that described in Subsection 2.6.1.9.

2.6.1.11 Auxiliary Electrical Equipment (Account 18)

Five subaccounts have been opened for major electrical equipment. These are:

- Motors, Miscellaneous Transformers, etc. (Subaccount 18.1)
- Switchgear and Motor Control Panels (Subaccount 18.2)
- Conduit Trays, Power and Control Cable (Subaccount 18.3)
- Isolated Phase Bus (Subaccount 18.4)
- Lighting and Communications (Subaccount 18.5)

Subaccounts 18.1, 18.2, and 18.5 are assumed to be proportional to the station power, MWe. Subaccounts 18.3 and 18.4 are a direct function of the number of feet estimated. Table 2.43 lists the cost assumptions.

Table 2.43 - Auxiliary Electric Equipment Material and Installation Cost

	Material	Installation
Motors, Miscellaneous Transformers, etc.	\$1.40/kWe	\$0.17/kWe
Switchgear and Motor Control Panels	\$1.95/kWe	\$0.45/kWe
Conduit, Trays, Power and Control Cable	\$1.32/ft	\$1.36/ft
Isolate Phase Bus	\$510.00/phase ft	\$450.00/phase ft
Lighting and Communication	\$0.35/kWe	\$0.43/kWe

The kWe equivalent or the quantity is determined by using the equivalent value techniques described in Subsection 2.6.1.9.

2.6.1.12 Controls and Metering (Account 19)

Computers (Subaccount 19.1). Computers for controlling and monitoring the power system are suggested as lump sum items based on the cost of computers used in coal-fired plants, plus the requirements for integrating the advanced cycle components subsystem computer segments with the overall plant control schemes.

Other Controls (Subaccount 19.2). This subaccount includes all other metering, monitoring, or supervisory equipment required and is also introduced as a lump sum item. A technique similar to the equivalent value method described in Subsection 2.6.1.9 was used to determine the lump sum value.

2.6.1.13 Process Waste Equipment (Account 20)

A wide variety of wastes from gasifiers, combustors, scrubbers, and other plant components had to be considered in Task I. Disposal of large quantities of waste is dependent on the type of material, land availability, state and local regulations, and utility preference. Waste disposal systems and cost were considered in two parts: removal of wastes from the point of origin, initial scrubbing or collection, and transport to storage silos, filters, or thickeners or decanting bins, and sluicing, conveying, trucking, or other secondary transport to on-site disposal areas. Return of wastes to mines, marketing of certain products such as fly ash and sulfur, and/or removal by disposal contractor was also considered but was not used as a basis for establishing cost.

Removal of molten ash (slag) from furnaces and boilers and molten ash and seed from MHD systems is accomplished by quenching, grinding where necessary, and transporting hydraulically to dewatering bins. The pulverizer reject and carry-over in the economizer hopper and other collections points are handled in a similar fashion. The resultant slurry is directed to a decant pond where the solids and water separate. The transport water is recycled.

Dry products such as fly ash, sulfur, limestone dust, and so on are transported by air to respective silos.

Spent slurry is hydraulically conveyed to a thickener and filter system which reduces the waste to 65% solids (slurry cake).

From these points the dewatered ash, dry ash, and/or slurry cake is transported to an on-site disposal area.

On-site storage was based on retention in a lined area surrounded by a dike. The area is determined by calculating the amount of waste [dry unit weight of 0.8009 Mg/m^3 (50 lb $/ft^3$) for coal by-products and 1.4417 Mg/m^3 (90 lb $/ft^3$) for limestone] converted to 65% solids. The dike is constructed initially to 60% of its ultimate planned height of 12.192 m (40 ft). During the first 473.04 Ms (15 years) material will be trucked or conveyed to the retention area and allowed to consolidate and partially dry. During the remaining 473.04 Ms (15 years) the height of the dike will be raised by using waste material placed earlier. Figure 2.41 shows the dike scheme.

The area required is calculated by using Equation 2.42:

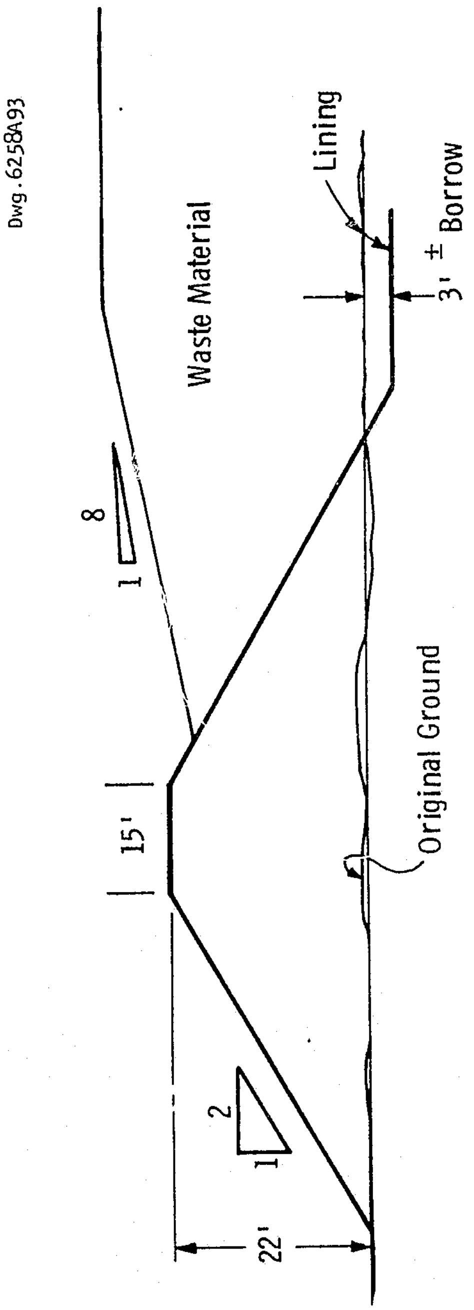


Fig. 2.41 — Solid waste disposal area

Curve 680010-A

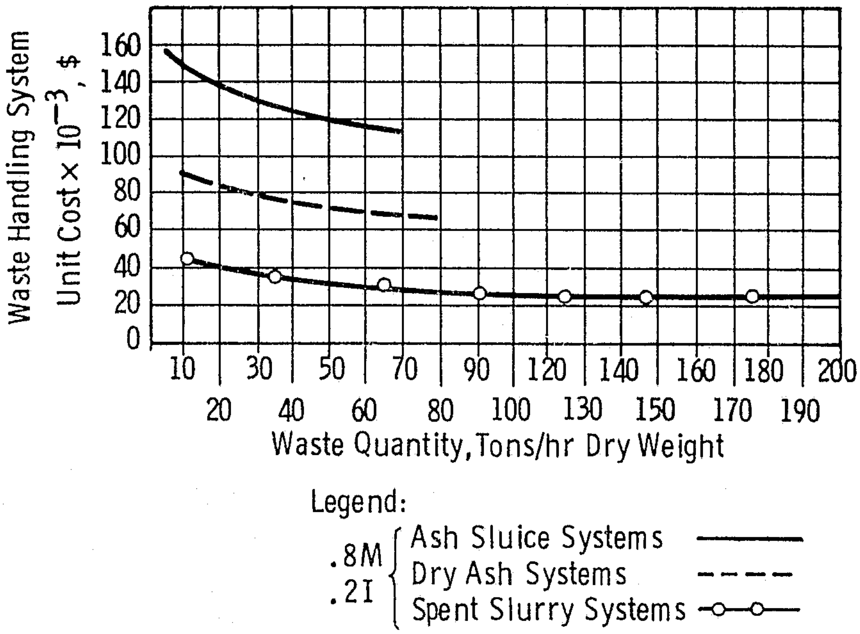


Fig. 2.42 — Waste-handling systems costs

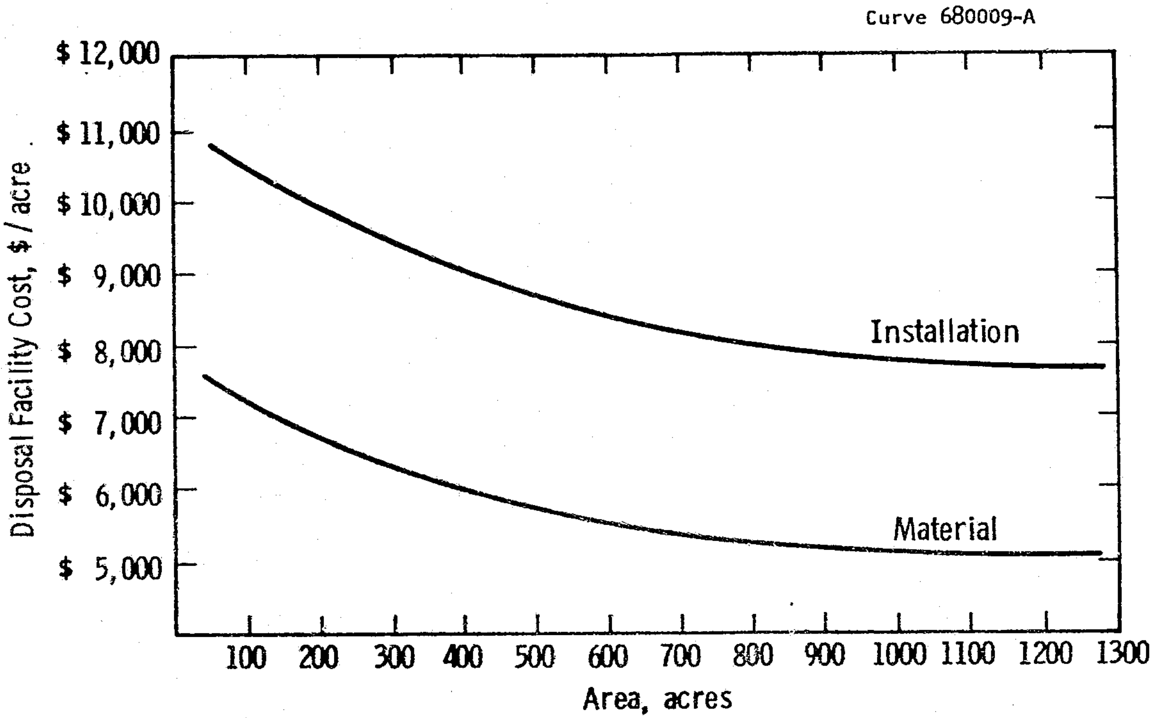


Fig. 2.43 —On-site disposal facility cost curve

$$\text{Area for Waste Disposal, Acres} = [(\text{sorb})(V_{sorb})$$

$$+ (\text{Ash})(V_{ash}) + (\text{Sul})(\text{Recovery Factor})\ V_{sulf}]\ 2000/[(40)(43560)] \qquad (2.42)$$

where Sorb, Ash, and Sul are the tons/hr of spent sorbent, ash and sulfur; and the specific gravities used were, V_{ash}=0.01861; V_{sorb}=0.01283; V_{sul}= 0.01129 and 83% of the sulfur is assumed to have been removed.

Bottom Ash (Subaccount 20.1). The bottom ash is assumed to be 20% of the total ash fired on the as received basis for a dry bottom pulverized coal-fired system, 80% for a single-stage cyclone, 90% for a two-stage cyclone, and 95% for a three-stage cyclone.

The cost of initial collection, transport, and storage is carried as a separate item and is based on Figure 2.42, showing unit costs for each type of system. In some cases costs are included with scrubbing systems and so noted. Costs are based on slag, ash, and sludge removal systems presently being installed at large coal-fired power plants.

Equation 2.43 is a fit of the curve in Figure 2.42 dealing with the cost of ash-handling equipment, showing the cost as a function of the tons of bottom ash handled (on a dry weight basis)

Installed Cost of Ash Sluice System =

$$[(161.15001 - 1.3391077(\text{Ash}) + 0.01008939(\text{Ash})]^2(1000)(\text{Ash}) \qquad (2.43)$$

Where Ash is the tons/hr of bottom ash.

This cost is assumed to be 80% material and 20% installation.

Dry Ash System (Subaccount 20.2). The dry ash-handling system installed cost given in Figure 2.43 is represented by Equation 2.44:

$$\text{Installed Cost of Dry Ash System} = [95.57162 - 0.72026769(D_{ash}) + 0.00437958(D_{ash})^2]\ (1000)(D_{ash}) \qquad (2.44)$$

where D_{ash} is the dry ash in tons/hr.

The installed cost is assumed to be 80% material cost.

Wet Slurry System (Subaccount 20.3). The cost of handling spent sorbent from a gasifier or scrubber is given by Equation 2.45

$$CSS = (1000)(Sorb)(43.600001 - 0.19535721(Sorb) + 0.00080357(Sorb)^2) \qquad (2.45)$$

CSS Installed Cost of the Spent Sorbent Handling System

Sorb = Sorbent use rate, tons/hr

The installed cost is assumed to be 80% material cost.

On-site Disposal (Subaccount 20.4). The cost of the on-site disposal facility is based on the construction of a dike using soil borrowed from inside the retention area and of sedimentation and treatment facility to control runoff. It includes the purchase of transportation equipment and the lining of the pond-dike bottom and sides at a cost of $\$1.615/m^2$ ($\$.15/ft^2$). Estimates were based on a cost of $\$1,371,435/km^2$ (\$5,550/acre) for material (including land cost) and $\$2,075,685/km^2$ (\$8,400/acre) installation for a $2.4281/km^2$ (600-acre) disposal facility, increasing to \$1,655,606 and $\$2,471,054/km^2$ (\$6,700 and \$10,000/acre) respectively for a 0.80938 km^2 (200-acre) disposal facility. This is shown in Figure 2.43. It should be noted that the $\$1.615/m^2$ ($\$0.15/ft^2$) used for the lining is optimistic and based on a technology not yet in existence but assumed to exist in the 1980s. Current cost would be approximately $\$2.69/m^2$ ($\$0.25/ft^2$).

The curves in Figure 2.43 are approximate polynomials in Equations 2.46 and 2.47:

$$DMC = (10^3)(A)(7.6764873 - 5.0107256 \times 10^{-3} A + 2.39707 \times 10^{-6} A^2) \quad (2.46)$$

$$DIC = 10^3(A)(11.070892 - 6.0003062 \times 10^{-3} A + 2.671501 \times 10^{-6} A^2) \quad (2.47)$$

where DMC = cost of on-site disposal facility material
DIC = cost of on-site disposal facility installation
A = site area, acres

2.6.14 Stack-Gas Cleaning (Account 21)

The items in this account may be misleading to the reader in that cleanup systems coupled with a process, for example a fluidized bed boiler or gasifier or the open-cycle MHD seed recovery system, may be lumped with the cost of that equipment and appear in another account. In the case of conventional boilers, however, precipitators and scrubbers were accounted for separately in Account 21. A detailed discussion of the stack cleanup equipment and cost is given in Section 4.

Electrostatic Precipitators (Subaccount 21.1). Electrostatic precipitators were assumed to have either a high or low efficiency, removing 99.5 and 90% of the incident particulate.

Precipitator equipment costs were assumed to be given by Equation 2.48:

$$\text{Precipitator equipment cost} = K\, W_c^{0.78} \quad (2.48)$$

where W_c is the coal-firing rate in tons/hr
K is cost factor (a function of coal type and excess air and given in Table 2.44 for excess air).

The cost of installation was assumed to be equal to 65% of the equipment cost. High-efficiency precipitators were used where none or only a part of the exhaust gas was scrubbed. Where all the exhaust gas was scrubbed, a low-efficiency precipitator was used.

Table 2.44 - Electrostatic Precipitator Costs

Coal Type	Moisture, %	Cost Factor K High Eff. (99.5%)	Cost Factor K Low Eff. (90%)
Illinois No. 6 Bituminous	13	6,930	3,510
Montana Subbituminous	24.3	7,020	3,550
North Dakota Lignite	36.7	5,750	2,910

Scrubber Costs (Subaccount 21.2). A scrubber system was assumed to cost \$27.70/kW for equipment and \$12.70/kW for installation. Based on data from Reference 2.5 (Figures 12 and 13) which treated a 500 MW plant burning a 3.5% sulfur coal with a HHV of 27.906 MJ/kg (12,000 Btu/lb) at a boiler efficiency of 89%, and a heat rate of cos τ, correction factors for plant size and coal type were generated as given in Equations 2.49, 2.50, and 2.51:

$$C = (0.89)(\text{Heat Rate})(\text{Station Power})/(9200\ \eta_B) \qquad (2.49)$$

where η_B is the boiler efficiency (see Section 4)

C is a station power normalize to coal firing rate.

$$CSIZE = (79.28824 - 0.077697 \cdot C + 3.5271656 \times 10^{-5}\ C^2)/48.4 \qquad (2.50)$$

where CSIZE is a limestone slurry scrubber cost multiplier based on normalized plant size variations.

$$CSULF = (38.165833 + 3.9832571 \cdot SUL - 0.15530297\ SUL^2)/51.339 \qquad (2.51)$$

where CSULF is a limestone slurry scrubber cost multiplier based on a normalized coal sulfur content

SUL is (% sulfur) (1200)/Coal heating value.

The resultant scrubber equipment unit cost was taken as the product of ($27.70/kW)(CSIZE)(CSULF).

2.6.2 Capital Cost

The direct capital cost, the sum of the material cost, and installation cost, was modified by the addition of the indirect cost to arrive at the total capital cost. These indirect costs include indirect construction costs, contingency, escalation, and interest during construction.

The direct capital cost estimates are based on the assumption that all concepts are proven technology and in commercial use. No research and development costs nor unusual engineering or construction risks have been included in the direct cost estimates.

2.6.2.1 Indirect Construction Costs

Indirect construction costs include such things as wage-related cost (overtime, etc.), payroll taxes, insurance, heavy construction equipment and small tools, construction facilities, expendable supplies, etc. Indirect construction costs were calculated by multiplying the direct installation cost estimated by 51%. The 51% number is based on the assumption given in Table 2.45.

2.6.2.2 Professional and Owners Costs

Professional costs refer to all costs of the engineer-constructor including his fee. This covers project management, engineering and design, start-up and testing, construction management, and supervision of construction. These costs typically run from 8-1/2% for a standard steam or combined cycle plant to as low as 5-1/2% for an advanced cycle system

Table 2.45 - Indirect Construction Costs

Indirect construction costs are calculated by multiplying the total direct installation cost by 0.51. The 51% multiplier is based on the following:

1.	Wage-related costs including foreman premiums, overtime and high pay premiums, show-up time, etc.	6%
2.	Payroll taxes and insurance.	16%
3.	Heavy construction equipment and small tools used by constructor	16%
4.	Construction buildings, facilities, guard service, and other service contracts.	6%
5.	Expendable supplies	3%
6.	Field hire nonmanual employees	4%
		51%

involving very high-cost components for high-temperature service. A mean of 7.5% was chosen for all concepts of the Task I study.

Owners costs include field operation costs, taxes during construction, capitalized start-up costs and insurance. It is recognized that a wide variance in utility practice exists in the treatment of these owner costs, and they may vary from as little as 0.5% to as much as 5%, depending on the amount of supervision during construction, start-up costs and degree of allocation of corporate expense. For Task I of this study a constant value of 0.5% was assumed, making a total professional and owner charge of 8% of the total direct capital cost.

2.6.2.3 Contingency

Contingency is an allowance made for additional costs likely to be encountered as a result of an incompletely specified design, estimating errors and omissions, unanticipated site conditions, design scope changes, inability to predict actual productivity and unforeseen construction problems. Forced station additions or modifications due to revised statatory requirements and unanticipated changes in escalation or interest during construction are not considered to be contingency costs. Recognizing that the complexity of a power plant is indirectly related to the time of construction, a contingency of 3% plus the time of construction in years was chosen as the base for this study. Fixed values of -5, 0, +5, and +20% were also used. The precentage contingency was multiplied by the total direct capital cost to calculate the contingency allowance.

2.6.2.4 Escalation

Escalation cost refers to the increase in capitalization due to increased costs of material and installation (the direct cost plus the indirect cost of construction; professional and owner costs; and contingency costs) because of inflationary pressures. The escalation cost is, therefore, not only a function of the escalation rate but also of the cash flow during the time of construction. The time of construction is assumed to include the engineering phase through the start-up phase to commercial operation. A typical cash flow curve is shown in Figure 2.44. This skewed "S" curve with a mean at about 63% was generated from the mean historical cost flows of several recent power plants. For purposes of calculation, this curve was divided into 20 equal time periods. The ordinates at the end of each period are given in Table 2.46.

From the given annual escalation rate, R_{esc}, a rate per period R_{esc}', was calculated which would give the annual rate when compounded over the number of periods per year from Equation 2.52:

$$R_{esc}' = (1 + R_{esc})^{D} - 1 \qquad (2.52)$$

Where D = time of construction/20 in years.

Curve 680372-B

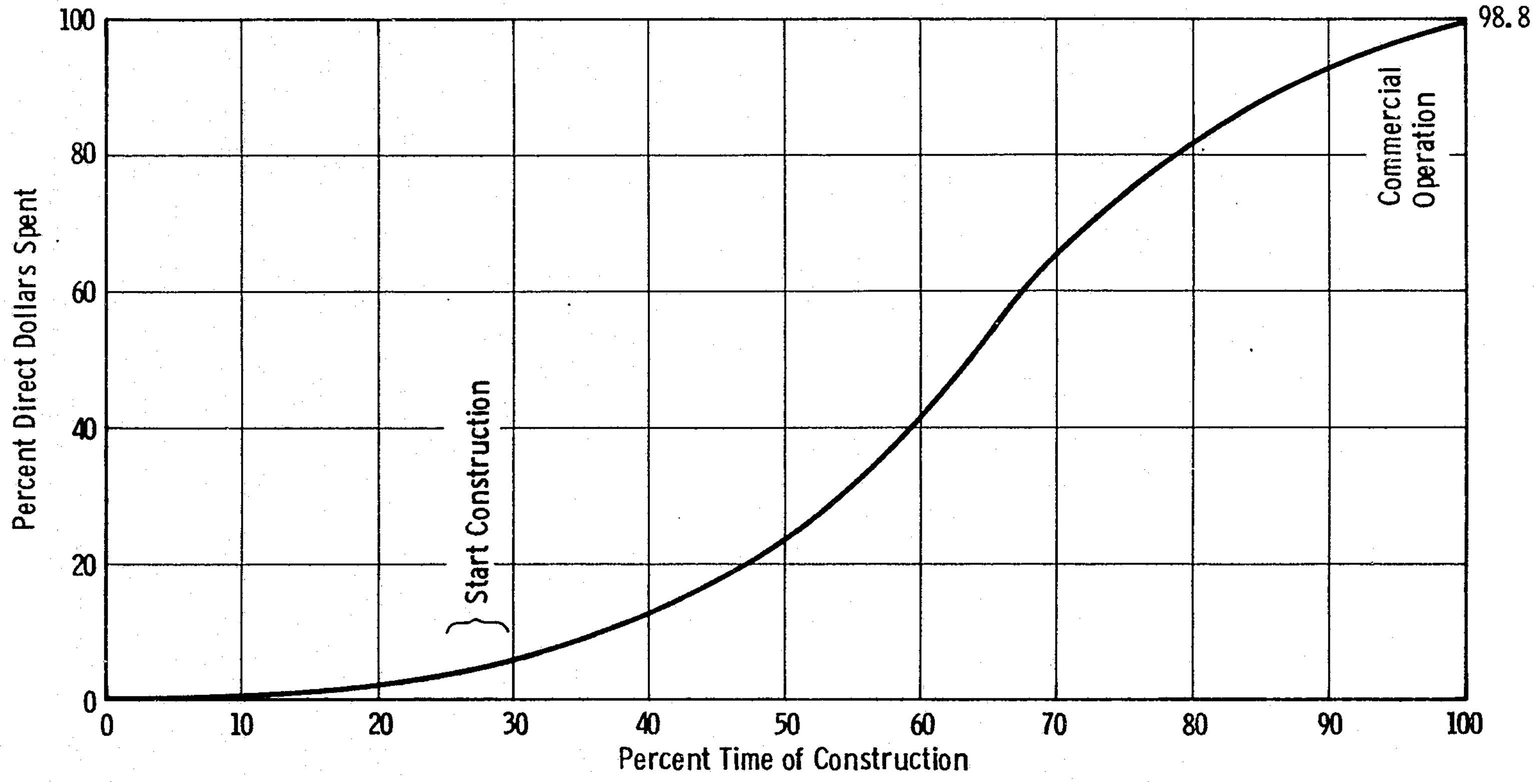


Fig. 2.44—Typical cash flow curve

Table 2.46 - Ordinate of the Assumed Cash Flow Curve

Index, i	Normalized Time of Construction	Normalized Total Cash Flow, C_i
1	0	0
2	0.05	0.002
3	0.10	0.008
4	0.15	0.015
5	0.20	0.020
6	0.25	0.033
7	0.30	0.060
8	0.35	0.092
9	0.40	0.130
10	0.45	0.180
11	0.50	0.240
12	0.55	0.322
13	0.60	0.425
14	0.65	0.540
15	0.70	0.650
16	0.75	0.743
17	0.80	0.820
18	0.85	0.880
19	0.90	0.928
20	0.95	0.960
21	1.00	0.988

In calculating the escalation cost, it was assumed that no escalation occurred until the end of the first period, i = 1, and then the ordinates of the remainder of the "S" curve, indices 2 to 21, were increased by multiplying by $(1 + R_{esc}')$. At the end of the second period, i = 2, all values of the modified "S" curve with indices greater than 3 were again modified. This was continued until the index was equal to 20 when C_{21} had been incremented 19 times.

In this way, an escalated "S" curve was generated whose normalized ordinates, C_i', are given by Equation 2.53:

$$C_{(i+1)}' = C_i' + (C_{(i+1)} - C_i)(1 + R_{esc}')^{(i-1)} \qquad (2.53)$$

where i goes from 1 to 20.

The escalation multiplier is given as the sum indicated in Equation 2.54 .

$$\text{Escalation multiplier} = \sum_{i=1}^{20} (C_{(i+1)} - C_i)\left[(1 + R_{esc}')^{(i-1)} - 1\right] \qquad (2.54)$$

The total increase in capitalization due to escalation is the product of the escalation multiplier and the subtotal of the direct capital costs plus the indirect cost of construction; professional and owner costs; and the contingency costs.

In addition to the NASA specified annual escalation rate of 6.5%, comparative calculations were also done for annual escalation rates of 0, 5, 8, and 10%.

2.6.2.5 Interest During Construction

The power plant construction cost includes the cost of the money which must be paid out during construction for goods, services and money. The cash flow curve, Figure 2.44, as modified by escalation as used to calculate the cost of interest during construction. Again, the time of construction was divided into 20 equal periods. From the given annual interest rate, R_{IDC}, a rate per period, R_{IDC}', was calculated which would give the annual rate compounded over the number of periods corresponding to one year from Equation 2.55:

$$R_{IDC}' = (1 + R_{IDC})^D - 1 \qquad (2.55)$$

where D is the time of construction/20 in years as before.

The escalation adjusted "S" curve with normalized ordinates C_i' can then be used to calculate the cost multiplier for interest during construction using Equation 2.56:

$$\text{Cost multiplier} = \sum_{i=1}^{20} \left(C_{(i+1)}' - C_i'\right)\left[\left(1 + R_{IDC}'\right)^{(20.5-i)} - 1\right] \tag{2.56}$$

The cost of interest during construction is then the product of the cost multiplier and the subtotal of the direct cost, plus the indirect construction costs; professional and owner costs; and contingency costs.

In addition to the NASA specified annual interest rate of 10%, some comparative calculations were made for rates of 6, 8, 12.5, and 15%.

The total capitalization of the plant is then the sum of the direct plant cost for material equipment and installation, the indirect construction cost; professional and owner costs; contingency costs; escalation costs; and the cost of interest during construction. The cost of escalation and interest during construction are sensitive to the assigned time of construction, which varied between 1.5 and 8 years for the concepts studied, being shortest for the fuel cell systems and longest for the MHD systems.

2.6.3 Cost of Electricity

The cost of electricity used in this study implies a cost at the power transformer high-voltage bushing and does not include any switchyard or distribution costs. For reasons of simplicity, these have been broken down into three components: fixed or capital costs, fuel costs, and operation and maintenance costs. The cost of electricity is very sensitive to the capacity factor [the ratio of the plant name plate rating (MW) times 8760 hr divided by the actual number of megawatt hours generated by the plant]. Plants with high capitalization and efficiency must run at high capacity factors to be economical.

2.6.3.1 Fixed Charges

Fixed charges include the cost of money; federal income tax; depreciation (30-year straight line); other taxes; insurance; and working capital. NASA specified a fixed charge rate of 18% of the total plant capital cost, a breakdown of which is shown in Table 2.47.

Table 2.47 - Fixed Charge Breakdown

Item	%
Cost of money	7.5
Federal income tax	4.1
Depreciation (30-year straight line)	3.3
Other tax	2.8
Insurance	0.1
Working capital	0.2
TOTAL	18.0

For comparative purposes, some calculations were also made for fixed charge rates of 10, 14.4, 21.6, and 25%.

2.6.3.2 Fuel Costs

Three delivered costs for each fuel in $\$/10^6$ Btu were specified by NASA. Westinghouse used these costs, plus a cost 20% higher than the suggested base rate and a higher cost beyond the three values suggested, to show the effect of fuel cost on the cost of electricity. These costs were detailed in Subsection 2.3.2.

2.6.3.3 Operation and Maintenance Costs

During Task I of this study, different approaches were taken in this area by some concept leaders. As a result, some differences may have resulted between concepts. Since the objective of Task I was an intra-concept comparison, and since the O&M charges were in general a small part of the cost of electricity (various fuel cell systems excepted), no attempt was made to rectify the situation. The general approach used to calculate auxiliary power requirements, operation costs, and maintenance costs is presented in subsequent subsections.

2.6.3.4 Operation Costs

Auxiliary Power Requirements. The auxiliaries of each concept were assumed to be either directly driven from the prime mover shaft, by a steam turbine, or by an electric motor. If a steam turbine drive was used the necessary steam was extracted from the main cycle (this is usual for the boiler feed pumps). If electric motor drives were used, the auxiliary electrical energy used was subtracted from the gross electrical output to provide a net plant electrical output on which the cost of electricity (mills/kWh) was based. Thus, no operation costs were generated to cover this auxiliary power. This avoided the question of the value of the electrical energy at the transformer high-voltage bushing. The auxiliary power requirements were limited to a few components and the values chosen based on a typical steam plant.

The wet cooling tower fan motor and circulating water pump power are treated in Subsection 2.4. The 149.2 shaft kW (200 hp) fan was assumed to be 92% efficient. The pump volume requirement at 0.29889 MPa (100 ft) head were a function of the heat rejection Q_c in Btu/hr and the circulating water range, R, in °F. A pump efficiency of 85% and a motor efficiency of 95% resulted in Equation 2.57 .

Wet cooling tower system auxiliary power requirements, MWe =

$$4.665854 \times 10^{-8} \; Q_c/R + 0.1621739 * N_w \qquad (2.57)$$

where N_w is the number of wet towers.

The dry cooling tower fan motor required 261.1 shaft kW (350 hp) and the circulating water pump developed a 0.22417 MPa (75 ft) head. The necessary dry cooling system auxiliary power is given by Equation 2.58 .

$$3.49939 \times 10^{-8}\ Q_c/R + 0.283804 * N_D \qquad (2.58)$$

where N_D is the number of dry towers required.

The once-through system utilizes circulation pump with a 0.14945 MPa (50 ft) head; and if a 2.78°K (5°F) mixing canal is used, a pump with a 0.02988 MPa (10 ft) head is also used. The auxiliary power required by the once-through system with and without the mixing canal is given by Equations 2.59 and 2.60 .

Once-through cooling system auxiliary power requirement, MWe =

$$2.33293 \times 10^{-8}\ Q_c/R \qquad (2.59)$$

Once-through cooling system with 5°F mixing canal auxiliary power

$$\text{Requirement, MWe} = 2.33293 \times 10^{-8}\ Q_c/R \left[\left(\frac{R}{5}\right) * 0.2 + 1\right] \qquad (2.60)$$

The auxiliary power required by the raw material handling system was assumed to be similar to a steam plant which required 0.009545 MWe per ton of raw material handled, which includes coal, dolomite, and limestone.

Where a gasifier is used, the auxiliary power requirements of the gasifier are given by Equation 2.61 .

$$\text{Gasifier auxiliary power, MWe} = (\text{Pow} + \text{Crush} + 2000 * \text{Yield}/3600)(W_c)/10^{-3} \qquad (2.61)$$

where W_c is the coal-firing rate in tons of as received coal/hr.
Yield is the pounds of product fuel gas per pound of coal

Pow is the booster compressor power required which is $\sim$ 33.5 kW/ton for the bituminous coal; 27.5 kW/ton for subbituminous coal, and 21.25 kW/ton of lignite coal assuming 150°F inlet air. Different values were used for different pressure ratio systems.

Crush is 1.3 kW/ton of coal.

For the Task I study, the combined-cycle gas turbine and those concepts using pressurized furnaces calculated their own booster compressor power, so Pow and Crush were omitted from the equation used for those calculations.

The boiler, when used, was assumed to have an auxiliary power amounting to 1.7% of the gross steam plant power, with an additional 1.7% added when a wet scrubber was present, to cover increased draft requirements and the scrubber circulating water pump. Atmospheric-pressure fluidized bed boilers were assumed to have a total auxiliary power requirement of 3.4% of the gross steam plant power.

The station auxiliary power requirements were assumed to be 0.5% of the gross station power; and for steam plants, or a plant with steam bottoming, an additional 0.6% of the steam plant gross output was assumed to be required for auxiliary power. The gas turbine systems made separate allowance for their auxiliary power. The numbers in this paragraph were not used for the open- and combined-cycle systems.

The auxiliary power requirement for the waste material handling system was assumed to be 274.06 MJ/Mg (0.0839 MW hr/ton) of waste material.

The required precipitator power was assumed to be 9.536, 7.936, and 6.238 MJ/Mg (2.92, 2.43 and 1.91 kW hr/ton) for as-fired bituminous, subbituminous, and lignite coals respectively.

The cost of makeup water for the wet cooling towers and the demineralizer and the polishing of the condensate are covered by Equations 2.62 , 2.63 , and 2.64 respectively.

$$\text{Cost of tower makeup water, \$/hr} = 1.6326 \times 10^{-8}\ Q_c \qquad (2.62)$$

where Q_c is the cooling tower heat lead in Btu/hr.

$$\text{Cost of demineralizing water, \$/hr} = 660\ \text{Gal}/8760 \qquad (2.63)$$

where Gal is the demineralizer load in gpm.

$$\text{Cost of polishing (where used), \$/hr} = 46 * \text{Powstp}/8760 \qquad (2.64)$$

where Powstp is the total steam plant power in MWe.

Equation (2.62) assumes a cost of \$0.0211/m^3 (\$0.08/1000 gal) of water for tower makeup. The demineralizing and polishing charges involve bed replacement and are a function of water throughput.

Another operating expense was \$4.85/Mg (\$5.35/ton) for sorbent. A charge for plant manning given by Equation 2.65 was used for all but the gas turbine plants.

$$\text{Maintenance cost, \$/hr} = \frac{X_x\ (\text{Power})}{[(0.004)\ (\text{Power}) + 0.6]} \left(\frac{15000}{8760}\right) \qquad (2.65)$$

where Power is the nominal plant power in MWe

X_x is a multiplier taken as 1 for conventional plants; greater than one for more complicated plants (MHD, etc.).

For the open recuperated gas turbine cycle which assumed interrupted duty, Equation 2.66 was used for maintenance costs.

$$\text{Maintenance cost, mills/kWh} \approx 3.6 \left(\frac{.12}{\text{cap}}\right)^{0.5823} \qquad \text{cap} < 0.5$$

$$= 0.84 \left(\frac{0.5}{\text{cap}}\right)^{0.61208} \qquad \text{cap} < 0.5 \quad (2.66)$$

where cap is the capacity factor.

For combined gas turbine plants, both open and closed as well as the closed recuperated plant Equation 2.67 was used.

$$\text{Maintenance cost, mills/kWh} = 0.945 \left(\frac{0.3}{\text{cap}}\right)^{0.70501} \qquad (2.67)$$

2.6.3.5 Unit Cost of Electricity

The unit cost of electricity per net kilowatt supplied to the switchyard was then calculated, not only for the NASA-specified capacity factor of 0.65 but also for capacity factors of 0.12, 0.45, 0.50, and 0.80.

2.7 Computer Output

Because of the large volume of data associated with the more than 630 parametric points studied in Task I and the difficulty in reporting these data by normal means in the time available, it was decided to print out some part of these data in the direct cost accounts for each parametric point so that they could be transmitted to NASA. A single copy of this detailed printout weighed more than 18.144 kg (40 lb) and was over 0.4064 m (16 in) thick. It was thus not practical to include the total printout as part of this report. The detailed printouts for the base cases and for the recommended points for the Task II study were included, however, as well as the summary sheets for each concept.

2.7.1 Detailed Direct Cost Accounting

The 21 accounts listed in Table 2.40 with appropriate subaccounts were reported for each parametric point. An example is given in Table 2.48. Five columns of numbers follow each subaccount name. The first is titled "amount" and must be interpreted with care. The amount is associated with the unit immediately to its left. In the case of Subaccount 1.1 of the example cited, the plant is assumed to be situated on a 134-acre site exclusive of access rail, right-of-way, and waste disposal areas. The unit cost of the land is given as a material cost, in this case, $1,000/acre, and there is no unit cost of installation. The direct material cost is the product of the amount and the material unit cost, or $134,000, and the installation cost is the product of the installation unit cost and the amount, in this case, $0. In some cases, where curves were used, unit costs have not been generated (Subaccount 1.7), and only material and installation cost totals were printed out. The zero in the amount column is, therefore, misleading and should be ignored. The direct cost of material and installation are totaled for each account and the percent of the total direct cost associated with that account printed out. Where "ea" or "each" appears in the unit column, the account lists only the cost of the equipment as a whole, and the amount is an integer value. Two exceptions were made in the gas turbine system. These were to add to the product of the unit cost and the amount a fixed number in Subaccounts 18.2 and 19.1, the fixed adder being supplied as part of the major component package and the product term being supplied as part of the balance of plant cost by the A/E.

For all concepts, the amount shown adjacent to a unit of kWe in Subaccounts 16.1, 16.2, 16.3, 18.2, and 18.4 are the product of the actual plant output, a size multiplier, and/or a factor to relate that concept to the reference steam plant used as a base.

The total direct material and installation cost is given immediately below Account 21.

Table 2.48

ADVANCED STEAM CYCLE WITH ATM BOILER ACCOUNT LISTING
PARAMETRIC POINT NO.21

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
SITE DEVELOPMENT						
1. 1 LAND COST	ACRE	134.0	1000.00	.00	134000.00	.00
1. 2 CLEARING LAND	ACRE	44.7	.00	600.00	.00	26797.32
1. 3 GRADING LAND	ACRE	134.0	.00	3000.00	.00	402000.00
1. 4 ACCESS RAILROAD	MILE	5.0	115000.00	110000.00	575000.00	550000.00
1. 5 LOOP RAILROAD TRACK	MILE	2.5	120000.00	70000.00	300000.00	175000.00
1. 6 SIDING R R TRACK	MILE	.0	125000.00	80000.00	.00	.00
1. 7 OTHER SITE COSTS	ACRE	.0	.00	.00	294612.43	294612.43
PERCENT TOTAL DIRECT COST IN ACCOUNT 1 = 2.051 ACCOUNT TOTAL,$					1303612.42	1448409.73
EXCAVATION & PILING						
2. 1 COMMON EXCAVATION	YD3	47400.0	.00	3.00	.00	142200.00
2. 2 PILING	FT	126400.0	6.50	8.50	821600.00	1074400.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 2 = 1.519 ACCOUNT TOTAL,$					821600.00	1216600.00
PLANT ISLAND CONCRETE						
3. 1 PLANT IS. CONCRETE	YD3	15800.0	70.00	80.00	1106000.00	1264000.00
3. 2 SPECIAL STRUCTURES	YD3	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 3 = 1.766 ACCOUNT TOTAL,$					1106000.00	1264000.00
HEAT REJECTION SYSTEM						
4. 1 COOLING TOWERS	EACH	33.0	.00	.00	8552500.00	.00
4. 2 CIRCULATING H2O SYS	EACH	1.0	.00	.00	726684.49	994019.45
4. 3 SURFACE CONDENSER	FT2	227387.9	.00	.00	1054263.56	159521.55
PERCENT TOTAL DIRECT COST IN ACCOUNT 4 = 8.650 ACCOUNT TOTAL,$					10453447.87	1153541.00
STRUCTURAL FEATURES						
5. 1 STAT. STRUCTURAL ST.	TON	1500.0	650.00	175.00	975000.00	262500.00
5. 2 SILOS & BUNKERS	TPH	221.4	1800.00	750.00	398440.67	166016.95
5. 3 CHIMNEY	FT	500.0	.00	.00	593557.88	890336.82
5. 4 STRUCTURAL FEATURES	EACH	1.0	374000.00	114000.00	374000.00	114000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 5 = 2.812 ACCOUNT TOTAL,$					2340998.53	1432853.77
BUILDINGS						
6. 1 STATION BUILDINGS	FT3	3750000.0	.16	.16	600000.00	600000.00
6. 2 ADMINSTRATION	FT2	5000.0	18.00	14.00	90000.00	70000.00
6. 3 WAREHOUSE & SHOP	FT2	10000.0	12.00	8.00	120000.00	80000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 6 = 1.155 ACCOUNT TOTAL,$					800000.00	750000.00
FUEL HANDLING & STORAGE						
7. 1 COAL HANDLING SYS	TPH	225.8	.00	.00	3344687.59	1542485.47
7. 2 DOLOMITE HAND. SYS	TPH	33.0	.00	.00	590708.02	349632.92
7. 3 FUEL OIL HAND. SYS	GAL	100000.0	.00	.00	20706.41	16770.13
PERCENT TOTAL DIRECT COST IN ACCOUNT 7 = 4.370 ACCOUNT TOTAL,$					3956102.00	1907888.47
FUEL PROCESSING						
8. 1 COAL DRYER & CRUSHER	TPH	.0	.00	.00	.00	.00
8. 2 CARBONIZERS	TPH	.0	.00	.00	.00	.00
8. 3 GASIFIERS	TPH	.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 8 = .000 ACCOUNT TOTAL,$					.00	.00

Table 2.48 Continued

ADVANCED STEAM CYCLE WITH ATM BOILER ACCOUNT LISTING
PARAMETRIC POINT NO.21

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
FIRING SYSTEM						
9. 1		.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 9 =			.000	ACCOUNT TOTAL,$	.00	.00
VAPOR GENERATOR (FIRED)						
10. 1 ATM STEAM BOILER	EACH	1.0	21167000.00	10583500.00	21167000.00	10583500.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 10 =			23.661	ACCOUNT TOTAL,$	21167000.00	10583500.00
ENERGY CONVERTER						
11. 1 STEAM TURBINE-GEN	EACH	1.0	17936842.00	1261002.05	17936842.00	1261002.06
11. 2 STEAM PIPING	EACH	1.0	1800000.00	700000.00	1800000.00	700000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 11 =			16.169	ACCOUNT TOTAL,$	19736842.00	1961002.06
COUPLING HEAT EXCHANGER						
12. 1		.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 12 =			.000	ACCOUNT TOTAL,$	.00	.00
HEAT RECOVERY HEAT EXCH.						
13. 1 FEED WATER HEATER STRING		1.0	1200000.00	36000.00	1200000.00	36000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 13 =			.921	ACCOUNT TOTAL,$	1200000.00	36000.00
WATER TREATMENT						
14. 1 DEMINERALIZER	GPM	80.0	2500.00	700.00	200000.00	56000.00
14. 2 CONDENSATE POLISHING	KWE	500000.0	1.25	.30	625000.00	150000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 14 =			.768	ACCOUNT TOTAL,$	825000.00	206000.00
POWER CONDITIONING						
15. 1 STD TRANSFORMER	KWE	511111.1	.00	.00	1258104.41	25362.09
PERCENT TOTAL DIRECT COST IN ACCOUNT 15 =			.964	ACCOUNT TOTAL,$	1268104.41	25362.09
AUXILIARY MECH EQUIPMENT						
16. 1 BOILER FEED PUMP &DR.	KWE	548189.2	1.67	.10	915475.99	54818.92
16. 2 OTHER PUMPS	KWE	603008.1	.88	.12	530647.16	72360.98
16. 3 MISC SERVICE SYS	KWE	548189.2	1.17	.73	641381.38	400178.12
16. 4 AUXILIARY BOILER	PPH	200000.0	4.00	.80	800000.00	160000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 16 =			2.664	ACCOUNT TOTAL,$	2887504.50	687358.01
PIPE & FITTINGS						
17. 1 CONVENTIONAL PIPING	TON	750.0	3000.00	1800.00	2250000.00	1350000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 17 =			2.683	ACCOUNT TOTAL,$	2250000.00	1350000.00
AUXILIARY ELEC EQUIPMENT						
18. 1 MISC MOTERS,ETC		548189.2	1.40	.17	767464.91	93192.17
18. 2 SWITCHGEAR & MCC PAN	KWE	548189.2	1.95	.45	1068968.97	246685.15
18. 3 CONDUIT,CABLES,TRAYS	FT	1950000.0	1.32	1.36	2573999.97	2651999.97
18. 4 ISOLATED PHASE BUS	FT	450.0	510.00	450.00	229500.00	202500.00
18. 5 LIGHTING & COMMUN	KWE	548189.2	.35	.43	191866.23	235721.36
PERCENT TOTAL DIRECT COST IN ACCOUNT 18 =			6.157	ACCOUNT TOTAL,$	4831800.00	3430098.62

Table 2-48 Continued

ADVANCED STEAM CYCLE WITH ATM BOILER ACCOUNT LISTING
PARAMETRIC POINT NO.21

ACCOUNT NO. & NAME,	UNIT	AMOUNT	MAT $/UNIT	INS $/UNIT	MAT COST,$	INS COST,$
CONTROL, INSTRUMENTATION						
19. 1 COMPUTER	EACH	1.0	400000.00	10000.00	400000.00	10000.00
19. 2 OTHER CONTROLS	EACH	1.0	400000.00	240000.00	400000.00	240000.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 19 = .782 ACCOUNT TOTAL,$					800000.00	250000.00
PROCESS WASTE SYSTEMS						
20. 1 BOTTOM ASH	TPH	4.3	539427.92	134856.98	539427.92	134856.98
20. 2 DRY ASH	TPH	17.3	1170850.64	292712.66	1170850.64	292712.66
20. 3 WET SLURRY	TPH	33.0	1003468.01	250867.00	1003468.01	250867.00
20. 4 ONSITE DISPOSAL	ACRE	177.9	6860.92	10087.84	1220735.69	1794913.95
PERCENT TOTAL DIRECT COST IN ACCOUNT 20 = 4.775 ACCOUNT TOTAL,$					3934482.25	2473350.59
STACK GAS CLEANING						
21. 1 PRECIPITATOR	EACH	1.0	2368526.34	1539542.11	2368526.34	1539542.11
21. 2 SCRUBBER	KWE	500000.0	28.01	12.84	14003197.67	6420238.69
21. 3 MISC STEEL & DUCTS		.0	.00	.00	.00	.00
PERCENT TOTAL DIRECT COST IN ACCOUNT 21 =18.132 ACCOUNT TOTAL,$					16371724.12	7959780.75
TOTAL DIRECT COSTS,$					96054217.00	38135744.50

2.7.2 Cost of Electricity Display

Following the detailed account listing are printouts of the cost of electricity as affected by labor rates, contingency, escalation, interest during construction, fixed charges, fuel cost, and capacity factor, an example of which is given in Table 2.49. The base case and four variations are given for each. Only one number was varied during each comparison, with the other six held at the base values.

The base case values chosen were a labor rate of $0.002944/s ($10.60/hr), a percent contingency of 3 plus the time of construction in years, an escalation rate of 6.5%, an interest during construction rate of 10%, a fixed charge rate of 18%, a fuel cost as the second number cited, and a capacity factor of 65%.

The effect of the field labor rate was calculated by multipling the total direct installation cost by the ratio of the labor rate divided by the base labor rate.

The numbers in the column headed "rate, percent" are correct with two exceptions. Identical zeros have no meaning, and the rate associated with a variable in that listing will be the last amount used. For example, the contingency cost shows a rate percent of 20 in the second case where the contingency is varied for values of -5, 0, 7, 5, and 20%. The 20% obviously does not apply here.

2.7.3 Internal Auxiliary Power Calculations

On the last page of the output (Table 2.50) associated with each parametric point are three groups of printout: a listing of the internally calculated auxiliary power and operation or maintanence costs associated with each account; numbers describing the system power output and heat rejection system; and the input data list.

The auxiliary power calculated internal to the program is shown for the account with which it is associated. This is not to be construed as the total auxiliary power since many other items were taken into account by the concept team in their calculations, e.g., boiler feed pump

Table 2.49

ADVANCED STEAM CYCLE WITH ATM BOILER COST OF ELECTRICITY,MILLS/KW.HR
PARAMETRIC POINT NO.21

ACCOUNT	RATE, PERCENT	LABOR RATE, $/HR 5.00	8.50	10.60	15.00	21.50
TOTAL DIRECT COSTS,$	.0	117640487.	126634766.	134189961.	150019892.	173405018.
INDIRECT COST,$	51.0	11338993.	15596080.	13449229.	27522494.	39448908.
PROF & OWNER COSTS,$	8.0	9411239.	10130781.	10735197.	12001591.	13872401.
CONTINGENCY COST,$	8.0	9411239.	10130781.	10735197.	12001591.	13872401.
SUB TOTAL,$	.0	147471960.	162492406.	175109582.	201545566.	240598726.
ESCALATION COST,$	6.5	30235020.	33314545.	35901345.	41321308.	49328071.
INTREST DURING CONST,$	10.0	34827085.	38374325.	41354006.	47597147.	56819570.
TOTAL CAPITALIZATION,$	.0	212534064.	234181274.	252364932.	290464020.	346746764.
COST OF ELEC-CAPITAL	18.0	14.72342	16.22304	17.48273	20.12207	24.02109
COST OF ELEC-FUEL	.0	9.07458	9.07458	9.07468	9.07468	9.07468
COST OF ELEC-OP & MAIN	.0	1.12727	1.12727	1.12727	1.12727	1.12727
TOTAL COST OF ELEC	.0	24.92537	26.42500	27.68468	30.32402	34.22304

ACCOUNT	RATE, PERCENT	CONTINGENCY, PERCENT -5.00	.00	8.00	5.00	20.00
TOTAL DIRECT COSTS,$	.0	134189961.	134189961.	134189961.	134189961.	134189961.
INDIRECT COST,$	51.0	19449229.	19449229.	19449229.	19449229.	19449229.
PROF & OWNER COSTS,$	8.0	10735197.	10735197.	10735197.	10735197.	10735197.
CONTINGENCY COST,$	20.0	-6709498.	0.	10735197.	6709498.	26837992.
SUB TOTAL,$	.0	157664888.	164374386.	175109582.	171083884.	191212378.
ESCALATION COST,$	6.5	32324797.	33700392.	35901345.	35075988.	39202775.
INTREST DURING CONST,$	10.0	37234254.	38818774.	41354006.	40403294.	45156854.
TOTAL CAPITALIZATION,$	.0	227223938.	236893552.	252364932.	246563166.	275572004.
COST OF ELEC-CAPITAL	18.0	15.74107	16.41094	17.48273	17.08091	19.09041
COST OF ELEC-FUEL	.0	9.07468	9.07468	9.07468	9.07468	9.07468
COST OF ELEC-OP & MAIN	.0	1.12727	1.12727	1.12727	1.12727	1.12727
TOTAL COST OF ELEC	.0	25.94302	26.61289	27.68468	27.28276	29.29237

ACCOUNT	RATE, PERCENT	ESCALATION RATE, PERCENT 5.00	6.50	8.00	10.00	.00
TOTAL DIRECT COSTS,$	.0	134189961.	134189961.	134189961.	134189961.	134189961.
INDIRECT COST,$	51.0	19449229.	19449229.	19449229.	19449229.	19449229.
PROF & OWNER COSTS,$	8.0	10735197.	10735197.	10735197.	10735197.	10735197.
CONTINGENCY COST,$	8.0	10735197.	10735197.	10735197.	10735197.	10735197.
SUB TOTAL,$	.0	175109582.	175109582.	175109582.	175109582.	175109582.
ESCALATION COST,$	.0	27153781.	35901345.	44937870.	57447529.	0.
INTREST DURING CONST,$	10.0	39908577.	41354006.	42837510.	44875863.	35357455.
TOTAL CAPITALIZATION,$	.0	242171938.	252364932.	262884962.	277432972.	210467036.
COST OF ELEC-CAPITAL	18.0	16.77660	17.48273	18.21151	19.21933	14.58023
COST OF ELEC-FUEL	.0	9.07458	9.07458	9.07468	9.07468	9.07468
COST OF ELEC-OP & MAIN	.0	1.12727	1.12727	1.12727	1.12727	1.12727
TOTAL COST OF ELEC	.0	26.97856	27.68458	28.41346	29.42128	24.78218

ACCOUNT	RATE, PERCENT	INT DURING CONST,PERCENT 6.00	8.00	10.00	12.50	15.00
TOTAL DIRECT COSTS,$	.0	134189961.	134189961.	134189961.	134189961.	134189961.
INDIRECT COST,$	51.0	19449229.	19449229.	19449229.	19449229.	19449229.
PROF & OWNER COSTS,$	8.0	10735197.	10735197.	10735197.	10735197.	10735197.
CONTINGENCY COST,$	8.0	10735197.	10735197.	10735197.	10735197.	10735197.
SUB TOTAL,$	.0	175109582.	175109582.	175109582.	175109582.	175109582.
ESCALATION COST,$	6.5	35901345.	35901345.	35901345.	35901345.	35901345.
INTREST DURING CONST,$	15.0	24167798.	32650795.	41354006.	52548896.	64102237.
TOTAL CAPITALIZATION,$	.0	235178724.	243661722.	252364932.	263559822.	275113160.
COST OF ELEC-CAPITAL	18.0	15.29214	16.87931	17.48273	18.25826	19.05863
COST OF ELEC-FUEL	.0	9.07468	9.07468	9.07468	9.07468	9.07468
COST OF ELEC-OP & MAIN	.0	1.12727	1.12727	1.12727	1.12727	1.12727
TOTAL COST OF ELEC	.0	26.49410	27.08176	27.68468	28.46021	29.26058

Table 2.49 Continued

ADVANCED STEAM CYCLE WITH ATM BOILER COST OF ELECTRICITY,MILLS/KW.HR
PARAMETRIC POINT NO.21

ACCOUNT	RATE, PERCENT	FIXED CHARGE RATE, PCT 10.00	14.40	18.00	21.60	25.00
TOTAL DIRECT COSTS,$	.0	134189961.	134189961.	134189961.	134189961.	134189961.
INDIRECT COST,$	51.0	19449229.	19449229.	19449229.	19449229.	19449229.
PROF & OWNER COSTS,$	8.0	10735197.	10735197.	10735197.	10735197.	10735197.
CONTINGENCY COST,$	8.0	10735197.	10735197.	10735197.	10735197.	10735197.
SUB TOTAL,$	.0	175109582.	175109582.	175109582.	175109582.	175109582.
ESCALATION COST,$	6.5	35901345.	35901345.	35901345.	35901345.	35901345.
INTREST DURING CONST,$	10.0	41354006.	41354006.	41354006.	41354006.	41354006.
TOTAL CAPITALIZATION,$	.0	252364932.	252364932.	252364932.	252354932.	252364932.
COST OF ELEC-CAPITAL	25.0	9.71263	13.98618	17.48273	20.97927	24.28157
COST OF ELEC-FUEL	.0	9.07458	9.07458	9.07468	9.07468	9.07468
COST OF ELEC-OP & MAIN	.0	1.12727	1.12727	1.12727	1.12727	1.12727
TOTAL COST OF ELEC	.0	19.91458	24.18814	27.68468	31.18123	34.48352

ACCOUNT	RATE, PERCENT	FUEL COST, $/10**6 BTU .50	.85	1.50	2.50	1.02
TOTAL DIRECT COSTS,$	.0	134189961.	134189951.	134189961.	134189961.	134189961.
INDIRECT COST,$	51.0	19449229.	19449229.	19449229.	19449229.	19449229.
PROF & OWNER COSTS,$	8.0	10735197.	10735197.	10735197.	10735197.	10735197.
CONTINGENCY COST,$	8.0	10735197.	10735197.	10735197.	10735197.	10735197.
SUB TOTAL,$	.0	175109582.	175109582.	175109582.	175109582.	175109582.
ESCALATION COST,$	6.5	35901345.	35901345.	35901345.	35901345.	35901345.
INTREST DURING CONST,$	10.0	41354006.	41354006.	41354006.	41354006.	41354006.
TOTAL CAPITALIZATION,$	.0	252364932.	252364932.	252364932.	252364932.	252364932.
COST OF ELEC-CAPITAL	18.0	17.48273	17.48273	17.48273	17.48273	17.48273
COST OF ELEC-FUEL	.0	5.33805	9.07468	16.01414	26.69023	10.88961
COST OF ELEC-OP & MAIN	.0	1.12727	1.12727	1.12727	1.12727	1.12727
TOTAL COST OF ELEC	.0	23.94805	27.68468	34.62414	45.30023	29.49962

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ACCOUNT	RATE, PERCENT	CAPACITY FACTOR, PERCENT 12.00	45.00	50.00	65.00	80.00
TOTAL DIRECT COSTS,$	.0	134189961.	134189961.	134189961.	134189961.	134189961.
INDIRECT COST,$	51.0	19449229.	19449229.	19449229.	19449229.	19449229.
PROF & OWNER COSTS,$	8.0	10735197.	10735197.	10735197.	10735197.	10735197.
CONTINGENCY COST,$	8.0	10735197.	10735197.	10735197.	10735197.	10735197.
SUB TOTAL,$	.0	175109582.	175109582.	175109582.	175109582.	175109582.
ESCALATION COST,$	6.5	35901345.	35901345.	35901345.	35901345.	35901345.
INTREST DURING CONST,$	10.0	41354006.	41354006.	41354006.	41354006.	41354006.
TOTAL CAPITALIZATION,$	.0	252364932.	252364932.	252364932.	252354932.	252364932.
COST OF ELEC-CAPITAL	18.0	94.69811	25.25283	22.72755	17.48273	14.20472
COST OF ELEC-FUEL	.0	9.07468	9.07458	9.07468	9.07468	9.07468
COST OF ELEC-OP & MAIN	.0	1.12731	1.12728	1.12728	1.12727	1.12727
TOTAL COST OF ELEC	.0	104.90010	35.45479	32.92950	27.68468	24.40667

Table 2.50 Auxilixry Output and Input List

ADVANCED STEAM CYCLE WITH ATM BOILER

ACCOUNT NO	AUX POWER, MWE	PERC PLANT POW	OPERATION COST	MAINTENANCE COST
4	12.48517	28.58694	.00000	19.87498
7	2.71717	6.22142	176.45388	.00000
10	8.50000	19.46220	.00000	.00000
14	.00000	.00000	8.65297	.00000
18	5.50000	12.59319	.00000	.00000
20	5.32569	12.19409	.00000	.00000
21	9.14636	20.94215	.00000	.00000
TOTALS	43.57439	9.57033	514.40083	13.87498

1	500.000	2	.000	3	.000	4	8421.000	5	5.000
5	500.000	7	9.000	8	2563000000.000	9	2.000	10	1.000
11	2.000	12	.000	13	1.000	14	.000	15	.000
15	2.000	17	134.000	18	3.000	19	5.000	20	2.500
21	.000	22	15800.000	23	.000	24	1500.000	25	500.000
25	3750000.000	27	5000.000	29	10000.000	29	100000.000	30	1.100
31	1.000	32	750.000	33	1.000	34	1.000	35	1.000
36	1950000.000	37	450.000	38	1.000	39	1.000	40	374000.000
41	114000.000	42	400000.000	43	10000.000	44	400000.000	45	240000.000
46	.000	47	.000	43	3.000	49	2.000	50	1.000
51	.000	52	5.350						
1	1.000	2	20900000.000	3	.500	4	1.000	5	1.000
6	1.000	7	14400000.000	8	.000	9	3800000.000	10	.000
11	1800000.000	12	700000.000	13	1.000	14	1200000.000	15	.030
16	.000	17	200000.000	18	1.000	19	1.000	20	1.000
21	.000	22	.000	23	.000	24	.000	25	.000
26	1.000	27	.000	28	.000	29	.000	30	.000

ADVANCED STEAM CYCLE WITH ATM BOILER BASE CASE INPUT

NOMINAL POWER, MWE	500.0000	NET POWER, MWE	456.3256
NOM HEAT RATE, BTU/KW-HR	9743.5430	NET HEAT RATE, BTU/KW-HR	10676.0925
OFF DESIGN HEAT RATE	9827.9375		
CONDENSER			
DESIGN PRESSURE, IN HG A	3.0000	NUMBER OF SHELLS	2.0000
NUMBER OF TUBES/SHELL	6385.4348	TUBE LENGTH, FT	68.1604
U, BTU/HR-FT2-F	591.4577	TERMINAL TEMP DIFF, F	6.2500
HEAT REJECTION			
DESIGN TEMP, F	93.0000	APPROACH, F	28.9353
RANGE, F	28.7500	OFF DESIGN TEMP, F	59.0000
OFF DESIGN PRES, IN HG A	3.8130	LP TURBINE BLADE LEN, IN	25.0000

power. Some concepts also made allowances for station power requirements of a general nature. Where the allowances were deemed adequate by the A/E, no additional auxiliary power was included here.

The second column gives the percent of the auxiliary power internally calculated for each account, and the bottom line includes the total internally calculated auxiliary power in megawatts and the percent of station power this represents.

The operation and/or maintenance cost associated with each account are also listed in \$/hour of operation. These charges include the cost of limestone (Account 7) and the cost of water (Accounts 4 and 14), among others.

The general plant manning charges are not included here.

2.7.4 Plant Power and Heat Rejection

The second grouping of printouts in Table 2.50 is entitled "Base Case Input." This is a misnomer in that only a part of those items listed are input. Nominal power, MWe, is the station power cited by the concept team. The net power, MWe, is the nominal power less the auxiliary power calculated internal to the program and is assumed to be the station power delivered to the grid. The nominal and net heat rates are the heat rates based on the nominal and net powers respectively.

The next item in the printout is called "steam turbine heat rate change" in some printouts and "off design heat rate" in these plants with steam turbines only. The number cited in the first case is the ratio of the steam turbine heat rate at the off-design ambient to the heat rate at the design ambient. In the second case, where only the steam turbine is present (no topping cycle or gas turbine pump up sets), the off-design heat rate is reported directly. The number cited should be viewed with care since, in some cases, due to an extrapolation beyond the range of a polynomial fit of the steam turbine back pressure heat rate correction curves the numbers are inane.

The condenser design pressure, number of shells, number of tubes per shell, tube length, overall heat transfer coefficient, and terminal temperature difference are self-explanatory. The total condenser surface area was given as the amount in Subaccount 4.3.

The heat rejection design ambient, range, and resultant approach appear next, followed by the off-design ambient temperature and the expected off-design condenser back pressure. The last item is the last-row blade length of the steam turbine LP ends used. Only data for 635, 723.9, and 787.4 mm (25, 28.5, and 31 in) LP ends were programmed into the computer, so in the few cases where an LP end with less than 635 mm (25 in) blades were used, the output will still show 25 in.

2.7.5 Input Data

The two groups of numbers in the center portion of Table 2.50 represent an "A" and "B" input matrix. The "A" matrix is similar for most systems and will be explained in detail. The "B" matrix is associated with the amounts and cost of major equipment defined by the concept team.

The "A" matrix always consisted of 52 inputs as defined in Table 2.51.

Table 2.51 - Definition of "A" Matrix Inputs

Item Number	Scaling Power, n	Description	Units
1		Nominal Station Power	MWe
2		Cycle Thermodynamic Efficiency	
3		Plant Efficiency	
4		Plant Heat Rate	Btu/kWh
5	0.175	Time of Construction	years
6		Steam Turbine Power	MWe
7		Condenser Pressure	in Hg abs

Table 2.51 (continued)

Item Number	Scaling Power, n	Description	Units
8		Cycle Heat Rejection to Cooling Water	Btu/hr or MWt
9		Number of Condenser Shells	
10		Steam Turbine Last-Row Blade Length 1 - 25 in , 2 - 28.5 in , 3 - 31 in	
11		Means of Heat Rejection 1 - Wet Tower, 2 - Dry Tower, 3 - Once-through	
12		Steam Turbine Drive Compressor Power	MWe
13		Coal Used 1 - bituminous, 2 - subbituminous, 3 - lignite	
14		Fuel Form as Used 0 - coal direct, 1 - clean distillate, 2 - high-Btu gas, 3 - medium-Btu gas, 4 - low-Btu gas, 5 - methanol, 6 - hydrogen	
15		Approximate Low-Btu Gasifier Air Inlet Temp. 1 - 750°F, 2 - 550°F, 3 - 350°F	
16		Steam Turbine Throttle Pressure 1 - 2400 psig or less, 2 - 3500 psig 3 - 5000 psig or more	
17		Site Size	
18		Site Type 3 - Middletown, 4 - Industrial, 5 - Commercial	
19	0.32	Access Railroad	miles

Table 2.51 (continued)

Item Number	Scaling Power, n	Description	Units
20	0.32	Loop Track	miles
21	0.73	Ladder Track	miles
22	0.73	Plant Island Concrete	yd^3
23	0.55	Special Concrete	yd^3
24	0.20	Station Structural Steel	tons
25	0.62	Chimney Height	ft
26	0.62	Station Buildings	ft^3
27	0.5	Administration Buildings	ft^2
28	0.8	Warehouse, shop, and garages	ft^2
29	0.62	Distillate storage	gal
30	0.62	Factor (Subaccount 16.2)	
31	0.62	Factor (Subaccount 16.3)	
32	0.62	Piping	tons
33		Coal Silos and Bumpers (Factor Sub-account 5.2)	
34	0.62	Factor (Subaccount 18.1)	
35	0.62	Factor (Subaccount 18.2)	
36	0.62	Conduit Trays and Cable	linear ft
37	0.32	Isolated Phase Bus and Leads	linear ft
38	0.25	Computer (number)	
39	0.32	Other Controls (number)	
40	0.4	Miscellaneous Structural Features Material Cost (Subaccount 5.4)	$
41	0.4	Miscellaneous Structural Features Installation Cost (Subaccount 5.4)	$
42	0.25	Computer Material Cost (Subaccount 19.1)	$
43	0.25	Computer Installation Cost (Subaccount 19.1)	$

Table 2.51 (continued)

Item Number	Scaling Power, n	Description	Units
44	0.32	Other Controls Material (Subaccount 19.2)	$
45	0.32	Other Controls Installation (Subaccount 19.2)	$
46		Coal Processing Equipment Type 1 - Low-Btu Gasifier (hot cleanup - 15 to 20 atm) 2 - Low-Btu Gasifier (hot cleanup - 5 atm) 3 - Low-Btu Gasifier (cold cleanup - 15 to 20 atm) 4 - Low-Btu Gasifier (cold cleanup - 5 atm) 5 - Carbonizer 6 - Crusher Only	
47		Carbonizer Input Coal Description 0 - no carbonizer, 1 - bituminous (0% moisture), 2 - subbituminous (20% moisture), 3 - subbituminous (16% moisture), 4 - lignite (27% moisture), 5 - lignite (18% moisture)	
48		Transmission Voltage 1 - 69 kV, 2 - 230 kV, 3 - 550 kV	
49		Ambient 1 - ISO day, 2 - 5% day	
50		Ash 0-0% bottom ash - 100% ash carry-over 1-20% bottom ash - 80% ash carry-over 2-80% bottom ash - 20% ash carry-over	

Table 2.51 (continued)

Item Number	Scaling Power, n	Description	Units
		3-90% bottom ash - 10% ash carry-over	
		4-95% bottom ash - 5% ash carry-over	
		5-100% bottom ash - 0% ash carry-over	
		6-0% bottom ash - 0% ash carry-over (a clean fuel)	
51		Dummy Variable	
52		Cost of Limestone	$/ton

In order to modify the base cases to fit power plants of other sizes for the various parametric points, a ratio was formed by dividing the nominal power of that plant by the nominal power of the base case plant. This ratio raised to the scaling power, n, (also given in Table 2.51), was then used as a multiplier to modify the values originally calculated for the base case. These scaling powers were used for all concepts except fuel cells.

The "B" matrix includes numbers or amounts followed by material and installation costs. The size of the "B" matrix varied from 25 to 182 elements. No attempt is made to cite the particulars of each concept here. The significance of this input can be found only by looking at the programs for each concept individually.

2.7.6 Summary Tables

Three different sets of summary sheets are found in the printout after the last parametric point detailed listing. The first of these includes the plant efficiency, cost of electricity, and construction time; the second, a more detailed breakdown of the cost of electricity, including the indirect cost breakdown; and the third, a table of the natural resources required (coal, sorbent, water, etc.).

2.7.6.1 Efficiencies and Cost of Electricity

Table 2.52 shows an example of the first set of summary tables. They were used to prepare the NASA-specified report forms and include values of plant efficiency, plant capital cost, and the cost of electricity for the base value of each of the seven cost variables treated and the time of construction in years.

2.7.6.2 Cost of Electricity

Table 2.53 is an example of the summary tables prepared which detail the material cost of some of the major component subsystems. All other direct material costs are lumped together as balance of plant costs. Site labor refers to the total installation cost and includes all labor-related direct cost, including some subcontractor, indirect, and profit. The indirect costs listed correspond to the base value of each of the seven cost variables. In addition, special cases involving the effect on the cost of electricity of a change in capacity factor from 0.65 to both 0.5 and 0.8 are separately listed. Also listed are the effects of 20% increases in the cost of fuel or capitalization and of zero contingency or escalation charges.

2.7.6.3 Resource Usage

Table 2.54 shows an example of a resource usage table. Included for each parametric point are the amount of coal, sorbent, and water used as well as a breakdown of total land use for the plant, access railroad, and disposal area for ash and spent sorbent.

Table 2.52 ADVANCED STEAM CYCLE WITH ATM BOILER SUMMARY PLANT RESULTS

PARAMETRIC POINT	1	2	3	4	5	6	7	8
THERMODYNAMIC EFF	.461	.449	.420	.457	.445	.445	.488	.475
POWER PLANT EFF	.369	.359	.332	.366	.356	.354	.390	.381
OVERALL ENERGY EFF	.369	.359	.332	.366	.356	.354	.390	.381
CAP COST MILLION $	730.110	729.501	740.668	252.422	251.474	212.223	396.319	355.989
CAPITAL COST,$/KWE	1572.383	1571.859	1519.128	544.229	542.298	455.068	852.397	851.778
COE CAPITAL	49.722	49.690	51.184	17.204	17.143	14.417	26.946	26.927
COE FUEL	7.861	8.073	8.739	7.930	8.146	8.202	7.429	7.618
COE OP & MAIN	1.132	1.145	1.111	1.138	1.152	2.068	1.106	1.118
COST OF ELECTRIC	58.715	58.908	61.035	26.273	26.441	24.687	35.481	35.663
EST TIME OF CONST	6.000	6.000	6.000	5.000	5.000	5.000	5.000	5.000

PARAMETRIC POINT	9	10	11	12	13	14	15	16
THERMODYNAMIC EFF	.475	.513	.501	.501	.477	.481	.453	.442
POWER PLANT EFF	.379	.411	.401	.399	.382	.385	.363	.354
OVERALL ENERGY EFF	.379	.411	.401	.399	.382	.385	.363	.354
CAP COST MILLION $	325.760	593.932	587.456	472.307	355.547	381.380	239.119	236.540
CAPITAL COST,$/KWE	598.435	1275.355	1251.193	1010.847	755.247	820.644	515.696	509.870
COE CAPITAL	22.079	40.307	39.869	31.955	24.191	25.942	16.302	16.118
COE FUEL	7.661	7.063	7.231	7.254	7.600	7.540	8.001	8.206
COE OP & MAIN	1.974	1.083	1.093	1.905	1.117	1.113	1.143	1.152
COST OF ELECTRIC	31.714	48.453	48.193	41.124	32.908	34.596	25.446	25.476
EST TIME OF CONST	5.000	5.000	5.000	5.000	5.000	5.000	5.000	5.000

PARAMETRIC POINT	17	18	19	20	21	22	23	24
THERMODYNAMIC EFF	.471	.459	.445	.434	.405	.474	.461	.461
POWER PLANT EFF	.377	.367	.356	.347	.320	.379	.369	.367
OVERALL ENERGY EFF	.377	.367	.356	.347	.320	.379	.369	.367
CAP COST MILLION $	353.086	318.517	232.825	231.064	252.365	289.976	288.105	252.570
CAPITAL COST,$/KWE	759.353	685.924	502.445	498.748	553.037	624.282	520.336	542.003
COE CAPITAL	24.021	21.684	15.883	15.767	17.483	19.735	19.610	17.134
COE FUEL	7.592	7.896	8.149	8.363	9.075	7.655	7.851	7.899
COE OP & MAIN	1.118	1.135	1.153	1.166	1.127	1.121	1.133	2.016
COST OF ELECTRIC	32.731	30.715	25.185	25.295	27.685	28.511	28.594	27.049
EST TIME OF CONST	5.000	5.000	5.000	5.000	5.000	5.000	5.000	5.000

PARAMETRIC POINT	25	26	27	28	29	30	31	32
THERMODYNAMIC EFF	.495	.483	.483	.458	.471	.485	.485	.475
POWER PLANT EFF	.397	.387	.385	.366	.377	.388	.386	.380
OVERALL ENERGY EFF	.397	.387	.385	.366	.377	.388	.386	.380
CAP COST MILLION $	378.948	378.156	347.011	263.428	296.284	324.880	288.986	345.269
CAPITAL COST,$/KWE	814.360	812.733	743.411	567.946	637.969	698.882	619.684	743.242
COE CAPITAL	25.744	25.692	23.501	17.954	20.168	22.093	19.590	23.496
COE FUEL	7.305	7.490	7.529	7.922	7.697	7.469	7.507	7.627
COE OP & MAIN	1.098	1.110	1.951	1.138	1.123	1.109	1.948	1.119
COST OF ELECTRIC	34.149	34.292	32.981	27.014	28.989	30.671	29.045	32.242
EST TIME OF CONST	5.000	5.000	5.000	5.000	5.000	5.000	5.000	5.000

PARAMETRIC POINT	33	34	35	36	37	38	39	40
THERMODYNAMIC EFF	.475	.434	.434	.434	.434	.434	.437	.426
POWER PLANT EFF	.378	.350	.335	.344	.350	.335	.349	.340
OVERALL ENERGY EFF	.378	.350	.335	.344	.350	.335	.349	.340
CAP COST MILLION $	311.543	232.818	245.135	203.015	223.288	231.889	223.003	220.183
CAPITAL COST,$/KWE	669.458	499.990	527.929	436.547	479.524	499.403	481.726	475.576
COE CAPITAL	21.131	15.806	16.689	13.803	15.159	15.787	15.228	15.034
COE FUEL	7.670	8.296	8.643	8.424	8.296	8.648	8.308	8.522
COE OP & MAIN	1.976	.894	.912	2.107	.894	.912	1.157	1.170
COST OF ELECTRIC	30.777	24.996	26.250	24.334	24.349	25.349	24.694	24.727
EST TIME OF CONST	5.000	5.000	5.000	5.000	5.000	5.000	5.000	5.000

Table 2.53 ADVANCED STEAM CYCLE WITH ATM BOILER SUMMARY PLANT RESULTS

PLANT RESULT BREAKDOWN

PARAMETRIC POINT		17	18	19	20	21	22	23	24
TOTAL CAPITAL COST	,M$	353.09	318.52	232.82	231.06	252.36	289.98	288.11	252.57
STM TURB-GEN & FEED STRING	,M$	56.354	52.793	17.044	15.525	13.137	38.444	36.525	36.525
STEAM BOILER	,M$	33.267	24.367	19.367	19.867	21.167	22.367	22.967	23.800
STEAM PIPING	,M$	.000	.000	.000	.000	.000	.000	.000	.000
TOT MAJOR COMPONENT COST	,M$	93.121	77.150	35.411	35.392	40.304	60.811	59.492	60.325
TOT MAJOR COMPONENT COST	,$/KWE	193.946	166.163	78.576	76.392	88.323	130.918	128.095	129.454
BALANCE OF PLANT COST	,$/KWE	113.239	111.738	104.559	104.355	122.172	117.154	117.269	98.299
SITE LABOR	,$/KWE	102.301	91.964	81.486	81.670	83.571	87.071	87.313	66.999
TOTAL DIRECT COST	,$/KWE	409.547	369.855	264.721	262.423	294.066	335.144	332.677	294.752
INDIRECT COSTS	,$/KWE	52.174	46.901	41.558	41.652	42.621	44.406	44.530	34.170
PROF & OWNER COSTS	,$/KWE	32.754	29.589	21.178	20.994	23.525	26.811	26.614	23.580
CONTINGENCY COST	,$/KWE	32.754	29.589	21.178	20.994	23.525	26.811	26.614	23.580
ESCALATION COST	,$/KWE	108.097	97.579	71.478	70.952	73.675	89.810	80.249	77.105
INT DURING CONSTRUCTION	,$/KWE	124.515	112.399	82.334	81.728	90.624	102.298	101.652	88.816
TOTAL CAPITALIZATION	,$/KWE	759.850	635.924	502.445	493.743	553.037	624.292	620.336	542.003
COST OF ELEC-CAPITAL	,MILLS/KWE	24.021	21.684	15.893	15.767	17.483	19.735	19.610	17.134
COST OF ELEC-FUEL	,MILLS/KWE	7.592	7.895	8.149	8.353	9.075	7.655	7.851	7.899
COST OF ELEC-OP&MAIN	,MILLS/KWE	1.118	1.135	1.153	1.166	1.127	1.121	1.133	2.016
TOTAL COST OF ELEC	,MILLS/KWE	32.931	30.715	25.195	25.295	27.685	28.511	28.594	27.049
COE 0.5 CAP. FACTOR	,MILLS/KWE	40.037	37.220	29.950	30.025	32.930	34.431	34.477	32.189
COE 0.8 CAP. FACTOR	,MILLS/KWE	28.327	25.549	22.207	22.339	24.407	24.811	24.917	23.836
COE 1.2XCAP. COST	,MILLS/KWE	37.635	35.052	28.362	28.448	31.181	32.458	32.516	30.476
COE 1.2XFUEL COST	,MILLS/KWE	34.363	32.234	25.815	25.353	29.500	30.042	30.165	28.629
COE (CONTINGENCY=0)	,MILLS/KWE	31.336	29.367	24.220	24.339	26.613	27.289	27.382	25.975
COE (ESCALATION=0)	,MILLS/KWE	28.843	27.115	22.548	22.673	24.782	25.234	25.339	24.204

PLANT RESULT BREAKDOWN

PARAMETRIC POINT		25	25	27	28	29	30	31	32
TOTAL CAPITAL COST	,M$	378.95	378.16	347.01	263.43	296.28	324.88	288.99	345.27
STM TURB-GEN & FEED STRING	,M$	70.056	59.437	58.437	27.556	33.144	48.756	48.756	52.444
STEAM BOILER	,M$	25.567	26.167	30.400	21.567	22.967	23.967	26.300	24.967
STEAM PIPING	,M$	.000	.000	.000	.000	.000	.000	.000	.000
TOT MAJOR COMPONENT COST	,M$	95.523	94.604	93.837	48.223	51.111	72.723	75.056	77.411
TOT MAJOR COMPONENT COST	,$/KWE	205.495	203.323	211.741	106.124	131.586	156.442	160.946	166.638
BALANCE OF PLANT COST	,$/KWE	143.212	143.744	123.525	110.636	123.690	129.057	108.887	143.370
SITE LABOR	,$/KWE	96.146	96.609	75.534	85.372	87.755	92.063	70.046	93.478
TOTAL DIRECT COST	,$/KWE	444.852	443.576	411.300	302.133	343.031	377.573	339.879	403.485
INDIRECT COSTS	,$/KWE	49.035	49.271	38.726	43.540	44.755	46.952	35.723	47.674
PROF & OWNER COSTS	,$/KWE	35.588	35.494	32.904	24.175	27.442	30.295	27.190	32.279
CONTINGENCY COST	,$/KWE	35.588	35.494	32.904	24.175	27.442	30.206	27.190	32.279
ESCALATION COST	,$/KWE	115.351	115.519	105.757	80.736	90.757	99.423	88.156	105.733
INT DURING CONSTRUCTION	,$/KWE	133.446	133.179	121.820	93.067	104.541	114.523	101.545	121.792
TOTAL CAPITALIZATION	,$/KWE	814.350	812.733	743.411	567.345	637.969	698.882	619.584	743.242
COST OF ELEC-CAPITAL	,MILLS/KWE	25.744	25.692	23.501	17.954	20.168	22.093	19.590	23.496
COST OF ELEC-FUEL	,MILLS/KWE	7.305	7.490	7.529	7.922	7.697	7.469	7.507	7.627
COST OF ELEC-OP&MAIN	,MILLS/KWE	1.098	1.110	1.951	1.138	1.123	1.109	1.948	1.119
TOTAL COST OF ELEC	,MILLS/KWE	34.143	34.292	32.981	27.014	28.988	30.671	29.045	32.242
COE 0.5 CAP. FACTOR	,MILLS/KWE	41.871	41.999	40.031	32.400	35.038	37.299	34.922	39.291
COE 0.8 CAP. FACTOR	,MILLS/KWE	29.321	29.474	23.574	23.547	25.206	26.528	25.372	27.837
COE 1.2XCAP. COST	,MILLS/KWE	39.296	39.430	37.681	30.605	33.021	35.090	32.963	36.941
COE 1.2XFUEL COST	,MILLS/KWE	35.609	35.790	34.437	28.538	30.527	32.155	30.546	33.767
COE (CONTINGENCY=0)	,MILLS/KWE	32.526	32.675	31.482	25.912	27.738	29.295	27.806	30.771
COE (ESCALATION=0)	,MILLS/KWE	29.874	30.026	29.079	24.033	25.540	27.003	25.792	28.341

Table 2.54 ADVANCED STEAM CYCLE WITH ATM BOILER NATURAL RESOURCE REQUIREMENTS

PARAMETRIC POINT	1	2	3	4	5	6	7	8
COAL, LB/KW-HR	.85723	.88039	.95306	.86483	.88837	.89441	.81015	.83082
SORBANT OR SEED, LB/KW-HR	.12488	.12833	.13911	.12601	.12351	.47323	.11787	.12095
TOTAL WATER, GAL/KW-HR	.966	1.013	.099	1.000	1.048	1.117	.885	.928
COOLING WATER	.875	.921	.000	.909	.955	.952	.800	.841
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP ,	.01034	.01034	.01049	.01035	.01035	.01032	.01032	.01032
WASTE HANDLING SLURRY	.0293	.0301	.0326	.0295	.0303	.1015	.0276	.0283
SCRUBBER WASTE WATER	.05029	.05158	.05602	.05074	.05215	.05355	.04746	.04870
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	101.75	94.86	190.04	106.04	99.15	142.35	99.78	92.78
MAIN PLANT	28.87	28.87	29.29	28.89	28.90	28.80	28.82	28.82
DISPOSAL LAND	33.72	34.65	37.54	34.03	34.96	78.38	31.85	32.67
LAND FOR ACCESS RR	39.17	31.34	123.21	43.12	35.29	35.17	39.11	31.29

PARAMETRIC POINT	9	10	11	12	13	14	15	16
COAL, LB/KW-HR	.93542	.77323	.78359	.79212	.82983	.82229	.87255	.89489
SORBANT OR SEED, LB/KW-HR	.44202	.11193	.11466	.41911	.12065	.11967	.12716	.13049
TOTAL WATER, GAL/KW-HR	.993	.804	.843	.905	.923	.910	1.015	1.026
COOLING WATER	.838	.722	.760	.757	.836	.823	.923	.932
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP ,	.01029	.01030	.01030	.01027	.01033	.01033	.01035	.01035
WASTE HANDLING SLURRY	.0348	.0262	.0269	.0399	.0283	.0280	.0298	.0306
SCRUBBER WASTE WATER	.05013	.04507	.04617	.04753	.04858	.04819	.05120	.05254
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	133.14	94.16	87.08	125.35	100.57	100.29	106.37	95.46
MAIN PLANT	28.73	28.77	28.77	28.68	28.84	28.83	28.90	28.88
DISPOSAL LAND	73.22	30.26	30.99	69.43	32.59	32.33	34.33	35.22
LAND FOR ACCESS RR	31.19	35.13	27.32	27.24	39.13	39.12	43.13	31.35

PARAMETRIC POINT	17	18	19	20	21	22	23	24
COAL, LB/KW-HR	.83880	.86110	.88867	.91200	.98963	.83482	.85621	.86145
SORBANT OR SEED, LB/KW-HR	.12213	.12545	.12955	.13303	.14455	.12154	.12473	.45579
TOTAL WATER, GAL/KW-HR	.893	.930	1.052	1.100	.103	.939	.984	1.051
COOLING WATER	.805	.830	.959	1.005	.000	.851	.894	.891
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP ,	.01033	.01034	.01035	.01035	.01052	.01033	.01034	.01030
WASTE HANDLING SLURRY	.0286	.0294	.0304	.0312	.0333	.0285	.0292	.0978
SCRUBBER WASTE WATER	.04318	.05052	.05217	.05357	.05821	.04894	.05023	.05169
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	100.35	94.06	107.05	100.15	199.84	100.32	93.85	135.47
MAIN PLANT	28.84	28.86	28.92	28.92	29.36	28.85	28.85	28.76
DISPOSAL LAND	32.39	33.88	34.98	35.91	38.99	32.33	33.68	75.51
LAND FOR ACCESS RR	39.13	31.32	43.16	35.32	131.49	39.14	31.32	31.21

PARAMETRIC POINT	25	26	27	28	29	30	31	32
COAL, LB/KW-HR	.79571	.81578	.82134	.85391	.83937	.81449	.81867	.83179
SORBANT OR SEED, LB/KW-HR	.11537	.11886	.43441	.12587	.12222	.11851	.43316	.12109
TOTAL WATER, GAL/KW-HR	.850	.902	.956	.999	.948	.897	.951	.934
COOLING WATER	.776	.816	.814	.909	.860	.811	.809	.847
GASIFIER PROCESS H2O	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
CONDENSATE MAKE UP ,	.01032	.01032	.01028	.01035	.01034	.01033	.01029	.01033
WASTE HANDLING SLURRY	.0272	.0278	.0932	.0295	.0286	.0279	.0929	.0294
SCRUBBER WASTE WATER	.04666	.04706	.04926	.05069	.04922	.04772	.04912	.04876
NOX SUPPRESSION	.00000	.00000	.00000	.00000	.00000	.00000	.00000	.00000
TOTAL LAND ACRES/100MWE	95.28	88.27	127.94	106.00	101.02	99.96	139.48	100.70
MAIN PLANT	28.80	28.80	28.71	28.89	28.85	28.83	28.73	28.85
DISPOSAL LAND	31.31	32.11	71.96	33.99	33.01	32.02	71.76	32.71
LAND FOR ACCESS RR	35.17	27.35	27.27	43.12	39.15	39.11	38.99	39.14

2.8 References

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