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Design Study of a 200 MW(e) Alkali Metal/Steam Binary Power Plant Using a Coal-Fired Fluidized Bed Furnace

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MASTER

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Engineering Technology Division

DESIGN STUDY OF A 200 MW(e) ALKALI METAL/STEAM BINARY POWER PLANT
USING A COAL-FIRED FLUIDIZED BED FURNACE

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ABSTRACT

This report describes the results of a study of 200 MW(e) alkali metal/steam binary power plant using a coal-fired fluidized bed furnace. Both cesium and potassium were evaluated for the topping cycle working fluid and cesium was selected. The fuel used was Illinois No. 6 coal, and limestone was used as the bed sorbent material. For the reference design, the furnace operated at atmospheric pressure and the cycle conditions for the power conversion system were 1500°F (1089°K) to 900°F (756°K) for the topping cycle and 2400 psi (16.5 MPa)/1000°F (811°K)/1000°F (811°K) to 1 1/2 in. Hg (5079 Pa) for the steam system. Several variations of the plant were briefly evaluated. These variations included using a supercritical steam system and using a pressurized furnace.

The principal conclusions of the study are as follows:

1. A satisfactory design of an atmospheric pressure fluidized bed furnace binary power plant was evolved. The concept uses a variation of the conventional binary cycle which permits utilizing the full potential of the alkali metal topping cycle.
2. The net plant efficiency (coal to busbar) of the reference system was 44.6%.
3. The net plant efficiency of a larger system with a 3500 psi (24.1 MPa)/1000°F (811°K)/1000°F (811°K) steam system was 46.8%.
4. An intermediate pressure turbocharged system with a furnace pressure of 4 atm (0.4 MPa) would have many advantages in comparison to the atmospheric pressure system, including 1) a plant efficiency about one percentage point higher than the reference design, 2) reduced limestone requirement and 3) potential capital cost saving.
5. Although cost estimates were not a part of the design study, a comparison of the design of this study to that of the Energy Conversion Alternative Study (ECAS) indicates plant costs 20 to 25% less than that of the final ECAS design.

1. SUMMARY

This study describes a 200 MW(e) power plant that superimposes an alkali metal (cesium) vapor Rankine topping cycle on a conventional steam cycle system. The cesium system operates with an average boiler temperature of 1535°F (1108°K), a turbine inlet temperature of 1500°F (1089°K) and a condenser temperature of 900°F (756°K). The steam cycle is a conventional 2400 psi (16.5 MPa) system with superheat and reheat of 1000°F (811°K) and condenser pressure of 1 1/2 in. Hg (5079 Pa). The objective of the alkali metal/steam binary system is to improve the efficiency of the plant. The system described here has a net plant efficiency of 44.6%, well above the 36% that is projected for the best conventional steam plants that can use high sulfur coal and meet EPA emission regulations.

1.1. Selection of the Topping Cycle Fluid

Cesium was selected as the topping cycle working fluid. Both potassium and cesium were analyzed throughout the design study and cesium was selected primarily because of its advantages in regard to the turbine design. For turbines of similar size and performance characteristics, the cesium turbine will operate at one-half the speed of the potassium turbine which reduces the blade and disk stresses and permits higher temperature operation. Also the high vapor pressure and density of the cesium at the lower temperatures in the turbine requires shorter blades in the low temperature stages and permits a lower condenser temperature.

The major disadvantages of cesium are its relative availability, cost, and lack of extensive materials compatibility data. The world's proven reserves of high grade cesium ore are about 100,000 tons (9.1×10^7 kg) with probable reserves of several hundred thousand tons. The proven reserves are thousands of times greater than current annual requirements and thus there is little incentive to find additional reserves or sources of cesium. The magnitude of these reserves in regard to large power systems can be put in perspective by the following example. Let us assume that the required inventory for a 1000 MW(e) plant is 850,000 lb (386,000 kg) [5 times the inventory required for the 200 MW(e) plant

of this study], and that five 1000 MW(e) plants are built each year for 20 years. The total inventory required would be 42,500 tons (38.5×10^6 kg) or about 40% of the world's proven reserves and perhaps 10% of probable reserves. A more serious disadvantage may be the lack of proven significant U.S. reserves.

Based on an extrapolation of rather limited data, the cost of cesium for production quantities of a few million pounds per year would be less than \$5/lb (\$11/kg), as compared to potassium costs of about \$1.7/lb (\$3.75/kg). Much more study of the availability and cost is needed.

Although there is considerable evidence that cesium is compatible with structural materials of interest, the amount of information available is quite small and additional tests are required.

The lower melting temperature and the higher vapor pressure of cesium are advantages for startup and for major maintenance operations. Also, the higher vapor pressure of cesium at normal operating temperatures is an advantage for a pressurized furnace which appears to be a better system than an atmospheric unit.

Although cesium was selected as the fluid for the final design for this study, this choice is somewhat speculative and more information is needed. If the alkali metal system is to receive serious consideration as a power system of the future, additional studies on production cost, availability and location of metal sources, and materials compatibility are needed for cesium. Also, a much more detailed study of turbine design and fabrication methods is needed.

1.2. System Description

1.2.1 Reference Design

The cycle arrangement selected for this study is shown in Figs. 1.1 and 1.2. Figure 1.1 shows the flow diagram for the combustion air and flue gas system which includes the coal and limestone drying system, the ash cooling system and the coal and limestone injection system. The furnace for the plant is an atmospheric pressure fluidized bed unit and operates at 1750°F (1228°K). Of the heat transferred from the furnace to

Fig. 1.1. Schematic flow diagram of the combustion air and flue gas systems.

Fig. 1.2. Schematic flow diagram of the power conversion system.

the power conversion systems, 71% is transferred to the cesium system through boiler tube bundles located in the fluidized bed. The remaining 29% is transferred from the flue gases to a steam superheater and reheater located in the upper part of the furnace. The flue gas temperature is reduced from 1750°F (1228°K) to 900°F (756°K) in passing through the steam superheater and reheater. The main combustion air stream enters the furnace at 800°F (700°K).

A schematic flow diagram of the power conversion system is shown in Fig. 1.2. The cycle arrangement used in this study differs from a conventional binary system in that prime heat, the heat from the combustion system, is added to both the cesium and the steam cycle. Normally, the heat flow through a binary cycle is such that all of the highest temperature energy passes first through the topping cycle and only the heat rejected from the topping cycle passes through the secondary cycle.

In this study, heat is added to the secondary (steam) cycle by that rejected from the topping cycle and also from the furnace. The addition of the higher temperature energy from the furnace increases the steam temperature to that of conventional steam plants [1000°F (811°K)] while allowing the topping cycle condenser to operate at 900°F (756°K). This feature results in an overlapping of the two temperature-entropy diagrams and improves the overall plant efficiency.

There are other advantages of the arrangement shown in Fig. 1.2 that are more significant than the improvement in plant efficiency. If all of the furnace energy were transferred to the topping cycle, it would be necessary to use the energy in the flue gases from the furnace to preheat the combustion air. This would require a high temperature heat exchanger with a very high effectiveness; i.e., a very expensive heat exchanger. The combustion air temperature to the furnace would be on the order of 1400 to 1500°F (1033 to 1089°K) which would make the design of the furnace far more difficult than with the 800°F (700°K) of the reference design.

Another advantage of the arrangement of Fig. 1.2 is that only 71% of the heat transferred to the power conversion system passes through the topping cycle. This reduces the size of the entire high temperature part

of the plant; i.e., the size of the boiler, piping, pumps, turbine, and the condenser.

In the conventional binary cycle it is important to minimize the temperature difference between the condensing alkali metal and the steam exit conditions. For most studies this is usually taken as 100°F (56°K). With the system of this study this is not necessary and a larger temperature difference was used. In this case the difference was about 135°F (75°K) which will reduce the heat transfer surface required for the steam generator.

The reference design of the topping system incorporates two independent power loops. Each loop receives heat from one side of the fluidized bed furnace and supplies heat to a double-flow turbine. The vapor flow from each end of the double-flow turbines passes directly to a condenser-steam generator. The two condensers in each power loop are interconnected; however, the two loops are not. The steam lines from all four condenser-steam generators are connected to form a single steam system for the plant.

The steam system for the plant is an adaptation of an existing design of a 2400 psi (16.5 MPa) system with superheat and reheat of 1000°F (811°K) and a condenser pressure of 1 1/2 in. Hg (5079 Pa).

1.2.2. Turbocharged System

During the initial phase of this study an intermediate pressure [3 to 6 atm (0.3 to 0.6 MPa)] and a high pressure [6 to 10 atm (0.6 to 1.0 MPa)] fluidized bed system were briefly studied. Although these systems were not analyzed in the same detail as the atmospheric pressure system, the intermediate pressure concept appears to be very attractive. This system would use a turbocharger to supply combustion air to the furnace with the turbine driven by flue gases at a temperature of about 650°F (617°K). The turbocharger would be used only to pressurize the furnace and would produce no electric power (see Fig. 1.3). Experience with gas turbines operating on gases containing relatively high particulate loads have shown very little blade damage with inlet temperatures below 1000°F

Fig. 1.3. Schematic flow diagram of the combustion air and flue gas system using a pressurized furnace.

(811°K). Currently available inertial type separators would be adequate for reducing the particulate load to an acceptable level.

Pressurized systems have several advantages as compared to atmospheric units. For the alkali metal topping cycle to be worth the added complexity and cost and to be an attractive addition to a conventional steam plant, it must operate at a peak temperature in the range of 1500°F (1089°K). This high working fluid temperature requires a bed operating temperature in the range of 1700 to 1800°F (1200 to 1256°K) — a temperature which is well above that for good sulfur retention for atmospheric pressure operation. However, with pressurized beds, the sulfur retention capability of limestone improves with temperature, and for temperatures of 1700 to 1800°F (1200 to 1256°K) a Ca/S ratio of 2.5 or less would be satisfactory.² This compares to a value of 4.0 used in this study for the atmospheric pressure system.

The lower limestone requirement results in a number of improvements to the system. In addition to reducing the operating cost for limestone and its disposal, it also reduces the size of the preparation and handling plant and the amount of energy required for this operation. It reduces the furnace energy loss due to calcination of the limestone, the amount of CO₂ formed from the calcination of the limestone (which in turn reduces the amount of flue gases and thus the plant stack gas losses), the amount of water that must be removed from the limestone, and the amount of solids rejected from the plant and thus the energy loss due to the heat capacity of these solids.

The fluidized bed operating pressure, in addition to its effect on sulfur retention, has an important effect on the design and cost of the furnace boiler. Higher furnace pressures permit a greater power output per unit area of bed surface and also a much deeper bed. The power output of a bed is limited by two factors; the rate at which coal can be burned, and the rate at which heat can be removed from the bed. The combustion is limited by the weight flow of air that can be fed to the unit. One interesting aspect of fluidized beds is that their fluidization characteristics are primarily dependent on the gas velocity through the bed and are rather insensitive to the bed pressure. Thus, the weight flow of air and the combustion rate can be increased in direct proportion

to the pressure with constant gas velocity and little change in fluidization characteristics.

The ability to use a deeper bed results from the fact that the pressure loss penalty or pumping power required to operate the system depends on the system $\Delta P/P$ (actually the summation of the ratio of the pressure drop to the pressure throughout the system) and not just the ΔP (pressure drop). The ΔP across the bed is equal to the weight per unit area of the bed and is therefore proportional to the bed depth. Also, to insure uniform flow through the bed, the pressure drop through the air distribution plates must be some fraction of (and proportional to) the pressure drop through the bed. Thus, the two largest pressure losses in the system are directly proportional to the bed depth. Both the bed depth and the combustion rate can be increased in proportion to the furnace pressure.

The turbocharged concept of this study takes advantages of these added design options by using a vertical tube natural circulation boiler system. This type of furnace is much smaller and the natural circulation boiler is much simpler than the atmospheric pressure furnace which requires a horizontal tube forced circulation boiler.

An additional advantage of the turbocharged furnace is that during normal operation the furnace pressure is slightly greater than the boiler pressure. If a small leak should develop in the boiler, the leak would be into the cesium system and could be readily detected. A small leak in the reverse direction would be very difficult to detect.

In addition to these design advantages, the turbocharged system shows a small improvement in plant efficiency. This results from the use of the turbocharger replacing the forced draft fans and reducing the plant auxiliary power requirements.

A higher pressure system (6-10 atm) using a high temperature gas turbine was given brief attention early in the project and appeared to offer higher efficiency than the turbocharged system. However, the additional problems associated with the hot gas cleanup system, creep buckling in the boiler tubes and other added complexities made this system less attractive.

1.3. Plant Performance

The overall plant power output and efficiency for the atmospheric pressure reference design are given in Table 1.1. Of the total gross electrical output of 209.18 MW, 31.2% is produced by the cesium system and 68.8% by the steam system. The auxiliary power losses for the plant, which are shown in Table 1.2, total 8.58 MW. The overall plant net efficiency is 44.58%.

Table 1.1. Overall Plant Power Output and Efficiency

Thermal Power, MW	450.00
Cesium System, MW	281.90
Steam System, MW	327.10
Gross Electrical, MW	209.18
Cesium System, MW	65.26
Steam System, MW	143.92
Auxiliary Power Requirements, MW	8.58
Net Electrical, MW	200.60
Gross Plant Efficiency, %	46.48
Net Plant Efficiency, %	44.58

Table 1.2. Plant Auxiliary Power Requirements

Equipment Item	Power Requirement kW
Forced Draft Fans	2640
Transport Air Fans	12
Ash Cooling System Fans	195
Coal and Limestone Handling and Drying	1390
Induced Draft Fans	1540
Cesium Boiler Recirculating Pumps	310
Plant Cooling System	420
Hotel Load	1570
Miscellaneous	500
	8577

Several variations of the reference system were also investigated. The plant performance given in the previous sections is based on a 200 MW(e) plant which is typical of a demonstration size unit rather than a conventional full size plant. A larger plant would permit the use of higher steam pressure [3500 psi (24.1 MPa)] rather than the 2400 psi (16.5 MPa) used in the study. This higher pressure would give a better overall plant efficiency.

Although no design work was done for a larger plant, it would probably be composed of modules similar to that selected for this study. However, the steam system would be combined into a larger, more conventional unit typical of the large steam plants being built today. Thus, the only basic difference in the larger plants would be the different conditions in the steam cycle.

In order to compare the efficiency of a 3500 psi (24.1 MPa) plant to a 2400 psi (16.5 MPa) unit, let us assume that all other plant conditions remain the same; i.e., assume that a 3500 psia (24.1 MPa) system is used with the 200 MW(e) plant. The only effect then is to increase the cycle efficiency of the steam system from 44.0% of this study to over 47% which is typical of the higher pressure system. For this comparison a cycle efficiency of 47.0% will be used. The results of this analysis are shown in Table 1.3. As can be seen in this table, the output of the steam system increases by almost 7% and the net plant efficiency increases by over 2 points in comparison to the reference design presented in Table 1.1.

Table 1.3. Overall Plant Power Output and Efficiency
with 3500/1000/1000 Steam Conditions

Thermal Power, MW	450.00
Cesium System, MW	281.90
Steam System, MW	327.10
Gross Electrical, MW	218.99
Cesium System, MW	65.26
Steam System, MW	153.73
Auxiliary Power Requirements, MW	8.58
Net Electrical, MW	210.41
Gross Plant Efficiency, %	48.66
Net Plant Efficiency, %	46.76

The efficiencies of two other cycle variations were also calculated. For the intermediate pressure system using a turbocharger to pressurize the furnace, the system efficiency was found to be 45.6 (one point greater than that of the reference design). One other variation was to reduce the cesium inlet temperatures by 100°F (56°K) to 1400°F (1033°K). This reduced the efficiency of the reference system by 0.8 points to 43.8%.

1.4. Comparison with the Results of the Energy Conversion Alternatives Study

The results of this study were compared to the results of the Energy Conversion Alternatives Study (ECAS). The ECAS is one of at least twelve studies undertaken since 1970 to compare the merits of advanced power conversion systems. The ECAS was the most detailed of these studies and indicated that the alkali metal system was a rather poor candidate.

There are several basic and significant differences between the reference design described in this report and the system used in the ECAS evaluation.

1. The ECAS system used a high pressure [10 atm (1.0 MPa)] furnace pressurized by a gas turbine operated with an inlet temperature of 1700°F (1200°K). Power was extracted from the gas turbine as well as from the alkali metal and steam system. The costs associated with this pressurizing system, including the gas turbine, gas cleanup system and the turbine exhaust feedwater heaters, were over 22% of the total plant cost. None of these components are required for this system.

2. The ECAS also limited the alkali metal turbine inlet temperature to 1400°F (1033°K) rather than the 1500°F (1089°K) of this study.

3. The overlapping temperatures used in this study permitted the alkali metal cycle to operate over the range of 1500°F (1089°K) to 900°F (756°K) while maintaining steam conditions of 1000°F (811°K). The ECAS alkali metal cycle operated only from 1400°F (1033°K) to 1100°F (867°K).

4. The ECAS system used potassium as the topping cycle fluid rather than cesium. This choice led to a larger turbine and would also limit the condensing temperature of the alkali metal system.

The ECAS plant output was approximately 1000 MW(e) and the steam conditions were 3500 psi/1000°F/1000°F (24.1 MPa/811°K/811°K).

A comparison of the efficiencies from the ECAS and from this study is given in Table 1.4. The General Electric results for phase 1 are too low because of high stack gas losses and excessive pumping power used for the boiler recirculating pumps. These were corrected in the phase 2 part of the study. Westinghouse evaluated the alkali metal system only in the preliminary or phase 1 part of the study. As can be seen in Table 1.4, for comparable steam conditions, the efficiency of the reference design of this study exceeds that of the final phase 2 G.E. results by 2 or 3 points.

Table 1.4. Comparison of ECAS Efficiencies to Those of This Study

	Net Plant Efficiency, %
ECAS Phase 1	
G.E.	39.6
Westinghouse	42.4*
ECAS Phase 2	
G.E.	44.4
This Study	
Atmospheric Furnace and	
2400 psi/1000°F/1000°F steam conditions	44.6
3500 psi/1000°F/1000°F steam conditions	46.8
Turbocharged Furnace and	
2400 psi/1000°F/1000°F steam conditions	45.6
3500 psi/1000°F/1000°F steam conditions	47.8

*This value was for potassium. Westinghouse gave a value of 42.9% for cesium.

Although plant cost estimates were not made for this study, a comparison of the system of this study to the ECAS system indicates a potential plant cost of about 25% less than that of the ECAS system. The main cost savings stem from eliminating the high temperature gas turbine, the gas cleanup system, and the turbine exhaust feed water heaters, and from the smaller alkali metal turbine. There are other areas with significant differences; however, these appear to offset one another. This comparison,

which is admittedly very rough, does indicate that the cost for the plant described in this study is lower than the cost estimated by the ECAS to be necessary for the system to be attractive.

2. INTRODUCTION AND SCOPE

The current fuel shortage is expected to get worse during the coming years and fuel prices are rising at a distressing rate. Furthermore, the rapid increase in the size of the electric utility industry is leading to increasing difficulties arising from the rejection of waste heat to the environment. Thus there is a strong incentive to increase the thermal efficiency of central station power plants which implies increasing the peak temperature in the thermodynamic power cycle. The thermal efficiency of central stations was increased during the period 1890-1950 by roughly a factor of ten, thus drastically cutting fuel requirements. This was accomplished by increasing the steam temperature, but since 1950 there has been essentially no further increase in this parameter due to an inherent upper temperature limit of about 1050°F (839°K) imposed on the steam cycle by basic materials characteristics. Iron-chrome-nickel alloys lose strength rapidly as one increases their temperature above about 1000°F (811°K), making it difficult to contain the hundreds of atmospheres of pressure inherent in the use of steam at high temperatures. Further, water begins to dissociate at around 1100°F (867°K), the hydrogen diffuses through the boiler tube walls, and the oxygen attacks the steel surfaces so that waterside corrosion becomes a serious problem for metal temperatures above about 1200°F (922°K) in low alloy steels and above about 1350°F (1006°K) in high alloy steels.^{1,2}

The approach discussed in this study entails raising the peak temperature of the thermodynamic cycle by employing an alkali metal vapor Rankine topping cycle that operates with a peak pressure of only a few atmospheres with a turbine inlet temperature of ~1500°F (1089°K) and condenser temperature of about 900°F (756°K). This topping cycle is superimposed on a conventional steam cycle with a turbine inlet temperature of ~1000°F (811°K) by transferring the waste heat rejected from the condensing alkali metal vapor to a steam generator. The combined cycle is similar to the mercury-steam binary vapor cycle employed in seven plants in the United States between 1925 and 1948. However, the operating temperature of these plants was limited to 900°F (756°K) because at higher temperatures severe corrosion of all types of steel by mercury made operation impractical.

The 900°F (756°K) mercury turbine inlet temperature was higher than the steam turbine inlet temperature of the 1920's, but by the latter 1930's, with the development of new materials, it was surpassed in new steam plants whose higher temperatures and higher thermal efficiencies made the mercury vapor system obsolete.³

Work on the potassium vapor cycle began in 1958 with a study at ORNL in which various systems suited to space power plants were examined critically and the potassium vapor cycle was found to be the most promising candidate.⁴ Work was then initiated at ORNL on what were considered to be the two most crucial questions; namely, the compatibility of boiling potassium with stainless steel in the temperature range of 1500 to 1600°F (1089 to 1144°K) and the heat transfer coefficient for boiling potassium. By 1961 both questions had been resolved favorably and a program to develop a potassium vapor cycle space power plant was initiated. Concurrently, a study was made of the possibility of coupling a potassium vapor cycle to a molten salt reactor and it was found that this should yield an overall thermal efficiency of the order of 54%.³ Subsequently, some 200,000 hr of boiling potassium system operation at ORNL and ~70,000 hr at G.E.-Evendale under the space power plant program resolved some 40 other important questions related to the design of boiling potassium systems.⁵ Unfortunately, cutbacks in the space program terminated this work before a complete power plant was built. However, as work proceeded on the development of thermonuclear reactors,⁶ high-temperature gas-cooled reactors,⁷ and LMFBR's⁸ the results of the space power plant work were applied to the design of potassium vapor cycles in a series of advanced system studies carried out at ORNL and by various organizations in Europe (Refs. 7, 9, 10, 11). These studies have indicated that thermal efficiencies of ~50% can be obtained with reasonable capital costs.

Studies of alkali metal vapor cycles for fossil fuel plants were not made prior to 1970 because of the serious corrosion problems posed by the relatively high sulfur content of the coal or residual fuel oil employed in most central stations. However, it became evident in 1970 that the impending fuel shortage coupled with restrictions on sulfur emissions provided new incentives to employ a potassium vapor topping cycle for central stations. The requirement for low sulfur emissions favored either

coal gasification to remove the sulfur or a combustion process in which the sulfur would be absorbed by a more active agent than iron-chromium-nickel alloys, and hence sulfur corrosion would be a much less serious problem than in conventional oil- or coal-fired plants. Design studies showed that a potassium vapor cycle is indeed attractive and that one of the principal uncertainties is associated with the potassium boiler.¹²⁻¹⁶ NSF support was obtained, and work was initiated at ORNL in 1972 on the design, construction, and test of a large potassium boiler tube bundle module and gas burner assembly. The boiler tube bundle and the burner were built, but testing was delayed for about a year because of funding delays stemming from shifts in the responsibility of various agencies associated with the creation of ERDA. These questions were resolved and funding was received from ERDA. The initial tests of the unit with water have been successfully completed and installation of the additional equipment for potassium testing is currently (September 1977) being completed. Initial tests with potassium are expected to start in October 1977.

Early in the NSF program it became evident that major savings in both fuel and capital costs could be effected by employing a fluidized bed coal combustion system rather than gas or oil as the fuel.¹⁴ Both the wisdom and the feasibility of this course of action have become increasingly apparent, and the current ORNL alkali metal study program is primarily concerned with this approach.

The objective of this study is to establish the reference design of a 200 MW(e) alkali metal vapor/steam Rankine cycle system employing a coal-fired fluidized bed furnace, to study the relative merits of cesium and potassium as the topping cycle working fluid, and to prepare a preliminary design of a fluidized bed furnace.

The initial effort on this study entailed parametric cycle analyses of various plant arrangements with both cesium and potassium as the working fluid and for three fluidized bed pressures — atmospheric pressure, intermediate pressure [3 to 5 atm (0.3 to 0.5 MPa)], and high pressure [6 to 10 atm (0.6 to 1.0 MPa)]. This effort included the effects of plant operating conditions on sulfur retention, combustion efficiency, and heat transfer coefficients as well as conceptual designs of the alkali-metal-fluidized bed furnace.¹⁷ Budgetary limitation required a curtailment of

the scope of the study, and a decision was made to concentrate the effort on an atmospheric pressure furnace. Although the atmospheric system does not appear to be as well suited to the alkali metal boiler as a turbo-charged system, it is the system receiving the most attention at the present time and is the reference system of this phase of the study.

The primary effort of the study was directed toward the fluidized bed furnace-alkali metal boiler and the alkali metal system. The effort on other parts of the plant was limited to that necessary to support the design of these systems. The major tasks addressed in this study are:

1. Parametric cycle analyses of the cesium and potassium power systems and of the overall plant.
2. Alkali metal vapor turbine analytical design and analysis.
3. Alkali metal condenser-steam generator conceptual designs for cesium and potassium.
4. Conceptual design of various fluidized bed furnace-alkali metal boiler arrangements.
5. Conceptual layouts of the overall system.
6. Selection of and preliminary design of an atmospheric pressure furnace-boiler system. This includes the coal and limestone feed system and the ash system.

The steam cycle selected for the plant was an existing design of a 2400 psi (16.5 MPa) system with superheat and reheat of 1000°F (811°K) and a condenser pressure of 1 1/2 in. Hg (5079 Pa).

Because of the limited scope of the study no attempt was made to estimate the capital or operating cost for the plant.

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3. SELECTION OF THE TOPPING CYCLE FLUID

The alkali metals have been studied as working fluids in Rankine cycle systems for ~20 years. Work in the late 1950's was related to small space power plants using potassium, rubidium, and cesium as the working fluids. For these plants, potassium appeared to be equal to or better than the other alkali metals, more readily available, and much less expensive. Thus, potassium was used almost exclusively in the space power plant program with cesium receiving a small amount of attention.

The relative advantages of the fluids for a large central power station using a binary vapor cycle are different from those for a small power system. In this study both potassium and cesium were used in the analysis of the system and both have advantages and disadvantages relative to each other. Although cesium was selected as the fluid for the final design for this study, this choice is somewhat speculative and more information is needed. If the alkali metal system is to receive serious consideration as a power system of the future, additional studies on production cost, availability and location of metal sources, and materials compatibility are needed for cesium. Also, a much more detailed study of turbine design and fabrication methods is needed.

The major advantages of cesium are rather subtle, being related to the turbine design and also to its lower melting point and much higher vapor pressure at low temperature. Potassium has the more obvious advantages of greater availability, extensive U.S. reserves, a lower cost and a higher latent heat of vaporization which reduces the amount of pumping work required in the boiler. The higher latent heat of vaporization also reduces the required size of the liquid lines and thus the total volume of working fluid in the system. There is little difference in the size of the vapor lines required for the two materials.

3.1. Effect of Working Fluid Properties on Turbine Design and Cycle Efficiency

The comparison of the turbine designs and efficiencies for cesium and potassium are based on the decision to make the units nearly identical

in size and number of stages. With cesium it is possible to take advantage of the smaller enthalpy change across the turbine to reduce the number of stages as compared to potassium. Although this would result in a somewhat less expensive cesium unit it would also reduce the efficiency of the unit by several points. Rather than take this approach, the speed of the cesium unit was made one-half that of the potassium unit which gives machines of nearly the same size and number of stages. This lower speed has a very significant effect on the materials requirements for the blades and disks and is discussed later.

In analyzing the turbines for potassium and cesium for the various cycle conditions it was assumed that the only difference in the turbine efficiency for the various cases was that associated with supersaturation, moisture, and exhaust losses. The moisture losses were assumed to be 0.6% for each percent of moisture and the exhaust loss was assumed to be the kinetic energy in the fluid from the last stage. The supersaturation losses were calculated as a function of the enthalpy drop per stage and were found to range between 2% for a stage enthalpy change of 11 Btu/lb (25.6×10^3 J/kg) to 4% from an enthalpy drop of 15 Btu/lb (34.9×10^3 J/kg) for cesium. Comparable results for potassium show a loss of 2% for a stage enthalpy drop of 35 Btu/lb (81.4×10^3 J/kg) and 4% for an enthalpy drop of 60 Btu/lb (140×10^3 J/kg). These results are based on the assumption that the return to equilibrium occurs along a constant pressure line which is probably somewhat conservative.

In carrying out both the turbine and the cycle analyses, it was assumed that vapor, for regenerative feed heating, would be bled from the turbine between all stages except the first and second. Furthermore it was assumed that moisture removal devices were located at each bleed point. The efficiency of the moisture removal was taken as 25%, i.e., 25% of the moisture in the working fluid at each bleed point is removed with the bleed flow. This value is well within the 20 to 35% efficiency demonstrated on a small potassium turbine tested by General Electric for the space power program.¹

Both the turbine and cycle analyses were carried out at turbine inlet temperatures from 1850°R (1028°K) to 2000°R (1111°K) and condenser temperatures from 1350°R (750°K) to 1550°R (861°K). The number of turbine

stages varies from 3 to 5 and the number of feed heaters used in the system was always 2 less than the number of turbine stages. The feed pump energy requirements were based on a pump efficiency of 70% and a head equal to twice the difference between the boiler and condenser pressures. The cycle efficiency includes no allowances for mechanical or electrical losses in the generator. The results of the analyses are summarized in Tables 3.1 and 3.2. The efficiency of the alkali metal cycles varied from 13% to 26% or from 77% to 80% of the Carnot efficiency. The turbine efficiency varied between 77% and 82%. These turbine efficiencies were calculated by subtracting the sum of the superheating losses, moisture losses and exhaust losses from a basic machine efficiency of 88%.

Table 3.2 gives information on the turbine design used in the analysis. For all cases the turbine hub radius was fixed at 20 in. (0.51 m). The most interesting column is that showing the height of the first stage blade for a power output of 2 MW for the stage. The blade heights are the same for a cesium unit operating at 1800 rpm as for a potassium unit operating at 3600 rpm. This means that the centrifugal stress on the first stage rotor blade, which is proportional to the square of the speed, is four times as great for the potassium turbine as for the cesium unit. Furthermore, as can be seen from Table 3.1, the required flow area and thus the blade height increases much more rapidly for the potassium unit than for the cesium unit with the last stage requiring about 1.5 times the area. As the centrifugal blade stress is directly proportional to the rotor annulus area, the centrifugal blade stress in the last stage rotor will be about six times greater for the potassium turbine.

The differences in stress have a significant effect on the blade and rotor creep rates and allowable operating temperatures. The relation between creep rate, stress and temperature is shown in the following equation.

$$\epsilon = A \sigma^m e^{-h/kT}$$

where

ϵ = creep rate

A = constant

Table 3.1. Summary of Alkali Metal Cycle Analysis

Turbine Inlet Temperature °R	Condenser Temperature °R	No. of Turbine Stages	No. of Feed Heaters	Cycle Efficiency %	% of Carnot Efficiency	Ratio of Turbine Exit Area to Inlet Area	Turbine Efficiency %
<u>Cesium</u>							
2000	1350	5	3	25.4	78.1	21.6	78.8
2000	1450	4	2	21.3	77.6	11.1	78.6
2000	1550	4	2	17.2	76.6	6.2	80.7
1900	1350	5	3	22.9	79.1	15.5	79.2
1900	1450	4	2	18.4	77.8	8.0	79.7
1900	1550	3	1	14.2	77.2	4.6	78.8
1850	1350	4	2	21.6	79.9	13.3	77.9
1850	1450	4	2	17.8	82.3	6.7	80.4
1850	1550	3	1	13.0	80.1	3.7	81.9
<u>Potassium</u>							
2000	1350	5	3	25.3	77.8	38.9	77.3
2000	1450	5	3	21.8	79.2	17.4	79.0
2000	1550	4	2	17.7	78.6	8.8	78.9
1900	1350	5	3	22.8	78.9	26.3	78.7
1900	1450	4	2	18.5	78.1	11.9	78.6
1900	1550	4	2	14.7	80.0	6.2	80.6
1850	1350	5	3	21.6	79.8	21.3	79.0
1850	1450	4	2	17.3	79.9	9.7	79.5
1850	1550	3	1	12.9	79.7	4.9	79.9

Table 3.2. Summary of Turbine Design Analysis

Working Fluid	Turbine Inlet Temperature °R	Condenser Temperature °R	Turbine Hub Radius in.	Turbine Speed rpm	First Stage Blade Height For 2MW Output in.	Rotor Inlet Temperature °R	Inlet Nozzle Mach No.
Cs	2000	1350	20	1800	3.8	1855	.80
Cs	1900	1450	20	1800	5.5	1780	.77
Cs	1850	1550	20	1800	7.0	1750	.75
K	2000	1350	20	3600	3.8	1850	.86
K	1900	1450	20	3600	5.5	1780	.84
K	1850	1550	20	3600	7.0	1740	.83

σ = stress

m = constant

h = activation energy

k = Boltzmann's constant

T = absolute temperature

The value of m is in the range of 4 to 8 while the value of the activation energy is normally between 80,000 and 100,000 for materials of interest. For illustration purposes a value $m = 6$ and $h = 80,000$ will be used. For these values, the creep rate for the first stage blades in the potassium unit is 4100 times that for the cesium unit. If one were to try to offset this higher creep rate with a lower temperature, then the blade temperature would have to be reduced from about 1850°F (1028°K) to 1550°R (861°K). However as the temperature is reduced, the flow area and thus the blade height would increase and the required temperature reduction would be even greater.

For the cesium turbine the centrifugal stress at the root of the first stage blade is about 1200 psi (8.3×10^6 Pa) for each MW of power output for a blade with a constant cross sectional area. A tapered blade would reduce this stress by about one-third.

Thus, in comparing the two fluids, the use of cesium gives the advantage of much lower turbine blade and disk stresses and also permits operation at a higher temperature than potassium. Also, as the flow area required for cesium is less in the lower temperature stages, these stages can be smaller than with potassium, or conversely, the exit temperature of the turbine and thus the lower cycle temperature can be reduced with cesium as compared to potassium. The problems of creep rate limitations in the high temperature stages and the large volumetric flow in the lower temperature stages can be avoided with potassium by the use of more turbines. However, this would increase the plant cost and add to the complexity of the system.

As can be seen in Table 3.1, there is very little, if any, difference in the cycle efficiency of cesium and potassium for equal turbine inlet and exit temperatures.

3.2. Availability and Costs

Potassium has the advantage of being relatively inexpensive [$\sim \$1.7/\text{lb}$ ($\$3.75/\text{kg}$)]^{2,3}, abundant, and with extensive U.S. reserves. In the Energy Conversion Alternative Studies (ECAS), General Electric used a cost of $\$1.71/\text{lb}$ ($\$3.77/\text{kg}$)³ while Westinghouse used $\$1.68/\text{lb}$ ($\$3.70/\text{kg}$)².

Cesium is a relatively rare element having a low abundance in the earth's crust. Fortunately, there are a few large and rich deposits of the metal. Most of the production and nearly all of the known reserves are located in Rhodesia, South West Africa, and Canada.⁴ Cesium is also concentrated in natural brines during crystallization of salts from sea water and two analyses of Salton Sea geothermal brine show 14 and 20 ppm cesium.⁴

Reference 4 also estimated U.S. and world production and consumption and states that

"The United States in 1964 used about 8 short tons of cesium salts and somewhat more than 3,000 pounds of cesium metal. In 1970, the United States imported 5,129 pounds of cesium compounds plus an unstated quantity of pollucite, and the U.S.S.R. imported 200 tons of pollucite containing about 40 tons of cesium. From these figures, the world's consumption now and in the immediate future may be gaged at only a few tons of cesium annually. Prices are high; the 5,129 pounds of cesium products imported in 1970 were valued at $\$189,056$, or about $\$37$ per pound.

The principal source has been the Bikita (Rhodesia) pegmatite, mainly from pollucite but also as a byproduct of the extraction of lithium from lepidolite, which resulted in an alkali carbonate residue containing about 2 percent cesium. South West Africa has been the order source in recent years, and Mozambique has also yielded cesium. Only about 100 tons of pollucite has been mined in the U.S., mostly from the Tin Mountain pegmatite, Black Hills, S. Dak., but also from Oxford County, Maine. All these sources are likely to be dwarfed by the Tanco (Canada) deposit during any expansion of the cesium market unless a now unanticipated source of cesium (for example, as a byproduct of brine processing) makes an appearance."

In regard to reserves and resources, Ref. 4 states

"Reserves of the Tanco deposit have most recently been placed at 293,863 tons of pollucite with 23.67 percent Cs_2O ,

or 66,000 tons of cesium. At Bikita the dimensions of the pollucite body imply about 25,000 tons of cesium. Reserves in South West Africa may also be sizable, but elsewhere in the world the known amount of pollucite is negligible. Nevertheless, the world's total reserves may be placed at very close to 100,000 tons of cesium, which is virtually infinite relative to the trivial annual consumption or to anticipated demand.

Estimates of conditional or hypothetical resources are not worthwhile in the absence of data and in the absence of any predictable need for a larger quantity of cesium. If one makes the plausible assumption that the ratio between cesium reserves and proved and probable reserves of lithium in pegmatites (about 1:7) will remain approximately the same as additional lithium deposits are developed, the world has resources of several hundred thousand tons of cesium."

Costs for producing high purity cesium metal by several processes were analyzed by Dean, et al.⁵ In this 1966 paper, cost estimates were based on a feed ratio of 3500 tons/yr (3.17×10^6 kg/yr) of pollucite ore containing 10% cesium. The cost of the ore was assumed to be \$20/ton (\$0.022/kg). Because of the small scale of the operation, the projected plants (for each of the four processes) were designed to operate only one shift per day and five days per week. The estimated cost for the cesium metal for the four processes was \$1.50, \$1.60, \$1.70, and \$3.80 per pound. Although this paper does not estimate costs for larger production rates, it is obvious that a greater utilization of plant facilities would reduce these estimates.

In the Phase I ECAS Study by Westinghouse,² the cost of cesium was estimated to be \$18/lb (\$39.7/kg) on the basis of a production rate of 100,000 lb/yr (45,000 kg/yr) — a production rate only 1/7 of that assumed in the paper by Dean. It is interesting to note that if one scales up the production rate of 100,000 lb/yr (45,000 kg/yr) to the 700,000 lb/yr (317,000 kg/yr) of the Dean study, with a scale factor of 0.6, the \$18/lb (\$39.7/kg) cost is reduced to \$5.60/lb (\$12.3/kg). Considering the approximately 100% escalation in cost from the time of the Dean paper to the ECAS Study, the difference in the cost estimates is not unreasonable.

One interesting aspect of cesium is that the ore contains about 5% rubidium, an alkali metal with properties very similar to those of cesium. The vapor pressure of rubidium is only 10% to 25% less than cesium over

the temperature range of 1540°F (1111°K) to 940°F (778°K) while the enthalpy change during an isentropic expansion between these two temperatures is only 50% greater for rubidium than for cesium. Thus, a mixture of the two metals containing 5% rubidium would be as satisfactory a working fluid as pure cesium, assuming no materials compatibility problems. Furthermore, in the production of high purity cesium, one of the more expensive steps (for at least one U.S. manufacturer) is the final separation of rubidium from the cesium. It is probable that this would not be necessary.

3.3. Compatibility with Structural Materials

Although there is general agreement or belief that there is little difference in the compatibility of cesium and potassium with high temperature materials, potassium does have the advantage of a few hundred thousand hr of operation at temperatures of 1500 to 1600°F (1089 to 1144°K) in stainless steel systems. An excellent review of this subject for boiling potassium is given by DeVan.⁶ The conclusion of Ref. 6 is listed below.

"1. The corrosion resistance of 300-stainless steels appears adequate for service in potassium recirculating-boiling systems to 840°C, which is the temperature limit set by mechanical property considerations. Nickel- and cobalt-base alloys could, on the basis of high-temperature strength, increase this limiting temperature to 900°C.

2. Forced-circulation loop tests of nickel-base alloys in sodium show higher mass transfer rates above 700°C than comparable austenitic stainless steel loops. However, corrosion rates of the nickel-base alloys appear less affected by oxygen contamination in the liquid metal.

3. Few data are available on the behavior of cobalt-base alloys in sodium or potassium. The maximum corrosion rate of Haynes alloy No. 25 in a boiling-potassium loop at 870°C approximated that of a similar type 316 stainless steel loop. The corrosion rate of Haynes alloy No. 25 in a boiling potassium at 980°C was not substantially higher than at 870°C.

4. The Rankine-cycle, boiling-potassium circuit is characterized by three basically different corrosion zones: (1) the recirculating boiler, (2) the condenser and condensate return lines, and (3) the vapor lines connecting to and including the turbine. In the first of these zones, the potassium tends to remain saturated to supersaturated with solute at all points. Hence this first zone is characterized by deposition rather than corrosion. In contrast, the second corrosion zone, which carries condensate, tends to be highly undersaturated and will be the principal region of dissolutive attack in the system. The third zone, since it is kept virtually free of liquid, will be the least important from a corrosion or deposition standpoint.

5. The degree to which corrosion occurs in a potassium Rankine-cycle power conversion system is strongly affected by the liquid recirculation. At high recirculation ratios, the subcooling introduced by the condensate return is relatively small, and the boiler circuit becomes effectively isothermal. Under such conditions, corrosion is controlled by the amount of solute carried in by the condensate return relative to that carried out by saturated vapor. Operating experience at an 80% recirculation rate indicates that the degree of supersaturation in the boiler liquid is not sufficient to cause significant solute deposition.

6. Since the major temperature drop in the Rankine cycle is taken in the vapor phase, the drop does not entail significant solute deposition. The extent of dissolutive corrosion in the cycle depends on the temperature attained by the condensate during its return to the boiler. At high boiler recirculation ratios, it is possible to return the condensate at temperatures well below the boiling temperature."

The data available for cesium are rather limited and were primarily obtained in tests at higher temperatures with refractory metals. Refluxing capsule experiments⁷ comparing the relative corrosiveness of sodium, potassium, rubidium and cesium in a Nb-1% Zr alloy capsule indicated little difference between these alkali metals operating at 2192°F (1473°K). Each showed excellent compatibility with Nb-1% Zr. Similar results were also obtained by Romano et al.⁸ in 6000 hr test of Nb-1% Zr alloy at 2102°F (1423°K).

An earlier survey paper on corrosion in liquid metal systems presented at the 1964 Geneva Conference⁹ presented some data on rubidium and cesium. The brief summary on these two metals is quoted below.

"The data available on corrosion of materials by rubidium and cesium are less extensive than the data on the other alkali metals and are primarily qualitative.

Rubidium corrosion has been studied up to 1100°C. In tantalum and Haynes 25 reflux capsules containing titanium-zirconium gettered rubidium, the tantalum capsules were attacked whereas inserted tabs of beryllium, vanadium, Nb-1% Zr, Mo-1/2% Ti and Haynes 25 resisted corrosion at 1100°C. Solubility experiments performed in tantalum crucibles were inconclusive. A Nb-1% Zr test conducted at 1000°C, in a forced convection boiling loop containing titanium-zirconium gettered rubidium, has not shown corrosion after 700 hours.

Cesium corrosion has been studied up to 1700°C. Below 1000°C, most of the common engineering alloys and the unalloyed refractory metals were moderately corrosion resistant; TD nickel was attacked at 1000°C. In a series of reflux capsule tests certain ceramics (Al_2O_3 , MgO , and BeO) were corrosion resistant up to 1100°C. Nb-1% Zr and Ta - 8% W - 2% Hf appeared to resist attack at 1370°C in non-gettered cesium; molybdenum alloys were attacked at this temperature and Nb-1% Zr showed intergranular penetration at 1000°C. W - 25% Re was satisfactory up to 1700°C in non-gettered cesium."

The most extensive survey on the effects of cesium on structural materials was by Klueth and Jansen.¹⁰ This 1967 report gives results for cesium corrosion of nickel - and iron - base alloys as well as for the refractory metals and alloys. Table 3.3 and 3.4, from Ref. 10, summarize the corrosion data for the nickel and iron based alloys. Reference 10 concludes that

"Although the corrosion data for cesium are nowhere so extensive as for potassium, our understanding of the behavior of cesium, in the writers' opinion, is at a stage where further simple screening tests can afford little new or useful information. Capsule tests, despite the shortcomings in chemical analysis, have verified that the corrosion properties of cesium, like those of sodium or potassium, are intimately associated with the effects of interstitial impurities and that, if these impurities are kept at low levels, corrosion studies of conventional and refractory metals give results that are qualitatively the same as for sodium and potassium."

Table 3.3. Corrosion of Nickel-Base Alloys in Cesium

Material	Temperature (°F)	Time (Hr)	Cesium Phase(s)	Results
Nickel	1112	100	Liquid and vapor	No attack
	1292	100	Liquid	No attack
	1832	100	Liquid	Intergranular attack
	1652	281	Vapor	Mixed bright and matte surface; reaction zone at surface ^a
"A" Nickel	500, 800, 1200	48	Vapor	No attack ^a
	1380	1000	Vapor	Surface grayed; no apparent attack
"L" Nickel	500, 800, 1200	48	Vapor	No attack ^a
Inconel X	500, 800, 1200	48	Vapor	No attack ^a
	1600	720	Liquid	Decarburization; no surface attack
Ni-Span C	1200	1000	Vapor	Loss of subsurface precipitates; no surface attack

^a Tests were carried out in a stainless steel container with several specimens being run simultaneously.

Table 3.4. Corrosion of Iron-Base Alloys in Cesium

Material	Temperature (°F)	Time (Hr)	Cesium Phase(s)	Results
Iron	1600	1000	Vapor	No attack
1020 steel	500, 800, 1200	48	Vapor	Tarnishing, probably due to oxygen in the cesium ^a
17-4 PH	750	500	Liquid and vapor	No attack
Fe-Cr-Al-Y alloy	1652	281	Vapor	Mild surface attack ^a
AM-350	750	500	Liquid and vapor	No attack
EN 53B stainless steel	1112	100	Liquid and vapor	No attack
Type 304 stainless steel	1600	1000	Vapor	No attack
	500, 800, 1200	48	Vapor	Tarnishing, probably due to oxygen in the cesium ^a
Type 310 stainless steel	1600	720	Liquid	Slight transfer of carbon from stainless steel to cesium
Type 321 stainless steel	1292	100	Liquid	Titanium leached from the stainless steel
Type 416 stainless steel	750	500	Liquid and vapor	No attack
Type 430 stainless steel	1652	281	Vapor	No attack ^a

^aTests were carried out in a stainless steel container with several specimens being run simultaneously.

It is interesting to note that one report¹¹ speculates that cesium may be less corrosive than potassium, especially when the nickel based alloys are involved.

Another item, although not strictly related to corrosion, that has received some attention is turbine bucket erosion with the alkali metals. An excellent survey of this subject is given in Ref. 12. Several of the conclusions and recommendations of this survey are quoted below.

Tests at ORNL of simulated turbine bucket specimens in potassium vapor systems have shown only trace amounts of erosion damage after a total of more than 23,000 hr with the test specimens impacted by jets of potassium vapor having velocities as high as 3000 ft/sec and nominal qualities as low as 82%. These tests have included both stainless steel at specimen temperatures of 1000 to 1500°F, and TZM at temperatures of 1300 to 1800°F. The initial tests with TZM showed severe damage, but this was thought to be the result of oxygen contamination. Subsequent tests have supported this conclusion. Thus, extensive tests indicate that, if the oxygen content of the potassium can be held below around 40 ppm, corrosion damage is minimal, and that micron-size droplets emerging from a nozzle and impinging on static specimens at jet velocities approaching 3000 ft/sec will not induce appreciable erosion.

Most of the erosion damage in a turbine is caused by relatively large droplets that dribble from the trailing edge of the stator and move at low absolute velocities into the rotor where the impact on the convex face of the rotor blades near the leading edge causes erosion. The bulk of the damage generally appears near the tips of the rotor blades because secondary flows in the stator lead to the accumulation of small liquid pools on the rear faces of the stator blades near the tips, and these pools tend to slough off in the form of rather large globs.

A review of the theoretically derived relations for the stress in a solid imposed by the impact of a high velocity liquid droplet indicates that the relevant physical properties of water, potassium, and cesium are similar, and that erosion should be somewhat less of a problem for cesium and potassium than for water. The meager test data available for erosion by wet potassium vapor appear to be consistent with this thesis, but more data are needed for a definitive conclusion.

Pending results of a detailed analysis following the above procedure and to avoid difficulties with erosion, the tip speeds in preliminary designs should be

kept below 1200 ft/sec in both the cesium and potassium reference turbines.

It should be noted that the tip speed of the last stage bucket for the reference design using cesium is only 810 ft/sec (247 m/s) and erosion damage would not be expected.

3.4. Other Considerations

There are other differences in the properties of cesium and potassium that influenced the selection of cesium. The melting point of cesium is 83°F (302°K) as compared to 146°F (337°K) for potassium. The lower melting point is advantageous during a start-up with a cold system as there will be less chance of the metal freezing and plugging part of the system. Trace heating of lines susceptible to freezing has been used in smaller experiments with potassium, however, this would be more difficult with large systems.

The higher vapor pressure and much higher vapor density of cesium at low temperatures may also be advantageous. For major maintenance operations, in which it is necessary to have the alkali metal system open for an extended period of time, it will probably be necessary to drain and evacuate the system. Although the system is designed so that the metal can be drained, there will be some material left on the walls and some pockets that will not drain. The time required to evacuate this material will depend on the vapor pressure and density of the material and to some extent on the latent heat of vaporization.

A comparison of the vapor pressure and density of cesium and potassium is given in Table 3.5. Over the temperature range of 100 to 500°F (311 to 533°K) the vapor pressure of cesium is from 50 to 9 times that of potassium and the vapor density of cesium is from 180 to 30 times that of potassium. Also, the heat of vaporization of cesium is only one-fourth that of potassium. This property can be important in that it determines the rate of cooling of the material being vaporized. The lower heat of vaporization means that the temperature of the material being vaporized will not be cooled as much. This in turn means that the temperature will

Table 3.5. Comparison of the Vapor Pressure and Vapor Density of Cesium and Potassium at Low Temperature

Temperature °F °K		Vapor Pressure				Vapor Density			
		Cesium		Potassium		Cesium		Potassium	
		psia	Pa	psia	Pa	lb/ft ³	kg/m ³	lb/ft ³	kg/m ³
100	311	8.0×10^{-8}	5.5×10^{-4}	1.6×10^{-9}	1.1×10^{-5}	1.8×10^{-9}	2.8×10^{-8}	1.0×10^{-11}	1.6×10^{-10}
200	367	6.5×10^{-6}	4.5×10^{-2}	2.4×10^{-7}	1.7×10^{-3}	1.2×10^{-7}	1.9×10^{-6}	1.3×10^{-9}	2.1×10^{-8}
300	422	1.6×10^{-4}	1.1	9.7×10^{-6}	6.7×10^{-2}	2.6×10^{-6}	4.2×10^{-5}	4.6×10^{-8}	7.4×10^{-7}
400	478	1.9×10^{-3}	13	1.6×10^{-4}	1.1	2.7×10^{-5}	4.3×10^{-4}	6.9×10^{-7}	1.1×10^{-5}
500	533	1.3×10^{-2}	90	1.5×10^{-3}	10	1.7×10^{-4}	2.7×10^{-3}	5.8×10^{-6}	9.3×10^{-5}

remain higher which will help maintain a higher vapor pressure and thus reduce the time required to evacuate the system.

The higher vapor pressure of cesium at operating temperatures of 1500 to 1600°F (1089 to 1144°K) may also be important for future systems. In the ECAS evaluations of the alkali metal system both General Electric and Westinghouse concluded that a pressurized fluidized furnace was better than an atmospheric unit. Also, during the initial part of this study, it was concluded that an intermediate pressure [3 to 6 atm (0.3 to 0.6 MPa)] fluidized bed system would have better efficiency, would probably cost less and would avoid several problems inherent in the atmospheric unit. The pressurized furnace will introduce different problems and one of these is creep-buckling of the large vapor header under an external pressure stress. This is at least partially alleviated by the higher internal pressure with a cesium system. For example, with a boiling temperature of 1540°F the vapor pressure of cesium is 57.7 psia (0.398 MPa) as compared to 29.6 psia (0.204 MPa) for potassium. Thus, cesium will reduce the buckling loads and for the intermediate pressure system will make the loads very small or even nonexistent.

3.5. Summary

Cesium was selected as the topping cycle working fluid for the reference design primarily because of its advantages in relation to the turbine design. For turbines of similar size and performance characteristics, the cesium turbine will operate at one-half the speed of the potassium turbine which reduces the blade and disk stresses and permits higher temperature operation. Also the high vapor pressure and density of the cesium at the lower temperatures in the turbine requires shorter blades in the low temperature stages and permits a lower condenser temperature.

The major disadvantages of cesium are its relative availability, cost, and lack of extensive materials compatibility data. The world's proven reserves of high grade cesium ore are about 100,000 tons (9.1×10^7 kg) with probable reserves of several hundred thousand tons. The proven reserves are thousands of times greater than current annual requirements

and thus there is little incentive to find additional reserves or sources of cesium. The magnitude of these reserves in regard to large power systems can be put in perspective by the following example. Let us assume that the required inventory for a 1000 MW(e) plant is 5 times the inventory required for the 200 MW(e) plant of this study or 850,000 lb (386,000 kg), and that five 1000 MW(e) plants are built each year for 20 years. The total inventory required would be 42,500 tons (38.5×10^6) or about 40% of the world's proven reserves and perhaps 10% of probable reserves. A more serious disadvantage may be the lack of proven significant U.S. reserves.

Based on the data of Refs. 2 and 5 the cost of cesium for production quantities of a few million pounds per year would be less than \$5/lb (\$11/kg). Much more study of the availability and cost are needed.

Although there is considerable evidence that cesium is compatible with structural materials of interest, the amount of information available is quite small and additional tests are required.

The lower melting temperature and the higher vapor pressure of cesium are advantages for startup and for major maintenance operations. Also, the higher vapor pressure of cesium at normal operating temperatures is an advantage for a pressurized furnace which appears to be a better system than an atmospheric unit.

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4. SYSTEM DESCRIPTION

This section describes the major systems of the plant and the ground rules and assumptions used in the study. The plant is a binary cycle system with power extracted from both the high temperature cesium topping cycle and the lower temperature steam cycle. The cycle arrangement is different from a conventional binary system in which only the heat rejected by the topping cycle is transferred to the secondary system. In this study the energy added to the secondary (steam) system includes that rejected from the cesium system and additional prime heat from the furnace to increase the steam temperature to that of conventional steam plants. This feature results in an overlapping of the cesium and steam temperature-entropy diagrams and improves the overall plant efficiency. Flow diagrams of the combustion air and the gas system and the power conversion system are shown in Figs. 4.1 and 4.2.

4.1. Design Assumptions and Data Sources

4.1.1. Coal

The plant design and analysis was based on the use of Illinois No. 6 coal as the fuel. This is the same coal being used in the Energy Conversion Alternatives (ECAS).¹ The analysis of this coal as reported in the ECAS study is as follows:

Proximate analysis, (as received) percent:

Moisture	13.0
Volatile	36.7
Fixed Carbon	40.7
Ash	9.6

Ultimate analysis (as received) percent:

Ash	9.6
Sulfur	3.9
Hydrogen	5.9
Carbon	59.6

Fig. 4.1. Schematic flow diagram of combustion air and flue gas system.

Fig. 4.2. Schematic flow diagram of the power conversion system.

Nitrogen	1.0
Oxygen	20.0

Higher heating value (as received)

Btu/lb	10,788
MJ/kg	25.09

Corresponding values for the ultimate analysis and the higher heating value of the coal on a dry basis is as follows:

Ultimate analysis (dry) percent:

Ash	11.0
Sulfur	4.5
Hydrogen	5.1
Carbon	68.5
Nitrogen	1.2
Oxygen	9.7

Higher Heating value (dry)

Btu/lb	12,400
MJ/kg	28.84

A detailed analysis of the ash shows that about one-half (46.6%) is SiO_2 and for analysis purposes, i.e., energy balances on the system, the ash was assumed to be all SiO_2 .

4.1.2. Limestone

Limestone was selected as the sorbent material for sulfur retention. The limestone was assumed to be composed of 95% CaCO_3 and 5% ash (SiO_2). This material was also assumed to contain 5% moisture when fed to the plant.

4.1.3. Bed Temperature

One of the most important plant design parameters is the operating temperature of the fluidized bed as it has a significant influence on

the efficiency and capital cost of the plant. The efficiency of the plant will depend on the temperature of the alkali metal vapor generated — a 100°F (56°K) change in boiler temperature results in about a one point change in overall plant efficiency over the temperature range of 1390 to 1540°F (1028 to 1111°K). Thus, for best plant efficiency the metal vapor temperature should be kept high. However, the cost of the furnace-boiler will depend primarily on the boiler tube material, the surface area required, and the general arrangement of the boiler. These costs will increase as the boiler temperature increases.

With an alkali metal boiling system, the upper bed temperature restriction of 1750 to 1800°F (1228 to 1256°K) places a rather tight limit on the temperature difference between the bed and the boiler which in turn limits the heat transfer rate to the boiler and becomes the controlling factor in determining the surface area requirements and cost of the furnace-boiler. Thus, for reasons of both efficiency and economy a fluidized bed temperature of about 1750°F (1228°K) was selected. A temperature of 1800°F (1256°K) would be better but there is little operating experience at temperatures above 1750°F (1228°K) and even 1750°F (1228°K) is beyond the normal operating temperature of atmospheric pressure beds. Atmospheric pressure systems normally operate in a temperature range of 1500 to 1550°F (1089 to 1117°K) where the sulfur retention capability of the calcium oxide is near its maximum. Higher temperature beds do have good sulfur retention characteristics when operated under pressurized conditions where the sulfur retention properties of the calcium oxide improves with higher temperatures.

4.1.4 Ca/S Ratio

This combination of atmospheric pressure and high temperature has not been adequately investigated, but it is apparent that the amount of limestone required will be significantly greater, by a factor of about 1.5 to 2.0, than that required at 1500 to 1550°F (1089 to 1117°K). The calcium-to-sulfur ratio required to retain sufficient sulfur to meet the EPA emissions standard of 1.2 lb (0.54 kg) of SO₂ per million Btu (1055 MJ) was estimated to be 4.0. This was based on an extrapolation of lower

temperature data. There are possible ways in which to improve the calcium utilization in the system.

Workers at Westinghouse Research Laboratories^{2,3} attribute this decrease in sulfur retention at high temperature and atmospheric pressure to too rapid a calcination of the stone. Their laboratory tests indicate that when calcination takes place in nitrogen, which permits fast rates of calcination, the resulting oxide has a low capacity for sulfate formation. However, when the calcination takes place under a high partial pressure of carbon dioxide, the resulting stone has a much better sulfur removal activity at higher temperatures. This would explain the improved sulfur retention at higher temperatures in pressurized systems where the partial pressure of the carbon dioxide is relatively high.

Their results suggest that the amount of limestone required may be reduced if the stone is calcined under controlled conditions prior to being fed into the furnace. It is also possible that the limestone can be fed to the bed through the freeboard of the furnace in such a manner that the carbon dioxide, released as the material is calcined, will provide the needed high partial pressure for the incoming stone. It should be noted that these Westinghouse results were obtained with laboratory apparatus and not in an actual operating bed. For this reason it is felt that additional experimental data are needed before these approaches are incorporated into a system design.

4.1.5. Motor and Fan Efficiencies

The electric motor efficiencies used in the analysis of power requirements were as follows:

<u>Motor Size (hp)</u>	<u>Efficiency (%)</u>
> 600	96
150 - 600	95
100 - 150	94
50 - 100	93
25 - 50	92

These efficiencies were selected on the basis of a very elaborate study carried out over 20 years ago for the gaseous diffusion plants.

The economic parameters used in determining the capital cost vs efficiency in the study were power costs of 4 mills/kWhr and a payoff period of 5 years. Considering the economic changes since this study was made and the projections for the future cost of electricity, one should easily justify the cost of the more efficient motors. The adiabatic efficiency of the fans and compressors was assumed to be 82%.

4.1.6. Furnace

In the analysis of the furnace it is assumed that all of the limestone is calcined. However, less than 25% of the lime formed reacts with the sulfur. To meet the EPA requirements of 1.2 lb (0.54 kg) of SO₂ per million Btu (1055 MJ) with the Illinois No. 6 coal requires that 83% of the sulfur be retained in the bed. A value of 85% was used for the design.

The heat of reaction of both the calcination and the sulfation reactions was included in the furnace energy balance. The energy required for the calcination reaction

and the energy released from the sulfation reaction

are shown in Table 4.1 as a function of temperature.

The combustion efficiency in both the main cells and the recycle cells was assumed to be 95%. Work at both BCURA⁴ and Exxon⁵ has demonstrated combustion efficiencies of 95% to 99.5% without recycle in beds operating in the range of 1650 to 1750°F (1172 to 1228°K). These results were with pressurized systems operating at lower fluidizing velocities. With the furnace conditions of this study an efficiency of 95% or more would be expected. With the furnace arrangement described in Section 4.2 cell efficiencies of 95% give an overall combustion efficiency of about 99%.

The amount of material elutriated from the bed is another quantity which is only approximately known. For this system, it was assumed that

Table 4.1. Heats of Reaction

Temperature (°F)	Heat of Reaction for			
	$\text{CaCO}_3 \rightarrow \text{CaO} + \text{CO}_2$		$\text{CaO} + \text{SO}_2 + 1/2 \text{O}_2 \rightarrow \text{CaSO}_4$	
	Btu/lb mole	MJ/kg mole	Btu/lb mole	MJ/kg mole
0	76,600	178.2	215,200	500.6
500	75,500	175.6	214,700	499.5
1000	73,900	171.9	212,200	493.7
1500	71,800	167.0	208,100	484.1
1750	70,700	164.5	204,000	474.6
2000	69,500	161.7	199,800	464.8

90% of the coal and limestone ash and one-third of the CaO and CaSO₄ was elutriated from the regular cells. For the recycle cells, the above assumptions were made in regard to fresh coal and limestone and all of the recycled material was assumed to be elutriated. It was also assumed that the bed during normal operation would contain less than 1% carbon and that the material withdrawn from the bottom of the bed will contain 3/4% carbon by weight.

The efficiency of the cyclone separators was also assumed to be 95% for all of the material elutriated from the bed except for the unburned carbon. The carbon carried from a fluidized bed combustion system is generally occluded inside larger ash particles and is readily removed by a cyclone separator. For this reason all of the unburned carbon was assumed to be removed in the separators.

4.1.7. Alkali Metal Properties

The basic thermodynamic data used for cesium and potassium were those published by Ewing, et al., in NRL Reports 6233 and 6246.^{6,7} However, these data do not cover the lower temperatures needed for this study, and it was necessary to prepare computer codes to extend the range of data to lower temperatures. These data are given in Appendixes A, D, and G. Tables of the specific heat, viscosity, and thermal conductivity for both

liquid and vapor, and a table of sonic velocities for both fluids are also given in Appendixes B, C, E, and F. A documentation of this work is given in Ref. 8.

4.1.8. Other Assumptions and Data Sources

The ambient air conditions were assumed to be a temperature of 60°F (289°K), a pressure of 14.7 psia (0.1 MPa), and a relative humidity of 60% which corresponds to a humidity ratio of 0.0066. The amount of combustion air used was 15% greater than that required for complete combustion of the coal. The stack gas temperature was assumed to be approximately 300°F (422°K).

The properties of air and the constituents of the combustion products were taken from Ref. 9 and were adjusted to have a base temperature of 60°F (289°K); i.e., the enthalpy was corrected to 0 for 60°F (289°K). The heat capacity of the solids as a function of temperature was taken from Ref. 10. The enthalpy of the solids was also corrected to a base temperature of 60°F (289°K). Steam properties were taken from Ref. 11 and were not adjusted for temperature.

4.2. Furnace-Boiler

The design of the furnace and boiler for the plant is shown in Fig. 4.3. The furnace consists of eight 18.17 ft by 18.50 ft (5.54 m by 5.64 m) main cells and two 9.92 ft by 18.50 ft (3.02 m by 5.64 m) recycle cells. The larger cells each contain two cesium boiler tube bundles (see Fig. 4.4) while each recycle cell contains only one. All of the cells operate at a bed temperature of 1750°F (1228°K). The combustion products on leaving the fluidized bed pass through the narrow upper section of the furnace which houses the steam superheater and reheater and then through cyclone separators. With this arrangement most of the particulate material carried from the eight regular cells will be removed by two large cyclone separators located at one end of the furnace. This material, which will contain a few percent of the carbon fed to the eight cells, is then recycled to the two small beds located at the end of the furnace near the cyclones. The particulate material carried from the recycle cells is

Fig. 4.3. Fluidized bed furnace.

removed from the combustion products from these two cells by a third cyclone separator and rejected from the plant. This arrangement requires a baffle in the upper part of the furnace between the combustion products from the regular cells and the recycle cells to prevent excessive mixing between the two product streams.

The amount of carbon carried from the eight regular cells is not known very well, but with the high bed temperature it will probably be less than 5% of the carbon fed to these cells. The recycle cells have sufficient heat transfer capacity to burn up to 14% of the carbon fed to the regular

Fig. 4.4. Cesium boiler tube bundle.

cells, thus, the recycle cells will operate primarily on coal and will be provided with a coal and limestone feed system.

There are several reason for using the arrangement of Fig. 4.3 rather than a high temperature carbon burnup cell or a more conventional recycle system. For the bed operating conditions of near atmospheric pressure, 1750°F (1228°K) and a coal (Illinois No. 6) with a sulfur content of 3.9%, it may be difficult to meet the EPA emission standard for SO₂ which for this case requires that 83% of the sulfur be retained in the bed. Most atmospheric pressure systems operate with a bed temperature of about 1550°F (1117°K) which is near the optimum for minimizing the required limestone feed rate. There is little data available for systems operating at atmospheric pressure and 1750°F (1228°K) and the design limestone feed rate (Ca/S ratio of 4) selected for this design is based on extrapolation of lower temperature results. If a high temperature [2000°F (1367°K)] carbon burnup cell is used, then sulfur carried into this bed will be released to the stack. This increased release from the burnup cell then requires a higher sulfur retention in the regular cells. This would be very difficult for the conditions of this study.

Small experimental fluidized beds often recycle the particulate carryover back to the bed. This permits the unburned carbon to make several or many passes through the bed and reduces the carbon loss from the system. However, this type of system also recycles the inert material along with the unburned carbon and leads to a very high particulate loading in the combustion products. With the plant arrangement of Fig. 4.3 the combustion products enter the steam superheater at a velocity of 35 to 40 ft/sec (10.7 to 12.2 m/s) and it was deemed prudent to limit the particulate loading in these products. Also, it is simpler to recycle the material to only two cells located near the cyclone separators than to transport and feed the material to all cells.

In order to obtain a high overall combustion efficiency (>98%) with this concept, it is necessary to obtain a combustion efficiency of over 90% in the regular cells and the recycle cells. With the system of Fig. 4.3 a combustion efficiency of 95% or more would be expected from the cells and an overall efficiency of 99% should be obtainable. It should be noted that the superficial velocity from the recycle cells is about

10% less than from the regular cells and this reduction should aid the efficiency of these cells.

4.2.1. Furnace Operating Conditions

The furnace operating conditions are given in Tables 4.2, 4.3, and 4.4. Table 4.2 gives the overall furnace conditions while Tables 4.3 and 4.4 give conditions for the main cells and the recycle cells respectively. Note that all of the data are based on dry coal and limestone. The combustion air does contain water vapor — 0.0066 lb of water per lb of air. A flow diagram of the plant combustion air and flue gas system is shown in Fig. 4.1.

The thermal input to the furnace is 450 MW and requires 61.93 tons of dry coal per hr (15.60 kg/s). With Illinois No. 6 coal and a calcium to sulfur feed ratio of 4.0, the limestone feed rate is 0.588 lb (containing 0.559 lb of CaCO_3) per lb of coal or a total of 36.42 ton/hr (9.18 kg/s). The solids rejection rate from the furnace is 68,450 lb/hr (8.62 kg/s). Of this total, 29,820 lb/hr (3.76 kg/s) are removed by the recycle cell cyclone separator at a temperature of 900°F (756°K), 35,610 lb/hr (4.49 kg/s) are drained from the beds at a temperature of 1750°F (1228°K), and 3020 lb/hr (0.38 kg/s) leave with the flue gases at a temperature of 900°F (756°K).

The energy lost from the furnace as unburned carbon was calculated to be less than 1% of the energy input. In a typical fluidized bed coal combustor, the bed contains less than 1% carbon and for this analysis it was assumed that the material drained from the bed would have a composition typical of the bed, i.e., would contain 3/4% carbon. The rate of carbon lost in the material drained from the bed is 267 lb/hr (0.034 kg/s). The carbon lost by elutriation was calculated from the total carbon feed rate. The dry coal contains 68.5% carbon and the carbon feed rate to the main cells is 78,017 lb/hr (9.83 kg/s). Five percent of this material is elutriated and recycled to the recycle cells which also receive and additional 6827 lb/hr (0.86 kg/s) of carbon with their coal feed. Of this total carbon feed rate of 10,728 (1.35 kg/s) to the recycle cells, 5% or 536 lb/hr (0.068 kg/s) is assumed to be carried to the cyclone separator and rejected from the plant. Thus, the total carbon lost

Table 4.2. Overall Furnace Operating Conditions

Heat Generation Rate, Btu/hr, MW	1.53×10^9	450
Fluidized Bed Temperature, °F, °K	1750	1228
Cesium Operating Conditions		
Weight Flow, lb/hr, kg/s	1.78×10^7	2242
Inlet Enthalpy, Btu/lb, MJ/kg	106.0	0.247
Exit Enthalpy, Btu/lb, MJ/kg	160.0	0.372
Average Boiling Temperature, °F, °K	1535	1108
Coal Feed Rate (Dry), lb/hr, kg/s	123,860	15.60
Limestone Feed Rate (Dry) lb/hr, kg/s	72,830	9.18
Combustion Air Flow Rate, lb/hr, kg/s	1,347,600	169.8
Fluidizing Air		
Flow Rate, lb/hr, kg/s	1,298,400	163.6
Temperature, °F, °K	800	700
Enthalpy, Btu/lb, MJ/kg	183.4	0.427
Transport Air		
Flow Rate, lb/hr, kg/s	49,200	6.20
Temperature, °F, °K	262	401
Enthalpy, Btu/lb, MJ/kg	48.9	0.114
Heat Transferred to Power System, Btu/hr, MW	1.345×10^9	394.1
Cesium, Btu/hr, MW	9.61×10^8	281.6
Steam Superheater, Btu/hr, MW	2.00×10^8	58.6
Steam Reheater, Btu/hr, MW	1.84×10^8	53.9
Products from Bed (to Steam Superheater)		
Solids		
Flow Rate, lb/hr, kg/s	64,970	8.18
Temperature, °F, °K	1750	1228
Enthalpy, Btu/lb, MJ/kg	452.3	1.05
Gases		
Flow Rate, lb/hr, kg/s	1,475,800	185.9
Temperature, °F, °K	1750	1228
Enthalpy, Btu/lb, MJ/kg	460.0	1.09
Products to Steam Reheater		
Solids		
Flow Rate, lb/hr, kg/s	64,970	8.18
Temperature, °F, °K	1318	988
Enthalpy, Btu/lb, MJ/kg	315.3	0.733
Gases		
Flow Rate, lb/hr, kg/s	1,475,800	185.9
Temperature, °F, °K	1318	988
Enthalpy, Btu/lb, MJ/kg	338.6	0.788
Products to Cyclones		
Solids		
Flow Rate, lb/hr, kg/s	64,970	8.18
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	193.4	0.450
Gases		
Flow Rate, lb/hr, kg/s	1,475,800	185.9
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	219.1	0.510
Solids Rejected from Recycle Cell Cyclone		
Flow Rate, lb/hr, kg/s	29,800	3.76
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	190.3	0.443
Material Passing Through Cyclones (to Bag Filters)		
Solids		
Flow Rate, lb/hr, kg/s	3020	0.380
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	193.2	0.449
Gases		
Flow Rate, lb/hr, kg/s	1,475,800	185.9
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	219.1	0.510
Solids Drained From Bed		
Flow Rate, lb/hr, kg/s	35,610	4.49
Temperature, °F, °K	1750	1228
Enthalpy, Btu/lb, MJ/kg	421.8	0.981
Steam Superheater Conditions		
Flow Rate, lb/hr, kg/s	907,100	114.3
Inlet Temperature, °F, °K	739	666
Inlet Enthalpy, Btu/lb, MJ/kg	1240.8	2.89
Exit Temperature, °F, °K	1000	811
Exit Enthalpy, Btu/lb, MJ/kg	1461.2	3.40
Steam Reheater Conditions		
Flow Rate, lb/hr, kg/s	800,200	100.8
Inlet Temperature, °F, °K	582	579
Inlet Enthalpy, Btu/lb, MJ/kg	1292.4	3.01
Exit Temperature, °F, °K	1000	811
Exit Enthalpy, Btu/lb, MJ/kg	1522.4	3.54

Table 4.3. Main Cells Operating Conditions

Heat Generation Rate, Btu/hr, MW	1.365×10^9	400
Fluidized Bed Temperature, °F, °K	1750	1228
Bed Superficial Velocity, ft/sec, m/s	6.55	2.97
Cesium Operating Conditions		
Weight Flow, lb/hr, kg/s	1.58×10^7	1990
Inlet Enthalpy, Btu/lb, MJ/kg	106.0	0.247
Exit Enthalpy, Btu/lb, MJ/kg	160.0	0.372
Average Boiling Temperature, °F, °K	1535	1108
Coal Feed Rate (Dry) lb/hr, kg/s	113,893	14.35
Limestone Feed Rate (Dry) lb/hr, kg/s	66,969	8.44
Coal and Limestone Feed, lb/hr, kg/s	180,862	22.78
Temperature, °F, °K	262	401
Enthalpy, Btu/lb, MJ/kg	44.0	0.102
Combustion Air		
Fluidizing Air		
Flow Rate, lb/hr, kg/s	1,154,146	145.4
Temperature, °F, °K	800	700
Enthalpy, Btu/lb, MJ/kg	183.4	0.427
Transport Air		
Flow Rate, lb/hr, kg/s	43,709	5.51
Temperature, °F, °K	262	401
Enthalpy, Btu/lb, MJ/kg	48.9	0.114
Products from Bed (to Steam Superheater)		
Solids		
Flow Rate, lb/hr	33,610	4.23
Temperature, °F, °K	1750	1228
Enthalpy, Btu/lb, MJ/kg	460.9	1.07
Gases		
Flow Rate, lb/hr, kg/s	1,311,798	165.3
Temperature, °F, °K	1750	1228
Enthalpy, Btu/lb, MJ/kg	468.0	1.09
Products to Steam Reheater		
Solids		
Flow Rate, lb/hr, kg/s	33,610	4.23
Temperature, °F, °K	1318	988
Enthalpy, Btu/lb, MJ/kg	320.7	0.746
Gases		
Flow Rate, lb/hr, kg/s	1,311,798	165.3
Temperature, °F, °K	1318	988
Enthalpy, Btu/lb, MJ/kg	338.6	0.788
Products to Cyclones		
Solids		
Flow Rate, lb/hr, kg/s	33,610	4.23
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	196.2	0.456
Gases		
Flow Rate, lb/hr, kg/s	1,311,798	165.3
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	219.1	0.510
Solids Separated in Cyclones		
Flow Rate, lb/hr, kg/s	32,129	4.05
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	196.2	0.456
Material Passing Through Cyclones (to Bag Filters)		
Solids		
Flow Rate, lb/hr, kg/s	1481	0.187
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	196.2	0.456
Gases		
Flow Rate, lb/hr, kg/s	1,311,798	165.3
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	219.1	0.510
Solids Drained From Bed		
Flow Rate, lb/hr, kg/s	32,741	4.13
Temperature, °F, °K	1750	1228
Enthalpy, Btu/lb	421.8	0.981

Table 4.4. Recycle Cells Operating Conditions

Heat Generation Rate, Btu/hr, MW	1.707×10^8	50
Fluidized Bed Temperature, °F, °K	1750	1228
Bed Superficial Velocity, ft/sec, m/s	5.99	2.72
Cesium Operating Conditions		
Weight Flow, lb/hr, kg/s	1.98×10^6	250
Inlet Enthalpy, Btu/lb, MJ/kg	106.0	0.247
Exit Enthalpy, Btu/lb, MJ/kg	160.0	0.372
Average Boiling Temperature, °F, °K	1535	1108
Coal Feed Rate (Dry) lb/hr, kg/s	9966	1.26
Limestone Feed Rate (Dry) lb/hr, kg/s	5860	0.738
Coal and Limestone Feed, lb/hr, kg/s	15,826	1.99
Temperature, °F, °K	262	401
Enthalpy, Btu/lb, MJ/kg	44.0	0.102
Carbon Recycle Flow Rate, lb/hr, kg/s	3901	0.491
Total Recycled Material		
Flow Rate, lb/hr, kg/s	32,129	4.05
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	196.2	0.456
Combustion Air		
Fluidizing Air		
Flow Rate, lb/hr, kg/s	144.268	18.17
Temperature, °F, °K	800	700
Enthalpy, Btu/lb, MJ/kg	183.4	0.427
Transport Air		
Flow Rate, lb/hr, kg/s	5464	0.688
Temperature, °F, °K	262	401
Enthalpy, Btu/lb, MJ/kg	48.9	0.114
Products from Bed (to Steam Superheater)		
Solids		
Flow Rate, lb/hr, kg/s	31,356	3.95
Temperature, °F, °K	1750	1228
Enthalpy, Btu/lb	443.0	1.03
Gases		
Flow Rate, lb/hr, kg/s	163,978	20.66
Temperature, °F, °K	1750	1228
Enthalpy, Btu/lb, MJ/kg	468.0	1.09
Products to Steam Reheater		
Solids		
Flow Rate, lb/hr	31,356	3.95
Temperature, °F, °K	1318	988
Enthalpy, Btu/lb, MJ/kg	309.5	0.720
Gases		
Flow Rate, lb/hr, kg/s	163,978	20.66
Temperature, °F, °K	1318	988
Enthalpy, Btu/lb, MJ/kg	338.6	0.788
Products to Cyclones		
Solids		
Flow Rate, lb/hr, kg/s	31,356	3.95
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	190.3	0.443
Gases		
Flow Rate, lb/hr, kg/s	163,978	20.66
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	219.1	0.510
Solids Separated in Cyclones		
Flow Rate, lb/hr, kg/s	29,815	3.76
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	190.3	0.443
Material Passing Through Cyclones (to Bag Filters)		
Solids		
Flow Rate, lb/hr, kg/s	1541	0.194
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	190.3	0.443
Gases		
Flow Rate, lb/hr, kg/s	163,978	20.66
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	219.1	0.510
Solids Drained from Bed		
Flow Rate, lb/hr, kg/s	2870	0.362
Temperature, °F, °K	1750	1228
Enthalpy, Btu/lb, MJ/kg	421.8	0.981

is 803 lb/hr (0.10 kg/s) which, with a heating value of 14,000 Btu/lb (32.56 MJ/kg), represents a loss of 1.12×10^7 Btu/hr (3.29 MW) or 0.76% of the heating value of the coal input.

4.2.2. Furnace Material and Energy Balance

Tables 4.5 and 4.6 give the material and energy balance for the furnace. The material balance of Table 4.5 assumes that all of the CaCO_3 feed to the furnace is calcined and that 85% of the sulfur is retained in the bed as CaSO_4 . The remaining 15% of the sulfur escapes as SO_2 .

Table 4.5. Furnace Material Balance

Item	Weight lb/hr	kg/s
Materials In		
Fluidizing Air	1,298,400	163.6
Transport Air	49,200	6.2
Limestone (Dry)		
CaCO_3	69,240	8.7
Ash	3,590	.5
Coal (Dry)	<u>123,860</u>	<u>15.6</u>
	1,544,290	194.6
Materials Out		
Gases	1,475,800	185.9
Solids		
From Cyclones	29,820	3.8
From Bed Drains	35,610	4.5
To Bag Filters	<u>3,020</u>	<u>.4</u>
	1,544,250	194.6

Table 4.6. Furnace Energy Balance

Item	Energy Btu/hr	MW
Energy In		
Coal Heating Value	1.536×10^9	450.0
Sensible Energy		
Solids	8.65×10^6	2.5
Fluidization Air	2.38×10^8	69.7
Transport Air	2.41×10^6	0.7
Sulfation Reaction	3.20×10^7	9.4
	1.817×10^9	532.3
Energy Out		
Cesium Boiler	9.61×10^8	281.6
Steam Superheater	2.00×10^8	58.6
Steam Reheater	1.84×10^8	53.9
Flue Gases	3.23×10^8	94.6
Solids		
From Cyclones	5.67×10^6	1.7
From Bed Drains	15.02×10^6	4.4
To Bag Filters	0.58×10^6	0.2
Radiation and Convection	3.07×10^6	0.9
Calcination Reaction	53.07×10^6	15.5
Latent Heat	60.13×10^6	17.6
Carbon Loss	11.24×10^6	3.3
	1.817×10^9	532.3

The total combustion air flow rate is 15% in excess of that required for complete combustion of the coal and for the formation of the CaSO_4 .

The energy balance of Table 4.6 includes radiation and convection losses of 0.2% of the energy input, a calcination loss of 765 Btu/lb (1.78 MJ/kg) of CaSO_4 which is equivalent to 428 Btu/lb (1.00 MJ/kg) of coal, and a latent heat loss of 1060 Btu/lb (2.47 MJ/kg) for the water vapor formed from combustion. The amount of water vapor formed in combustion is .458 lb per lb of coal or a total of 56,700 lb/hr (7.14 kg/s). It should be noted that the water in the "as received" coal does not enter

this calculation as this water is removed before the coal enters the furnace. The energy balance also includes a heat input of 6360 Btu/lb (14.8 MJ/kg) of sulfur reacted which for the system is equivalent to 256 Btu/lb (0.59 MJ/kg).

4.2.3. Alkali Metal Boiler

The alkali metal boiler, which transfers about 71% of the total heat from the furnace to the power conversion system, is located within the fluidized bed. The boiler consists of a total of 18 individual tube bundles (see Fig. 4.4) with two bundles located in each of the main furnace cells and one bundle located in each of the two recycle cells. These units require forced circulation of the alkali metal and are recirculating boilers as opposed to once-through boilers.

The tube bundle headers are 16 in. (40.64 cm) O.D. pipes. The liquid header contains a centrally located 12 in. pipe [12.75 in. (32.4 cm) O.D.] which is blanked off to minimize the liquid inventory. Each bundle contains 588 Incoloy 800 tubes which are 1 in. (2.54 cm) O.D. by 0.8 in. (2.03 cm) I.D. and located on a triangular array on 2 in. (5.08 cm) centers. The average length of the tubes is 42 ft (12.8 m) with 36 ft (11.0 m) of the length in the active part of the bed.

The length of each header is 104.3 in. (2.65 m) with the tubes extending 96.3 in. (2.45 m) along this length. The header end caps are inverted to minimize both the clearance between the tube banks and the cesium inventory in the headers. The selection of the bundle dimensions was based on several considerations. The header diameter was sized to allow sufficient perimeter for the tube connections while the number of tubes per length of header is limited by the tube spacing and the bed depth. The bed depth is limited by the bed pressure drop and the resulting air pumping requirements of the combustion system. For atmospheric systems, the bed depth is normally limited to 3 to 4 ft (0.91 to 1.22 m).

The length of the individual tube bundles is limited by the diameter of the vapor headers because the heat load, the vapor flow rate and the vapor pressure loss increases with increasing bundle length. The pressure loss is an important consideration in this system as the pressure

loss also results in a temperature loss. For example, with cesium a 5 psi (34.5 kPa) pressure drop between the boiler and the turbine will result in a temperature decrease of about 25°F (14°K). A large vapor header would permit a longer tube bundle, however, there are also economic and maintenance considerations that favor shorter lengths.

One obvious requirement of the cesium boiler is very high quality fabrication and quality assurance. However, in the event of a leak, it is anticipated that the bundle would be replaced by a spare unit and the failed unit would be repaired in a special shop or plant shop area where the quality of the repair work can be assured. A shorter tube bundle length is preferred in order to minimize the difficulty and time required to replace a unit.

The boiler was analyzed for operation with both cesium and potassium and with a range of values of exit quality. A detailed description of the analytical method is presented elsewhere¹², only a brief description being included in this report. Determination of the pressure drop due to the momentum exchange and fluid friction phenomena associated with two-phase flow generally involves a series of nontrivial calculations. The common approach in quantifying the two-phase friction pressure drop is to apply a so-called two-phase multiplier to the conventional Fanning or Darcy friction factor utilized in determining the pressure loss in single phase flow. This same approach was adopted in this study.

An integral part of the analysis is the determination of the local flow quality in the boiler tubes. This, of course, requires knowledge of the heat transfer from the fluidized bed to the alkali metal. Since the fireside heat transfer coefficient is perhaps two orders of magnitude less than the boiling heat transfer coefficient, the bed-tube coefficient is nearly the same as an overall coefficient between the bed and alkali metal. An overall coefficient of 45 Btu/hr-ft²-°F (255 W/m²·°C) was used in the analysis. The boiling temperature of the substance was assumed constant along the tube length. Other simplifying assumptions in the analysis are listed below.

1. The kinetic and potential energy changes were negligible in comparison with the heat flux and were not included in the energy

equation used to arrive at the flow quality. The viscous heating terms were also neglected.

2. The vapor compressibility and its effect on the two-phase pressure drop was found to be negligible.

The above simplifications are not uncommon in boiler analyses, especially in view of the uncertainty associated with the two-phase friction multiplier. The alkali metal properties were taken from a recent documentation⁸ and are tabulated in Appendixes A through G.

In the absence of large static heads, the two-phase friction pressure drop is dominant in the total pressure loss around the boiling loop. The resulting pumping power required to circulate the alkali metal through the boiler is shown in Figs. 4.5 and 4.6. Although the shorter tubes would give smaller pumping power, a compromise length resulted from considerations previously discussed in this section. A adiabatic pump efficiency of 80% was utilized for calculating the data presented in Figs. 4.5 and 4.6. The pumping power is higher for cesium since, for a designated exit quality, the cesium must be circulated at approximately twice the velocity necessary for potassium.

Since the tubes in the boiler loops are of slightly different lengths due primarily to the variation in the length of the vertical section, the exit quality will vary somewhat from tube to tube. Investigation of this behavior provided the results in Fig. 4.7. The variation is small enough that no related problems are anticipated.

Although it is unlikely that a potential tube burnout condition exists since the temperature in a fluidized bed is limited, values for the critical heat flux were obtained from the literature as a precaution. The critical heat fluxes cited in the companion report¹² were greater than the anticipated furnace heat flux of 10,500 Btu/hr-ft² (33,100 W/m²) by at least one order of magnitude at qualities on the order of 20%.

A summary of the cesium boiler design and operating conditions are given in Table 4.7. The different boiling conditions given reflect the pressure drop through the boiler. The pressure and thus the temperature at the boiler exit was determined by the pressure drop between the centrally located tube bundle in the furnace and the turbine inlet. The turbine inlet pressure was 49.38 psia (.3405 MPa) which corresponds to

Fig. 4.5. Pumping power for cesium boiler loop at 1540°F (1111°K) and for a 36 in. (91 cm) vertical section.

Fig. 4.6. Pumping power for potassium boiler loop at 1540°F (1111°K) and for a 36 in. (91 cm) vertical section.

Fig. 4.7. Variation of quality at the exit of horizontal boiler tubes as a function of the height of the vertical section with a 12 ft (3.66 m) horizontal section and equal ΔP through tubes.

a temperature of 1500°F (1089°K). Thus, the data of Table 4.7 represent the temperature conditions in the average bundle. The total pressure loss in the headers external to the furnace is only 1.40 psia (9.6 kPa) which corresponds to a temperature differential of 8°F (4.44°K). Thus, the temperatures in the tube bundle in the recycle cell will be 4°F (2.22°K) higher while those at the opposite end will be 4°F (2.22°K) lower.

Table 4.7. Cesium Boiler Design and Operating Conditions

Heat Load, Btu/hr, MW	9.61×10^8	281.6
Weight Flow, lb/hr, kg/s	1.78×10^7	2245
Inlet Pressure, psia, MPa	62.31	0.430
Inlet Temperature, °F, °K	1437	1054
Inlet Enthalpy, Btu/lb, MJ/kg	105.97	0.246
Exit Pressure, psia, MPa	51.71	0.357
Exit Temperature, °F, °K	1508	1093
Exit Enthalpy, Btu/lb, MJ/kg	160.03	0.372
Exit Quality, %	25	
Bed Temperature, °F, °K	1750	1228
Average Boiling Temperature, °F, °K	1535	1108
Overall Heat Transfer Coefficient, Btu/hr-ft ² -°F, W/m ² -°C	45	255
Surface Area, ft ² , m ²	9.97×10^4	9.2×10^3
Number of Tubes	10,584	
Cesium Inventory		
Liquid Header, lb, kg	4672	2119
Boiler Tubes, lb, kg	39,160	17,800
Vapor Header, lb, kg	3241	1470

4.2.4. Steam Superheater and Reheater

The steam superheater and reheater are located in the upper narrow section of the furnace (see Fig. 4.3). The flue gases pass first through the superheater and then the reheater. No analysis was made to investigate the possible advantages of splitting the superheater section. The design of these units is identical except that the reheater has 2.15 times the surface area of the superheater. Both units are vertical serpentine coils made of 2 in. (5.08 cm) O.D. tubes. The width of the coils is 7 ft (2.13 m) and both units run the full length of the furnace.

The particulate loading in the gases passing through the superheater and reheater sections above the recycle cells is much greater than that

passing through the coils above the main cells; 0.19 lb per lb of gas from the recycle cells as compared to 0.026 lb per lb of gas from the main cells. As noted previously the width of the recycle cells is 54.5% that of the main cells which results in a 9% lower velocity through these cells. In addition to improving the combustion efficiency in these cells, the reduced velocity also helps offset the increased pressure loss due to the higher particulate loading and also reduces the possibility of tube erosion. The pressure loss across the steam coils above the main and recycle cells should be nearly the same since the flow streams through these coils are separated only by a baffle in the upper part of the coils and it is desirable to minimize mixing of particulates between the two streams.

The design and operating conditions of the steam superheater and reheater are given in Tables 4.8 and 4.9. Note that the pressure loss calculated for the main cells assumes that there is no mixing between the two streams in passing through the steam coils. Actually, unless the pressure losses are identical, there will be some mixing in order to equalize the loss. For the values shown, there would be a small cross flow from the main cells to the recycle section of the coils. This would amount to about 0.1% of the flow from the main cells. Thus, the amount of unburned carbon from the main cells that is not recycled is very small. It should be noted that this value of 0.1% cross flow is based on the assumption that the pressure loss in the cyclone separators for the main cells is the same as that for the cyclone in the recycle cell stream.

Cross flow between the cells in the steam coils will make a small change in the gas temperature to the cyclones and the steam temperature from the superheater and reheater. For the 0.1% cross flow from the main to the recycle cells, the gas and steam temperatures from the main cells would be reduced by less than 1°F (0.56°K). The gas and steam conditions from the recycle cells would increase by a few degrees.

4.2.5. Air Nozzles and Solid Feed Ports

The air nozzles and the coal and limestone feed ports for the main cells and the recycle cells are shown in Figs. 4.8 and 4.9. The air

Table 4.8. Steam Superheater Design and Operating Conditions

Heat Load, Btu/lb, MW	2.00×10^8	58.6
Steam Conditions		
Weight Flow, lb/hr, kg/s	907,100	114.3
Inlet Pressure, psia, MPa	2430	16.72
Inlet Temperature, °F, °K	739	666
Inlet Enthalpy, Btu/lb, MJ/kg	1240.8	2.89
Exit Pressure, psia, MPa	2415	16.65
Exit Temperature, °F, °K	1000	811
Exit Enthalpy, Btu/lb, MJ/kg	1461.2	3.40
Gas Side Conditions (Main Cells)		
Gas Weight Flow, lb/hr, kg/s	1,311,798	165.3
Solids Weight Flow, lb/hr, kg/s	33,610	4.23
Inlet Velocity, ft/sec, m/s	38.7	11.9
Inlet Temperature, °F, °K	1750	1228
Inlet Enthalpy		
Gas, Btu/lb, MJ/kg	468.0	1.09
Solids, Btu/lb, MJ/kg	460.9	1.07
Exit Temperature, °F, °K	1318	988
Exit Enthalpy		
Gas, Btu/lb, MJ/kg	338.6	0.788
Solids, Btu/lb, MJ/kg	320.7	0.746
Pressure Loss,* psi, Pa	0.0868	598
Gas Side Conditions (Recycle Cells)		
Gas Weight Flow, lb/hr, kg/s	163,978	20.66
Solids Weight Flow, lb/hr, kg/s	31,356	3.95
Inlet Velocity, ft/sec, m/s	35.0	10.7
Inlet Temperature, °F, °K	1750	1228
Inlet Enthalpy		
Gas, Btu/lb, MJ/kg	468.0	1.09
Solids, Btu/lb, MJ/kg	443.0	1.03
Exit Temperature, °F, °K	1318	988
Exit Enthalpy		
Gas, Btu/lb, MJ/kg	338.6	0.788
Solids, Btu/lb, MJ/kg	309.5	0.720
Pressure Loss,* psi, Pa	0.0850	586
Log Mean Temperature Difference, °F, °K	661	367
Overall Heat Transfer Coefficient, Btu/hr-ft ² -°F, W/m ² -°K	14.4	81.7
Surface Area, ft ² , m ²	21,000	1950
Tube O.D., in., cm	2.0	5.08
Vertical Tube Spacing, in., cm	6.0	15.24
Transverse Tube Spacing, in. cm	4.0	10.16
Height of Unit, ft, m	10.7	3.26
Gas Side Pressure Drop, psi, Pa	0.0866	597

*Pressure losses calculated by assuming no cross flow between main cells and recycle cells.

Table 4.9. Steam Reheater Design and Operating Conditions

Heat Load, Btu/hr, MW	1.84×10^8	53.9
Steam Conditions		
Weight Flow, lb/hr, kg/s	800,200	100.8
Inlet Pressure, psia, MPa	446	3.08
Inlet Temperature, °F, °K	582	579
Inlet Enthalpy, Btu/lb, MJ/kg	1292.4	3.01
Exit Pressure, psia, MPa	402	2.77
Exit Temperature, °F, °K	1000	811
Exit Enthalpy, Btu/lb, MJ/kg	1522.4	3.54
Gas Side Conditions (Main Cells)		
Gas Weight Flow, lb/hr, kg/s	1,311,798	1653
Solids Weight Flow, lb/hr, kg/s	33610	4.23
Inlet Velocity, ft/s, m/s	31.4	9.57
Inlet Temperature, °F, °K	1318	988
Inlet Enthalpy		
Gas, Btu/lb, MJ/kg	338.6	0.788
Solids, Btu/lb, MJ/kg	320.7	0.746
Exit Temperature	900	756
Exit Enthalpy		
Gas, Btu/lb, MJ/kg	219.1	0.510
Solids, Btu/lb, MJ/kg	196.2	0.456
Pressure Loss,* psi, Pa	0.148	1020
Gas Side Conditions (Recycle Cells)		
Gas Weight Flow, lb/hr, kg/s	163,978	20.66
Solids Weight Flow, lb/hr, kg/s	31,356	3.95
Inlet Velocity, ft/sec, m/s	28.4	8.66
Inlet Temperature, °F, °K	1318	988
Inlet Enthalpy		
Gas, Btu/lb, MJ/kg	338.6	0.788
Solids, Btu/lb, MJ/kg	309.5	0.720
Exit Temperature, °F, °K	900	756
Exit Enthalpy		
Gas, Btu/lb, MJ/kg	219.1	0.510
Solids, Btu/lb, MJ/kg	190.3	0.443
Pressure Loss,* psi, Pa	0.144	993
Log Mean Temperature Difference, °F, °K	318	432
Overall Heat Transfer Coefficient, Btu/hr-ft ² -°F, W/m ² -°K	12.8	72.7
Surface Area, ft ² , m ²	45,200	4200
Tube O.D., in., cm	2.0	5.08
Vertical Tube Spacing, in., cm	6.0	15.24
Transverse Tube Spacing, in., cm	4.0	10.16
Height of Unit, ft, m	23.7	7.21
Gas Side Pressure Drop, psi, Pa	0.147	1014

*Pressure losses calculated by assuming no cross flow between main cells and recycle cells.

Fig. 4.8. Air nozzles and solids feed ports to the main cells.

Fig. 4.9. Air nozzles and solids feed ports to the recycle cells.

nozzles are capped-off 2.5 in. pipes [2.875 in. (7.30 cm) O.D.] located on 9 in. (22.86 cm) centers. Each nozzle contains four 0.50 in. (1.27 cm) holes from which the fluidizing air enters the bed. The location of the nozzles was based on the penetration distance of the jets from the 0.50 in. (1.27 cm) holes in each nozzle. The criterion used was that the distance between nozzles, using a square array, would be approximately equal to the jet penetration distance. With the nozzles turned at an angle of 45° from the adjacent nozzle, no distribution should be obtained without any overlapping of the jet.

As will be noted in the following section, the pressure drop allowed for the nozzles is 0.68 psi (4620 Pa) which results in a jet velocity of about 340 ft/sec (104 m/sec). The jet penetration distance¹³ for this case was found to be about 6 times the hole diameter or about 3 in. (7.62 cm). Thus, the total span of the jets is twice the penetration distance plus the O.D. of the nozzles or 8.75 in. (22.3 cm).

Figure 4.8 also shows the vertical location of the air holes, feed ports and boiler tube bundle. The air holes are located 6 in. (15.24 cm) above the bed support plate. This allows a stagnant unfluidized 6 in. (15.24 cm) layer of material which serves to insulate the bed plate from the high temperature of the bed. The coal and limestone feed ports are located 4 in. (10.16 cm) above the air holes and the lower tubes of the boiler bundles are 8 in. (20.32 cm) above the feed ports. These distances were chosen to allow some distance for mixing and dispersion of the coal before passing into the tube bundles. These distances add to the bed pressure losses but are believed to be useful in the mixing and will help prevent reducing combustion conditions within the tube bundles.

The recycle cells require only $1/2$ the air flow rate of the main cells and have 54.5% of the area. Thus, the air flow per nozzle and the jet penetration distance will be slightly smaller. However, this difference is very small and the same 9 in. (22.86 cm) spacing was used in these cells.

The number of coal feed ports required is an item that needs more experimental investigation for fluidized bed systems. A general "rule-of-thumb" that is being used is to locate one feed port for each 9 ft^2

(0.836 m²) of bed. For the main cells a total of 36 feed ports are provided which gives one for each 9.34 ft² (0.867 m²) of bed. As can be seen in Figs. 4.8 and 4.9, the coal and limestone feed ports are located in a triangular array which results in a slightly uneven feed area for some ports. This resulted from the coal and limestone feed system using a four to three splitter arrangement (see Section 4.6) which was selected to give nearly equal line lengths for the feed system to the furnace. The final location of the feed ports was selected by overlaying the optimum triangular array over the air nozzle pattern and moving the feed ports to the nearest air nozzle location. It should be noted that the coal and limestone feed lines are not welded to the bottom bed plate. This will allow some leakage of air through these clearances. However, as each feed port replaces an air nozzle and the transport air entering the bed through a feed port is less than the air flow through an air nozzle, the leakage should improve the air distribution.

The location of the feed ports in the recycle cells are identical to those of the main cells. However, only 1/2 as many ports are used and these are offset from the center of the cells. This leaves a gap along the outer wall of the cells which is fed with the recycle material through three lances which penetrate the outer furnace wall.

Figures 4.8 and 4.9 also show the bed drain ports for removing material from the beds. The ash removal system is discussed later in Section 4.7.

4.2.6. Bed Depth and Furnace Pressure Losses

One of the disadvantages of the fluidized bed combustor is the higher pressure losses associated with the combustion system. The major loss is in the bed and the air distribution system; i.e., the air nozzles. For this system the total operating bed depth required is the 12 in. (30.48 cm) from the air nozzles to the boiler bundle tubes plus the 43 in. (109 cm) dept of the boiler tubes or a total of 55 in. (140 cm). The pressure loss across the bed depends on the weight or density of the bed which at full operating conditions is expected to be about one-half of the static

bed density. Typical static bed densities are about 85 lb/ft^3 (1361 kg/m^3). Thus, the pressure loss across the bed is equivalent to 27.5 in. (70 cm) of material with a density of 85 lb/ft^3 (1361 kg/m^3) or 1.35 psi (9.31 kPa).

The second largest pressure loss in the system is that across the air nozzles. In order to maintain good stability in a fluidized bed the pressure drop across the air nozzles should be a minimum of 10 to 15% of the bed pressure drop. This, in turn, means that the design pressure drop for the nozzles will depend on the amount of turndown required and the manner in which the turndown is achieved. For example, if one wants to get a 50% turndown by reducing the fuel feed rate to all cells while maintaining a fixed air-to-fuel ratio, then the air flow must be reduced to one-half of full design flow and the nozzle pressure drop will be one-fourth of that at full design flow. As the pressure at this reduced flow must be 10 to 15% of the bed pressure drop (which is a constant independent of air flow rate), then the pressure drop at full power conditions must be about 50% of the pressure drop across the bed.

There are other control methods that do not require as large a nozzle pressure drop. It is possible to keep air flow constant and reduce the fuel feed. This would increase stack loss and reduce plant efficiency. It would also be possible to cut off all air and coal feed to one cell and maintain normal air and fuel flow to the other cells.

For this study the higher pressure drop of 50% of that of the bed will be used. This permits a large, relatively easy, turndown without slumping a bed and without excessive stack gas losses. As will be noted in the discussion of plant control in Section 4.8.2, the high temperature boiler of this system is very amenable to this type of turndown.

Table 4.10 lists the pressure losses associated with the furnace and includes the steam superheater and reheater and the cyclone separators. The pressure loss in high efficiency cyclones normally range between 2 and 6 in. of water (498 and 1494 Pa) and for this study the higher value of 6 in. (1494 Pa) was used.

Table 4.10. Furnace Pressure Losses

Item	Pressure Loss	
	psi	Pa
Air Nozzles	0.675	4,654
Bed	1.350	9,308
Steam Superheater	0.087	600
Steam Reheater	0.147	1,014
Cyclones	<u>0.217</u>	<u>1,496</u>
	2.476	17,072

4.3. Combustion Air and Flue Gas System

The flow diagram of the combustion air and flue gas system is shown in Fig. 4.1. A total of 1,347,600 lb/hr (169.8 kg/s) of air enters the system of 60°F (289°K) and, after being compressed by the forced draft fan, is split into two streams. The fluidizing air stream [1,298,400 lb/hr (163.6 kg/s)] passes through the air preheater where its temperature is increased to 800°F (700°K) before entering the furnace. The transport air stream [49,200 lb/hr (6.2 kg/s)] passes through a second fan where its pressure is raised to 17.05 psia (0.118 MPa) and is then used to transport the coal and limestone to the furnace. Because of the higher temperature of the coal and limestone [316°F (431°K)] the temperature of this stream enters the furnace at 262°F (401°F).

The flue gases which leave the furnace cyclones at 900°F (756°K) are joined by the ash cooling flow at 900°F (756°K). This combined flow of 1,580,200 lb/hr (199.1 kg/s) is then separated into two streams. One stream of 194,800 lb/hr (24.5 kg/s) is used to dry the coal and limestone (see Section 4.6). After passing through the dryers the gases pass through bag filters, an fan, and then to the plant stack at 272°F (407°K). It should be noted that the weight flow and energy balances with the flue gases are based on the gases only and neglect the the particulates which amount to only a few tenths of 1% of the total flow. The second flue

gas stream [1,385,200 lb/hr (174.5 kg/s)] from the furnace passes through the air preheater where its temperature is reduced to 275°F (408°K). This stream is then split with the larger flow [1,281,000 lb/hr (161.4 kg/s)] passing through the bag filters, the exhaust fan and then the stack at 291°F (417°K). The other stream [104,400 lb/hr (13.2 kg/s)] after being compressed to 15.4 psia (0.106 MPa) is used to transport the ash drained from the beds and also to cool the ash rejected from the furnace. This stream then rejoins the main flue gases flow from the furnace. The ash handling and cooling system is described in Section 4.7.

Bag filters were selected to clean the stack gases. There are several reasons for the selection of bag filters rather than electrostatic precipitations. First, there are questions as to the effectiveness of precipitators with the composition of the particulate matter from a fluidized bed furnace. A second reason is that the current technology hot precipitators perform best over a narrow temperature range and with a maximum operating temperature of 750°F (672°K). This requires that the flue gas temperature from the furnace be limited to 750°F (672°K) which penalizes the plant efficiency, or requires that the heat exchanger that preheats the combustion air be split into two sections with the precipitators located between sections — an added complication. A third reason is that bag filters will probably be less expensive than a high temperature precipitator.

The total amount of particulates shown carried to the bag filters in the furnace analysis (see Table 4.2) is somewhat greater than the actual case. The quantity of solids carried from the cyclones was based on an efficiency of 95% for the cyclones. This is probably an underestimation of the efficiency of the recycle cell cyclone. Most of this material was of sufficient size to be removed from the flue gases from the main cells and, neglecting attrition on passing through the recycle cells, should be removed from the recycle cell flue gases. Also, the much higher particulate loading in the flue gases from the recycle cells will tend to improve the efficiency of the cyclone. However, as will be seen in Section 4.7, some additional particulates will be added to the flue gases by the ash cooling system. Thus, the particulate load of 3020 lb/hr (0.38 kg/s) to the bag filters (shown in Table 4.2) is a good estimate for this flow.

The particulate emissions from the plant will be very low. The efficiency of the bag filters is over 99%. Thus, taking the value of particulate loading from Table 4.2 of 3020 lb/hr, the amount leaving the plant will be less than 30.2 lb/hr. This is only 0.02 lb (0.001 kg) per million Btu (1055 MJ) and well below the EPA limit of 0.1 lb (0.045 kg) per million Btu (1055 MJ).

A compilation of the weight flows and operating conditions throughout the combustion system is presented in Table 4.11. The pressure losses for the system and the fan pressure ratios and power requirements are shown in Table 4.12. The pressure losses in the air preheater were estimated from those of a unit designed for operating conditions very similar to those of this system. The stack loss includes an exit loss corresponding to an exit velocity of 90 ft/sec (40.8 m/s). Other losses include 4 in. of water (1000 Pa) for the air ducts, 5 in. of water (1248 Pa) for the flue gas ducts, and 4 in. of water for the bag filters. Note that Table 4.12 includes the losses for the furnace that were given in Section 4.2.6. The fan pressure ratios are based on maintaining the freeboard of the furnace at 14.7 psia (0.101 MPa).

4.4. Power Conversion System

The plant is a binary cycle system with power extracted from both the high temperature topping cycle (cesium) and the lower temperature steam cycle. The cycle arrangement used in this study differs from a conventional binary system in that prime heat, the heat from the combustion system, is added to both the cesium and the steam cycle. Normally, the heat flow through a binary cycle is such that all of the highest temperature energy passes first through the topping cycle and only the heat rejected from the topping cycle passes through the secondary cycle.

In this study heat is added to the secondary (steam) cycle by that rejected from the topping cycle and also from the furnace (see Fig. 4.2). The addition of the higher temperature energy from the furnace increases the steam temperature to that of conventional steam plants [1000°F (811°K)] while allowing the topping cycle condenser to operate at 900°F (756°K). This feature results in an overlapping of the two temperature-entropy

Table 4.11. Combustion Air and Flue Gas System Operating Conditions

Air Inlet		
Weight Flow, lb/hr, kg/s	1,347,600	169.8
Temperature, °F, °K	60	289
Enthalpy, Btu/lb, MJ/kg	0	0
Air From Forced Draft Fans		
Weight Flow, lb/hr, kg/s	1,347,600	169.8
Temperature, °F, °K	86	303
Enthalpy, Btu/lb, MJ/kg	6.4	0.015
Fluidizing Air From Preheater		
Weight Flow, lb/hr, kg/s	1,298,400	163.6
Temperature, °F, °K	800	700
Enthalpy, Btu/lb, MJ/kg	183.4	0.427
Transport Air to Coal and Limestone Feeder		
Weight Flow, lb/hr, kg/s	49,200	6.2
Temperature, °F, °K	90	306
Enthalpy, Btu/lb, MJ/kg	7.2	0.017
Transport Air to Furnace		
Weight Flow		
Air, lb/hr, kg/s	49,200	6.2
Solids, lb/hr, kg/s	196,690	24.8
Temperature, °F, °K	262	401
Enthalpy		
Gas, Btu/lb, MJ/kg	48.9	0.114
Solids, Btu/lb, MJ/kg	44.0	0.102
Flue Gases From Furnace		
Weight Flow, lb/hr, kg/s	1,475,800	185.9
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	219.1	0.510
Flue Gases to Drying System		
Weight Flow, lb/hr, kg/s	194,800	24.5
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	219.1	0.510
Flue Gases From Drying System		
Weight Flow, lb/hr, kg/s	217,100	27.4
Temperature, °F, °K	272	407
Enthalpy, Btu/lb, MJ/kg	58.2	0.135
Flue Gases From Preheater		
Weight Flow, lb/hr, kg/s	1,385,400	174.5
Temperature, °F, °K	275	408
Enthalpy, Btu/lb, MJ/kg	53.5	0.124
Flue Gases to Ash System (After Fan)		
Weight Flow, lb/hr, kg/s	104,400	13.2
Temperature, °F, °K	299	
Enthalpy, Btu/lb, MJ/kg	59.5	0.138
Flue Gases From Ash System		
Weight Flow, lb/hr, kg/s	104,400	13.2
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	219.1	0.510

Table 4.11 (Continued)

Flue Gases From Preheater to Stack		
Weight Flow, lb/hr, kg/s	1,281,000	161.4
Temperature, °F, °K	291	417
Enthalpy, Btu/lb, MJ/kg	57.4	0.134
Flue Gases to Stack		
Weight Flow, lb/hr, kg/s	1,498,100	188.7
Temperature, °F, °K	288	416
Enthalpy, Btu/lb, MJ/kg	57.5	0.134

Table 4.12. Combustion Air and Flue Gas System Pressure Losses, Fan Pressure Ratios and Fan Power Requirements

Pressure Losses		
Air Side	Loss	
	psi	Pa
Air Preheater	0.108	745
Ducts	0.145	1000
Air Distributor Plate	0.675	4654
Fluidized Bed	1.350	9308
	2.278	15707
Gas Side		
Superheater	0.087	600
Reheater	0.147	1014
Cyclones	0.217	1496
Ducts	0.181	1248
Air Preheater	0.054	372
Bag Filters	0.145	1000
Stack	0.070	483
	0.901	6213

Fan Pressure Ratios and Power Requirements

Fan	Pressure Ratio	Power Required (kW)
Forced Draft	1.155	2640
Transport Air	1.017	12
Ash Cooler System	1.107	195
Dryer System	1.048	203
Induced Draft	1.065	1540

diagrams and improves the overall plant efficiency. Also, the transfer of heat to the steam from the high temperature [1750°F (1228°K)] flue gases eliminates the need for a high temperature gas-to-gas heat exchanger and reduced the combustion air temperature entering the furnace. This feature also reduced the size of the cesium boiler and heat transfer surface area in the fluidized bed by 29% for the conditions of the study. This in turn reduced the fluidized bed surface area and furnace size.

4.4.1 Alkali Metal Topping Cycle

The alkali metal topping cycle was analyzed for both cesium and potassium. The range of parameters studied included turbine inlet temperatures of 1390 to 1540°F (1028 to 1111°K) and condenser temperatures of 890 to 1090°F (750 to 861°K). The turbine analysis and the resulting turbine efficiencies used in the parametric cycle analysis appear in Section 3.1.

4.4.1.1. Parametric Cycle Analysis

The results of the topping cycle parametric analysis are given in Tables 4.13, 4.14, and 4.15.

In carrying out both the turbine and the cycle analyses, it was assumed that vapor, for regenerative feed heating, would be bled from the turbine between all stages except the first and second. Furthermore, it was assumed that moisture removal was taken as 25%, i.e., 25% of the moisture in the working fluid at each bleed point is removed with the bleed flow. This value is well within the 20 to 35% efficiency demonstrated on a small potassium turbine tested by General Electric for the space power program.¹⁴

The alkali metal feed heating arrangement is shown in Fig. 4.10. The number of feed heaters used varied from 1 to 3 depending on the difference between the turbine inlet and the condenser temperature. The bleed flow to the respective feed heaters was considered to be composed of the moisture removed plus a quantity of the remaining working fluid. The bleed enthalpy listed in Tables 4.13 and 4.14 is that of the portion of the working fluid which was bled. The enthalpy of the moisture removed

Table 4.13. Cesium Cycle State Points

Turbine Inlet Temp. °F	Condenser Temp. °F	Feed Heater	Bleed Pressure psia	Bleed Temp. °F	Bleed Enthalpy	T ₁ °F	T ₂ °F	T ₃ °F	T ₄ °F	h ₁ Btu/lb	h ₂ Btu/lb	h ₃ Btu/lb	h ₄ Btu/lb	Bleed Flow	Moisture Flow
1540	850	3	3.92	1020	281.0	1005	900	1000	890	159.8	75.7	79.7	75.3	.0180	.0276
		2	8.30	1135	285.5	1120	1010	1115	1000	190.7	80.2	85.9	79.7	.0291	.0265
		1	16.69	1250	290.8	1240	1125	1235	1115	215.3	86.4	93.8	85.9	.0379	.0238
1540	950	2	8.06	1130	284.8	1115	1000	1110	990	188.1	79.7	85.6	79.3	.0285	.0272
		1	16.27	1255	290.3	1235	1120	1230	1110	213.2	86.1	93.5	85.6	.0377	.0246
1540	1050	2	11.12	1185	288.0	1165	1100	1160	1090	182.9	85.1	89.4	84.6	.0218	.0249
		1	19.44	1290	292.0	1270	1170	1265	1160	198.6	90.0	95.6	89.4	.0301	.0227
1440	850	3	3.64	1010	282.8	995	900	990	890	175.4	74.3	79.2	73.9	.0225	.0254
		2	6.90	1135	286.8	1090	1000	1085	990	182.8	79.7	84.3	79.2	.0227	.0242
		1	12.78	1210	291.5	1190	1095	1185	1085	210.8	84.8	90.9	84.3	.0311	.0212
1440	990	2	6.69	1150	286.7	1085	1000	1080	990	180.4	79.7	84.0	79.3	.0215	.0240
		1	12.10	1230	291.5	1185	1090	1180	1080	210.9	84.5	90.6	84.0	.0310	.0210
1440	1090	1	11.12	1185	290.0	1165	1100	1160	1090	190.4	85.1	89.4	84.6	.0230	.0230
1390	890	2	3.92	1020	283.0	1005	900	1000	890	187.6	74.3	79.7	73.9	.0300	.0268
		1	8.30	1135	288.5	1120	1010	1115	1000	197.5	80.2	85.9	79.7	.0288	.0237
1390	990	2	6.08	1085	287.2	1070	1000	1065	990	174.1	79.7	83.2	79.3	.0177	.0223
		1	10.81	1180	291.7	1165	1075	1160	1065	209.6	83.7	89.4	83.2	.0290	.0200
1390	1090	1	11.45	1190	292.1	1170	1100	1165	1090	198.3	85.3	89.7	84.6	.0226	.0198

Table 4.14. Potassium Cycle State Points

Turbine Inlet Temp. °F	Condenser Temp. °F	Feed Heater No.	Bleed Pressure psia	Bleed Temp. °F	Bleed Enthalpy	T ₁ °F	T ₂ °F	T ₃ °F	T ₄ °F	h ₁ Btu/lb	h ₂ Btu/lb	h ₃ Btu/lb	h ₄ Btu/lb	Bleed Flow	Moisture Flow
1540	890	3	1.30	1020	1082.1	1005	900	1000	890	620.5	265.4	283.7	263.8	.0210	.0291
		2	3.10	1135	1099.1	1120	1010	1115	1000	676.7	285.5	305.2	283.7	.0242	.0278
		1	6.99	1260	1119.9	1245	1125	1240	1115	762.5	307.1	329.7	305.2	.0294	.0245
1540	990	3	2.16	1085	1096.8	1070	1000	1065	990	587.3	283.7	295.8	282.0	.0148	.0262
		2	4.50	1190	1111.5	1175	1075	1170	1065	701.7	297.6	316.1	295.8	.0231	.0248
		1	8.83	1300	1129.6	1280	1180	1275	1170	761.1	318.0	336.5	316.1	.0246	.0216
1540	1090	2	4.50	1190	1111.5	1175	1100	1170	1090	649.6	302.3	316.1	300.6	.0177	.0248
		1	8.83	1300	1129.6	1280	1180	1275	1170	761.1	318.0	336.5	316.1	.0246	.0216
1440	890	3	1.01	990	1086.6	980	900	975	890	585.2	265.4	279.1	263.7	.0164	.0271
		2	2.24	1090	1102.4	1075	985	1070	975	650.2	280.9	296.7	279.1	.0198	.0256
		1	4.96	1205	1121.6	1190	1080	1185	1070	753.6	298.6	319.0	296.7	.0264	.0225
1440	990	2	2.24	1090	1102.4	1075	1000	1070	990	629.4	283.7	297.6	281.9	.0178	.0256
		1	4.96	1205	1121.6	1190	1080	1185	1070	745.6	298.6	319.0	297.6	.0253	.0225
1440	1090	2	3.68	1160	1119.8	1145	1100	1140	1090	583.6	302.3	309.9	300.6	.0106	.0211
		1	6.57	1250	1135.3	1235	1150	1230	1140	764.0	312.2	327.2	309.9	.0213	.0183
1390	890	3	0.93	980	1090.0	970	900	965	890	571.5	265.4	277.2	263.7	.0145	.0258
		2	2.00	1075	1105.0	1060	975	1055	965	648.8	279.1	293.9	277.2	.0187	.0243
		1	4.08	1175	1123.2	1165	1065	1160	1055	744.3	295.8	314.2	293.9	.0239	.0212
1390	990	2	2.00	1075	1105.0	1060	1000	1055	990	586.6	283.7	293.9	281.9	.0136	.0244
		1	4.08	1175	1123.2	1160	1065	1155	1055	735.7	295.8	313.2	293.9	.0228	.0211
1390	1090	1	4.08	1175	1123.2	1160	1100	1155	1090	652.0	302.3	313.2	300.5	.0150	.0211

Table 4.15. Summary of Alkali Metal Cycle Analysis

Turbine Inlet Temp. °F	Condenser Temp. °F	No. of Turbine Stages	No. of Feed Heaters	Turbine Efficiency %	Cycle Efficiency %	Net System Efficiency %
<u>Cesium</u>						
1540	890	5	3	78.8	25.4	24.6
1540	990	4	2	78.6	21.3	20.7
1540	1090	4	2	80.7	17.2	16.7
1440	890	5	3	79.2	22.9	22.2
1440	990	4	2	79.7	18.4	17.8
1440	1090	3	1	78.8	14.2	13.8
1390	890	4	2	77.9	21.6	21.0
1390	990	4	2	80.4	17.8	17.3
1390	1090	3	1	81.9	13.0	12.6
<u>Potassium</u>						
1540	890	5	3	77.3	25.3	24.5
1540	990	5	3	79.0	21.8	21.1
1540	1090	4	2	78.9	17.7	17.2
1440	890	5	3	78.7	22.8	22.1
1440	990	4	2	78.6	18.5	17.9
1440	1090	4	2	80.6	14.7	14.3
1390	890	5	3	79.0	21.6	21.0
1390	990	4	2	79.5	17.3	16.8
1390	1090	3	1	79.9	12.9	12.5

Fig. 4.10. Alkali metal system feed heating arrangement.

would be that of saturated liquid at the bleed temperature. The pressure loss between and bleed points and feed heaters was assumed to be a $\Delta P/P$ of 10%. The effectiveness of the feed heat exchangers was taken into account by imposing the following limits on the temperature changes across any one heater

$$T_4 = T_2 - 10^\circ\text{F}$$

$$T_3 = T_1 - 5^\circ\text{F}$$

where the temperatures are defined as shown in Fig. 4.10.

Tables 4.13 and 4.14 give the cycle state points for the cases considered for cesium and potassium. The vapor at the turbine inlet was saturated for all cases. The bleed flow shown is the fraction of the turbine inlet flow extracted for feed heating. The moisture flow is also the fraction of the turbine inlet flow and represents the moisture separated from the working fluid on passing through the turbine. The enthalpy identified as h_1 is the enthalpy of the combined bleed and moisture on entering the feed heater. The cesium pressure and enthalpy for turbine inlet temperatures of 1540, 1440, and 1390°F are 57.65, 38.67, and 31.14 psia and 309.5, 308.3, and 307.6 Btu/lb. Corresponding values for potassium inlet temperatures of 1540, 1440, and 1390°F are 29.58, 18.58, and 14.4 psia and 1194.7, and 1090.5 and 1188.4 Btu/lb. The cesium pressure and enthalpy for condenser temperatures of 890, 990, and 1090°F are 1.44, 3.16, and 6.28 psia and 73.9, 79.2 and 84.5 Btu/lb. Corresponding values for potassium condenser temperatures of 890, 990, and 1090°F are 0.40, 1.01, and 2.24 psia and 263.6, 281.8, and 300.5 Btu/lb.

The results of the alkali metal system analysis are summarized in Table 4.15. The number of turbine stages varies from 3 to 5 and the number of feed heaters used in the system was always 2 less than the number of turbine stages. The feed pump energy requirements were based on a pump efficiency of 70% and a head equal to twice the difference between the boiler and condenser pressures. The net system efficiency was based on assumed mechanical and electrical losses of 3%, i.e., 97% of the turbine shaft power. The efficiency of the system varies between 12 and 25%. However, expressed as a percent of the ideal Carnot cycle the variation was from 74 to 78%. The turbine efficiency varied between 77 and 82%.

In the overall plant analysis, the power output from the alkali metal system is the heat input from the furnace-boiler multiplied by the net system efficiency. The amount of heat rejected from the alkali metal system to the steam system is the input from the furnace boiler multiplied by one minus the cycle efficiency (the use of cycle efficiency rather than the net system efficiency assumes that the mechanical and electrical losses are lost to the plant cycle). It should be noted that this analysis was made to apply to both natural circulation as well as forced circulation systems and does not include the energy loss by the boiler recirculation pumps. The losses for these pumps were listed in Section 4.2.3 and the manner in which these losses are applied to the plant efficiency is given in Section 4.4.1.3.

4.4.1.2 Alkali Metal Turbine Design for the Reference System

The alkali metal vapor turbine analysis was performed to provide information as to the turbine efficiency and its basic configuration and dimensions. A detailed analysis to determine the optimum turbine blade twist, pitch/chord ratio, etc., was not performed for this study. A comparison between cesium and potassium turbines is presented in Section 3.1. This section primarily is concerned with the cesium turbine for the reference design.

The number of stages required for the turbine is a function of the turbine pressure ratio, speed and blading. Due to the high density of cesium vapor a machine of reasonable size can operate at 1800 rpm and produce considerable power. The relatively low speed greatly relieves problems associated with blade and disk stresses and creep. If the turbine speed and size are tentatively specified, the blading severity and sonic velocity essentially establish a maximum enthalpy drop per stage. Specifically, blades which require extreme deflection of the fluid velocity vector generally have high losses. In addition sonic velocity of the relative to the blade, although not disastrous, is usually avoided.¹⁵

From the above considerations, a five-stage turbine was selected. Vapor bleed ports will exist between all stages except the first and second with moisture removal devices located at each bleed location. The

moisture removal devices were assumed to be capable of removing 25% of the liquid alkali metal entrained in the flow through the turbine. This moisture removal capability has been demonstrated experimentally.¹⁴

The power required from the turbine determines its hub diameter and blade height, the speed having been previously designated. This power must include an allowance for the stage efficiency. The primary losses associated with a turbine stage are the result of the following.

1. Stage exit kinetic energy (utilization factor).
2. Blade tip leakage.
3. Fluid flow losses in nozzles.
4. Fluid flow losses in rotor blade passage.
5. Moisture churning.
6. Supersaturation.

The kinetic energy leaving a stage is recoverable by the following stage to a large degree. Hence, only the last stage need be seriously penalized for an exhaust loss. Experimental data are available in the literature^{15,14} to provide information concerning nozzle and blade efficiencies. Blade leakage and other secondary flow losses are apparently more difficult to quantify but have an effect of approximately the same magnitude as the blade loss.¹⁶

Utilizing the data from Refs. 15 and 16, a base machine efficiency of 88% was derived excluding the moisture, supersaturation and exhaust losses. It was desirable to treat the moisture and supersaturation losses as variables since they were somewhat different for cesium and potassium and also varied from stage to stage. A collection of experimental results reported in Ref. 17 indicated that moisture would account for about 0.6 percentage points lost in turbine efficiency for every 1.0% moisture in the flow. The expansion of supersaturated vapor was treated as the expansion of a perfect gas using the isentropic expansion indices reported in Ref. 8. The resulting losses were calculated by assuming that reversion to equilibrium would occur at constant pressure. The stage enthalpy drop chosen for the cesium turbine resulted in about 3.6 percentage points lost in stage efficiency due to the metastable equilibrium expansion.

A detailed velocity diagram and thermodynamic analysis verified that the simplified stage efficiency calculation gave reasonable results which were also found to be in agreement with typical values for large steam turbines.

The annulus area (blade height) was calculated using the vapor density leaving the turbine nozzle to arrive at the volumetric flow rate through a stage. The axial velocity through the turbine was assumed to remain constant except for the final stage where it was increased to avoid extremely long blades. This resulted in a small additional exhaust penalty but it was believed to be necessary from a mechanical standpoint.

The basic features of the turbine are provided in Table 4.16. Although no detailed attention was given to blade profile other than the fluid turning angles, the work and the flow coefficients for the turbine stages are within a practical range indicating that the blading should not be of an uncommon design.¹⁶ Two double flow turbines of this configuration are required for the 200 MW(e) plant. A potassium turbine of comparable size and power output would have to operate at 3600 rpm and would require a somewhat larger annulus area on the later stages.

The stage inlet temperature is that temperature which the nozzle sees. The vapor temperature passing through the rotor is actually closer to the stage exit temperature since the stages are anticipated to be primarily impulse. The blade temperature runs slightly higher than the vapor temperature due to the recovery of temperature as the vapor stagnates on the blade surface. This temperature recovery is proportional to the relative velocity between vapor and blade.

Although a rigorous stress analysis for the turbine was not attempted, based on preliminary results the stress levels were found to lie below the design stresses of current technology materials. The primary materials concern appears to be the maximum disk size which can be produced from the more exotic alloys such as Astroloy which might be required for the initial high temperature turbine stages. The 52 in. (132.1 cm) diameter wheel specified in this design is a small extrapolation of the current maximum of 40 in. (101.6 cm). However, the disk size for the first and second stages of this reference design turbine could have been limited to 40 in. (101.6 cm) without greatly affecting its overall design.

Table 4.16. Cesium Turbine Design

	Stage				
	1	2	3	4	5
Stage inlet temperature, °F	1500	1365	1235	1115	1005
°K	1089	1014	942	875	814
Stage exit temperature, °F	1365	1235	1115	1005	900
°K	1014	942	875	814	756
Stage enthalpy drop, Btu/lbm	14	14	14	14	14.6
kJ/kg	32.56	32.56	32.56	32.56	33.96
Exit quality	.9508	.9302	.9172	.908	.868
Relative mass flow	1.0	1.0	.9396	.8902	.8440
Stage inlet pressure, psia	49.38	27.82	14.64	7.35	3.53
MPa	0.340	0.192	0.101	0.0507	0.0243
Hub diameter, inch	52.0	52.0	52.0	52.0	52.0
cm	132.1	132.1	132.1	132.1	132.1
Blade height, inch	4.25	7.0	11.6	19.0	25.5
cm	10.80	17.78	29.46	48.26	64.77
Tip diameter, inch	60.5	66.0	75.2	90.0	103.0
cm	153.67	167.64	191.01	228.6	261.6
Mean blade speed, ft/sec	442	463	499	557	609
m/sec	135	141	152	170	186
Root centrifugal stress, ¹ psi	3,317	5,729	10,233	18,711	27,412
MPa	22.9	39.5	70.56	129.0	189.0
Max disk stress, ² psi	10,351	12,319	15,992	22,906	30,000
MPa	71.37	84.9	110.3	157.9	206.9
Stage efficiency	0.844	0.8145	0.802	0.794	0.667
Cesium specific volume - nozzle exit	4.56	8.06	15.05	29.27	59.32
Stage power, kW	3,525	3,596	3,425	3,355	2,758

¹Based on untapered blades²Based on uniform disk and blade tip diameter

4.4.1.3. Cesium Reference System Design

The reference design of the cesium system is shown in Figs. 4.11, 4.12, and 4.13. There are two independent cesium power systems. Each system receives heat from one side of the furnace and supplies vapor to a double-flow turbine. The vapor flow from each end of the double-flow turbine passes directly to a cesium condenser-steam generator. The liquid metal sides of the condenser-steam generators in each loop are interconnected; however, the two loops are interconnected. Thus, the topping

Fig. 4.11. Alkali metal power conversion system layout - plan view.

Fig. 4.12. Alkali metal power conversion system layout - elevation view.

Fig. 4.13. Alkali metal boiler recirculating system piping arrangement.

cycle system consists of two independent loops, each supplied vapor from nine cesium boiler modules. Each loop in turn has its own double-flow turbine and two interconnected condenser-steam generators. It should be noted that the steam from the four steam generators is interconnected to provide a single steam system for the plant.

The cesium piping arrangement along one side of the furnace is shown in Fig. 4.13. The two-phase flow from the boiler tube bundle modules feeds a vapor header that runs the length of the furnace. The liquid is separated from the mixture with the vapor passing to the turbine and the liquid returned to the drum that also receives the condensate return from the condenser hotwell. The liquid from the drum is pumped to the nine boiler modules. There is also a small line (identified as a drain line in Fig. 4.13) that is used only for startup. The function of this line is described in Section 4.8.

The design and operating conditions for the system are itemized in Table 4.17. The analytical methods used to obtain these results are the same as for the parametric analysis except that the effect of the boiler recirculation pumps are included. The power input to the system by these pumps is added to the energy input from the furnace. The electrical energy required to drive the motors is not deducted from the cycle efficiency but is included in the overall auxiliary power requirements for the plant.

For the conditions shown in Table 4.17, the turbine work output is 67.28 MW (cycle efficiency of 23.87%) and the generator output is 65.26 MW for a net system efficiency of 23.15%.

4.4.2. Steam System

The steam cycle selected for the plant is an adaptation of an existing design of a 2400 psi (16.5 MPa) system with superheat and reheat of 1000°F (811°K) and a condenser pressure of 1 1/2 in. Hg (5079 Pa). This cycle was designed for a larger system [~700 MW(e)] and was scaled down for this study. The system has six stages of feedwater heating with the feedwater temperature raised to 450°F (506°K). The boiler feed pumps, which pressurize the feedwater to 3140 psia (21.7 MPa), are driven by steam turbines

Table 4.17. Topping Cycle Design and Operating Conditions

Power input		
Furnace, Btu/hr, MW	9.61×10^8	281.6
Recirculating pumps, Btu/hr, MW	1.00×10^6	0.3
Cycle conditions		
Boiler		
Weight flow, lb/hr, kg/sec	1.78×10^7	2245
Inlet pressure, psia, MPa	62.31	0.430
Inlet temperature, °F, °K	1435	1053
Inlet enthalpy, Btu/lb, MJ/kg	105.97	0.246
Exit pressure, psia, MPa	51.71	0.357
Exit temperature, °F, °K	1508	1093
Exit enthalpy, Btu/lb, MJ/kg	160.03	0.372
Exit quality, %	25	
Separator		
Inlet pressure, psia, MPa	51.00	0.352
Inlet temperature, °F, °K	1508	1093
Inlet enthalpy, Btu/lb, MJ/kg	160.03	0.372
Liquid exit flow rate, lb/hr, kg/sec	1.34×10^7	1683
Liquid exit pressure, psia, MPa	51.00	0.352
Liquid exit temperature, °F, °K	1508	1093
Liquid exit enthalpy, Btu/lb, MJ/kg	110.34	0.257
Vapor exit flow rate, lb/hr, kg/sec	4.45×10^6	561
Vapor exit pressure, psia, MPa	50.5	0.348
Vapor exit temperature, °F, °K	1505	1092
Vapor exit enthalpy, Btu/lb, MJ/kg	309.07	0.719
Pump exit conditions		
Pressure, psia, MPa	70.61	0.487
Temperature, °F, °K	1437	1054
Enthalpy, Btu/lb, MJ/kg	105.97	0.246
Turbine		
Inlet pressure, psia, MPa	49.38	0.340
Inlet temperature, °F, °K	1500	1089
Inlet enthalpy, Btu/lb, MJ/kg	309.01	0.719
First bleed weight flow	2.69×10^5	
First bleed pressure, psia, MPa	14.64	0.101
First bleed temperature, °F, °K	1235	942
First bleed enthalpy, Btu/lb, MJ/kg	215.16	0.500
Second bleed weight flow	2.20×10^5	
Second bleed pressure, psia, MPa	7.34	0.051
Second bleed temperature, °F, °K	1115	875
Second bleed enthalpy, Btu/lb, MJ/kg	181.01	0.421
Third bleed weight flow	2.05×10^5	
Third bleed pressure, psia, MPa	3.525	0.024
Third bleed temperature, °F, °K	1005	814
Third bleed enthalpy, Btu/lb, MJ/kg	162.78	0.379

Table 4.17 (continued)

Exit Weight Flow, lb/hr, kg/s	3.76×10^6	473.7
Exit Pressure, psia, MPa	1.56	0.011
Exit Temperature, °F, °K	900	756
Exit Enthalpy, Btu/lb, MJ/kg	269.45	0.627
Exit Quality, %	86.8	
Condenser		
Heat Load, Btu/hr, MW	7.324×10^8	214.6
Temperature, °F, °K	900	756
Third Stage Feed Heater		
Inlet Temperature, °F, °K	902	757
Inlet Enthalpy, Btu/lb, MJ/kg	74.45	0.173
Exit Temperature, °F, °K	985	803
Exit Enthalpy, Btu/lb, MJ/kg	78.90	0.183
Second Stage Feed Heater		
Exit Temperature, °F, °K	1095	864
Exit Enthalpy, Btu/lb, MJ/kg	84.81	0.197
First Stage Feed Heater		
Exit Temperature, °F, °K	1215	931
Exit Enthalpy, Btu/lb, MJ/kg	92.64	0.215
Cycle Efficiency, %	23.87	
Turbine Work, Btu/hr	2.30×10^8	67.28
Net System Efficiency, %	23.15	
Electrical Output, Btu/hr, MW	2.23×10^8	65.26
System Inventory (Total)		
Boiler		
Liquid Headers, lb, kg	4670	2120
Boiler Tubes, lb, kg	39160	17760
Vapor Headers	3240	1470
Recirculating System		
Liquid Lines, lb, kg	29,210	13250
Vapor Lines, lb, kg	11650	5280
Drum, lb, kg	25,000	11340
Vapor Piping and Turbines, lb, kg	2000	910
Condenser and Hotwell, lb, kg	10,000	4540
Liquid Lines and Feed Heaters, lb, kg	15,000	6800
Miscellaneous Lines, lb, kg	5000	2270
Inventory Storage Tanks, lb, kg	25,000	11,340
Total Inventory, lb, kg	169,930	77,080
Inventory Storage Volume, ft ³ , m ³	2000	186

supplied by steam from the discharge of the high pressure turbine. A summary of the key features of this cycle that influence the design of the overall binary plant is given in Table 4.18 and a simplified flow diagram of the system is shown in Fig. 4.14. Note that the values of flow in Table 4.18 are given in lb. of steam per kW hr of output. The heat rate for the cycle is 7760 Btu/kW hr (8.19 MJ/kW hr).

Table 4.18. System Parameters for the Steam Cycle Selected for the Binary Plant

Steam Generator Inlet Conditions		
Feedwater		
Weight Flow, lb/kW hr, kg/kW hr	6.300	2.858
Temperature, °F, °K	450.3	505.7
Enthalpy, Btu/lb, MJ/kg	432.3	1.006
Reheat Steam		
Weight Flow, lb/kW hr, kg/kW hr	5.558	2.521
Pressure, psia, MPa	446	3.08
Temperature, °F, °K	582	579
Enthalpy, Btu/lb, MJ/kg	1292.4	3.006
Steam Generator Outlet Conditions		
High Pressure Steam		
Weight Flow, lb/kW hr, kg/kW hr	6.300	2.858
Pressure, psia, MPa	2415	16.65
Temperature, °F, °K	1000	811
Enthalpy, Btu/lb, MJ/kg	1461.2	3.399
Reheat Steam		
Weight Flow, lb/kW hr, kg/kW hr	5.558	2.521
Pressure, psia, MPa	402	2.78
Temperature, °F, °K	1000	811
Enthalpy, Btu/lb, MJ/kg	1522.4	3.541

Although the basic steam cycle for the binary system of this study is the same as that shown in Fig. 4.14, the equipment arrangement is different. For the binary system the heat rejected from the alkali metal provides the energy to convert the feedwater to steam with about 100°F (56°K) of superheat. The remaining superheat and the reheat energy is obtained from the combustion products in the furnace. The alkali metal condenser-steam generator provides steam at 764°F (680°K) while the

Fig. 4.14. Flow diagram of the steam system selected for the binary plant.

conditions selected for the furnace require that the steam supplied to the steam generator be 739°F (666°K). This is accomplished by bypassing part of the feedwater around the condenser-steam generator to attemperate the steam temperature from the steam generator.

The flow diagram for the steam system for the binary plant is shown in Fig. 4.15 and the key cycle conditions are listed in Table 4.19. Note that the flow rates given here are the total for the plant.

Table 4.19. System Parameters for the Binary Plant Steam Cycle

Alkali Metal Condenser-Steam Generator		
Feedwater In		
Weight Flow, lb/hr, kg/s	871,240	109.8
Temperature, °F, °K	450.3	505.7
Enthalpy, Btu/lb, MJ/kg	432.3	1.006
Steam Out (to attemperator)		
Weight Flow, lb/hr, kg/s	871,240	109.8
Pressure, psia, MPa	2430	16.75
Temperature, °F, °K	764	680
Enthalpy, Btu/lb, MJ/kg	1274.2	2.964
Attemperator		
Feedwater In		
Weight Flow, lb/hr, kg/s	35,860	4.52
Temperature, °F, °K	450.3	505.7
Enthalpy, Btu/lb, MJ/kg	432.3	1.006
Steam Out (to superheater)		
Weight Flow, lb/hr, kg/s	907,100	114.3
Pressure, psia, MPa	2425	16.72
Temperature, °F, °K	739	666
Enthalpy, Btu/lb, MJ/kg	1240.8	2.886
Furnace Superheater Outlet		
Weight Flow, lb/hr, kg/s	907,100	114.3
Pressure, psia, MPa	2415	16.65
Temperature, °F, °K	1000	811
Enthalpy, Btu/lb, MJ/kg	1461.2	3.399
Furnace Reheater		
Reheat Steam In		
Weight Flow, lb/hr, kg/s	800,200	100.8
Pressure, psia, MPa	446	3.08
Temperature, °F, °K	582	579
Enthalpy, Btu/lb, MJ/kg	1292.4	3.006
Reheat Steam Out		
Weight Flow, lb/hr, kg/s	800,200	100.8
Pressure, psia, MPa	402	2.78
Temperature, °F, °K	1000	811
Enthalpy, Btu/lb, MJ/kg	1522.4	3.541

Fig. 4.15. Flow diagram of the binary plant steam system.

4.5 Alkali Metal Condenser - Steam Generator

4.5.1. The Reentry Tube Concept

In order to couple the two portions of the alkali metal vapor/steam binary cycle, a special heat exchanger is required. This unit must act as both a condenser for the alkali metal vapor and a boiler for the steam. Because of the advantages listed below, the reentry tube shown in Fig. 4.16 seems to be ideal for this purpose. The concept of using a double tube arrangement in steam boilers is not new. The reentry tube design was patented in 1831¹⁸ and was successfully used in marine boilers and in the mercury/steam binary cycle power plants of the late 1920's and 1930's.¹⁹ This type of heat exchanger has been proposed for use in other types of power systems (Refs. 20, 21), and the concept has been analyzed and evaluated by several organization (Refs. 22, 23, 24).

4.5.2. Basic Reentry Tube Operation

During normal steady-state operation of the reentry tube, feedwater is admitted to the bottom of the central tube where it boils as it rises to the top. The superheated steam emerging from the top of the central tube then flows vertically downward in the annulus between the inner and outer tubes and exits at the bottom as superheated vapor. The alkali metal vapor will condense on the outside of the outer tube, and a film of liquid metal will flow downward to a sump at the base of the condenser.

The concentric tube design has several advantages over single tube systems. Since boiling occurs in the inside tube, the high thermal stresses in the boiling region will not be aggravated by the presence of pressure stresses. The reentry tube also places a "steam blanket" between the boiling feedwater and the condensing alkali metal, thus avoiding the thermal stresses which would be present if both of these processes took place on opposite sides of a single wall. Installation and support of the concentric tube will also present fewer problems. There will be only one entry point through the header sheets, and the tubes will be free to expand axially.

Fig. 4.16. Section through a reentry tube for the alkali metal condenser-steam boiler.

Because of the high heat transfer rates associated with condensation, the difference between the alkali metal condensing temperature and the steam exit temperature will be small. This will result in a smaller surface area requirement than for the tubes used in a conventional fossil-fuel fired steam generator. The reentry tube is also capable of providing a uniformly high heat flux to the steam with no danger of tube burnout, since the tube walls can get no hotter than the condensing alkali metal.

The steam temperature available from the condenser steam generator will be about 25°F (14°K) higher than that required at the inlet of the steam superheater in the furnace. This excess can be reduced by bypassing part of the feedwater around the condenser steam generator and attenuating the temperature of the steam going to the superheater. This bypass flow can be used to control the furnace energy input distribution between the steam superheater and reheater and the alkali metal boiler, which, in turn, determines the combustion product temperatures downstream of the furnace.

The reentry tube will also provide a means of detecting and controlling leaks between the alkali metal system and the steam system. Since the steam is the high pressure fluid, any reentry tube leak will result in steam entering the tank containing the alkali metal vapor. Since any reaction would be occurring in a large volume of low pressure alkali metal vapor, no serious pressure buildup should occur. Any leak should be detectable as a result of hydrogen generation long before a large amount of reaction has occurred. Once a leak is discovered, the feedwater flow to that bank of reentry tubes can be controlled to prevent further leakage. Even if a leak is not quickly detected, the inventory of steam per tube is so small that the reaction should not occur at too rapid a rate.

4.5.3. Reentry Tube Parametric Analysis

Since the reentry tube is a heat exchange, its operation is governed by a set of heat balances. Equations were developed to analyze the heat loads and to generate temperature profiles for the reentry tube. These equations were coded into a computer program. A complete description of the analysis procedure and the computer code is presented in Ref. 25.

The temperature profile throughout the tube, the tube length and the exit steam temperature were obtained from the computer analysis.

Three different reentry tubes were analyzed in order to determine the effects of design parameter changes on the performance, geometry, and cost of the complete condenser/boiler unit. The dimensions for the three reentry tubes in the table below correspond to schedule 40 pipe.

Size	Inner Tube		Outer Tube		Hydraulic Diameter For Annulus
	ID	OD	ID	OD	
1-1/2 × 2	1.610 in.	1.900 in.	2.067 in.	2.375 in.	0.167 in.
1 × 1-1/2	1.049 in.	1.315 in.	1.610 in.	1.900 in.	0.295 in.
3/8 × 3/4	0.493 in.	0.675 in.	0.824 in.	1.050 in.	0.149 in.

Figure 4.17 shows a typical set of output data for the 3/8 × 3/4 in. tube with a feedwater inlet temperature of 450°F (506°K) and cesium vapor condensing at 900°F (756°K). It should be evident from the figure that the steam exit temperature is relatively constant, regardless of tube length. While this may at first seem surprising, such behavior is not unreasonable since the tube must operate between the temperature limits set by the condensing alkali metal and the inlet feedwater. These temperature limits and flow conditions essentially define the steam exit temperature. As can be seen in Fig. 4.17, any increase in length (and thus surface area) is offset by a smaller temperature difference between the alkali metal and the steam at the top end of the tube. In other words, the degree of superheat at the top of the tube and the tube length are uniquely related.

For the sake of consistency all three of the above tube configurations were analyzed by assuming 20°F (11°K) superheat at the top of each tube. The computer results are summarized in Figs. 4.18 and 4.19. Fig. 4.18 shows only a slight variation of steam exit temperature with changes in the feedwater flow rate. Again, the steam exit temperature is defined by the temperature limits and flow conditions. Any increase in feedwater flow rate must then result in increased tube length (as shown in Fig. 4.19)

Fig. 4.17. Reentry tube temperature profiles for pipe sizes of 3/8 x 3/4 in. and a flow rate of 600 lb/hr.

Fig. 4.18. Reentry tube performance for 20°F (11°K) superheat from central tube, 450°F (506°K) feedwater and 900°F (756°K) condensing alkali metal.

Fig. 4.19. Reentry tube geometry requirements for 20°F (11°K) superheat from central tube and a heat load of 7.5×10^8 Btu/hr.

in order to allow enough time for the steam to reach the "defined" exit temperature.

In an effort to identify the relative cost for the complete condenser/boiler unit, the weight of the tube matrix was computed. Based on the cross-sectional area of the tubes, A_x , the tube length, L , and the number of tubes, N , Fig. 4.20 was developed. The density of the tube material was taken as 494 lb/ft^3 (7914 kg/m^3) so that the weight for the total unit was expressed as

$$W = 494 A_x L N.$$

From the figure it is more economical to use smaller diameter tubes with higher flow rates, even though the higher flow rates require longer tubes.

Fig. 4.20. Reentry tubing total weight for 20°F (11°K) superheat from central tube and a heat load of $7.5 \times 10^8 \text{ Btu/hr}$.

4.5.4. Cesium Condenser - Steam Boiler Reference Design

For ease of fabrication, installation, maintenance and replacement, the condenser/boiler unit uses a modular design. The headering arrangement for the reentry tubes is shown in Fig. 4.21. Each one of the header drums contains 31 reentry tubes. The arrangement of seven of these drums to form one 9 ft (2.75 m) diameter condenser/boiler unit is also shown in Fig. 4.21. Since 4 of these units are to be incorporated into the overall plant design, there will be 868 individual reentry tubes. For the $3/8 \times 3/4$ in. reentry tube, this yields a feedwater flow rate of 1004 lb/hr (0.126 kg/s) for each tube. The length corresponding to this tube configuration and flow rate is 46 ft (14 m). The temperature profile through one reentry tube is shown in Fig. 4.22. Other relevant parameters for the above reference design are listed in Table 4.20.

From considerations of tube stresses and operating temperatures, Croloy should be a satisfactory material for the outer tubes. The only question in regard to its use in the possibility of problems with carbon depletion. It is interesting to note that the Liquid Metal Fast Breeder Reactor program allows the use of Croloy for up to 900°F in their liquid sodium system.

Since there is a lack of pressure stress on the inner tube, carbon steel could be used for the tube material. Any problems with thermal stresses on either the inner or outer tube could be effectively eliminated by the addition of insulation sleeves.

4.5.5. Reentry Tube Operating Conditions

Operation of the binary cycle power plant will not be limited entirely to the reference design conditions, particularly during a plant turndown. As the flow of cesium vapor from the turbine is reduced, the flow of steam must also be cut back. Since there is a distinct relation between these two flow rates and the heat load dissipated through the condenser/boiler, the operating temperatures of the reentry tube are also likely to change.

Figure 4.23 gives the variation in steam exit temperature as operating conditions of the reentry tube change. Once again there is limited

Fig. 4.21. Incorporation of the reentry tube header drum into a complete condenser-steam boiler unit.

Fig. 4.22. Reentry tube temperature profile for $3/8 \times 3/4$ in. tube at reference design conditions.

Fig. 4.23. Reentry tube steam exit temperature as a function of steam flow rate per tube and cesium condensing temperatures.

Table 4.20. Cesium Condenser-Steam Generator Reference Design

Reentry Tube Configuration		
Tube Size	3/8x3/4 in., Schedule 40 pipe	
Inner Tube, ID, in., cm	0.493	1.25
Inner Tube, OD, in., cm	0.675	1.71
Outer Tube, ID, in., cm	0.824	2.09
Outer Tube, OD, in., cm	1.050	2.67
Hydraulic Diameter of Annulus, in., cm	0.149	0.378
Tube length, ft, m	46.08	14.05
Number of Reentry Tubes per Header Drum	31	
Number of Header Drums per Unit	7	
Number of Units for Entire Plant	4	
Total Number of Reentry Tubes	868	
Individual Reentry Tube Weight, lb, kg	100.5	45.6
Total Weight for All Tubes, lb, kg	87,280	39,580
Condensing Cesium Conditions		
Cesium Condensing Pressure, psi, MPa	1.56	0.011
Cesium Condensing Temperature, °F, °K	900	756
Cesium Heat of Vaporization, Btu/lb, MJ/kg	224.81	0.523
Total Cesium Inlet Flow, lb/hr, kg/s	3.76×10^6	473.7
Cesium Inlet Quality, %	86.8	
Condensate Mass Flow per Reentry Tube, lb/hr, kg/s	3,748	0.472
Total Condensate Mass Flow, lb/hr, kg/s	3.254×10^6	410
Maximum Temperature Drop Through Condensate Film, °F, °K	12.9	7.2
Location,* ft, m	46.08	14.05
Minimum Temperature Drop Through Condensate Film, °F, °K	6.3	3.5
Location,* ft, m	12.0	3.6
Maximum Condensate Film Thickness, in, mm	0.0203	0.515
Average Condensate Film Thickness, in, mm	0.0162	0.411
Condensate Inventory per Tube, lb, kg	1.72	0.78
Total Condensate Inventory, lb, kg	1,494	677.6
Individual Reentry Tube Heat Load, Btu/hr, W	842,700	246,800
Total Condenser/Boiler Heat Load, Btu/hr, W	7.324×10^8	214
Feedwater/Steam Conditions		
Feedwater Flow Rate per Tube, lb/hr, kg/s	1004	0.126
Total Feedwater Flow Rate, lb/hr, kg/s	871,240	109.8
Feedwater Inlet Temperature, °F, °K	450.3	505.7
Feedwater Inlet Enthalpy, Btu/lb, MJ/kg	432.3	1.006
Feedwater Inlet Velocity, ft/s, m/s	4.0	1.2
Temperature at Top of Tube, °F, °K	680	633
Superheat at Top of Tube, °F, °K	17.9	9.9
Enthalpy at Top of Tube, Btu/lb, MJ/kg	1153.4	2.683
Feedwater Velocity at Top of Inner Tube, ft/s, m/s	34.4	10.5
Annulus Steam Velocity at Top of Tube, ft/s, m/s	37.4	11.4
Steam Exit Temperature, °F, °K	764	690
Steam Exit Enthalpy, Btu/lb, MJ/kg	1274.2	2.964
Steam Exit Velocity, ft/s, m/s	50.8	15.5
Maximum Steam Temperature, °F, °K	813	707
Location,* ft, m	23	7
Location for Start of Feedwater Saturated Boiling,* ft, m	32.0	9.8
Location of Boiling Feedwater Dryout,* ft, m	23.0	7.0
Feedwater Flow Quality at Dryout (Percent Steam)	40	
Location for Finish of Saturated Boiling,* ft, m	7.0	2.1
Pressure Drops Through Reentry Tube		
Steam Exit Pressure, psi, MPa	2400	16.5
Pressure Drop in Annulus, psi, MPa	63.4	0.44
Pressure Drop at Top of Tube Turn-Around, psi, MPa	4.1	0.03
Pressure Drop in Central Tube, psi, MPa	13.1	0.09
Combined Pressure Drop for Tube Inlet and Exit, psi, MPa	1.0	0.007
Total Pressure Drop Through Reentry Tube, psi, MPa	81.6	0.56
Temperature Drops Through Tube Walls		
Inner Tube Wall Maximum Temperature Drop, °F, °K	96.9	53.8
Location,* ft, m	46.08	14.05
Inner Tube Wall Minimum Temperature Drop, °F, °K	4.1	2.3
Location,* ft, m	0.0	0.0
Outer Tube Wall Maximum Temperature Drop, °F, °K	114.5	63.6
Location,* ft, m	0.0	0.0
Outer Tube Wall Minimum Temperature Drop, °F, °K	36.0	20.0
Location,* ft, m	22.0	6.7

*Location refers to distance from top of reentry tube

influence of feedwater flow rate on the temperature of the steam at the reentry tube exit, even if the cesium condensing temperature shifts.

The heat loads for a single reentry tube are shown in Fig. 4.24. Based on these heat loads, Fig. 4.25 shows the relation between cesium condensate flow and steam flow rate. As stated above, any cutback in cesium vapor flow from the turbine must be accompanied by a reduction in steam flow rate. The temperature of the condensing alkali metal will be determined largely by the topping cycle flow conditions and the potential for heat transfer defined by the steam flow conditions.

Operation of the reentry tube at steam flow rates higher than those indicated by the dashed line in Figs. 4.23, 4.24 and 4.25 will result in incomplete boiling of the feedwater inside the central tube. This causes wet steam to "spillover" into the annulus. Because one advantage of using the reentry tube is the reduction of tube wall thermal stress in the boiling region, "spillover" should be avoided. The emergence of wet steam from

Fig. 4.24. Individual reentry tube heat load as a function of steam flow rate per tube and cesium condensing temperature.

Fig. 4.25. Cesium condensate flow rate per tube as a function of steam flow rate per tube and cesium condensing temperature.

the top of the central tube presents such a unique set of flow patterns and heat transfer rates that an accurate prediction of reentry tube performance during "spillover" may not be possible.

4.5.6. Comparison of Condensing Cesium and Potassium

Use of potassium instead of cesium makes very little difference as far as the steam generation capabilities of the reentry tube are concerned. Since liquid potassium has twice the thermal conductivity as cesium at the 900°F (756°K) condensing temperature, the outside tube wall temperature is

higher for the condensing potassium. This gives slightly better heat transfer to the steam in the annulus. In comparison to the 764°F (690°K) steam exit temperature for the 46.08 ft (14.05 m) reentry tube of the reference design, a condensing potassium system gives a steam exit temperature of 767°F (682°K) and a tube length of 45.5 ft (13.9 m). Both the cesium and potassium lose almost the same amount of heat during condensation at reference design conditions, but due to the differences in their respective heats of vaporization, 3750 lbm/hr (0.472 kg/s) of cesium are condensed compared to only 950 lbm/hr (0.120 kg/s) of potassium. While these are significant differences with respect to the alkali metal portion of the system, the steam portion of the binary cycle is unaffected by the choice of the condensing alkali metal.

4.6. Coal and Limestone Preparation and Feed System

The coal and limestone preparation and feed system is designed to process run-of-the-mine material to the desired particle size and to convey this feed to the atmospheric fluidized bed combustor. The handling and preparation process graphically illustrated in Fig. 4.26 includes the receiving, storage, crushing, and conveying of the coal and limestone to

Fig. 4.26. Coal and limestone preparation plant.

the plant storage hoppers. The coal and limestone arrive at the plant by unit trains, are discharged into receiving hoppers, and transferred over magnetic head pulleys (to remove tramp iron) to an open pit storage system. The storage consists of a 3-day active pile and a 60-day full-load dead storage pile.

The coal and limestone are conveyed from the stockpile to the transfer house where the run-of-the-mine material is reduced in size by a crusher prior to being conveyed to either the coal or the limestone dryer. From the dryer the material is conveyed to a hammer mill where it is ground to minus 1/8 in. (0.32 cm) + 0 size. The product from the hammer mill is classified to insure that particles larger than 1/8 in. (0.32 cm) are separated from the finished product is conveyed to storage tanks for distribution to the furnace cell feed tanks. Each of the coal and limestone storage tanks is sized to hold a 1-day full-load supply.

The complete coal and limestone preparation and feed system is shown schematically in Fig. 4.26. The system is supplied with flue gas from the fluidized bed combustor at a temperature of approximately 900°F (756°K) which is used to dry the coal and limestone sufficiently to allow pneumatic transport of the feed materials. The system is also supplied with air from the inlet air compressor at a temperature of approximately 90°F (306°K) which is used to pneumatically transport the coal and limestone feed to the fluidized bed combustor. The integration of the coal and limestone preparation and feed system is shown in Fig. 4.1.

The coal and limestone are fed to a common feed line from which the mixture is fed to the furnace cell feed tanks as shown in Fig. 4.27. The ratio of limestone to coal in the mixture is determined by the sulfur analysis of the coal feed stock. From the furnace cell feed tanks the coal and limestone mixture is pneumatically transported to the individual furnace cell feed points. The main furnace cells each have 36 feed points and the burnup cells have 18 feed points. The division of the feed streams is made in a series of machined splitter blocks that either divide the stream into 2, 3, or 4 parts. For the main cells a 4-3-3 series split is used to subdivide the main flow into 36 equal parts and a 2-3-3 series split is used to supply the 18 combustor cell feed points. Preliminary

Fig. 4.27. Schematic diagram of the complete coal and handling, preparation, and feed system.

testing of prototype splitter geometries indicate that a flow split of $\pm 5\%$ can be consistently obtained under operating conditions suitable for the feeding of the fluidized bed furnace.

The carbon combustor cell requires that the elutriated material from the main furnace cells be recycled through the cell in order to reduce the overall carbon loss from the system. This material is separated from the main cell flue gas stream with cyclone separators and pneumatically fed to the burnup cell through lances that penetrated the side wall of the cell at locations between the air tuyeres and the heat transfer matrix. The burnup cell is also fed with sufficient coal and limestone mixture to maintain the cell at its operating temperature level.

Shown in Table 4.21 are the power requirements for the coal and limestone preparation and feed system. The thermal feedback column in Table 4.21 represents that part of the electrical power requirements that enter the system. This feedback is the thermal input to the fuel and limestone feed system and thus heat input to the furnace. The operating conditions for the system are given in Table 4.22.

4.7 Ash Handling and Cooling System

The ash and spent sorbent are removed from the atmospheric fluidized bed combustor by cyclone dust collectors located in the bed exit flue gas stream, by bottom ash withdrawal ports located in each of the furnace cells and by bag filters located in flue gas exhaust lines feeding the plant exhaust stack.

The elutriated material from the main furnace cells is separated from the main cell flue gas stream with cyclone separators and is fed to the recycle cells to reduce the overall carbon loss from the system. The elutriated material from the recycle cells is separated from the recycle cell flue gas stream by a cyclone separator. The solid material from this separator is fed to an ash cooler as shown in Fig. 4.28. The solids in the flue gas streams from the cyclone separators are removed by the bag filters located upstream of the plant exhaust stack.

The cell bottom ash withdrawal stream is also shown in Fig. 4.28. Ash is withdrawn from each furnace cell at a rate sufficient to maintain a constant bed height in the cell as determined by the bed pressure drop.

Table 4.21. Power Requirements for the Coal and Limestone Preparation and Feed System

	Motor Size hp	Power Requirements kW(e)	Thermal Feedback* kW(t)
Track hopper, collecting and reclaim conveyers			
Coal	20	14.9	
Limestone	12	9.0	
Transfer house			
Coal	20	14.9	
Limestone	10	7.5	
Dryer			
Coal	87	64.9	45
Limestone	50	37.3	26
Hammer mill, classifier and storage pneumatic conveyer			
Coal	394	293.9	237
Limestone	228	170.1	138
Reclaiming cyclones and dust control			
Coal	175	130.6	
Limestone	65	48.5	
Transport Pumps			
Coal	150	111.9	104
Limestone	90	67.1	62
Furnace cell feed circulation fan and rotary valves			
	290	216.3	204
Exhaust fan			
	272	203.0	—
TOTAL	1863	1390.0	816

*This column represents the thermal energy absorbed by the coal and limestone and returned to the furnace via the feed system.

Table 4.22. Coal and Limestone Preparation, Handling, and Feed System Operating Conditions

Coal and Limestone In		
Weight (Dry), lb/hr, kg/s	196,690	24.78
Temperature, °F, °K	60	289
Enthalpy, Btu/lb, MJ/kg	0	0
Water In		
Weight, lb/hr, kg/s	22,340	2.81
Temperature, °F, °K	60	289
Enthalpy, Btu/lb, MJ/kg	-1060	2.466
Flue Gases In		
Weight, lb/hr, kg/s	194,800	24.5
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	219.1	0.510
Thermal Input to System		
Btu/hr, kW	3.67×10^6	1075
Coal and Limestone to Transport System		
Weight, lb/hr, kg/s	196,690	24.78
Temperature, °F, °K	316	431
Enthalpy, Btu/lb, MJ/kg	54.37	0.127
Exhaust Fan Inlet		
Weight, lb/hr, kg/s	217,100	27.4
Temperature, °F, °K	261	401
Enthalpy	55.14	0.128
Exhaust to Stack		
Weight, lb/hr, kg/s	217,100	27.4
Temperature, °F, °K	272	407
Enthalpy, Btu/lb, MJ/kg	58.18	0.135

Fig. 4.28. Ash and Spent Sorbent Removal System.

The ash withdrawal port contains an internal spiral ramp whose pitch is designed so that the angle of repose of the bed material will not allow the bed material to flow down the ramp. The bed material is caused to flow at a modulated rate by a countercurrent fluidizing air stream supplied by the feed injection fan. The cell ash is collected in a cell ash hopper and fed through a rotary valve to a pneumatic transport system for delivery through a cyclone separator to the ash cooler. Flue gas at approximately 275°F (408°K) from the combustion air preheater is used as the transport gas for the bottom ash removal system and as a coolant for the total removed solids. The transport flue gas and the flue gas from the ash cooler are combined upstream of the ash cooler cyclone separator. The combined streams leave the separator at a temperature of approximately 900°F (756°K) and join the flue gas stream from the recycle cell cyclone separator.

The ash and spent sorbent from the ash cooler at approximately 400°F (478°K) and from the bag filters at approximately 275°F (408°K) are transported by covered conveyor belts to surge bins at the rail spur for off-site disposal.

The operating conditions for the system are given in Table 4.23. As noted in Section 4.3 the weight flow and energy balances with the flue

Table 4.23. Ash Handling and Cooling System Operating Conditions

Flue Gas In		
Weight, lb/hr, kg/s	104,400	13.2
Temperature, °F, °K	275	408
Enthalpy, Btu/lb, MJ/kg	53.5	0.124
Flue Gases from Fan		
Weight, lb/hr, kg/s	104,400	13.2
Temperature, °F, °K	299	422
Enthalpy, Btu/lb, MJ/kg	59.5	0.138
Solids in (From Cyclone)		
Weight, lb/hr, kg/s	29,000	3.76
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	190.3	0.443
Solids in (From Bed Drain)		
Weight, lb/hr, kg/s	35,610	4.49
Temperature, °F, °K	1750	1228
Enthalpy, Btu/lb, MJ/kg	421.8	0.981
Particulate in Flue Gases		
Weight, lb/hr, kg/s	3,020	0.38
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	190.3	0.443
Flue Gases Out		
Weight, lb/hr, kg/s	104,400	13.2
Temperature, °F, °K	900	756
Enthalpy, Btu/lb, MJ/kg	219.1	0.510
Solids Out		
Weight, lb/hr, kg/s	68,410	8.62
Temperature, °F, °K	400	478
Enthalpy, Btu/lb, MJ/kg	67.5	0.157

gases were based on the gases only and neglected the particulate which amounted to only a few tenths of 1% of the total flow. In order to get a heat balance on the total plant, the ash cooling system is assumed to cool all of the ash to 400°F (478°K). Thus, the energy of the 3020 lb/hr of particulates is added to this loop.

4.8 Plant Startup and Control

4.8.1. Startup Procedure

Although there are numerous details concerning the starting procedure for the entire power system which have not been examined in detail, a tentative procedure is outlined for firing the fluidized bed furnace. Determinations of the size of the startup system and its degree of complexity in addition to gaining an understanding of the transient phenomena of startup were judged to be significant elements in the reference design.

4.8.1.1. Outline of Startup Procedure

Startup will be accomplished by fluidizing one of the main cells for each half of the furnace and heating these startup cells via kerosene burners until an adequate temperature has been reached to initiate a self-sustaining coal combustion. The startup cells are located at the end of the furnace opposite the recycle cells and cyclone separators (see Fig. 4.3). Liquid alkali metal will be circulated through the boiler tubes in this cell only, all other liquid feed valves being closed as shown in Fig. 4.29. The liquid metal temperature (and pressure) will increase as the bed is heated and boiling or flashing of this liquid should occur as it dumps into the large vapor line. The vapor line to the alkali metal turbine is closed during startup. The pressure difference established in the alkali metal lines as a result of the temperature gradient between the startup cell and cold static beds will cause any vapor formed to flow to the colder parts of the furnace. The subsequent condensation of the vapor in the extremities of the boiler system will effect the heating of the remainder of the boiler and furnace. The tight tube spacing in the beds practically insures that the temperature of the bed material (limestone) will closely follow that of the boiler tubes. The condensate is

Fig. 4.29. Alkali metal boiler recirculation system piping arrangement showing valve positions for startup.

free to drain back into the liquid drum through a drain line (see Fig. 4.29).

After coal ignition has been achieved in the startup cell, the remaining beds will be individually fluidized and injected with coal. The combustion air will be preheated via the system air preheater to aid in igniting the remaining beds in addition to the startup cell.

4.8.1.2. Summary of Analytical Startup Model

A simplified analysis was performed to both quantify the startup and verify that this procedure has merit. The heating of the startup cell is to be accomplished by preheating the combustion air to about 800°F (700°K) via a gas or kerosene burner. The air mass flow would be adjusted to provide a superficial velocity of 3.0 ft/sec (0.91 m/sec) at the tuyere plate at 800°F (700°K). When the material in the startup cell reaches about 600°F (589°K), which is well above the kerosene boiling point, kerosene will be injected directly into this fluidized bed through tubes located at the bottom of the bed. The kerosene would self-ignite [the ignition temperature of kerosene is less than 500°F (533°K)] and would complete the temperature rise in the startup bed up to about 1100°F (866°K) at which temperature coal can be fired in the bed.

For analytical purposes, the heat from the kerosene burners can be visualized to be distributed as follows:

1. The heat loss to the environment by the four main cells and one recycle cell for one-half of the furnace.
2. The hot flue gas leaving the single fluidized startup cells.
3. The heat capacity of the boiler and bed material in the main cells and recycle cell.

A schematic representation of the startup cell and energy flow is shown in Fig. 4.30. The primary heat loss mechanism in the static beds was resolved to be radiation. The losses associated with the flue gases were based on air properties and the assumption that the gases leave the

Fig. 4.30. Schematic diagram of startup cell.

bed at the bed temperature. The temperature rise of the boiler and limestone was calculated by using a lumped heat capacity approach. The principal internal resistances to heat transfer are the condensation rate of the alkali metal vapor and the conductivity of the packed limestone bed. Since condensation rates are generally high and the tube spacing in the beds is tight, the lumped capacity approach was judged to be valid. The rate of condensation per unit area of boiler tube surface was found to be well within the realm of feasibility.

Preliminary calculations showed that a sufficient pressure difference could exist through the vapor line between the startup cell and cold beds to create a sonic block (choked flow) of vapor, thus limiting the rate at which energy could be transported to the cold beds. One-dimensional, isentropic, compressible flow relations were used to model this transient phenomena. It was found that the vapor line would be choked for a considerable time during the startup period. In the analytical model, the

heat to the combustion air was assumed to be instantaneously applied and constant. The heat supplied by the kerosene burners in the startup cell, also a constant, was modeled as being instantaneously added when the temperature of the cell reached 600°F (589°K).

The transient response of the furnace during startup is shown in Fig. 4.31. The IBM CSMP computer package was utilized in obtaining the solution to the differential equations which described the response of the system. Although the properties of cesium vapor were used to obtain the results presented in Fig. 4.31 similar calculations using potassium properties revealed that the response did not deviate significantly.

Fig. 4.31. Transient temperatures during furnace startup (for cesium).

The temperature of the startup cell is seen to rise quite rapidly during the initial heating. The extremely low density of the alkali metal vapor at low temperatures results in a relatively small transport of energy away from the startup cell. Hence, there is little heating accomplished in the packed beds since only a small amount of vapor is available for condensation. The limited energy supply from the kerosene burners causes the startup cell temperature to remain constant for a time until the static beds "catch up" in temperature. The sonic block in the vapor line is not lost until the temperature difference between the static cells and startup cell is less than about 60°F (33°K). The temperature increase of the furnace wall is the result of the radiation heat exchange modeled between the wall and the static beds.

4.8.1.3. Fuel Supply for Startup

The supply of fuel required to bring the startup cell to the coal ignition temperature was calculated by integrating the heat supplied by the burners with respect to time. Since there is some question as to the minimum temperature for coal ignition, the results of the calculations are shown parametrically in Fig. 4.32. The probable required temperature range is 1000 to 1200°F (811 to 922°K).

The fuel supply shown here is an upper limit since, in the calculation, the heat recovered from the flue gas via the air preheater was not included. Further reduction in fuel usage could result from using the smaller recycle bed to start the system since the flue gas losses would be reduced. However, even greater uncertainties than with the present system arise regarding the flow of vapor and subsequent energy flow throughout the boiler.

4.8.2. Plant Control

This section reports the turndown characteristics of the plant and some control options available. The results shown represent a first approximation of the system response and are not intended to represent a complete and rigorous analysis.

Fig. 4.32. Kerosene required to heat startup cell to coal ignition temperature.

4.8.2.1. Major Control Variables

A better understanding of the control of the system and the major variables which affect the control can be gained by first examining the means by which operation at design conditions is established. The operating saturation temperature of the alkali metal boiler is established by adjusting the coal and air feed rates until the desired operating point is reached. The alkali metal turbine can be visualized as a converging nozzle operating between the boiler pressure and condensing pressure. The condenser pressure is established by the specific design of the reentry tube unit and the mass flow rate of the water circulated through the condenser-steam generator. Hence, the heat demand of the steam system actually controls the alkali metal condensing temperature. The design of the turbine is such that the proper amount of cesium vapor will be provided at the condenser to satisfy the energy balance at the condenser temperature.

The steam turbine inlet temperature is set by attemperating the steam exiting the steam generator. The exit pressure of the steam turbine (steam condenser pressure) is established by the temperature and flow rate of the cooling water circulated through the steam condenser.

A significant reduction in power output can be achieved by decreasing the fuel flow to the furnace. A reduction in bed combustion temperature and the saturation temperature will follow with less vapor subsequently flowing to the alkali metal turbine. The alkali metal condenser operating temperature would still be somewhat flexible since the water flow rate can be varied as discussed later. It would be undesirable to allow the condenser temperature to drop significantly due to the extremely large volumetric flow of cesium vapor which would result. Turndown of the steam system is accomplished in part by a reduction of feedwater flow as mentioned above. Power reduction is also achieved by throttling the steam turbine.

There are two methods for varying the steam flow rate through the alkali metal condenser. For small changes in steam generator exit conditions part of the feedwater can bypass the reentry tube unit. For turndown applications, the feed pump flow must be reduced. Both variables

may require change in order to arrive at a desirable combination of alkali metal condensing temperature and steam turbine inlet temperature. The superheat-reheat conditions can be controlled somewhat by attemperation, but the furnace turndown will have a direct effect since most reheat and superheat are done by the furnace gases. In general, the control of the steam cycle is anticipated to be accomplished satisfactorily by conventional methods.

The sulfur retention characteristics of the fluidized bed dictate that the temperature of the bed be no lower than about 1400°F (760°C), hence further power reduction may require that the fuel flow to a section of the bed be stopped entirely. The alkali metal will continue to circulate through the tubes in this section at a reduced rate to assure that the bed temperature remains high enough that restarting would present little difficulty. Both slumping a bed section and simply reducing the fuel flow to the entire bed have the same effect of lowering the alkali metal saturation temperature, i.e., the turbine inlet temperature, and the alkali metal flow rate.

4.8.2.2. Analytical Approach to Control

The fundamental energy and heat transfer relations can be utilized to quantify the variables used to control the system and to provide perhaps a clearer picture of how the system responds to changes in major variables.

In the alkali metal condenser-steam generator there exists a relation between the alkali metal temperature, the reentry tube geometry, and the steam exit conditions which may be hypothetically expressed as

$$\int_A U_{oa} (T_c - T_s) dA \propto \dot{m}_s (T_{ex} - T_i) . \quad (1)$$

where

U_{oa} = overall heat transfer coefficient

T_c = alkali metal condensing temperature

T_s = steam temperature

A = surface area

$\dot{m}_s$ = steam flow rate

T_{ex} = exit steam temperature

T_i = inlet feedwater temperature to reentry tube

The left-hand side of the equation represents the heat transfer over the reentry tube surface area and the right-hand side is the energy change of the steam flowing through the tube. The computer analysis for the reentry tube steam generator actually provides this relationship. There must also exist an overall energy balance between the condensing alkali metal and the steam which may be expressed as

$$\dot{m}_{cs} h_{fg}(T_c) \propto \dot{m}_s (T_{ex} - T_i) . \quad (2)$$

where

$\dot{m}_{cs}$ = mass flow of cesium condensate

h_{fg} = cesium enthalpy of vaporization, a function of T_c

which indicates that there must be a certain flow rate of cesium condensate to satisfy the energy balance in addition to the condition set forth by Eq. (1).

Furthermore, the cesium flow rate is dictated by the turbine design and the pressure ratio across the turbine. This function may be written simply as

$$\dot{m}_{cs} = f(P_i, P_c) .$$

where

P_i = cesium turbine inlet pressure

P_c = cesium condensing pressure.

However, since the inlet pressure depends primarily on the energy input to the alkali metal boiler and the condenser operates in the saturation region,

$$\dot{m}_{cs} = f(\dot{Q}, T_c) . \quad (3)$$

where

$\dot{Q}$ = boiler heat input

Now for given values of $\dot{Q}$, $\dot{m}_s$, and h_i there are the unknown quantities T_c , T_{ex} , and $\dot{m}_{cs}$. Hence, with the three relations above, a unique solution must exist which would describe the off-design operating conditions. It is apparent that a change in one of the independent variables can produce a change in any or all of the dependent variables. These three relations can be presented graphically to facilitate describing the off-design operation of the system.

4.8.2.3. Results of Preliminary Off-Design Analysis

The effects of turbine inlet temperature and condenser temperature for the alkali metal system were parametrically investigated. The topping cycle operation can be completely described if these two points are known.

The alkali metal condenser temperature was found to have a very small effect on the cesium flow rate for a given turbine inlet temperature. It is the boiler heat flux which has the most pronounced effect on the cesium flow since the boiler heat input directly affects the turbine inlet temperature. The relation between boiler power and cesium flow is illustrated in Fig. 4.33. No notable difference was found for the range of condenser temperatures listed.

Having the boiler power and turbine inlet temperature, the necessary fluidized bed combustion temperature can be calculated. The turndown of the system power can also be approximated if the following assumptions are employed.

1. The energy distribution between steam and alkali metal in the furnace is held constant during turndown.
2. The efficiency of the steam cycle is not greatly affected during off-design operation.

The resulting relation between the turndown and furnace temperature is shown in Fig. 4.34. An allowance was made for the pressure drop between the cesium boiler and turbine inlet. Of significance is the fact that over 50% turndown can be accomplished by a simple reduction in the coal feed rate to the furnace as a whole. Hence, shutdown of a section of the

Fig. 4.33. Energy to the alkali metal system as a function of the turbine inlet temperature and cesium flow rate.

fluidized bed should not be required during normal operation. The bed combustion temperature at 50% load is still such that high sulfur retention can be accomplished. Since the steam system efficiency will drop during off-design operation, the actual turndown would be slightly greater than that shown in Fig. 4.34. Apparently, turndown of well over 50% is possible without lowering the bed temperature below 1400°F (760°C). To maintain proper fluidization, however, the air-fuel ratio may have to be increased.

Fig. 4.34. Fluidized bed and turbine inlet temperatures as a function of turndown.

The condenser behavior and performance during off-design operation can be visualized by referring to Fig. 4.23 and 4.25 in section 4.5. The relation between the cesium vapor flow rate leaving the turbine and the steam mass flow rate circulating through the reentry tube unit is shown

in Fig. 4.25. The steam flow rate must be reduced accordingly as the furnace power and, subsequently, the cesium flow rate are reduced. Otherwise either the condenser temperature will drop or liquid (water) will spill over into the annular region of the reentry tube unit. Of course, if this occurs, the principal advantage of the reentry tube design will have been lost.

The effect of the reduction of steam mass flow on the exit steam temperature is illustrated in Fig. 4.23. The exit steam temperature is seen to be practically independent of the steam flow rate provided that the cesium condensing temperature and feedwater temperature do not vary significantly. However, for a designated tube length, the steam exit temperature is not highly sensitive to changes in the feedwater temperature and large variations in feedwater temperature are not likely to result from a system turndown.

4.8.2.4. Summary of System Control

The control of the binary vapor cycle does not appear to present difficulties significantly greater than with a conventional steam power plant. The fluidized bed furnace fuel and air feed rates are the primary control variables for the system. The cesium condenser-steam generator temperature can be controlled to a degree but has little effect on the off-design output of the system. The steam flow rate through the reentry tube unit must be trimmed during turndown to insure proper operation of the unit and to sustain the condenser temperature at about 900°F (482°C).

From the first approximation presented in this study a turndown of over 50% can be accomplished without either jeopardizing the sulfur retention performance or slumping a section of the fluidized bed furnace. Large turndown may necessitate an increase in the furnace air-fuel ratio to maintain fluidization.

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5. OVERALL PLANT ANALYSIS

With the selection of the basic plant system shown in Figs. 4.1. and 4.2, the overall plant efficiency becomes a function of four factors, i.e., the combustion efficiency, the conditions of the two power cycles, the split or ratio of the prime heat (that from the combustion system) between the two cycles, and the auxiliary or parasitic plant power requirements.

5.1. Plant Efficiencies

The combustion efficiency is reflected in the furnace energy balance given in Section 4.2. A total of 394.1 MW is transferred to the power systems. Of this total, 281.6 MW goes to the topping cycle, 58.6 MW to the steam superheater and 53.9 MW to the steam reheater. These values would imply a combustion efficiency of 87.6%. This value is slightly greater than the actual value as the furnace energy balance includes that part of the auxiliary power that is returned to the system as heat.

In the overall plant analysis, the power output from the cesium system is the heat input from the furnace-boiler multiplied by the net system efficiency. The amount of heat rejected from the cesium system to the steam system is the input from the furnace multiplied by one minus the cycle efficiency. The difference between the cycle efficiency and net system efficiency is the mechanical and electrical losses associated with the generator which are lost to the plant. For the conditions selected for the topping system, the cycle efficiency is 23.87% and the net system efficiency is 23.15%. Thus, 76.13% of the heat input to the system is rejected to the steam system. It should be noted that the total energy input to the cesium system includes that from the boiler recirculation pumps as well as that from the furnace.

The steam system receives heat from both the topping cycle and the furnace. This combined heat input system allows for operating the topping system condenser at 900°F (756°K) and still maintain conventional steam temperatures of 1000°F (811°K). For the steam system selected, the heat rate for the system is equivalent to a system efficiency of 44.0%.

The split in the amount of furnace energy transferred to the two power systems has a significant effect on the plant power output and efficiency. Once the cycle conditions of the two power systems are fixed, it is obvious that the more energy transferred to the topping cycle the larger the plant power output. However, as the amount of energy to the topping cycle is increased, the combustion air and flue gas temperatures throughout the system also increase as does the required effectiveness of the air preheater. These effects are shown in Figs. 5.1 and 5.2.

For both Figs. 5.1 and 5.2, it was assumed that the temperature of the flue gases to the stack would remain constant, i.e., the system combustion efficiency remained constant. As can be seen in Fig 5.1, higher plant efficiencies can be obtained as the heat transferred to the topping cycle is increased. However, the temperatures also increase to the point that more expensive piping and ducting is required. Also, as can be seen in Fig. 5.2 the heat load and required effectiveness of the air preheater is increased.

For this study, the temperature of the flue gases leaving the furnace was fixed at 900°F (756°K). This allows the use of low alloy steel in the duct system and preheater and also limits the temperature of the combustion air to the furnace to 800°F (700°K). This is one area of the design where a more detailed study of the trade-offs between plant cost and efficiency are needed.

5.2. Plant Auxiliary Losses

The auxiliary losses for the plant are listed in Table 5.1. In addition to those losses associated with the systems discussed in the previous section, additional losses are included for the plant cooling system (0.2% of the gross plant output), hotel load (0.75% of the plant gross output), and an additional miscellaneous loss of 0.5 MW. The auxiliary power requirements for the plant total 8.58 MW or 4.1% of the gross output. This is almost twice that of conventional steam plants and is caused by the higher pressure loss in the fluidized bed.

Fig. 5.1. System efficiency and key furnace gas temperatures as a function of the furnace energy distribution.

Fig. 5.2. Air preheater effectiveness and temperatures as a function of the furnace energy distribution.

Table 5.1. Plant Auxiliary Power Requirements

Equipment Item	Power Requirement kW
Forced Draft Fans	2640
Transport Air Fans	12
Ash Cooling System Fans	195
Coal and Limestone Handling and Drying	1390
Induced Draft Fans	1540
Cesium Boiler Recirculating Pumps	310
Plant Cooling System	420
Hotel Load	1570
Miscellaneous	<u>500</u>
	8577

5.3. Overall Plant Performance

The overall plant power output and efficiency are given in Table 5.2. Of the total gross electrical output of 209.18 MW, 31.2% is produced by the cesium system and 68.8% by the steam system. The overall plant net efficiency is 44.58%.

Table 5.2. Overall Plant Power Output and Efficiency

Thermal Power, MW	450.0
Cesium System, MW	281.9
Steam System, MW	327.1
Gross Electrical, MW	209.18
Cesium System, MW	65.26
Steam System, MW	143.92
Auxiliary Power Requirements, MW	8.58
Net Electrical, MW	200.60
Gross Plant Efficiency, %	46.48
Net Plant Efficiency, %	44.58

5.4. Plant Emissions

The only EPA emission standard that will be difficult to meet is the SO₂ limit of 1.2 lb (0.54 kg) per million Btu (1055 MJ). The furnace operating conditions of atmospheric pressure and 1750°F (1228°K) form a combination that has not been adequately investigated, but it is apparent that the amount of limestone required will be greater than that required for lower furnace temperatures. For this reason, a calcium-to-sulfur ratio of 4.0 was used for design purposes. This value was based on an extrapolation of lower temperature data.

The EPA requirements for NO_x emissions are not difficult with fluidized bed combustors. For the furnace conditions of this study, emission of about one-half the EPA standard of 0.7 lb (.32 kg) per million Btu (1055 MJ) would be expected.

Particulate emissions should be far below EPA standards with the bag filters used in this study. The actual plant emissions would be less than 0.02 lb (0.009 kg) per million Btu (1055 MJ) or less than 20% of the EPA maximum of 0.1 lb (0.045 kg) per million Btu (1055 MJ).

5.5. Projected Performance of Larger Plant Size

The plant performance given in the previous sections is based on a 200 MW(e) plant which is typical of a demonstration size unit rather than a conventional full size plant. A larger plant would permit the use of higher steam pressure [3500 psi (24.1 MPa)] rather than the 2400 psi (16.5 MPa) used in this study. This higher pressure would give a much better overall plant efficiency.

Although no design work was done for a larger plant, it would probably be composed of modules similar to that selected for this study. However, the steam system would be combined into a larger, more conventional unit typical of the large steam plants being built today. Thus, the only basic difference in the larger plants would be the different conditions in the steam cycle.

In order to compare the efficiency of a 3500 psi (24.1 MPa) plant to a 2400 psi (16.5 MPa) unit, let us assume that all other plant conditions remain the same; i.e., assume that a 3500 psia (24.1 MPa) system is used

with the 200 MW(e) plant. The only effect then is to increase the cycle efficiency of the steam system from 44.0% of this study to over 47% which is typical of the higher pressure system. For this comparison a cycle efficiency of 47.0% will be used. The results of this analysis are shown in Table 5.3. As can be seen in this table, the output of the steam system increases by almost 7% and the net plant efficiency increases by over 2 points.

Table 5.3. Overall Plant Power Output and Efficiency
with 3500/1000/1000 Steam Conditions

Thermal Power, MW	450.0
Cesium System, MW	281.9
Steam System, MW	327.1
Gross Electrical, MW	218.99
Cesium System, MW	65.26
Steam System, MW	153.73
Auxiliary Power Requirements, MW	8.58
Net Electrical, MW	210.41
Gross Plant Efficiency, %	48.66
Net Plant Efficiency, %	46.76

6. SPECIAL PROBLEMS AND R&D REQUIREMENTS

The problem areas associated with the fluidized bed-alkali metal topping cycle are primarily those of scale and not basic technology. The technology status of both the combustion system and the alkali metal power conversion system are similar. Both have extensive operating experience in small sizes but none in sizes of interest for large commercial power plants. There are no technical feasibility questions concerning either the operation of fluidized bed systems or alkali metal system. There are a number of engineering and design problem areas in both systems that will require an extensive research and development program before large scale plants can be built. Also, the combination of the two systems does raise one problem that could have serious economics consequences and this is the fireside corrosion of the boiler tubes in the fluidized bed. For the alkali metal system to be economically attractive, it must operate at high temperature and the boiler must have a long life while operating at temperatures of 1500 to 1600°F (1089 to 1144°K).

Any large new system such as the alkali metal system will require a substantial development program. Even if all of the basic technology is available, a logical development sequence would include a process development unit (PDU) and/or pilot plant. This would be followed by a large demonstration unit of perhaps 200 MW(e) that would represent a module of the full size commercial plant and use essentially full size alkali metal components. This is the basic approach now being planned for the coal-fired fluidized bed steam power system.

6.1. Combustion System

A fluidized bed combustion system is not a new idea or concept. There are hundreds of units in operation for roasting ores or incineration of solid wastes. Work on the fluidized bed combustion of coal for power generation began in the 1950's. There were two objectives of this early work; to burn low quality or high ash coal that did not burn well in conventional systems and to reduce the cost of steam boilers by taking advantage of the higher heat transfer coefficient in a fluidized bed.

The principal reason for the increasing interest in these systems today is to control emission of sulfur dioxide and oxides of nitrogen.

The current status of the coal-fired fluidized bed combustion system has reached the point where reasonable approximation and "rule-of-thumb" are available. However, during the course of a design study such as that described in the previous section of this report, it becomes apparent that a considerable number of design assumptions are necessary and that a considerable amount of development work is necessary. The current national emphasis being placed on the fluidized bed concept should lead to the successful development of the system.

Combining a coal-fired fluidized bed with a high temperature heat exchanger does pose the problem of fireside corrosion and the selection of the heat exchanger material. Unfortunately, experimental data are not available to establish the allowable temperature for the principal candidate materials that might be used for this application. It is interesting to note that in the ECAS study¹ General Electric chose Hastelloy X, a very expensive material, whereas Westinghouse chose Incoloy 800, the cost of which is about half that of Hastelloy X. One important difference between the ECAS systems of both General Electric and Westinghouse and this study is that the ECAS systems used high pressure furnaces in which the external pressure on the boiler headers and manifolds was much higher than the potassium working fluid. This led to high stresses and potential creep buckling problems of these components. Thus, the General Electric choice of Hastelloy X, which has better high temperature strength than Incoloy 800, may be justified for the higher pressure system. However, the Westinghouse ECAS system also used a high pressure furnace and they chose Incoloy 800. Their estimates of boiler cost were far below those of General Electric.¹

The problems of fireside corrosion are of concern for all coal-fired fluidized bed systems and experimental studies are being carried out in this country and abroad. ERDA is currently sponsoring several programs to study this problem. One program which is especially applicable to the furnace conditions typical of alkali metal systems is the Technology Test Unit (TTU) currently being designed at the Oak Ridge National Laboratory.² This unit uses an atmospheric fluidized bed to heat air for an open or

closed cycle gas turbine system. All heat supplied to the power cycle is through the heat exchanger located in the bed. The turbine inlet conditions for the system are 1500°F (1089°K) and the tubes in the bed operate in the range of 1550 to 1575°F (1117 to 1131°K). This temperature is the same as required for the alkali metal system. Furthermore, materials which appear best suited to withstand fireside corrosion appear to be compatible with the alkali metals.

One of the primary purposes of the TTU is to test a variety of tube materials. The principal material will be Incoloy 800; however, stainless steel 304 and 316 will also be used. Special readily removable test sections will also be included. Materials being considered for the test sections include Hastelloy X, HK 40, Incoloy 800 and the Inconels 600, 617, 713, 718, and 825. Funds for proceeding with the design and construction of this unit were provided in August 1977 and the results of this test should provide the information needed for the alkali metal boiler.

There are other areas of the combustion system that will require development and the current work underway should be applicable to the design of this system. The method of coal feed proposed in this study has been successful in small laboratory tests. However, larger scale tests would be required. Also, development or testing of the ash removal system would be required.

6.2. Alkali Metal System

It is interesting to note that there is more experience with the alkali metal systems operating at high temperatures than with coal-fired fluidized bed combustion systems. An enormous amount of information applicable to alkali metal Rankine cycle systems is available from the space program and an equally impressive amount of information on large sodium equipment is available from the breeder reactor program.

The information development for the space program was for small systems—thermal ratings of a few hundred kW to a few MW rather than the several hundred MW typical of modules for a central station power system. Thus, the problems of the large systems are those of scaling the equipment to larger size and insuring the quality control necessary for liquid metal systems in these large sizes.

6.2.1. Materials Compatibility

There are little, if any, questions concerning the compatibility of potassium with materials of interest for the system so long as the temperature difference in the boiler recirculation loop is small. For the reference design this temperature gradient is only about 100°F (56°K) and should be acceptable. Although the data for cesium indicates that it should be the same as potassium, the data are very limited. The selection of cesium as the working fluid does require additional testing to confirm its use. With either cesium or potassium, oxygen and other impurities must be limited. Both monitoring and purification techniques have been developed and are available for this purpose.

6.2.2. System Components

Most of the major components of the system will need some development and/or proof testing for the scaleup to larger size.

6.2.2.1. Boiler

Additional development work on the alkali metal boiler will probably not be necessary. However, in a logical overall plan to develop the alkali metal system, a process development unit and/or pilot plant should be included. Thus, proof testing of a prototype boiler would be part of the overall development.

The problems associated with the alkali metal side of the boiler are related to boiling stability and the tendency for the alkali metal liquids to superheat which can lead to thermal cycling of the tubes. This problem would be more severe in a natural circulation boiler than in a forced circulation unit. Also, since the potential for superheat in the liquid is inversely proportional to the vapor pressure, this would be less a problem with cesium than with potassium.

The Oak Ridge National Laboratory currently has under construction a large, natural circulation potassium boiler.³ This gas-fired unit employs a tube bundle module that consists of two annular rows of 1-in. (2.54 cm) diameter stainless steel tubes arranged around a 22-in. (55.9 cm) diameter, long, vertical, cylindrical chamber. The full-scale test

unit, which contains 134 tubes, is shown in Fig. 6.1. The flame from the burner directs hot gas upward through this combustion chamber (see Fig. 6.2). The tubes are bent at the top to provide flow passages between them so that the hot gas from the top of the combustion chamber can flow radially outward and then downward through an outer annulus over the outer row of tubes. To reduce the amount of heat transfer surface area required, the furnace and combustion chambers are designed for operation at a pressure up to 5 atm (0.5 MPa); i.e., substantially above the 2 atm (0.2 MPa) design operating pressure of the potassium vapor system at 1540°F (1111°K). Potassium circulates under natural thermal convection from a header tank at the top of the tube bundle through two downcomers, one on either side, to two ring-shaped manifolds at the bottom, and then vertically upward through the tubes back to a header drum and vapor separator at the top of the header tank.

The thermal rating of the unit is about 2×10^6 Btu/hr (586 kW) per atm of operating pressure. Shakedown tests of the unit are scheduled for October 1977.

6.2.2.2. Turbine

Shaft seals for the turbine will require some development and demonstration. Successful seals have been tested with a small high speed potassium turbine for the space program and this technology should be applicable to the larger, lower speed seals for a commercial size unit.

Moisture removal devices to separate and bleed moistures from the turbine were also developed under the space program. Adaptions of these should be satisfactory for larger units.

The turbine rotor disks which led to some controversy in the ECAS study should not be a problem for this design. In ECAS phase 2, G.E. specified 80 in. (2.03 m) diameter Astroloy disk which are twice the size of those available with today's technology.⁴ The disk size specified in this study was 52 in. (1.32 m). However, this size was chosen primarily for the last stage [which operates at 900°F (756°K)] where the volumetric flow rate is very large. It would be possible to keep the first and second stage disks to 40 in. (1.02 m) and flare out the rotor diameter in the

Fig. 6.1. Photo of the ORNL Potassium Boiler under construction.

Fig. 6.2. Section through the gas-fired potassium boiler tube bundle module, burner and furnace assembly.

lower temperature stages. Also, with cesium, the blade speed required for a given stage output is only half that with a potassium turbine and lower strength disks could be used.

Except for the special shaft seal requirements and requirements for high temperature materials, the technology developed for steam turbines should be applicable to alkali metal turbines. Some additional bucket erosion tests should be made to prove the acceptability of cesium.

The turbine control system will probably require development. The principal problem will be to provide overspeed protection in the event of a loss-of-load. Either large fast-acting stop valves or perhaps a water brake system could be used. This is an area that requires additional study and probably an R&D program.

6.2.2.3. Condenser-Steam Generator

Although the reentry tube heat exchanger is not new (it was originally patented in 1831) its application in this system is sufficiently different that a small module of full length tubes should be tested during the initial phase of the development program. Larger scale tests would be obtained in the PDU or pilot plant phase of the program.

6.2.2.4. Other Components

The amount of development effort required for valves will depend on the system design details; i.e., are the valves to restrict flow or to provide absolute seals? If one designs a system that has large high temperature valves that must be leak tight then one may have not only an expensive development program but also a feasibility problem. The philosophy adapted for this study was that valves were to restrict the flow to some small but tolerable leakage rate. No valves were used in the vapor lines in the furnace-boiler system. The only places that vapor valves may be needed are for turbine control and as a dump valve to bypass vapor around the turbine to the condenser. Neither of these would need to be absolutely leak tight.

Although one would prefer to avoid bellows in large high temperature alkali metal systems, they will probably be necessary. The system designed

for this study requires bellows in the boiler recirculation lines and in the lines to the turbine and condenser. Additional design studies should be devoted to possible means of eliminating this feature.

Large liquid metal pumps have been developed for the LMFBR program and should not require an R&D program. The pumps would receive adequate testing in the PDU or pilot plant.

A liquid metal purification and monitoring system will be required. The technology for this is available and the testing of the system could be provided by the PDU or pilot plant.

6.3. Leak Detection

Leak detection is a special problem with any liquid metal system and is as much, if not more, a design problem as an R&D problem. A good quality assurance program coupled to a well and conservatively designed system should eliminate catastrophic failures. The leak detection system should then be designed to detect a very small leak that can develop over a long period of operation before it can develop into a serious incident.

The two system components most susceptible to leaks are the boiler and the condenser-steam generator. The condenser-steam generator is of primary concern because of the metal-water reaction. Fortunately, a small steam leak into the condenser can be readily detected by monitoring the condenser for hydrogen which is formed from the metal-steam reaction. This would locate the condenser unit which has the leak. Conventional helium leak detection methods could then be used to locate the leaking module which would be replaced.

Detecting a small leak in the furnace-boiler will be very difficult with the atmospheric furnace since for normal operation the boiler pressure is greater than the furnace pressure. With a small leak, the amount of material entering the bed or stack gases will be so small that detection is questionable. Another possibility is to detect loss of material by monitoring the system inventory. However, a small leak would not show up until a considerable amount of material had been lost. The most feasible method may be to periodically reduce the power output and thus the pressure in the alkali metal system which would allow air or combustion

products to leak into the metal system. A leak into the metal system could be detected in several ways. One method and probably the easiest and quickest would be to monitor the condenser for the presence of nitrogen. A confirmation of the boiler leak could be obtained by checking the oxygen build-up in the drum in the boiler recirculation loop.

This is an area that requires a considerable amount of additional study. It is obvious that a system in which the furnace pressure normally operates slightly above the boiler pressure would be much more amenable to the detection of small boiler leaks. Another boiler design feature that would be very beneficial would be to use a number of small modules in such a manner that the oxides formed would be concentrated in a small part of the boiler. Once a leak is detected by the presence of nitrogen in the condenser, the liquid metal in the individual boiler modules could be checked for oxygen content. This would isolate the leak in a small part of the system that could be replaced and then later repaired.

The design of a system with these features is described in the following section.

References

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7. ADVANTAGES OF HIGHER FURNACE PRESSURE

The atmospheric pressure fluidized bed system described in the previous sections does not appear to be the best system for the alkali metal topping cycle. Both Westinghouse and General Electric in their evaluations¹ of the alkali metal system concluded that pressurized fluidized beds were superior to atmospheric units. Their systems were based on furnace pressures of 10 to 15 atm (1.0 to 1.5 MPa) and used high temperature [1700 to 1800°F (1200 to 1256°K)] gas turbines to extract power from the combustion products.

During the initial phase of this study three furnace pressures were considered. These were atmospheric pressure, intermediate pressure [3 to 6 atm (0.3 to 0.6 MPa)] and high pressure [6 to 10 atm (0.6 to 1.0 MPa)]. From these initial studies, it was concluded that the intermediate pressure (turbocharged system) was superior to the atmospheric system and was probably better than the higher pressure systems which were similar to those used in the Westinghouse and General Electric evaluations. These initial parametric analyses were performed on a more conventional topping cycle approach which did not include the overlapping temperatures used in the atmospheric system described in the previous sections of this report. However, the relative merits of the pressurized and atmospheric systems are changed very little by this feature.

7.1. Ca/S Ratio

Pressurized systems have several advantages as compared to atmospheric units. For the alkali metal topping cycle to be worth the added complexity and cost and to be an attractive addition to the conventional steam plant, it must operate at a peak temperature in the range of 1500°F (1089°K). This high working fluid temperature requires a bed operating temperature in the range of 1700 to 1800°F (1200 to 1256°K) — a temperature which is well above that for good sulfur retention for atmospheric pressure operation. However, with pressurized beds, the sulfur retention capability of limestone improves with temperature, and for temperatures of 1700 to 1800°F (1200 to 1256°K) a Ca/S ratio of 2.5 or less would

be satisfactory.² This compares to a value of 4.0 used in this study for the atmospheric pressure system.

The lower limestone requirements result in a number of improvements to the system. In addition to reducing the cost of the limestone feed material and its disposal, it also reduces the size of the preparation and handling plant and the amount of energy required for this operation. It reduces the furnace energy loss due to calcination of the limestone, it reduces the amount of CO₂ formed from the calcination of the limestone which reduces the amount of flue gases and thus the plant stack gas losses, it reduces the amount of water that must be removed from the limestone, and it also reduces the amount of solids rejected from the plant and thus the energy loss due to the heat capacity of these solids.

The thermal energy savings from the reduced amount of limestone would reduce the thermal input requirements for the plant by 1 to 1 1/2%. Also the lower feed rate through the preparation and feed system would reduce the power requirements of this system by at least 200 kW. Furthermore, if the feeder system were pressurized, additional savings would result. The total energy savings in this circuit would be in the range of 500 kW. These factors, the reduced thermal input and power savings, would add about 1/2 of a point to the net plant efficiency. It should be noted that the reduced power requirements for the pressurized system are not all gain since a large part of this energy was fed back to the atmospheric system as thermal input.

7.2. Furnace Size and Design

The fluidized bed operating pressure, in addition to its effect on sulfur retention, has an important effect on the design and cost of the furnace boiler. For example, the bed pressure will be the controlling factor in determining the bed surface area required and the type of boiler used.

The power output of a bed is limited by two factors; the rate at which coal can be burned, and the rate at which heat can be removed from the bed. The combustion is limited by the weight flow of air that can be fed to the unit. One interesting aspect of fluidized beds is that

their fluidization characteristics are primarily dependent on the gas velocity through the bed and rather insensitive to the bed pressure. Thus, the weight flow of air and the combustion rate can be increased in direct proportion to the pressure with constant gas velocity and little change in fluidization characteristics.

In order to utilize the higher combustion rate, the heat transfer rate from the bed to the boiler must also be increased. Since the heat transfer does not increase in proportion to the pressure, the boiler surface area in the bed must be increased. This, in turn, requires either a closer tube spacing in the bed or a deeper bed. The limits in regard to the minimum tube spacing that can be used in a fluidized bed is not known. However, as the spacing is decreased the amount of mixing within the bed will be decreased and it appears prudent to maintain a minimum spacing of twice the tube diameter. Since this is the same spacing as used for the atmospheric system described in the previous sections, the bed depth would have to be increased to take advantage of the higher combustion rate.

Fortunately, one of the advantages of a higher pressure system is that it permits operation with a deep bed. This results from the fact that the pressure loss penalty or pumping power required to operate the system depends on the system $\Delta P/P$ (actually the summation of the ratio of the pressure drop to the pressure throughout the system) and not just the ΔP (pressure drop). The ΔP across the bed is equal to the weight per unit area of the bed and is therefore proportional to the bed depth. Also, to insure uniform flow through the bed, the pressure drop through the air distribution plates must be some fraction of (and proportional to) the pressure drop through the bed. Thus, the two largest pressure losses in the system are directly proportional to the bed depth. In view of this dependency, a general rule-of-thumb is that the maximum depth of a bed should not exceed about 4 ft (1.2 m) per atm (0.1 MPa) of operating pressure. Thus, the bed depth and the surface area can be increased in proportion to the combustion rate.

The maximum allowable operating bed depth will also influence the type of boiler used. The boiler designed for a pressurized furnace during the initial phase of this study was a vertical tube natural circulation

unit (see Fig. 7.1). For this unit to be practical a bed depth of about 10 ft (3.1 m) is required to minimize the number of tubes, the tube to header welds, and headering and manifold problems. The furnace concept shown in Fig. 7.1 consists of vertical tubes arranged into bundles or modules. The modular concept was used for headering purposes, to facilitate assembly of the units, and for ease of maintenance and repair if necessary. The Incoloy 800 boiler tubes are 1.0 in. (2.54 cm) OD, 20 to 25 ft (6.1 to 7.6 m) long and located on 2.0 in. (5.08 cm) centers throughout the furnace. Figure 7.1 shows only the alkali metal boiler part of the furnace. It would be possible to locate the steam superheater and reheater in the upper part of the vessel or in a separate vessel if this proved to be advantageous.

There are many advantages to the unit shown in Fig. 7.1 as compared to the furnace shown in Fig. 4.3 of Section 4. In addition to reducing the size of the unit, the smaller furnace surface area also reduces the number of coal feed lines and feed ports and the number of air nozzles required. According to the heat transfer correlations of Gelperin,³ both the increased pressure and the vertical tube arrangement improve the heat transfer coefficient and reduce the boiler surface area required. The natural circulation system also eliminates the need for the recirculation pumps and their power requirements. This feature also eliminates the external recirculating system piping and drum and would reduce the required alkali metal inventory. The elimination of the external recirculating loop would also alleviate the problem associated with thermal expansion of this loop and the need for bellows inherent in the design of the furnace shown in Fig. 4.3.

The increased bed depth also increases the residence time of both gas and solids and would be expected to improve both the sulfur retention and combustion efficiency. The National Coal Board⁴ investigated the effects of the bed height in a few tests and found a small increase in sulfur retention as the bed height is increased. With Pittsburgh coal and Dolomite 1337 at a velocity of 8 fps (2.44 m/s), Ca/S ratio of 5.4 and a bed temperature of 1560°F (1122°K) the sulfur retention increased from 88 to 90% when the bed height was increased from 2.7 to 4 ft (0.82 to 1.22 m). With Illinois coal and Limestone 1359 at a velocity of 3 fps

Fig. 7.1. Conceptual design of a pressurized fluidized bed furnace-alkali metal boiler unit.

(0.91 m/s), Ca/S ratio of 2.2 and a bed temperature of 1470°F (1072°K), a change in bed height from 2 to 3 ft (0.61 to 0.91 m) resulted in an increase in the sulfur retention from 64 to 73%.

7.3. Plant Concept for a Pressurized System

The work on the overall plant concept for a pressurized furnace system was very limited. However, the advantages of this system in comparison to the atmospheric unit are significant, and a brief description of this system is included here.

7.3.1. Combustion Air and Flue Gas System

A simplified flow diagram of the combustion air and flue gas system is shown in Fig. 7.2. The combustion air, after passing through the compressor, is split into three streams. The main stream passes through the air preheater and then enters the furnace through air nozzles to fluidize the bed. One of the two side streams is used to cool the ash from the furnace and cyclones. This stream then rejoins the main air stream before entering the furnace. The second side stream passes through a booster compressor and is then used to transport the coal and limestone to the furnace. The combustion products from the furnace pass first through cyclone separators, the air preheater, and the turbine, and then go to the coal and limestone preparation system where they are used to dry these materials. The combustion products then pass through bag filters, an exhaust fan and then the stack. The system is designed to operate with 15% excess air.

The turbocharger is used only to pressurize the furnace and not to supply shaft power for electrical generation. This has several advantages over that of using a power turbine. First, the temperature required to drive the turbocharger need only be approximately 650°F (617°K). Currently available inertial type separators should be sufficiently effective in removing particulates from the combustion gas to prevent serious damage to the turbine for turbine inlet temperatures up to about 1100°F (867°K).⁵

A second advantage is that, by not driving a generator, a simple variable speed unit can be used. This will allow the turbocharger speed to be

Fig. 7.2. Schematic flow diagram of the combustion air and flue gas system using a pressurized furnace.

varied in order to match the combustion air flow rate to the furnace power output and thus maintain a fixed fuel-to-air ratio over a wide range of power output. This, in turn, keeps the stack losses down when the power output is reduced.

A third advantage of the turbocharged system is that the turndown ratio (ratio of design power output to actual output) of the furnace can be increased. This is due to the fact that the fluidization characteristics of the furnace bed are very weak functions of the bed pressure; i.e., they depend primarily on the velocity of the gases through the bed. With the turbocharger, the system power is reduced by decreasing the turbine inlet temperature which in turn decreases the turbocharger speed and the pressure of the air delivered to the furnace. Thus, most of the turndown can be achieved by a reduction in furnace pressure with little change in the furnace bed velocity or fluidization characteristics. It should be noted that, as the speed of the turbocharger is reduced, it will eventually reach an operating point at which the turbine does not deliver sufficient power to drive the compressor. At this point, auxiliary shaft power from a variable speed motor or a steam turbine will be needed. This auxiliary power source will also be needed to start the system. However, in this analysis, only the full-power-output case is analyzed.

Another interesting aspect of this system is that the efficiency of the turbocharger has little effect on the efficiency of the plant, and, in regard to cycle efficiency, can be considered a heat exchanger that replaces part of the surface area of the air preheater. The efficiency of the turbocharger does influence the turbine inlet temperature required to drive the unit and also the compressor outlet temperature. If the component efficiencies become too low, part of the energy in the turbine exhaust would be used for feedwater heating in addition to its use for drying the coal and limestone. The use of the turbine exhaust energy for feedwater heating will result in a small loss in plant efficiency.

7.3.2. Power Conversion System

The power conversion system for the pressurized furnace plant is identical to that used for the atmospheric system (see Fig. 4.2) except

that no boiler recirculation pumps are required. The same turbine inlet temperatures, condenser temperatures and efficiencies were used.

7.4. Plant Operating Parameters and Efficiencies

For the purposes of comparison to the atmospheric system, a turbocharged system operating at 4.0 atm (0.4 MPa) was chosen for evaluation of the key plant operating parameters and efficiencies. The furnace operating pressure of 4.0 atm (0.4 MPa) was chosen to allow the flue gas to leak into the alkali metal system in the event of a leak in the alkali metal boiler.

The power cycle chosen for the plant was the same cycle used in the atmospheric plant; i.e., a cesium-steam topping cycle. The operating temperatures, pressures, and electrical efficiencies of the component cycles were unchanged. The major change in the system that affects the overall plant performance is the replacement of the electric motor drive for the combustion air compressor with a turbine drive powered by the furnace flue gas which results in a reduction of the plant auxiliary power requirements.

Rather than speculate on what component efficiencies should be used for the turbocharger, actual operating data from a gas turbine-compressor set tested by the Locomotive Development Committee (LDC) of Bituminous Coal Research, Inc. was used. This unit, which was installed in 1951, was part of a program to develop a direct coal-fired turbine drive for railroad use. From idling to full load, the turbine efficiency averaged 87.5% and the compressor efficiency ranged from 82 to 86%. The unit operated with little or no blade damage so long as the turbine inlet did not exceed 1060°F (844°K). However, at temperatures in the range of 1250°F (950°K) the unit experienced some erosion damage of the blading. Inertial type separators were used to minimize particulate loadings entering the turbine. An excellent summary of this work is given in Ref. 5.

For the analyses of this section, compressor and turbine efficiencies of 86 and 87.5% are used. Both cyclone and multicyclone separators are used to remove particulates in the flue gas stream to the turbine.

The key operating parameters and efficiencies for the pressurized plant are given in Table 7.1.

Table 7.1. Intermediate Pressure Plant Operating Parameters and Efficiencies

Thermal Power, Btu/hr, MW	1.53×10^9	450
Cesium System, Btu/hr, MW	9.71×10^8	284.45
Steam System, Btu/hr, MW	1.11×10^9	325.09
Gross Electrical, Btu/hr, MW	7.13×10^8	208.89
Cesium System, Btu/hr, MW	2.25×10^8	65.85
Steam System, Btu/hr, MW	4.88×10^8	143.04
Auxiliary Power Requirements		
Solids Transport and Ash Cooling		
System Fans, Btu/hr, kW	4.1×10^5	120
Coal and Limestone Handling and Drying, Btu/hr, kW	3.17×10^6	930
Plant Cooling System, Btu/hr, kW	1.43×10^6	420
Hotel Load, Btu/hr, kW	5.36×10^6	1570
Miscellaneous, Btu/hr, kW	1.71×10^6	500
Net Electrical, Btu/hr, MW	7.01×10^8	205.35
Gross Plant Efficiency, %	46.42	
Net Plant Efficiency, %	45.63	
Furnace Pressure, atm, MPa	4.0	0.4
Fluidized Bed Temperature, °F, °K	1750	1228
Coal Feed Rate (dry), lb/hr, kg/s	123,860	15.6
Limestone Feed Rate (dry), lb/hr, kg/s	45,520	5.74
Combustion Air Flow Rate, lb/hr, kg/s	1,347,600	169.8
Fluidizing Air		
Flow Rate lb/hr, kg/s	1,305,200	164.5
Temperature, °F, °K	707	648
Transport Air (to furnace)		
Flow Rate, lb/hr, kg/s	42,368	5.34
Temperature, °F, °K	275	408
Compressor Discharge Temperature, °F, °K	375	464
Air Preheater Outlet Temperature, °F, °K	694	641
Flue Gas Flow Rate, lb/hr, kg/s	1,464,500	184.5
Flue Gas Temperature, °F, °K		
Furnace Outlet	900	756
Air Preheater Outlet	658	621
Turbine Outlet	383	468
Stack Outlet	303	424
Compressor Adiabatic Efficiency, %	86	
Turbine Adiabatic Efficiency, %	87.5	
Solids Temperature from Ash Cooler, °F, °K	500	533

References

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8. COMPARISON WITH ECAS RESULTS

Since 1970 there have been at least twelve studies¹ undertaken to compare the merits of advanced power conversion systems. The most detailed of these studies was the Energy Conversion Alternatives Study (ECAS).^{2,3} The fact that there have been twelve attempts in seven years to define the best system or systems gives a good indication of the difficulty of the task.

The ECAS was conducted in two phases. Phase 1 consisted of parametric analysis of the following ten energy conversion systems by General Electric and Westinghouse:

1. Advanced steam
2. Open-cycle gas turbine
3. Combined-cycle gas turbine/steam turbine
4. Closed-cycle gas turbine
5. Open-cycle magnetohydrodynamic (MHD)
6. Closed-cycle, inert-gas MHD
7. Liquid-metal MHD
8. Supercritical carbon dioxide
9. Liquid-metal Rankine topping cycle
10. Fuel cells

The G. E. phase 1 final results are reported in Ref. 4 and the Westinghouse phase 1 final results in Ref. 5. A comparative evaluation of phase 1 results is given in Ref. 2.

The phase 2 part of the study consisted of a more comprehensive study of the following systems:

1. Advanced steam systems
2. Combined-cycle gas turbine/steam turbine systems
3. Closed-cycle gas turbine/organic system
4. Potassium/steam system
5. Open-cycle MHD/steam system
6. Molten-carbonate fuel cell/steam system

The responsibility for conducting the different evaluations was shifted during the second phase. While both G.E. and Westinghouse studied the alkali metal system and considered both potassium and cesium during phase 1, only G.E. studied the system in phase 2 and considered only potassium. Both G.E. and Westinghouse concluded from the phase 1 study that a pressurized furnace system was better suited to the alkali metal system than an atmospheric unit. Also, from the phase 1 results G.E. concluded the potassium was the best working fluid while Westinghouse found some advantages for cesium and recommended further study of the two fluids.

The purpose of the following discussion is to compare the results of the ECAS evaluation of the alkali metal system to the results of this study and to point out significant differences in the results.

8.1. ECAS System Description

During the phase 1 study both G.E. and Westinghouse concluded that the pressurized furnace system was superior to the atmospheric unit. Their pressurized systems had three power conversion systems, a high pressure gas turbine-generator system operating on the flue gases from the furnace, an alkali metal cycle operating between 1400°F (1033°K) and 1100°F (867°K) and a conventional 3500 psi (24.13 MPa) steam system with 1000°F (811°K) turbine inlet temperature and one 1000°F (811°K) reheat (see Fig. 8.1). Fluidized bed temperatures and gas turbine inlet temperatures of 1600 to 1800°F (1144 to 1256°K) were studied. In all cases, all of the energy to convert the feedwater to superheated steam and to reheat the steam was supplied by the condensing alkali metal.

There were several differences between the G.E. and Westinghouse phase 1 results. G.E. used the flue gas exhaust from the gas turbine only in the recuperator (see Fig. 8.1). This resulted in stack gas temperatures over 600°F (589°K) and a significant energy loss. Westinghouse used the turbine exhaust for both air preheating and feedwater heating with resulting stack gas temperatures of 250 to 300°F (394 to 422°K).

Both G.E. and Westinghouse used forced circulation alkali metal boilers. However, G.E. used a recirculation ratio of 67 or an exit quality of 1.5% while Westinghouse used a ratio of 2.5 or an exit quality of 40%.

Fig. 8.1. Simplified schematic diagram of potassium pressurized fluidized-bed boiler powerplant.

This increased boiler flow in the G.E. case added about 60 MW(e) to the pumping power requirements for the plant. In phase 2, G.E. reduced the ratio to 10 for an exit quality of 10%. For the boiler heat flux typical of fluidized beds, the Westinghouse value is reasonable while the G.E. values are very conservative.

A notable difference which influenced the boiler cost was the selection of materials. G.E. used Hastelloy X, a very expensive alloy, while Westinghouse used Incoloy 800 for the boiler tubes and headers.

The G.E. system for the phase 2 study is shown in Fig. 8.2. Note that the stack gas temperature here has been reduced to 300°F (422°K).

Two variations of this G.E. configuration were examined by NASA.³

The first variation was designed to improve the system efficiency by using as much as possible of the pressurizing-gas-turbine exhaust energy to preheat the furnace combustion air before it is used for steam feedwater heating. This portion of the gas-turbine exhaust energy will then be converted into electricity at the efficiency of the combined potassium/steam system portion of the plant as compared with the lower efficiency of the steam bottoming cycle alone. Although the overall efficiency could be increased by about

Fig. 8.2. Simplified schematic diagram of potassium topping cycle - General Electric reference system.

0.5 percentage point, the change is not likely to be economical. It increases the power produced by the potassium cycle but at relatively high incremental capital cost as compared with the substantially lower capital cost of the steam system.

The second variation was a concept that has been proposed in the literature for ameliorating the potential hot-corrosion and erosion problems of the pressurizing gas turbine operating in the PFB exhaust gases. This is accomplished by reducing the turbine inlet temperature by inserting an air preheater between the PFB exit and the gas turbine inlet. This approach also permits final filtering of small particles at the reduced temperature at the gas-turbine inlet with a "hot" electrostatic precipitator, thus eliminating the requirement for the costly, high-temperature, granular bed filters. Turbine inlet temperature was reduced to 1000°F through this technique, resulting in a drop in efficiency of approximately 2 percentage points from the reference system value of 44.4 percent. An additional loss in efficiency of 1 percentage point occurred as the turbine inlet temperature was further reduced to 800°F. Slightly higher efficiencies might be obtainable by optimizing the steam feedwater heating system as the gas-turbine inlet temperature is reduced. The specific cost of the modified system was estimated to be about \$1002/kWe at 1000°F, increasing to about \$1090/kWe at 800°F, as compared with \$934/kWe for the reference system. The COE of the modified system was estimated to be 42.4 mills/kW-hr at 800°F, as compared with 39.9 mills/kW-hr for the reference system. Although reducing the inlet temperature to the pressurizing gas turbine relieves the problem of hot corrosion in the turbine, it is replaced by a materials problem in the air preheater, which must now operate in the high-temperature gases leaving the PFB. In addition, the potential problem of hot corrosion in the potassium boiler tubes in the PFB still remains.

It should be noted that neither the "hot" electrostatic precipitator nor the high temperature air preheater are necessary to operate the pressurizing turbine at low temperature as demonstrated in Section 7.

The basic difference in the ECAS systems and that described in this study is that in the ECAS the alkali metal operating range was restricted to an upper temperature of 1400°F (1033°K) and a lower temperature of 1100°F (867°K) while in this study the range was between 1500°F (1089°K) and 900°F (756°K). Furthermore, in this study, the steam conditions were maintained at 1000°F (811°K) by using the heat in the 1750°F (1228°K) flue gas to provide most of the steam superheat and all of the reheat. This in turn eliminated the high temperature preheater referred to above.

The reasons for the small temperature range used for the alkali metal system in the ECAS is noted in the following paragraph.³

The results of the study indicate that the efficiency of an advanced steam plant with a PFB furnace can be increased from about 40 percent to 44.4 percent by adding a 1400°F potassium topping cycle. The resulting additional power is very costly, however, not only in terms of capital cost and COE, but also in terms of complexity and the additional failure modes and hazards associated with the use of an alkali-metal working fluid. The potassium system is expensive relative to the power it generates. A basic problem is that the potassium cycle is constrained to operate over a temperature range that does not exploit its full thermodynamic potential. The potassium turbine inlet temperature is limited by fluidized-bed and material constraints, and the condensing temperature is limited by large potassium volume flows and associated high turbine, ducting, and condenser costs. As a result, the cycle temperature difference available for turbine work is small, and the potassium cycle must handle a large energy flow without extracting much power. High-volume flows and costly materials then result in a high cost for each kilowatt of power generated by the potassium cycle.

The upper temperature constraints do not appear to be reasonable. Most of the work with boiling potassium systems for the space program were at temperatures above 1500°F (1089°K). There is some question as to fire-side corrosion as a function of tube temperature but no firm answer is available as yet. However, the use of a 1500°F (1089°K) turbine inlet temperature represents a very modest extrapolation or degree of optimism as compared to that applied to other advanced conversion systems considered in the ECAS. The high vapor volume of the potassium at lower temperatures is a problem. However, with cesium, lower temperatures can be used. Unfortunately, cesium was dropped from consideration in the phase 2 study.

8.2. Plant Efficiency

A comparison of the plant efficiencies from the ECAS and from this study is given in Table 8.1. As noted before, the results of the G.E. phase 1 were generally recognized as too low because of the high stack gas losses and the excessive pumping power used for the boiler recirculating pumps. The higher efficiencies of this study result from the different temperature conditions for the alkali metal system and the use of

Table 8.1. Comparison of ECAS Efficiencies to Those of This Study

	Net Plant Efficiency, %
ECAS Phase 1	
G.E.	39.6
Westinghouse	42.4*
ECAS Phase 2	
G.E.	44.4
This Study	
Atmospheric Furnace and	
2400 psi/1000°F/1000°F steam conditions	44.6
3500 psi/1000°F/1000°F steam conditions	46.8
Pressurized Furnace and	
2400 psi/1000°F/1000°F steam conditions	45.6
3500 psi/1000°F/1000°F steam conditions	47.8

*This value was for potassium. Westinghouse gave a value of 42.9% for cesium.

the overlapping temperatures of the two power conversion systems. The turbine inlet temperature for this study is 100°F (56°K) higher than that for ECAS. The validity of using this higher temperature will depend on the fireside corrosion of the boiler tubes and this data is not available. The use of cesium as the working fluid permits the lower condenser temperature [900°F (756°K) as compared to 1100°F (867°K)]. Also, the cycle arrangement with the addition of heat to the steam system from the furnace gives conventional steam conditions of 1000°F (811°K) with the 900°F (756°K) alkali metal condensing condition. This arrangement which improves the overall plant efficiency was not considered in the ECAS.

NASA investigated two variations of the G.E. phase 2 system (see Section 8.1). The first variation used more of the gas turbine exhaust energy for preheating the combustion air. This would add about 0.5% to the plant efficiency. The second variation reduced the gas turbine inlet temperature and used more air preheating. As would be expected, this variation reduced the plant efficiency. The results of the analysis of this variation are shown in Fig. 8.3. It should be noted that for this case power is still extracted from the turbine. One can also conclude

Fig. 8.3. Effect of reduced PFB pressurized-gas-turbine inlet temperature on PFB/potassium/steam system performance.

from Fig. 8.3 that the gas turbine adds about 3 points to the plant efficiency of the G.E. phase 2 system as compared to a gas turbine used only for pressurizing the furnace.

The effect of lowering the cesium turbine inlet temperature from 1500°F (1089°K) to 1400°F (1033°K) for the system of this study was estimated. In making this estimate, the split in the furnace heat transferred to alkali metal system and the steam system was assumed to be the same as was the plant auxiliary losses. For both the 2400 psi/1000°F/1000°F (16.5 MPa/811°K/811°K) and the 3500 psi/1000°F/1000°F (24.1 MPa/811°K/811°K) steam conditions the 100°F (56°K) reduction in turbine inlet conditions resulted in an overall plant efficiency loss of only 0.8%.

This reduction in turbine inlet temperature would also reduce the boiler temperature by about the same amount and significantly reduce the surface area requirement and cost of the boiler. However, the reduced

temperature would require that the weight flow of the cesium would have to be increased by 15 to 20% while the pressure and thus the density of the cesium would be reduced by about 1/3. Thus the volume flow of vapor would be almost doubled, which would require larger headers and piping. Also, the larger weight flow would increase the size of the turbine and may give problems in the last stage and exhaust system at the 900°F (756°K) condensing temperature.

It should be noted that increasing the condenser temperature is more detrimental to the cycle efficiency than decreasing the turbine inlet temperature; i.e., a 100°F (56°K) increase in the condenser temperature reduces the topping cycle efficiency by about 4.0 points while a 100°F (56°K) reduction in turbine inlet temperature decreases the topping cycle efficiency by 2.5 points. This fact is especially important in an alkali metal system in which the condenser temperature is limited by the volume flow from the turbine. By increasing the temperature range of the working fluid, the volume flow of fluid per unit of heat transferred is reduced. This in turn permits a lower condenser temperature for a given volume flow.

8.3. Plant Costs

Plant cost estimates were not made for the system of this study and a direct comparison to the ECAS cost for every subsystem is not possible. However, an examination of the ECAS costs and a comparison of some subsystems will provide some insight as to the probable cost range for the system.

Economic ground rules were established for ECAS for calculation of plant capital cost and cost of electricity. These ground rules specified such items as site labor costs, escalation rates, interest rates, and other fixed charges.² The escalation and interest charges are high enough that even small differences in equipment and materials cost are notable in the overall plant cost. Hence it is quite difficult to pinpoint a value for plant cost and cost of electricity and, therefore, a congruent value from the different contractors in the ECAS should not necessarily be anticipated since each contractor was free to use his own "rules-of-thumb" in estimating the component costs and other direct costs. An

example of the above is seen in the NASA review of ECAS 1.² Of particular interest here is the comparison of contractor estimates for the liquid metal topping cycle plant capital costs as presented in Table 8.2 (from Ref. 2). Westinghouse effected a lower boiler cost by implementing Incoloy 800 tubing instead of Hastelloy X but the other differences are less easily discerned.

In fact, in ECAS 1, Westinghouse consistently predicted costs of electricity either equal to or less than G.E. with considerable differences in some cases which emphasizes the difficulty in costing advanced power systems. However, some valuable information may be obtained by comparing the system described in this report to the final topping cycle configuration derived by G.E. and reported in the ECAS 2.^{3,6} A summary description of the G.E. system was provided previously in Section 8.1. The major component costs of the G.E. system are summarized in Table 8.3 (from Ref. 6).

Table 8.2. Cost Comparisons of Pressurized-Fluidized-Bed Potassium Topping Cycle

	General Electric case 9 ($\eta = 0.396$)	Westinghouse case 49 ($\eta = 0.424$)	General Electric case 9 (adjusted to $\eta = 0.424$)
<u>Capital cost, \$/kW(e)</u>			
Major Components:	240.7	158.4	202.3
Furnace/boiler	144.4	53.2	112.4 ^a
Potassium turbogenerator	29.9	21.1	27.9
Condenser/steam generator	2.2	9.2	2.1
Steam turbogenerator	20.8	22.4	19.4
Gas turbine/compressor/generator	37.3	26.7	34.8
Other (pumps, dump tank, etc.)	6.1	25.8	5.7
Other Direct Costs:	319.9	256.2	294.4
Balance of plant	177.3	137.1	165.6
Site labor	49.2	83.0	46.0
Contingency	93.4	29.1	82.8
Escalation and Interest	357.9	252.4	317.1
Total	918.5	667.0	813.8
<u>Cost of electricity, mills/kW-hr</u>			
Capital costs	29.0	21.1	25.7
Operating and maintenance costs	3.3	1.7	3.3
Fuel costs	7.3	6.8	6.8
Total	39.6	29.6	35.8

^aIncludes reduction in recirculation pump costs.

Table 8.3. Major Component Costs for G.E. Alkali Metal Topping Cycle Plant

Major Component or Subsystem	No. of Units	Cost/Unit (FOB) (M\$)	Component or Subsystem Costs (FOB) (M\$)	BOP + Materials (M\$)	Site Labor (Direct + Indirect) (M\$)	Total Installed Cost (M\$)	\$/kW(e)
Fuel Handling & Preparation							
Coal & Dolomite Dryers, Crushers	3		5.568				
Petrocarb Injectors	4		21.680	9.100	5.023	44.371	44.560
Ash Handlers	2		1.551	3.570	2.143	7.264	7.290
High Pressure Air System & Controls	2		2.821			2.821	2.830
Prime Cycle Metal Vapor Topping Cycle							
Pressurized Fluidized Bed Modules	4		61.548	0.770	1.094	63.412	63.680
Potassium Turbine-Generators	4		33.090	0.740	2.009	35.839	35.990
Condenser-Steam Generators	4		10.182	0.090	0.446	10.718	10.760
Dump Tanks, Vac Pumps, Safety	4		5.594	2.340	1.229	9.163	9.200
Potassium Inventory	1		1.004			1.004	1.008
Pressurizing System							
Pressurizing Gas Turbine-Gen.	4		26.223	0.100	0.893	27.216	27.330
Hot Gas Filter Trains	2		63.280	2.040	2.210	67.530	67.810
Feed Water Economizers	4		7.272	0.930	1.563	9.765	9.806
Bottoming Cycle-Steam Cycle							
Steam Turbine-Generator	1		23.000	0.060	1.786	24.846	24.950
Condenser, Feedwater Heaters	1			4.210	0.558	4.768	4.790
Cooling Towers & Circ. Water Pumps	19			4.116	2.829	6.945	6.970
			262.800				

As noted by G.E., the hot gas filter is an extremely costly item.⁶ This subsystem, the gas turbine-generator, and the feedwater economizers are non-existent in the atmospheric pressure unit of this study, therefore reducing its cost below that of the ECAS reference system by about 105 \$/kW(e).

Another area where cost differences are likely is the alkali metal turbine. As illustrated in Section 3 of this report, a cesium turbine would be considerably smaller and therefore less expensive than a potassium turbine of comparable output. It is likely that the cost of the turbine is proportional to the rotor diameter or perhaps diameter squared.

The other subsystems itemized in Table 8.3 have costs which should be comparable to those of the system proposed in this study. The boiler in this study, although it employs Incoloy 800 and operates at a low furnace pressure which eliminates problems of creep buckling of the alkali metal headers, may not offer a significant cost savings because its surface area requirements are greater than the G.E. system. In contrast, the higher boiler exit quality (25% vs 10% for G.E. ECAS) and the much higher topping cycle temperature range [600°F (333°K) vs 300°F (167°K) for ECAS] will reduce the size and cost of the boiler recirculation system. In addition, the lower alkali metal condenser temperature with this system also permits the use of Croloy tubing rather than stainless steel in the steam generator. However, most of these later potential savings will probably be offset by the steam superheater and reheater and other system differences.

The effect of the cost differences can be seen by referring to the plant cost breakdown of the G.E. system in Table 8.4 (from Ref. 6).

Elimination of the pressurizing system reduces the prime cycle plant equipment by 105 \$/kW(e). An additional savings of 10 \$/kW(e) on the turbine would be a reasonable assumption in light of the comparison shown in Section 3. This would result in a 24.5% reduction in the subtotal shown in Table 8.4. Since the A-E service and contingency are simply fixed percentages of this subtotal, those charges would be reduced proportionally. Finally if for this simplified comparison the same interest and escalation percentage is used, the total plant cost becomes about

Table 8.4. Overall Plant Costs for G.E. Topping Cycle System

	Major Components Materials (M\$)	Balance of Plant Materials (M\$)	Site Labor (Direct & Indirect) (M\$)	Total (M\$)	\$/kW(e)
1.0 Land Improvements & Structures (Land, Plant Area 108 Acres) (Land, 30-Year Disposal 9.6 Acres)	0.0	19.2	27.1	46.3	46.5
2.0 Coal Handling	31.6	12.7	7.2	51.5	51.7
3.0 Prime Cycle Plant Equipment Potassium Cycle 164.4 MW(e) Pressurizing Gas Turbine 218.3 MW(e)	208.2	26.4	22.6	257.2	258.3
4.0 Bottom Cycle Steam 650.0 MW(e)	23.0	19.1	18.0	60.1	60.4
5.0 Electrical Plant & Instrumentation	0.0	23.2	31.4	54.6	54.8
Subtotal	262.8	100.5	106.2	469.7	471.7
6.0 A-E Service & Contingency				131.2	131.7
7.0 Escalation & Interest During Construction				329.3	330.7
Total M\$				930.2	
Plant Output MW(e)				995.8	
Total \$/kW(e)				934.1	

703 M\$ (706 \$/kW(e)). This value lies between the estimates of G.E. and Westinghouse in ECAS 1.

In their ECAS 2 summary, NASA performed an economic analysis to determine the maximum plant cost for the G.E. topping cycle system which would provide a competitive COE. Their conclusion is presented here.³

It is also informative to calculate the maximum specific incremental cost due to the addition of the potassium cycle that would provide a PFB/potassium/steam system COE no higher than the 34.1 mills/kW-hr of the untopped steam system. Under these conditions, the capital contribution to the COE for the PFB/potassium/steam system could be a maximum of 23.8 mills/kW-hr. This corresponds to a total capital cost for the PFB/potassium/steam system of about \$750 million as compared with the \$930 million estimated by G.E. Based on these values, a specific incremental cost of not more than \$880/kWe (gross) must be realized for the power produced by the potassium turbine-generator in order to achieve a capital cost of \$750 million for the entire PFB/potassium/steam system and COE of 34.1 mills/kW-hr.

Since the plant efficiencies are practically identical for the system in this study and the G.E. ECAS 2 system, the NASA analysis and conclusion is applicable here. It appears highly probable that the atmospheric pressure system presented here can provide a COE lower than the 34.1 mills/kW-hr cited above. Furthermore, the implementation of the higher efficiency systems (see Section 8.2) could result in even lower COE.

The problems of the ECAS-2 design are defined by NASA in the concluding remarks of their summary.³

A basic problem is that the potassium cycle is constrained to operate over a temperature range that does not exploit its full thermodynamic potential. The turbine inlet temperature is limited by fluidized-bed and material constraints, and the condensing temperature is limited by large potassium volume flows and associated high turbine, ducting, and condenser costs. As a result, the cycle temperature difference available for turbine work is small, and the potassium cycle must handle a large energy flow without extracting much power. High volume flows and costly materials then result in a high cost for each kilowatt of power produced.

In conclusion, although a complete and highly detailed comparison to ECAS was not possible, new data has been presented which adds merit to the topping cycle concept. The combination of high efficiency and low cost

for this system is a direct result of efforts to avoid the problem in potassium topping cycle systems as defined in the preceding paragraph.

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Appendix A

THERMODYNAMIC PROPERTIES OF SATURATED CESIUM

TEMP-R	PSAT-ATM	SPEC VOL	FT**3/LBM	VF	VG	HF	HF _G	HG	ENTHALPY BTU/LBM	ENTROPY BTU/LBM-R	SF	SG
960.0000	0.00089	0.00932	5928.207	0.00932	5928.207	54.3396	234.8780	289.2175	0.1987	0.4433		
970.0000	0.00105	0.00934	5056.297	0.00934	5056.297	54.7605	234.5925	289.3530	0.1991	0.4410		
980.0000	0.00124	0.00935	4327.582	0.00935	4327.582	55.1941	234.3245	289.5186	0.1996	0.4387		
990.0000	0.00146	0.00937	3716.308	0.00937	3716.308	55.6373	234.0702	289.7075	0.2000	0.4364		
1000.0000	0.00171	0.00939	3201.702	0.00939	3201.702	56.0879	233.8274	289.9153	0.2005	0.4343		
1010.0000	0.00200	0.00940	2766.968	0.00940	2766.968	56.5471	233.5913	290.1384	0.2009	0.4322		
1020.0000	0.00233	0.00942	2398.507	0.00942	2398.507	57.0119	233.3613	290.3733	0.2014	0.4302		
1030.0000	0.00270	0.00944	2085.194	0.00944	2085.194	57.4820	233.1357	290.6177	0.2018	0.4282		
1040.0000	0.00313	0.00946	1817.974	0.00946	1817.974	57.9550	232.9151	290.8701	0.2023	0.4262		
1050.0000	0.00361	0.00947	1589.351	0.00947	1589.351	58.4341	232.6941	291.1282	0.2027	0.4244		
1060.0000	0.00416	0.00949	1393.186	0.00949	1393.186	58.9168	232.4741	291.3909	0.2032	0.4225		
1070.0000	0.00477	0.00951	1224.393	0.00951	1224.393	59.4033	232.2542	291.6575	0.2037	0.4207		
1080.0000	0.00546	0.00953	1078.754	0.00953	1078.754	59.8930	232.0338	291.9268	0.2041	0.4190		
1090.0000	0.00624	0.00954	952.778	0.00954	952.778	60.3811	231.8166	292.1978	0.2046	0.4172		
1100.0000	0.00711	0.00956	843.491	0.00956	843.491	60.8762	231.5938	292.4700	0.2050	0.4156		
1110.0000	0.00808	0.00958	748.461	0.00958	748.461	61.3742	231.3689	292.7432	0.2055	0.4139		
1120.0000	0.00916	0.00960	665.624	0.00960	665.624	61.8747	231.1419	293.0166	0.2059	0.4123		
1130.0000	0.01036	0.00962	593.244	0.00962	593.244	62.3773	230.9124	293.2898	0.2064	0.4107		
1140.0000	0.01169	0.00964	529.871	0.00964	529.871	62.8749	230.6881	293.5630	0.2068	0.4092		
1150.0000	0.01317	0.00965	474.228	0.00965	474.228	63.3823	230.4529	293.8352	0.2072	0.4076		
1160.0000	0.01481	0.00967	425.276	0.00967	425.276	63.8920	230.2144	294.1064	0.2077	0.4061		
1170.0000	0.01661	0.00969	382.123	0.00969	382.123	64.4041	229.9728	294.3770	0.2081	0.4047		
1180.0000	0.01859	0.00971	343.999	0.00971	343.999	64.9184	229.7274	294.6458	0.2086	0.4032		
1190.0000	0.02077	0.00973	310.252	0.00973	310.252	65.4343	229.4787	294.9131	0.2090	0.4018		
1200.0000	0.02317	0.00975	280.336	0.00975	280.336	65.9385	229.2407	295.1792	0.2094	0.4005		
1210.0000	0.02579	0.00977	253.734	0.00977	253.734	66.4589	228.9842	295.4431	0.2099	0.3991		
1220.0000	0.02866	0.00979	230.050	0.00979	230.050	66.9814	228.7240	295.7053	0.2103	0.3978		
1230.0000	0.03179	0.00981	208.922	0.00981	208.922	67.5057	228.4597	295.9653	0.2107	0.3965		
1240.0000	0.03520	0.00982	190.043	0.00982	190.043	68.0326	228.1913	296.2239	0.2111	0.3952		
1250.0000	0.03892	0.00984	173.144	0.00984	173.144	68.5609	227.9189	296.4797	0.2116	0.3939		
1260.0000	0.04295	0.00986	158.010	0.00986	158.010	69.0648	227.6691	296.7339	0.2120	0.3927		
1270.0000	0.04733	0.00988	144.399	0.00988	144.399	69.5977	227.3881	296.9858	0.2124	0.3914		
1280.0000	0.05208	0.00990	132.154	0.00990	132.154	70.1326	227.1028	297.2354	0.2128	0.3902		
1290.0000	0.05721	0.00992	121.121	0.00992	121.121	70.6690	226.8134	297.4824	0.2132	0.3890		
1300.0000	0.06276	0.00994	111.165	0.00994	111.165	71.2081	226.5195	297.7275	0.2136	0.3879		
1310.0000	0.06875	0.00996	102.166	0.00996	102.166	71.7484	226.2213	297.9697	0.2141	0.3867		
1320.0000	0.07520	0.00998	94.021	0.00998	94.021	72.2908	225.9189	298.2097	0.2145	0.3856		
1330.0000	0.08214	0.01000	86.657	0.01000	86.657	72.7871	225.6611	298.4482	0.2148	0.3845		
1340.0000	0.08961	0.01002	79.953	0.01002	79.953	73.3335	225.3496	298.6831	0.2153	0.3834		
1350.0000	0.09762	0.01005	73.860	0.01005	73.860	73.8821	225.0336	298.9158	0.2157	0.3824		
1360.0000	0.10621	0.01007	68.314	0.01007	68.314	74.4325	224.7135	299.1460	0.2161	0.3813		
1370.0000	0.11542	0.01009	63.260	0.01009	63.260	74.9849	224.3889	299.3738	0.2165	0.3803		
1380.0000	0.12527	0.01011	58.647	0.01011	58.647	75.5392	224.0597	299.5989	0.2169	0.3792		
1390.0000	0.13579	0.01013	54.432	0.01013	54.432	76.0951	223.7265	299.8215	0.2173	0.3782		
1400.0000	0.14702	0.01015	50.577	0.01015	50.577	76.6532	223.3888	300.0420	0.2177	0.3772		

TEMP-R	PSAT-ATM	SPEC VOL FT**3/LBM		ENTHALPY BTU/LBM			ENTROPY BTU/LBM-R	
		VF	VG	HF	HFG	HG	SF	SG
1410.0000	0.15900	0.01017	47.063	77.1261	223.1348	300.2610	0.2180	0.3763
1420.0000	0.17177	0.01019	43.823	77.6881	222.7882	300.4763	0.2184	0.3753
1430.0000	0.18535	0.01021	40.848	78.2523	222.4369	300.6892	0.2188	0.3744
1440.0000	0.19979	0.01024	38.114	78.8176	222.0816	300.8992	0.2192	0.3734
1450.0000	0.21512	0.01026	35.598	79.3852	221.7218	301.1069	0.2196	0.3725
1460.0000	0.23139	0.01028	33.280	79.9555	221.3568	301.3123	0.2200	0.3716
1470.0000	0.24864	0.01030	31.143	80.5260	220.9889	301.5149	0.2204	0.3707
1480.0000	0.26691	0.01032	29.171	81.0982	220.6169	301.7151	0.2208	0.3698
1490.0000	0.28625	0.01035	27.348	81.6721	220.2407	301.9128	0.2212	0.3690
1500.0000	0.30669	0.01037	25.680	82.0965	220.0159	302.1123	0.2215	0.3681
1510.0000	0.32828	0.01039	24.117	82.6744	219.6308	302.3052	0.2218	0.3673
1520.0000	0.35107	0.01042	22.669	83.2544	219.2414	302.4958	0.2222	0.3665
1530.0000	0.37510	0.01044	21.326	83.8365	218.8478	302.6843	0.2226	0.3656
1540.0000	0.40042	0.01046	20.079	84.4197	218.4505	302.8701	0.2230	0.3648
1550.0000	0.42709	0.01048	18.920	85.0082	218.0452	303.0535	0.2234	0.3640
1560.0000	0.45514	0.01051	17.842	85.5944	217.6404	303.2349	0.2237	0.3633
1570.0000	0.48463	0.01053	16.838	86.1819	217.2319	303.4138	0.2241	0.3625
1580.0000	0.51562	0.01056	15.903	86.7706	216.8200	303.5906	0.2245	0.3617
1590.0000	0.54814	0.01058	15.050	87.0992	216.6745	303.7737	0.2247	0.3610
1600.0000	0.58226	0.01060	14.235	87.6903	216.2562	303.9465	0.2251	0.3602
1610.0000	0.61803	0.01063	13.441	88.7906	215.3083	304.0989	0.2257	0.3595
1620.0000	0.65549	0.01065	12.732	89.3817	214.8049	304.2666	0.2261	0.3588
1630.0000	0.69472	0.01068	12.068	89.9754	214.4570	304.4324	0.2265	0.3580
1640.0000	0.73575	0.01070	11.446	90.5701	214.0258	304.5959	0.2268	0.3573
1650.0000	0.77865	0.01072	10.864	91.1662	213.5914	304.7576	0.2272	0.3566
1660.0000	0.82347	0.01075	10.317	91.7634	213.1538	304.9172	0.2276	0.3560
1670.0000	0.87028	0.01077	9.807	92.3235	212.7529	305.0764	0.2279	0.3553
1680.0000	0.91912	0.01080	9.326	92.9230	212.3092	305.2322	0.2283	0.3546
1690.0000	0.97005	0.01083	8.874	93.5235	211.8627	305.3862	0.2286	0.3540
1700.0000	1.02315	0.01085	8.448	94.1251	211.4130	305.5381	0.2290	0.3533
1710.0000	1.07845	0.01088	8.048	94.7273	210.9609	305.6882	0.2293	0.3527
1720.0000	1.13602	0.01090	7.672	95.3304	210.5063	305.8367	0.2297	0.3521
1730.0000	1.19593	0.01093	7.317	95.9348	210.0488	305.9836	0.2300	0.3514
1740.0000	1.25824	0.01095	6.983	96.5396	209.5891	306.1287	0.2304	0.3508
1750.0000	1.32300	0.01098	6.667	97.1451	209.1269	306.2720	0.2307	0.3502
1760.0000	1.39027	0.01101	6.369	97.7516	208.6625	306.4141	0.2311	0.3496
1770.0000	1.46012	0.01103	6.088	98.3585	208.1959	306.5544	0.2314	0.3490
1780.0000	1.53262	0.01106	5.822	98.9661	207.7273	306.6934	0.2317	0.3484
1790.0000	1.60780	0.01109	5.573	99.5084	207.3261	306.8345	0.2320	0.3479
1800.0000	1.68576	0.01111	5.335	100.1172	206.8535	306.9707	0.2324	0.3473
1810.0000	1.76654	0.01114	5.110	100.7264	206.3791	307.1055	0.2327	0.3467
1820.0000	1.85022	0.01117	4.897	101.3361	205.9031	307.2393	0.2331	0.3462
1830.0000	1.93684	0.01120	4.695	101.9461	205.4255	307.3716	0.2334	0.3456
1840.0000	2.02648	0.01122	4.503	102.5564	204.9465	307.5029	0.2337	0.3451
1850.0000	2.11919	0.01125	4.321	103.1668	204.4662	307.6331	0.2340	0.3446

TEMP-R	PSAT-ATM	SPEC VOL FT**3/LBM		ENTHALPY BTU/LBM			ENTROPY BTU/LBM-R	
		VF	VG	HF	HFG	HG	SF	SG
1860.0000	2.21506	0.01128	4.149	103.7778	203.9847	307.7625	0.2344	0.3440
1870.0000	2.31412	0.01131	3.985	104.3883	203.5021	307.8904	0.2347	0.3435
1880.0000	2.41646	0.01134	3.829	104.9992	203.0186	308.0178	0.2350	0.3430
1890.0000	2.52213	0.01137	3.681	105.6100	202.5342	308.1443	0.2354	0.3425
1900.0000	2.63119	0.01140	3.541	106.2209	202.0491	308.2700	0.2357	0.3420
1910.0000	2.74370	0.01142	3.407	106.8314	201.5634	308.3948	0.2360	0.3415
1920.0000	2.85975	0.01145	3.279	107.4423	201.0770	308.5193	0.2363	0.3410
1930.0000	2.97938	0.01148	3.160	107.9445	200.7062	308.6506	0.2366	0.3406
1940.0000	3.10267	0.01151	3.044	108.5548	200.2191	308.7739	0.2369	0.3401
1950.0000	3.22965	0.01154	2.934	109.1655	199.7313	308.8967	0.2372	0.3396
1960.0000	3.36042	0.01157	2.828	109.7760	199.2431	309.0190	0.2375	0.3392
1970.0000	3.49503	0.01160	2.728	110.3859	198.7550	309.1409	0.2378	0.3387
1980.0000	3.63353	0.01163	2.632	110.9954	198.2668	309.2622	0.2381	0.3383
1990.0000	3.77600	0.01167	2.541	111.6047	197.7791	309.3833	0.2384	0.3378
2000.0000	3.92249	0.01170	2.453	112.2136	197.2913	309.5043	0.2387	0.3374
2010.0000	4.07306	0.01173	2.370	112.8218	196.8042	309.6260	0.2390	0.3369
2020.0000	4.22779	0.01176	2.290	113.4295	196.3173	309.7468	0.2393	0.3365
2030.0000	4.38673	0.01179	2.214	114.0370	195.8309	309.8679	0.2396	0.3361
2040.0000	4.54994	0.01182	2.141	114.6439	195.3448	309.9888	0.2399	0.3357
2050.0000	4.71747	0.01185	2.071	115.2500	194.8599	310.1099	0.2402	0.3353
2060.0000	4.88940	0.01189	2.004	115.8559	194.3750	310.2310	0.2405	0.3349
2070.0000	5.06577	0.01192	1.940	116.4613	193.8910	310.3523	0.2408	0.3345
2080.0000	5.24667	0.01195	1.880	116.8994	193.5891	310.4885	0.2410	0.3341
2090.0000	5.43211	0.01198	1.821	117.5022	193.1087	310.6108	0.2413	0.3337
2100.0000	5.62220	0.01202	1.765	118.1074	192.6258	310.7332	0.2416	0.3333
2110.0000	5.81696	0.01205	1.710	118.7119	192.1435	310.8555	0.2419	0.3329
2120.0000	6.01646	0.01208	1.658	119.3161	191.6621	310.9783	0.2422	0.3326
2130.0000	6.22077	0.01212	1.608	119.9203	191.1815	311.1018	0.2425	0.3322
2140.0000	6.42993	0.01215	1.560	120.5236	190.7017	311.2253	0.2427	0.3318
2150.0000	6.64401	0.01219	1.514	121.1269	190.2227	311.3496	0.2430	0.3315
2160.0000	6.86305	0.01222	1.470	121.7301	189.7443	311.4744	0.2433	0.3311
2170.0000	7.08711	0.01226	1.427	122.3324	189.2670	311.5994	0.2436	0.3308
2180.0000	7.31626	0.01229	1.386	122.9347	188.7901	311.7249	0.2438	0.3304
2190.0000	7.55052	0.01233	1.347	123.5372	188.3139	311.8511	0.2441	0.3301
2200.0000	7.78997	0.01236	1.309	124.1395	187.8381	311.9775	0.2444	0.3296
2210.0000	8.03465	0.01240	1.273	124.7573	187.3660	312.1033	0.2447	0.3294
2220.0000	8.28462	0.01243	1.239	125.1021	187.1557	312.2578	0.2448	0.3291
2230.0000	8.53993	0.01247	1.205	125.6909	186.6970	312.3879	0.2451	0.3288
2240.0000	8.80063	0.01251	1.173	126.2952	186.2209	312.5151	0.2454	0.3285
2250.0000	9.06677	0.01254	1.141	126.9014	185.7441	312.6455	0.2456	0.3282
2260.0000	9.33834	0.01258	1.111	127.5079	185.2670	312.7749	0.2459	0.3279
2270.0000	9.61555	0.01262	1.082	128.1150	184.7896	312.9045	0.2462	0.3276
2280.0000	9.89831	0.01266	1.053	128.7234	184.3118	313.0352	0.2464	0.3273
2290.0000	10.18667	0.01270	1.026	129.3658	183.7961	313.1619	0.2467	0.3270
2300.0000	10.48075	0.01273	1.000	129.9438	183.3521	313.2959	0.2469	0.3267

TEMP-R	PSAT-ATM	SPEC VOL FT**3/LBM VF	VF	HF	ENTHALPY BTU/LBM HFG	HG	ENTROPY BTU/LBM-R SF	SG
2310.0000	10.78053	0.01277	0.974	130.5869	182.8364	313.4233	0.2472	0.3264
2320.0000	11.08607	0.01281	0.950	131.2030	182.3512	313.5542	0.2475	0.3261
2330.0000	11.39744	0.01285	0.926	131.8169	181.8679	313.6848	0.2477	0.3258
2340.0000	11.71464	0.01289	0.903	132.4324	181.3835	313.8159	0.2480	0.3255
2350.0000	12.03776	0.01293	0.881	133.0499	180.8972	313.9470	0.2483	0.3252
2360.0000	12.36683	0.01297	0.859	133.6690	180.4086	314.0776	0.2485	0.3250
2370.0000	12.70188	0.01301	0.839	134.2899	179.9182	314.2080	0.2488	0.3247
2380.0000	13.04293	0.01305	0.820	134.9402	179.4438	314.3384	0.2489	0.3245
2390.0000	13.39008	0.01309	0.799	135.4948	178.9784	314.4731	0.2493	0.3242
2400.0000	13.74332	0.01313	0.782	135.7914	178.8524	314.6438	0.2494	0.3239
2410.0000	14.10266	0.01317	0.764	136.3575	178.4233	314.7808	0.2497	0.3237
2420.0000	14.46826	0.01322	0.746	136.9802	177.9299	314.9102	0.2499	0.3234
2430.0000	14.84001	0.01326	0.729	137.6169	177.4199	315.0369	0.2502	0.3232
2440.0000	15.21800	0.01330	0.713	138.2595	176.9036	315.1631	0.2504	0.3229
2450.0000	15.60232	0.01334	0.696	138.9048	176.3823	315.2871	0.2507	0.3227
2460.0000	15.99295	0.01339	0.681	139.5543	175.8561	315.4104	0.2509	0.3224
2470.0000	16.38992	0.01343	0.666	140.2067	175.3251	315.5317	0.2512	0.3222
2480.0000	16.79327	0.01348	0.651	140.8628	174.7888	315.6516	0.2515	0.3219
2490.0000	17.20302	0.01352	0.636	141.5221	174.2474	315.7695	0.2517	0.3217
2500.0000	17.61929	0.01356	0.623	142.1851	173.7004	315.8855	0.2520	0.3215

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Appendix B

TRANSPORT PROPERTIES OF SATURATED CESIUM

HEAT CAP BTU/LBM-R		VISCOSITY LBM/HR-FT		THERM COND BTU/HR-FT-F	
TEMP-R	LIQUID	LIQUID	VAPOR	LIQUID	VAPOR
960.0000	0.0568	0.6653	0.0409	12.3848	0.0022
965.0000	0.0568	0.6616	0.0410	12.3648	0.0022
970.0000	0.0568	0.6581	0.0411	12.3448	0.0022
975.0000	0.0568	0.6546	0.0412	12.3249	0.0022
980.0000	0.0568	0.6511	0.0413	12.3049	0.0022
985.0000	0.0568	0.6477	0.0414	12.2849	0.0022
990.0000	0.0568	0.6443	0.0415	12.2649	0.0022
995.0000	0.0568	0.6409	0.0416	12.2450	0.0022
1000.0000	0.0568	0.6376	0.0418	12.2250	0.0022
1005.0000	0.0568	0.6344	0.0419	12.2050	0.0023
1010.0000	0.0568	0.6311	0.0420	12.1850	0.0023
1015.0000	0.0568	0.6279	0.0421	12.1651	0.0023
1020.0000	0.0568	0.6246	0.0422	12.1451	0.0023
1025.0000	0.0568	0.6217	0.0423	12.1251	0.0023
1030.0000	0.0568	0.6186	0.0424	12.1051	0.0023
1035.0000	0.0568	0.6156	0.0425	12.0852	0.0023
1040.0000	0.0568	0.6126	0.0426	12.0652	0.0023
1045.0000	0.0568	0.6096	0.0427	12.0452	0.0023
1050.0000	0.0568	0.6067	0.0428	12.0252	0.0023
1055.0000	0.0568	0.6038	0.0429	12.0053	0.0023
1060.0000	0.0568	0.6009	0.0431	11.9853	0.0023
1065.0000	0.0568	0.5981	0.0432	11.9653	0.0023
1070.0000	0.0568	0.5953	0.0433	11.9453	0.0023
1075.0000	0.0568	0.5925	0.0434	11.9254	0.0023
1080.0000	0.0568	0.5898	0.0435	11.9054	0.0023
1085.0000	0.0568	0.5870	0.0436	11.8854	0.0023
1090.0000	0.0568	0.5844	0.0437	11.8654	0.0023
1095.0000	0.0568	0.5817	0.0438	11.8455	0.0024
1100.0000	0.0568	0.5791	0.0439	11.8255	0.0024
1105.0000	0.0568	0.5765	0.0440	11.8055	0.0024
1110.0000	0.0568	0.5739	0.0441	11.7855	0.0024
1115.0000	0.0568	0.5714	0.0442	11.7656	0.0024
1120.0000	0.0568	0.5689	0.0443	11.7456	0.0024
1125.0000	0.0568	0.5664	0.0444	11.7256	0.0024
1130.0000	0.0568	0.5639	0.0445	11.7056	0.0024
1135.0000	0.0568	0.5615	0.0446	11.6857	0.0024
1140.0000	0.0568	0.5591	0.0447	11.6657	0.0024
1145.0000	0.0568	0.5567	0.0448	11.6457	0.0024
1150.0000	0.0568	0.5543	0.0449	11.6257	0.0024
1155.0000	0.0568	0.5520	0.0450	11.6058	0.0024
1160.0000	0.0568	0.5497	0.0451	11.5858	0.0024
1165.0000	0.0568	0.5474	0.0452	11.5658	0.0024
1170.0000	0.0568	0.5451	0.0453	11.5458	0.0024
1175.0000	0.0568	0.5429	0.0454	11.5259	0.0024
1180.0000	0.0568	0.5407	0.0455	11.5059	0.0024

HEAT CAP BTU/LBM-R		VISCOSITY LBM/HR-FT		THERM COND BTU/HR-FT-F	
TEMP-R	LIQUID	LIQUID	VAPOR	LIQUID	VAPOR
1185.0000	0.0568	0.5385	0.0456	11.4859	0.0024
1190.0000	0.0568	0.5363	0.0457	11.4659	0.0025
1195.0000	0.0568	0.5341	0.0458	11.4460	0.0025
1200.0000	0.0568	0.5320	0.0459	11.4260	0.0025
1205.0000	0.0568	0.5299	0.0460	11.4060	0.0025
1210.0000	0.0568	0.5278	0.0461	11.3860	0.0025
1215.0000	0.0568	0.5257	0.0462	11.3661	0.0025
1220.0000	0.0568	0.5237	0.0463	11.3461	0.0025
1225.0000	0.0568	0.5216	0.0464	11.3261	0.0025
1230.0000	0.0568	0.5196	0.0465	11.3061	0.0025
1235.0000	0.0568	0.5176	0.0466	11.2862	0.0025
1240.0000	0.0568	0.5157	0.0467	11.2662	0.0025
1245.0000	0.0568	0.5137	0.0468	11.2462	0.0025
1250.0000	0.0568	0.5118	0.0469	11.2262	0.0025
1255.0000	0.0568	0.5098	0.0470	11.2063	0.0025
1260.0000	0.0568	0.5079	0.0471	11.1863	0.0025
1265.0000	0.0568	0.5060	0.0472	11.1663	0.0025
1270.0000	0.0568	0.5042	0.0473	11.1463	0.0025
1275.0000	0.0568	0.5023	0.0474	11.1264	0.0025
1280.0000	0.0568	0.5005	0.0475	11.1064	0.0025
1285.0000	0.0568	0.4986	0.0476	11.0864	0.0025
1290.0000	0.0568	0.4968	0.0477	11.0664	0.0026
1295.0000	0.0568	0.4951	0.0478	11.0465	0.0026
1300.0000	0.0568	0.4933	0.0479	11.0265	0.0026
1305.0000	0.0568	0.4915	0.0480	11.0065	0.0026
1310.0000	0.0568	0.4898	0.0481	10.9865	0.0026
1315.0000	0.0568	0.4881	0.0482	10.9666	0.0026
1320.0000	0.0568	0.4863	0.0483	10.9466	0.0026
1325.0000	0.0568	0.4846	0.0484	10.9266	0.0026
1330.0000	0.0568	0.4830	0.0485	10.9066	0.0026
1335.0000	0.0568	0.4813	0.0486	10.8867	0.0026
1340.0000	0.0568	0.4796	0.0487	10.8667	0.0026
1345.0000	0.0568	0.4780	0.0488	10.8467	0.0026
1350.0000	0.0568	0.4764	0.0489	10.8267	0.0026
1355.0000	0.0568	0.4748	0.0490	10.8068	0.0026
1360.0000	0.0568	0.4732	0.0491	10.7868	0.0026
1365.0000	0.0568	0.4716	0.0492	10.7668	0.0026
1370.0000	0.0568	0.4700	0.0492	10.7468	0.0026
1375.0000	0.0568	0.4684	0.0493	10.7269	0.0026
1380.0000	0.0568	0.4669	0.0494	10.7069	0.0026
1385.0000	0.0568	0.4653	0.0495	10.6869	0.0026
1390.0000	0.0568	0.4638	0.0496	10.6669	0.0027
1395.0000	0.0568	0.4623	0.0497	10.6470	0.0027
1400.0000	0.0568	0.4608	0.0498	10.6270	0.0027
1405.0000	0.0568	0.4593	0.0499	10.6070	0.0027

HEAT CAP BTU/LBM-R		VISCOSITY LBM/HR-FT		THERM COND BTU/HR-FT-F	
TEMP-R	LIQUID	LIQUID	VAPOR	LIQUID	VAPOR
1410.0000	0.0568	0.4579	0.0500	10.5870	0.0027
1415.0000	0.0568	0.4564	0.0501	10.5671	0.0027
1420.0000	0.0568	0.4549	0.0502	10.5471	0.0027
1425.0000	0.0568	0.4535	0.0503	10.5271	0.0027
1430.0000	0.0568	0.4521	0.0504	10.5071	0.0027
1435.0000	0.0568	0.4506	0.0505	10.4872	0.0027
1440.0000	0.0568	0.4492	0.0506	10.4672	0.0027
1445.0000	0.0568	0.4478	0.0506	10.4472	0.0027
1450.0000	0.0568	0.4465	0.0507	10.4272	0.0027
1455.0000	0.0568	0.4451	0.0508	10.4073	0.0027
1460.0000	0.0568	0.4437	0.0509	10.3873	0.0027
1465.0000	0.0568	0.4424	0.0510	10.3673	0.0027
1470.0000	0.0568	0.4410	0.0511	10.3473	0.0027
1475.0000	0.0568	0.4397	0.0512	10.3274	0.0027
1480.0000	0.0568	0.4384	0.0513	10.3074	0.0027
1485.0000	0.0568	0.4371	0.0514	10.2874	0.0027
1490.0000	0.0568	0.4357	0.0515	10.2674	0.0027
1495.0000	0.0568	0.4345	0.0516	10.2475	0.0027
1500.0000	0.0568	0.4332	0.0516	10.2275	0.0028
1505.0000	0.0568	0.4319	0.0517	10.2075	0.0028
1510.0000	0.0568	0.4306	0.0518	10.1875	0.0028
1515.0000	0.0568	0.4294	0.0519	10.1676	0.0028
1520.0000	0.0568	0.4281	0.0520	10.1476	0.0028
1525.0000	0.0568	0.4269	0.0521	10.1276	0.0028
1530.0000	0.0568	0.4257	0.0522	10.1076	0.0028
1535.0000	0.0568	0.4244	0.0523	10.0877	0.0028
1540.0000	0.0568	0.4232	0.0524	10.0677	0.0028
1545.0000	0.0568	0.4220	0.0524	10.0477	0.0028
1550.0000	0.0568	0.4208	0.0525	10.0277	0.0028
1555.0000	0.0568	0.4196	0.0526	10.0078	0.0028
1560.0000	0.0568	0.4185	0.0527	9.9878	0.0028
1565.0000	0.0568	0.4173	0.0528	9.9678	0.0028
1570.0000	0.0568	0.4161	0.0529	9.9478	0.0028
1575.0000	0.0568	0.4150	0.0530	9.9279	0.0028
1580.0000	0.0568	0.4138	0.0531	9.9079	0.0028
1585.0000	0.0568	0.4127	0.0532	9.8879	0.0028
1590.0000	0.0568	0.4116	0.0532	9.8679	0.0028
1595.0000	0.0568	0.4104	0.0533	9.8480	0.0028
1600.0000	0.0568	0.4093	0.0534	9.8280	0.0028
1605.0000	0.0568	0.4082	0.0535	9.8080	0.0028
1610.0000	0.0568	0.4071	0.0536	9.7880	0.0029
1615.0000	0.0568	0.4060	0.0537	9.7681	0.0029
1620.0000	0.0568	0.4049	0.0538	9.7481	0.0029
1625.0000	0.0568	0.4039	0.0539	9.7281	0.0029
1630.0000	0.0568	0.4028	0.0539	9.7081	0.0029

HEAT CAP BTU/LBM-R		VISCOSITY LBM/HR-FT		THERM COND BTU/HR-FT-F	
TEMP-R	LIQUID	LIQUID	VAPOR	LIQUID	VAPOR
1635.0000	0.0568	0.4017	0.0540	9.6882	0.0029
1640.0000	0.0568	0.4007	0.0541	9.6682	0.0029
1645.0000	0.0568	0.3996	0.0542	9.6482	0.0029
1650.0000	0.0568	0.3986	0.0543	9.6282	0.0029
1655.0000	0.0568	0.3975	0.0544	9.6083	0.0029
1660.0000	0.0568	0.3965	0.0545	9.5883	0.0029
1665.0000	0.0568	0.3955	0.0545	9.5683	0.0029
1670.0000	0.0568	0.3945	0.0546	9.5483	0.0029
1675.0000	0.0568	0.3935	0.0547	9.5284	0.0029
1680.0000	0.0568	0.3925	0.0548	9.5084	0.0029
1685.0000	0.0568	0.3915	0.0549	9.4884	0.0029
1690.0000	0.0568	0.3905	0.0550	9.4684	0.0029
1695.0000	0.0568	0.3895	0.0551	9.4485	0.0029
1700.0000	0.0568	0.3885	0.0551	9.4285	0.0029
1705.0000	0.0568	0.3875	0.0552	9.4085	0.0029
1710.0000	0.0568	0.3866	0.0553	9.3885	0.0029
1715.0000	0.0568	0.3856	0.0554	9.3686	0.0029
1720.0000	0.0568	0.3847	0.0555	9.3486	0.0029
1725.0000	0.0568	0.3837	0.0556	9.3286	0.0030
1730.0000	0.0568	0.3828	0.0557	9.3086	0.0030
1735.0000	0.0568	0.3818	0.0557	9.2887	0.0030
1740.0000	0.0568	0.3809	0.0558	9.2687	0.0030
1745.0000	0.0568	0.3800	0.0559	9.2487	0.0030
1750.0000	0.0568	0.3791	0.0560	9.2287	0.0030
1755.0000	0.0568	0.3782	0.0561	9.2088	0.0030
1760.0000	0.0568	0.3772	0.0562	9.1888	0.0030
1765.0000	0.0568	0.3763	0.0562	9.1688	0.0030
1770.0000	0.0568	0.3755	0.0563	9.1488	0.0030
1775.0000	0.0568	0.3746	0.0564	9.1289	0.0030
1780.0000	0.0568	0.3737	0.0565	9.1089	0.0030
1785.0000	0.0568	0.3728	0.0566	9.0889	0.0030
1790.0000	0.0568	0.3719	0.0567	9.0689	0.0030
1795.0000	0.0568	0.3710	0.0567	9.0490	0.0030
1800.0000	0.0568	0.3702	0.0568	9.0290	0.0030
1805.0000	0.0568	0.3693	0.0569	9.0090	0.0030
1810.0000	0.0568	0.3685	0.0570	8.9890	0.0030
1815.0000	0.0568	0.3676	0.0571	8.9691	0.0030
1820.0000	0.0568	0.3668	0.0572	8.9491	0.0030
1825.0000	0.0568	0.3659	0.0572	8.9291	0.0030
1830.0000	0.0568	0.3651	0.0573	8.9091	0.0030
1835.0000	0.0568	0.3643	0.0574	8.8892	0.0030
1840.0000	0.0568	0.3634	0.0575	8.8692	0.0031
1845.0000	0.0568	0.3626	0.0576	8.8492	0.0031
1850.0000	0.0568	0.3618	0.0576	8.8292	0.0031
1855.0000	0.0568	0.3610	0.0577	8.8093	0.0031

HEAT CAP BTU/LBM-R		VISCOSITY LBM/HR-FT		THERM COND BTU/HR-FT-F	
TEMP-R	LIQUID	LIQUID	VAPOR	LIQUID	VAPOR
1860.0000	0.0568	0.3602	0.0578	8.7893	0.0031
1865.0000	0.0568	0.3594	0.0579	8.7693	0.0031
1870.0000	0.0568	0.3586	0.0580	8.7493	0.0031
1875.0000	0.0568	0.3578	0.0581	8.7294	0.0031
1880.0000	0.0568	0.3570	0.0581	8.7094	0.0031
1885.0000	0.0568	0.3562	0.0582	8.6894	0.0031
1890.0000	0.0568	0.3554	0.0583	8.6694	0.0031
1895.0000	0.0568	0.3547	0.0584	8.6495	0.0031
1900.0000	0.0568	0.3539	0.0585	8.6295	0.0031
1905.0000	0.0568	0.3531	0.0585	8.6095	0.0031
1910.0000	0.0568	0.3524	0.0586	8.5895	0.0031
1915.0000	0.0568	0.3516	0.0587	8.5696	0.0031
1920.0000	0.0568	0.3509	0.0588	8.5496	0.0031
1925.0000	0.0568	0.3501	0.0589	8.5296	0.0031
1930.0000	0.0568	0.3494	0.0589	8.5096	0.0031
1935.0000	0.0568	0.3486	0.0590	8.4897	0.0031
1940.0000	0.0568	0.3479	0.0591	8.4697	0.0031
1945.0000	0.0568	0.3471	0.0592	8.4497	0.0031
1950.0000	0.0568	0.3464	0.0593	8.4297	0.0031
1955.0000	0.0568	0.3457	0.0593	8.4098	0.0031
1960.0000	0.0568	0.3450	0.0594	8.3898	0.0031
1965.0000	0.0568	0.3442	0.0595	8.3698	0.0032
1970.0000	0.0568	0.3435	0.0596	8.3498	0.0032
1975.0000	0.0568	0.3428	0.0597	8.3299	0.0032
1980.0000	0.0568	0.3421	0.0597	8.3099	0.0032
1985.0000	0.0568	0.3414	0.0598	8.2899	0.0032
1990.0000	0.0568	0.3407	0.0599	8.2699	0.0032
1995.0000	0.0568	0.3400	0.0600	8.2500	0.0032
2000.0000	0.0568	0.3393	0.0600	8.2300	0.0032
2005.0000	0.0568	0.3386	0.0601	8.2100	0.0032
2010.0000	0.0568	0.3379	0.0602	8.1900	0.0032
2015.0000	0.0568	0.3373	0.0603	8.1701	0.0032
2020.0000	0.0568	0.3366	0.0604	8.1501	0.0032
2025.0000	0.0568	0.3359	0.0604	8.1301	0.0032
2030.0000	0.0568	0.3352	0.0605	8.1101	0.0032
2035.0000	0.0568	0.3346	0.0606	8.0902	0.0032
2040.0000	0.0568	0.3339	0.0607	8.0702	0.0032
2045.0000	0.0568	0.3332	0.0608	8.0502	0.0032
2050.0000	0.0568	0.3326	0.0608	8.0302	0.0032
2055.0000	0.0568	0.3319	0.0609	8.0103	0.0032
2060.0000	0.0568	0.3313	0.0610	7.9903	0.0032
2065.0000	0.0568	0.3306	0.0611	7.9703	0.0032
2070.0000	0.0568	0.3300	0.0611	7.9503	0.0032
2075.0000	0.0568	0.3293	0.0612	7.9304	0.0032
2080.0000	0.0568	0.3287	0.0613	7.9104	0.0032

HEAT CAP BTU/LBM-R		VISCOSITY LBM/HR-FT		THERM COND BTU/HR-FT-F	
TEMP-R	LIQUID	LIQUID	VAPOR	LIQUID	VAPOR
2085.0000	0.0568	0.3281	0.0614	7.8904	0.0032
2090.0000	0.0568	0.3274	0.0614	7.8704	0.0033
2095.0000	0.0568	0.3268	0.0615	7.8505	0.0033
2100.0000	0.0568	0.3262	0.0616	7.8305	0.0033
2105.0000	0.0568	0.3256	0.0617	7.8105	0.0033
2110.0000	0.0568	0.3249	0.0618	7.7905	0.0033
2115.0000	0.0568	0.3243	0.0618	7.7706	0.0033
2120.0000	0.0568	0.3237	0.0619	7.7506	0.0033
2125.0000	0.0568	0.3231	0.0620	7.7306	0.0033
2130.0000	0.0568	0.3225	0.0621	7.7106	0.0033
2135.0000	0.0568	0.3219	0.0621	7.6907	0.0033
2140.0000	0.0568	0.3213	0.0622	7.6707	0.0033
2145.0000	0.0568	0.3207	0.0623	7.6507	0.0033
2150.0000	0.0568	0.3201	0.0624	7.6307	0.0033
2155.0000	0.0568	0.3195	0.0624	7.6108	0.0033
2160.0000	0.0568	0.3189	0.0625	7.5908	0.0033
2165.0000	0.0568	0.3183	0.0626	7.5708	0.0033
2170.0000	0.0568	0.3177	0.0627	7.5508	0.0033
2175.0000	0.0568	0.3171	0.0627	7.5309	0.0033
2180.0000	0.0568	0.3166	0.0628	7.5109	0.0033
2185.0000	0.0568	0.3160	0.0629	7.4909	0.0033
2190.0000	0.0568	0.3154	0.0630	7.4709	0.0033
2195.0000	0.0568	0.3148	0.0630	7.4510	0.0033
2200.0000	0.0568	0.3143	0.0631	7.4310	0.0033
2205.0000	0.0568	0.3137	0.0632	7.4110	0.0033
2210.0000	0.0568	0.3131	0.0633	7.3910	0.0033
2215.0000	0.0568	0.3126	0.0633	7.3711	0.0033
2220.0000	0.0568	0.3120	0.0634	7.3511	0.0034
2225.0000	0.0568	0.3115	0.0635	7.3311	0.0034
2230.0000	0.0568	0.3109	0.0636	7.3111	0.0034
2235.0000	0.0568	0.3104	0.0636	7.2912	0.0034
2240.0000	0.0568	0.3098	0.0637	7.2712	0.0034
2245.0000	0.0568	0.3093	0.0638	7.2512	0.0034
2250.0000	0.0568	0.3087	0.0639	7.2312	0.0034
2255.0000	0.0568	0.3082	0.0639	7.2113	0.0034
2260.0000	0.0568	0.3076	0.0640	7.1913	0.0034
2265.0000	0.0568	0.3071	0.0641	7.1713	0.0034
2270.0000	0.0568	0.3066	0.0642	7.1513	0.0034
2275.0000	0.0568	0.3060	0.0642	7.1314	0.0034
2280.0000	0.0568	0.3055	0.0643	7.1114	0.0034
2285.0000	0.0568	0.3050	0.0644	7.0914	0.0034
2290.0000	0.0568	0.3045	0.0645	7.0714	0.0034
2295.0000	0.0568	0.3039	0.0645	7.0515	0.0034
2300.0000	0.0568	0.3034	0.0646	7.0315	0.0034
2305.0000	0.0568	0.3029	0.0647	7.0115	0.0034

HEAT CAP BTU/LBM-R		VISCOSITY LBM/HR-FT		THERM COND BTU/HR-FT-F	
TEMP-R	LIQUID	LIQUID	VAPOR	LIQUID	VAPOR
2310.0000	0.0568	0.3024	0.0648	6.9915	0.0034
2315.0000	0.0568	0.3019	0.0648	6.9716	0.0034
2320.0000	0.0568	0.3014	0.0649	6.9516	0.0034
2325.0000	0.0568	0.3008	0.0650	6.9316	0.0034
2330.0000	0.0568	0.3003	0.0651	6.9116	0.0034
2335.0000	0.0568	0.2998	0.0651	6.8917	0.0034
2340.0000	0.0568	0.2993	0.0652	6.8717	0.0034
2345.0000	0.0568	0.2988	0.0653	6.8517	0.0034
2350.0000	0.0568	0.2983	0.0653	6.8317	0.0034
2355.0000	0.0568	0.2978	0.0654	6.8118	0.0035
2360.0000	0.0568	0.2973	0.0655	6.7918	0.0035
2365.0000	0.0568	0.2969	0.0656	6.7718	0.0035
2370.0000	0.0568	0.2964	0.0656	6.7518	0.0035
2375.0000	0.0568	0.2959	0.0657	6.7319	0.0035
2380.0000	0.0568	0.2954	0.0658	6.7119	0.0035
2385.0000	0.0568	0.2949	0.0659	6.6919	0.0035
2390.0000	0.0568	0.2944	0.0659	6.6719	0.0035
2395.0000	0.0568	0.2939	0.0660	6.6520	0.0035
2400.0000	0.0568	0.2935	0.0661	6.6320	0.0035
2405.0000	0.0568	0.2930	0.0661	6.6120	0.0035
2410.0000	0.0568	0.2925	0.0662	6.5920	0.0035
2415.0000	0.0568	0.2921	0.0663	6.5721	0.0035
2420.0000	0.0568	0.2916	0.0664	6.5521	0.0035
2425.0000	0.0568	0.2911	0.0664	6.5321	0.0035
2430.0000	0.0568	0.2906	0.0665	6.5121	0.0035
2435.0000	0.0568	0.2902	0.0666	6.4922	0.0035
2440.0000	0.0568	0.2897	0.0666	6.4722	0.0035
2445.0000	0.0568	0.2893	0.0667	6.4522	0.0035
2450.0000	0.0568	0.2888	0.0668	6.4322	0.0035
2455.0000	0.0568	0.2883	0.0669	6.4123	0.0035
2460.0000	0.0568	0.2879	0.0669	6.3923	0.0035
2465.0000	0.0568	0.2874	0.0670	6.3723	0.0035
2470.0000	0.0568	0.2870	0.0671	6.3523	0.0035
2475.0000	0.0568	0.2865	0.0671	6.3324	0.0035
2480.0000	0.0568	0.2861	0.0672	6.3124	0.0035
2485.0000	0.0568	0.2857	0.0673	6.2924	0.0035
2490.0000	0.0568	0.2852	0.0674	6.2724	0.0035
2495.0000	0.0568	0.2848	0.0674	6.2525	0.0036
2500.0000	0.0568	0.2843	0.0675	6.2325	0.0036

Appendix C

SONIC VELOCITY AND RELATED PROPERTIES OF
SATURATED CESIUM VAPOR

Values of the equilibrium sonic velocity are tabulated. GAMM-0 is the zero-pressure specific heat ratio. The instability in GAMM-0 is of numerical origin. Data for temperatures less than 1200°R should not be used.

TEMP-R	EQ SCNIC VEL-FT/SEC	CP/CV	GAMM-O	CP-BTU/LbM-R
960.00000	537.36178	0.79697	0.80256	-0.03761
970.00000	511.74744	0.71700	0.72100	-0.02615
980.00000	471.33300	0.60326	0.60586	-0.01669
990.00000	400.69319	0.43240	0.43377	-0.00880
1000.00000	238.83224	0.15235	0.15267	-0.00217
1010.00000	379.00940	-0.38048	-0.38093	0.00345
1020.00000	819.42134	-1.76374	-1.76424	0.00824
1030.00000	2278.03458	-13.51888	-13.51140	0.01236
1040.00000	1628.23420	6.84971	6.84045	0.01592
1050.00000	1202.71912	3.70688	3.69903	0.01904
1060.00000	1061.93778	2.86647	2.85827	0.02178
1070.00000	991.54144	2.47897	2.47008	0.02420
1080.00000	949.83693	2.25674	2.24704	0.02636
1090.00000	922.76718	2.11319	2.10262	0.02830
1100.00000	904.19017	2.01318	2.00170	0.03005
1110.00000	890.98096	1.93975	1.92734	0.03164
1120.00000	881.30637	1.88344	1.87007	0.03309
1130.00000	874.23791	1.83944	1.82509	0.03442
1140.00000	869.08839	1.80437	1.78902	0.03566
1150.00000	865.17867	1.77510	1.75873	0.03681
1160.00000	862.40290	1.75101	1.73359	0.03788
1170.00000	860.45785	1.73072	1.71224	0.03888
1180.00000	859.09372	1.71314	1.69357	0.03983
1190.00000	858.33803	1.69031	1.67763	0.04073
1200.00000	858.01833	1.68548	1.66366	0.04158
1210.00000	858.05599	1.67432	1.65133	0.04240
1220.00000	858.30019	1.66420	1.64002	0.04317
1230.00000	858.88235	1.65561	1.63022	0.04392
1240.00000	859.57606	1.64768	1.62102	0.04464
1250.00000	860.64736	1.64137	1.61343	0.04534
1260.00000	861.66765	1.63505	1.60581	0.04601
1270.00000	862.92217	1.62980	1.59922	0.04667
1280.00000	864.17589	1.62471	1.59278	0.04730
1290.00000	865.63875	1.62059	1.58725	0.04792
1300.00000	867.06381	1.61648	1.58172	0.04853
1310.00000	868.81236	1.61373	1.57751	0.04913
1320.00000	870.36861	1.61042	1.57272	0.04970
1330.00000	871.97211	1.60744	1.56822	0.05026
1340.00000	873.76080	1.60529	1.56451	0.05082
1350.00000	875.44362	1.60289	1.56054	0.05137
1360.00000	877.31184	1.60132	1.55734	0.05191
1370.00000	879.21515	1.60001	1.55438	0.05244
1380.00000	881.15059	1.59895	1.55163	0.05297
1390.00000	882.93730	1.59749	1.54847	0.05348
1400.00000	884.73596	1.59621	1.54545	0.05398
1410.00000	886.73354	1.59579	1.54323	0.05448
1420.00000	888.55347	1.59485	1.54049	0.05497
1430.00000	890.37652	1.59405	1.53786	0.05545
1440.00000	892.62653	1.59492	1.53680	0.05595
1450.00000	894.46515	1.59443	1.53442	0.05642
1460.00000	896.30083	1.59406	1.53212	0.05689
1470.00000	898.13193	1.59381	1.52989	0.05735
1480.00000	900.45026	1.59540	1.52942	0.05783
1490.00000	902.28494	1.59540	1.52738	0.05828
1500.00000	904.37633	1.59642	1.52631	0.05873
1510.00000	906.20181	1.59663	1.52442	0.05917
1520.00000	908.01633	1.59692	1.52258	0.05961
1530.00000	909.81833	1.59728	1.52078	0.06004
1540.00000	911.91255	1.59878	1.52004	0.06048
1550.00000	913.69796	1.59931	1.51836	0.06090
1560.00000	915.46811	1.59991	1.51671	0.06131

TEMP-R	EQ SONIC VEL-FT/SEC	CP/CV	GAMM-O	CP-BTU/LBM-R
1570.00000	917.90037	1.60292	1.51732	0.06176
1580.00000	919.31056	1.60248	1.51464	0.06215
1590.00000	921.40568	1.60454	1.51429	0.06257
1600.00000	923.13254	1.60542	1.51280	0.06296
1610.00000	925.22869	1.60768	1.51261	0.06338
1620.00000	926.93142	1.60869	1.51119	0.06376
1630.00000	929.02947	1.61117	1.51114	0.06417
1640.00000	930.70689	1.61229	1.50979	0.06455
1650.00000	932.80691	1.61498	1.50988	0.06494
1660.00000	934.45861	1.61620	1.50859	0.06531
1670.00000	936.08992	1.61746	1.50730	0.06567
1680.00000	937.70030	1.61873	1.50602	0.06603
1690.00000	939.79195	1.62176	1.50635	0.06641
1700.00000	941.89413	1.62492	1.50677	0.06678
1710.00000	943.47179	1.62635	1.50558	0.06713
1720.00000	945.02808	1.62779	1.50438	0.06746
1730.00000	947.13204	1.63120	1.50499	0.06783
1740.00000	948.66140	1.63272	1.50384	0.06815
1750.00000	950.19777	1.63434	1.50276	0.06848
1760.00000	951.68476	1.63586	1.50159	0.06879
1770.00000	953.79160	1.63960	1.50244	0.06913
1780.00000	955.91563	1.64346	1.50339	0.06948
1790.00000	957.37326	1.64510	1.50230	0.06977
1800.00000	958.80870	1.64673	1.50119	0.07007
1810.00000	960.22203	1.64834	1.50008	0.07035
1820.00000	961.61326	1.64995	1.49894	0.07063
1830.00000	963.74888	1.65417	1.50018	0.07095
1840.00000	965.11665	1.65580	1.49908	0.07121
1850.00000	966.45004	1.65747	1.49792	0.07148
1860.00000	967.77296	1.65907	1.49677	0.07173
1870.00000	969.93284	1.66358	1.49827	0.07202
1880.00000	971.23464	1.66519	1.49715	0.07226
1890.00000	973.42411	1.66989	1.49881	0.07255
1900.00000	974.70675	1.67151	1.49772	0.07278
1910.00000	976.93061	1.67641	1.49956	0.07305
1920.00000	977.20709	1.67465	1.49548	0.07321
1930.00000	978.42528	1.67616	1.49432	0.07341
1940.00000	980.66224	1.68119	1.49631	0.07367
1950.00000	981.86401	1.68270	1.49517	0.07386
1960.00000	984.14439	1.68794	1.49736	0.07410
1970.00000	985.33162	1.68944	1.49625	0.07428
1980.00000	987.66001	1.69488	1.49864	0.07451
1990.00000	987.64358	1.69230	1.49395	0.07461
2000.00000	989.98950	1.69783	1.49645	0.07483
2010.00000	991.12351	1.69922	1.49531	0.07498
2020.00000	992.23676	1.70056	1.49414	0.07511
2030.00000	993.32929	1.70183	1.49294	0.07524
2040.00000	995.75273	1.70768	1.49576	0.07544
2050.00000	996.83550	1.70895	1.49459	0.07555
2060.00000	997.89775	1.71015	1.49338	0.07566
2070.00000	1000.39810	1.71630	1.49650	0.07584
2080.00000	1001.45204	1.71749	1.49531	0.07593
2090.00000	1002.48515	1.71862	1.49409	0.07601
2100.00000	1003.49674	1.71969	1.49283	0.07608
2110.00000	1006.09614	1.72622	1.49631	0.07624
2120.00000	1005.51194	1.72185	1.49036	0.07621
2130.00000	1008.14217	1.72851	1.49397	0.07635
2140.00000	1009.10355	1.72946	1.49265	0.07639
2150.00000	1010.04195	1.73036	1.49130	0.07643
2160.00000	1012.76772	1.73741	1.49523	0.07656
2170.00000	1013.69796	1.73833	1.49389	0.07658

TEMP-R	EQ SONIC VEL-FT/SEC	CP/CV	GAMM-O	CP-BTU/LBM-R
2180.00000	1014.60370	1.73920	1.49250	0.07659
2190.00000	1015.48456	1.74002	1.49106	0.07660
2200.00000	1016.33915	1.74078	1.48957	0.07660
2210.00000	1019.19956	1.74847	1.49398	0.07670
2220.00000	1020.04366	1.74930	1.49250	0.07669
2230.00000	1020.86018	1.75009	1.49097	0.07667
2240.00000	1021.64785	1.75085	1.48938	0.07665
2250.00000	1024.63328	1.75921	1.49422	0.07674
2260.00000	1025.40723	1.76007	1.49264	0.07670
2270.00000	1026.15084	1.76090	1.49100	0.07666
2280.00000	1026.86196	1.76173	1.48930	0.07661
2290.00000	1027.53932	1.76254	1.48753	0.07656
2300.00000	1028.18336	1.76334	1.48571	0.07651
2310.00000	1028.79216	1.76415	1.48381	0.07644
2320.00000	1031.95700	1.77387	1.48932	0.07652
2330.00000	1032.54416	1.77488	1.48743	0.07645
2340.00000	1033.09315	1.77591	1.48547	0.07638
2350.00000	1033.60366	1.77697	1.48344	0.07631
2360.00000	1036.89038	1.78796	1.48942	0.07639
2370.00000	1037.37660	1.78930	1.48740	0.07631
2380.00000	1037.82201	1.79070	1.48532	0.07623
2390.00000	1038.22549	1.79216	1.48316	0.07616
2400.00000	1038.50621	1.79370	1.48094	0.07608
2410.00000	1038.90403	1.79532	1.47864	0.07600
2420.00000	1042.41119	1.80823	1.48549	0.07608
2430.00000	1042.80023	1.81060	1.48350	0.07601
2440.00000	1043.05716	1.81278	1.48120	0.07593
2450.00000	1043.27077	1.81508	1.47885	0.07586
2460.00000	1043.44242	1.81752	1.47644	0.07578
2470.00000	1043.57192	1.82011	1.47399	0.07571
2480.00000	1047.35831	1.83579	1.48193	0.07583
2490.00000	1047.48378	1.83899	1.47959	0.07577
2500.00000	1047.57310	1.84238	1.47723	0.07571

Fig. C.1. Comparison of frozen and equilibrium sonic velocities for saturated cesium vapor.

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Appendix D

THERMODYNAMIC PROPERTIES OF SATURATED POTASSIUM

TEMP-R	PSAT-ATM	SPEC VOL FT**3/LBM		ENTHALPY BTU/LBM			ENTROPY BTU/LBM-R	
		VF	VG	HF	HFG	HG	SF	SG
960.0000	0.00012	0.02045	151939.438	-577.2363	922.0571	344.8208	0.5505	1.5109
970.0000	0.00014	0.02049	126236.875	198.9058	921.5974	1120.5032	0.5521	1.5022
980.0000	0.00017	0.02052	105294.938	200.5225	921.1265	1121.6489	0.5538	1.4937
990.0000	0.00021	0.02055	88162.313	202.1421	920.6460	1122.7881	0.5554	1.4854
1000.0000	0.00025	0.02059	74090.188	203.7666	920.1541	1123.9207	0.5571	1.4772
1010.0000	0.00030	0.02062	62487.402	205.3945	919.6516	1125.0461	0.5587	1.4692
1020.0000	0.00036	0.02066	52884.945	207.0271	919.1377	1126.1648	0.5603	1.4614
1030.0000	0.00043	0.02069	44909.113	208.6626	918.6135	1127.2761	0.5619	1.4537
1040.0000	0.00051	0.02072	38260.824	210.3049	918.0752	1128.3801	0.5635	1.4462
1050.0000	0.00060	0.02076	32700.543	211.9502	917.5264	1129.4766	0.5650	1.4389
1060.0000	0.00070	0.02079	28034.430	213.5999	916.9656	1130.5654	0.5666	1.4317
1070.0000	0.00082	0.02083	24106.270	215.2529	916.3936	1131.6465	0.5681	1.4246
1080.0000	0.00096	0.02086	20788.879	216.9128	915.8069	1132.7197	0.5697	1.4177
1090.0000	0.00112	0.02090	17978.711	218.5776	915.2070	1133.7847	0.5712	1.4109
1100.0000	0.00131	0.02093	15591.180	220.2466	914.5950	1134.8416	0.5727	1.4042
1110.0000	0.00152	0.02097	13555.945	221.9194	913.9705	1135.8899	0.5743	1.3977
1120.0000	0.00176	0.02101	11818.762	223.5977	913.3320	1136.9297	0.5758	1.3912
1130.0000	0.00203	0.02104	10329.488	225.2810	912.6802	1137.9612	0.5773	1.3849
1140.0000	0.00233	0.02108	9050.098	226.9697	912.0142	1138.9839	0.5787	1.3788
1150.0000	0.00268	0.02111	7948.184	228.6626	911.3352	1139.9978	0.5802	1.3727
1160.0000	0.00307	0.02115	6996.742	230.3582	910.6445	1141.0027	0.5817	1.3667
1170.0000	0.00350	0.02119	6173.168	232.0618	909.9370	1141.9988	0.5832	1.3609
1180.0000	0.00399	0.02122	5458.570	233.7715	909.2144	1142.9858	0.5846	1.3551
1190.0000	0.00454	0.02126	4837.082	235.4868	908.4768	1143.9636	0.5861	1.3495
1200.0000	0.00516	0.02130	4295.363	237.2068	907.7253	1144.9321	0.5875	1.3439
1210.0000	0.00584	0.02134	3822.141	238.9229	906.9685	1145.8914	0.5889	1.3385
1220.0000	0.00660	0.02137	3407.814	240.6536	906.1875	1146.8411	0.5903	1.3331
1230.0000	0.00745	0.02141	3044.294	242.3899	905.3914	1147.7813	0.5918	1.3278
1240.0000	0.00838	0.02145	2724.718	244.1311	904.5806	1148.7117	0.5932	1.3227
1250.0000	0.00942	0.02149	2443.188	245.8782	903.7546	1149.6328	0.5946	1.3176
1260.0000	0.01056	0.02152	2194.733	247.6155	902.9290	1150.5444	0.5960	1.3126
1270.0000	0.01182	0.02156	1974.984	249.3730	902.0732	1151.4463	0.5973	1.3076
1280.0000	0.01321	0.02160	1780.290	251.1370	901.2014	1152.3384	0.5987	1.3028
1290.0000	0.01473	0.02164	1607.470	252.9050	900.3157	1153.2207	0.6001	1.2980
1300.0000	0.01641	0.02168	1453.805	254.6797	899.4136	1154.0933	0.6015	1.2933
1310.0000	0.01824	0.02172	1316.926	256.4592	898.4966	1154.9558	0.6028	1.2887
1320.0000	0.02024	0.02176	1194.755	258.2449	897.5637	1155.8086	0.6042	1.2842
1330.0000	0.02243	0.02180	1085.679	260.0363	896.6458	1156.6521	0.6055	1.2797
1340.0000	0.02481	0.02184	987.960	261.8030	895.6821	1157.4851	0.6069	1.2753
1350.0000	0.02741	0.02188	900.346	263.6042	894.7041	1158.3083	0.6082	1.2709
1360.0000	0.03023	0.02191	821.667	265.4109	893.7107	1159.1218	0.6095	1.2667
1370.0000	0.03329	0.02196	750.907	267.2227	892.7024	1159.9250	0.6109	1.2625
1380.0000	0.03661	0.02200	687.169	269.0405	891.6782	1160.7188	0.6122	1.2583
1390.0000	0.04021	0.02204	629.713	270.8093	890.6941	1161.5034	0.6135	1.2543
1400.0000	0.04410	0.02208	577.778	272.6377	889.6399	1162.2776	0.6148	1.2502

TEMP-R	PSAT-ATM	SPEC VOL	FT**3/LBM	ENTHALPY BTU/LBM			ENTROPY BTU/LBM-R	
		VF	VG	HF	HFG	HG	SF	SG
1410.0000	0.04831	0.02212	530.797	274.4712	888.5708	1163.0420	0.6161	1.2463
1420.0000	0.05284	0.02216	498.241	276.3091	887.4873	1163.7964	0.6174	1.2424
1430.0000	0.05773	0.02220	449.643	278.1528	886.3884	1164.5413	0.6187	1.2385
1440.0000	0.06298	0.02224	414.588	280.0020	885.2747	1165.2766	0.6200	1.2347
1450.0000	0.06864	0.02228	382.710	281.8552	884.1470	1166.0022	0.6212	1.2310
1460.0000	0.07471	0.02232	353.686	283.7139	883.0044	1166.7183	0.6225	1.2273
1470.0000	0.08122	0.02237	327.264	285.4802	881.9468	1167.4270	0.6237	1.2237
1480.0000	0.08820	0.02241	303.112	287.3484	880.7759	1168.1243	0.6250	1.2201
1490.0000	0.09567	0.02245	281.042	289.2214	879.5906	1168.8120	0.6263	1.2166
1500.0000	0.10365	0.02249	260.851	291.1006	878.3899	1169.4905	0.6275	1.2131
1510.0000	0.11218	0.02254	242.360	292.9827	877.1775	1170.1602	0.6288	1.2097
1520.0000	0.12129	0.02258	225.405	294.8694	875.9512	1170.8206	0.6300	1.2063
1530.0000	0.13099	0.02262	209.842	296.7607	874.7112	1171.4719	0.6313	1.2030
1540.0000	0.14133	0.02267	195.543	298.6567	873.4578	1172.1145	0.6325	1.1997
1550.0000	0.15233	0.02271	182.427	300.3806	872.3723	1172.7529	0.6336	1.1964
1560.0000	0.16403	0.02275	170.314	302.2859	871.0923	1173.3782	0.6348	1.1932
1570.0000	0.17646	0.02280	159.148	304.1951	869.7996	1173.9946	0.6361	1.1901
1580.0000	0.18964	0.02284	148.848	306.1084	868.4944	1174.6028	0.6373	1.1870
1590.0000	0.20363	0.02289	139.335	308.0254	867.1772	1175.2026	0.6385	1.1839
1600.0000	0.21844	0.02293	130.542	309.9463	865.8481	1175.7944	0.6397	1.1808
1610.0000	0.23413	0.02297	122.362	312.1738	864.1943	1176.3682	0.6411	1.1778
1620.0000	0.25072	0.02302	114.837	314.0532	862.8918	1176.9451	0.6422	1.1749
1630.0000	0.26825	0.02306	107.856	315.9839	861.5291	1177.5129	0.6434	1.1720
1640.0000	0.28677	0.02311	101.380	317.9165	860.1565	1178.0730	0.6446	1.1691
1650.0000	0.30631	0.02316	95.367	319.8535	858.7722	1178.6257	0.6458	1.1662
1660.0000	0.32692	0.02320	89.778	321.7930	857.3779	1179.1709	0.6469	1.1634
1670.0000	0.34864	0.02325	84.581	323.7356	855.9736	1179.7092	0.6481	1.1607
1680.0000	0.37151	0.02329	79.743	325.6814	854.5591	1180.2405	0.6493	1.1579
1690.0000	0.39557	0.02334	75.236	327.6299	853.1350	1180.7649	0.6504	1.1552
1700.0000	0.42088	0.02339	71.035	329.5808	851.7017	1181.2825	0.6516	1.1526
1710.0000	0.44747	0.02343	67.114	331.5347	850.2593	1181.7939	0.6527	1.1499
1720.0000	0.47539	0.02348	63.454	333.4912	848.8076	1182.2988	0.6539	1.1473
1730.0000	0.50469	0.02353	60.039	335.3655	847.4358	1182.8013	0.6549	1.1448
1740.0000	0.53542	0.02358	56.840	337.3271	845.9673	1183.2944	0.6561	1.1423
1750.0000	0.56763	0.02362	53.846	339.2903	844.4915	1183.7817	0.6572	1.1398
1760.0000	0.60136	0.02367	51.043	341.2554	843.0085	1184.2639	0.6583	1.1373
1770.0000	0.63668	0.02372	48.416	343.2222	841.5183	1184.7405	0.6594	1.1349
1780.0000	0.67362	0.02377	45.952	345.1907	840.0212	1185.2119	0.6605	1.1325
1790.0000	0.71225	0.02382	43.640	347.1611	838.5176	1185.6787	0.6616	1.1301
1800.0000	0.75261	0.02387	41.470	349.1335	837.0073	1186.1409	0.6627	1.1277
1810.0000	0.79476	0.02392	39.430	351.1072	835.4912	1186.5984	0.6638	1.1254
1820.0000	0.83876	0.02397	37.512	353.0818	833.9697	1187.0515	0.6649	1.1231
1830.0000	0.88466	0.02402	35.707	355.0581	832.4426	1187.5007	0.6660	1.1209
1840.0000	0.93252	0.02407	34.009	357.0354	830.9109	1187.9463	0.6671	1.1187
1850.0000	0.98239	0.02412	32.415	358.8647	829.5315	1188.3962	0.6681	1.1165

TEMP-R	PSAT-ATM	SPEC VOL FT**3/LBM		ENTHALPY BTU/LBM			ENTROPY BTU/LBM-R	
		VF	VG	HF	HFG	HG	SF	SG
1860.0000	1.03432	0.02417	30.907	360.8440	827.9907	1188.8347	0.6691	1.1143
1870.0000	1.08839	0.02422	29.484	362.8247	826.4456	1189.2703	0.6702	1.1121
1880.0000	1.14465	0.02427	28.142	364.8054	824.8975	1189.7029	0.6713	1.1100
1890.0000	1.20316	0.02432	26.874	366.7866	823.3462	1190.1328	0.6723	1.1079
1900.0000	1.26397	0.02437	25.677	368.7688	821.7915	1190.5603	0.6733	1.1059
1910.0000	1.32716	0.02442	24.545	370.7510	820.2346	1190.9856	0.6744	1.1038
1920.0000	1.39277	0.02448	23.475	372.7332	818.6760	1191.4092	0.6754	1.1018
1930.0000	1.46088	0.02453	22.462	374.7161	817.1145	1191.8306	0.6764	1.0998
1940.0000	1.53154	0.02458	21.503	376.6982	815.5522	1192.2505	0.6775	1.0979
1950.0000	1.60482	0.02463	20.595	378.6812	813.9883	1192.6694	0.6785	1.0959
1960.0000	1.68077	0.02469	19.734	380.6650	812.4219	1193.0869	0.6795	1.0940
1970.0000	1.75948	0.02474	18.924	382.6498	811.1270	1193.5208	0.6804	1.0921
1980.0000	1.84100	0.02479	18.149	384.6362	809.5605	1193.9368	0.6814	1.0903
1990.0000	1.92538	0.02485	17.414	386.6256	807.9932	1194.3528	0.6824	1.0884
2000.0000	2.01270	0.02490	16.716	388.6170	806.4268	1194.7688	0.6834	1.0866
2010.0000	2.10303	0.02496	16.052	390.6107	804.8599	1195.1846	0.6844	1.0848
2020.0000	2.19643	0.02501	15.422	392.6057	803.2952	1195.6008	0.6854	1.0830
2030.0000	2.29297	0.02507	14.822	394.6029	801.7305	1196.0173	0.6863	1.0813
2040.0000	2.39271	0.02512	14.252	396.6026	800.1672	1196.4348	0.6873	1.0795
2050.0000	2.49572	0.02518	13.709	398.6046	798.6055	1196.8530	0.6883	1.0778
2060.0000	2.60207	0.02523	13.192	400.6088	797.0457	1197.2725	0.6892	1.0761
2070.0000	2.71181	0.02529	12.700	402.6153	795.4878	1197.6931	0.6902	1.0745
2080.0000	2.82504	0.02535	12.230	404.6241	793.9324	1198.1155	0.6911	1.0728
2090.0000	2.94179	0.02540	11.783	406.6352	792.3792	1198.5396	0.6921	1.0712
2100.0000	3.06216	0.02546	11.356	408.6487	790.8291	1198.9648	0.6930	1.0696
2110.0000	3.18620	0.02552	10.955	409.6647	789.7231	1199.4253	0.6938	1.0681
2120.0000	3.31399	0.02557	10.566	411.6830	788.1777	1199.8557	0.6947	1.0665
2130.0000	3.44559	0.02563	10.194	413.6550	786.6333	1200.2883	0.6956	1.0649
2140.0000	3.58107	0.02569	9.839	415.6318	785.0923	1200.7241	0.6966	1.0634
2150.0000	3.72050	0.02575	9.500	417.6074	783.5544	1201.1619	0.6975	1.0619
2160.0000	3.86394	0.02581	9.176	419.5825	782.0203	1201.6028	0.6984	1.0604
2170.0000	4.01147	0.02587	8.865	421.5574	780.4893	1202.0466	0.6993	1.0590
2180.0000	4.16316	0.02593	8.568	423.5305	778.9631	1202.4937	0.7002	1.0575
2190.0000	4.31908	0.02599	8.284	425.5039	777.4397	1202.9436	0.7011	1.0561
2200.0000	4.47927	0.02605	8.012	427.4773	775.9199	1203.3972	0.7020	1.0547
2210.0000	4.64383	0.02611	7.751	429.4497	774.4048	1203.8545	0.7029	1.0533
2220.0000	4.81282	0.02617	7.501	431.4224	772.8931	1204.3154	0.7038	1.0519
2230.0000	4.98631	0.02623	7.262	433.3931	771.3865	1204.7795	0.7047	1.0506
2240.0000	5.16436	0.02629	7.032	435.3652	769.8826	1205.2478	0.7055	1.0492
2250.0000	5.34705	0.02635	6.818	436.7031	769.0747	1205.7778	0.7062	1.0480
2260.0000	5.53443	0.02641	6.607	438.6689	767.5859	1206.2549	0.7070	1.0467
2270.0000	5.72659	0.02648	6.404	440.6462	766.0886	1206.7349	0.7079	1.0454
2280.0000	5.92359	0.02654	6.209	442.6243	764.5947	1207.2190	0.7088	1.0441
2290.0000	6.12549	0.02660	6.021	444.6023	763.1045	1207.7068	0.7096	1.0429
2300.0000	6.33235	0.02666	5.841	446.5815	761.6174	1208.1990	0.7105	1.0416

TEMP-R	PSAT-ATM	SPEC VOL FT**3/LBM		ENTHALPY BTU/LBM			ENTROPY BTU/LBM-R	
		VF	VG	HF	HFG	HG	SF	SG
2310.0000	6.54426	0.02673	5.668	448.5613	760.1340	1208.6953	0.7113	1.0404
2320.0000	6.76128	0.02579	5.502	450.5417	758.6536	1209.1953	0.7122	1.0392
2330.0000	6.98347	0.02586	5.342	452.5239	757.1755	1209.6995	0.7130	1.0380
2340.0000	7.21088	0.02592	5.188	454.5063	755.7017	1210.2080	0.7139	1.0368
2350.0000	7.44361	0.02699	5.040	456.4905	754.2295	1210.7200	0.7147	1.0357
2360.0000	7.68172	0.02705	4.897	458.4761	752.7600	1211.2361	0.7156	1.0345
2370.0000	7.92525	0.02712	4.760	460.4634	751.2932	1211.7566	0.7164	1.0334
2380.0000	8.17428	0.02718	4.627	462.5071	749.7678	1212.2749	0.7172	1.0323
2390.0000	8.42887	0.02725	4.499	464.5007	748.3018	1212.8025	0.7181	1.0312
2400.0000	8.68910	0.02732	4.383	465.5090	747.9253	1213.4343	0.7185	1.0302
2410.0000	8.95499	0.02738	4.264	467.4568	746.5188	1213.9756	0.7193	1.0291
2420.0000	9.22666	0.02745	4.150	469.4609	745.0532	1214.5142	0.7202	1.0280
2430.0000	9.50415	0.02752	4.040	471.4727	743.5825	1215.0552	0.7210	1.0270
2440.0000	9.78752	0.02759	3.933	473.4873	742.1128	1215.6001	0.7218	1.0260
2450.0000	10.07683	0.02766	3.830	475.5061	740.6414	1216.1475	0.7226	1.0249
2460.0000	10.37212	0.02772	3.731	477.5295	739.1687	1216.6982	0.7235	1.0239
2470.0000	10.67351	0.02779	3.634	479.6772	737.5608	1217.2380	0.7243	1.0229
2480.0000	10.98098	0.02786	3.542	481.5938	736.2122	1217.8059	0.7251	1.0219
2490.0000	11.29469	0.02793	3.452	483.7441	734.6067	1218.3508	0.7259	1.0210
2500.0000	11.61459	0.02800	3.365	485.7971	733.1128	1218.9099	0.7268	1.0200

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Appendix E

TRANSPORT PROPERTIES OF SATURATED POTASSIUM

HEAT CAP BTU/LBM-R		VISCOSITY LBM/HR-FT		THERM COND BTU/HR-FT-F	
TEMP-R	LIQUID	LIQUID	VAPOR	LIQUID	VAPOR
960.0000	0.1863	0.6777	0.0487	26.2713	0.0091
965.0000	0.1862	0.6734	0.0489	26.2134	0.0091
970.0000	0.1861	0.6691	0.0490	26.1558	0.0091
975.0000	0.1860	0.6649	0.0491	26.0984	0.0092
980.0000	0.1859	0.6608	0.0492	26.0412	0.0092
985.0000	0.1858	0.6567	0.0494	25.9843	0.0092
990.0000	0.1857	0.6527	0.0495	25.9275	0.0092
995.0000	0.1856	0.6488	0.0496	25.8710	0.0093
1000.0000	0.1855	0.6449	0.0497	25.8147	0.0093
1005.0000	0.1855	0.6411	0.0499	25.7585	0.0093
1010.0000	0.1854	0.6374	0.0500	25.7026	0.0093
1015.0000	0.1853	0.6337	0.0501	25.6469	0.0093
1020.0000	0.1852	0.6300	0.0502	25.5914	0.0094
1025.0000	0.1851	0.6265	0.0504	25.5361	0.0094
1030.0000	0.1850	0.6229	0.0505	25.4809	0.0094
1035.0000	0.1849	0.6195	0.0506	25.4260	0.0094
1040.0000	0.1848	0.6161	0.0507	25.3712	0.0095
1045.0000	0.1848	0.6127	0.0508	25.3166	0.0095
1050.0000	0.1847	0.6094	0.0510	25.2622	0.0095
1055.0000	0.1846	0.6061	0.0511	25.2080	0.0095
1060.0000	0.1845	0.6029	0.0512	25.1540	0.0096
1065.0000	0.1844	0.5997	0.0513	25.1001	0.0096
1070.0000	0.1844	0.5966	0.0515	25.0464	0.0096
1075.0000	0.1843	0.5935	0.0516	24.9929	0.0096
1080.0000	0.1842	0.5905	0.0517	24.9395	0.0096
1085.0000	0.1841	0.5875	0.0518	24.8863	0.0097
1090.0000	0.1841	0.5846	0.0519	24.8333	0.0097
1095.0000	0.1840	0.5817	0.0520	24.7804	0.0097
1100.0000	0.1839	0.5788	0.0522	24.7276	0.0097
1105.0000	0.1839	0.5760	0.0523	24.6750	0.0098
1110.0000	0.1838	0.5732	0.0524	24.6226	0.0098
1115.0000	0.1837	0.5704	0.0525	24.5704	0.0098
1120.0000	0.1837	0.5677	0.0526	24.5182	0.0098
1125.0000	0.1836	0.5650	0.0528	24.4663	0.0098
1130.0000	0.1835	0.5624	0.0529	24.4144	0.0099
1135.0000	0.1835	0.5598	0.0530	24.3627	0.0099
1140.0000	0.1834	0.5572	0.0531	24.3112	0.0099
1145.0000	0.1833	0.5547	0.0532	24.2598	0.0099
1150.0000	0.1833	0.5522	0.0533	24.2085	0.0099
1155.0000	0.1832	0.5497	0.0535	24.1574	0.0100
1160.0000	0.1832	0.5473	0.0536	24.1064	0.0100
1165.0000	0.1831	0.5434	0.0537	24.0555	0.0100
1170.0000	0.1831	0.5401	0.0538	24.0047	0.0100
1175.0000	0.1830	0.5367	0.0539	23.9541	0.0101
1180.0000	0.1830	0.5335	0.0540	23.9037	0.0101
1185.0000	0.1829	0.5302	0.0541	23.8533	0.0101
1190.0000	0.1829	0.5270	0.0543	23.8031	0.0101
1195.0000	0.1828	0.5239	0.0544	23.7529	0.0101
1200.0000	0.1828	0.5208	0.0545	23.7029	0.0102
1205.0000	0.1827	0.5178	0.0546	23.6531	0.0102
1210.0000	0.1827	0.5148	0.0547	23.6033	0.0102
1215.0000	0.1826	0.5118	0.0548	23.5537	0.0102
1220.0000	0.1826	0.5089	0.0549	23.5041	0.0102
1225.0000	0.1826	0.5060	0.0551	23.4547	0.0103
1230.0000	0.1825	0.5031	0.0552	23.4054	0.0103
1235.0000	0.1825	0.5003	0.0553	23.3562	0.0103

HEAT CAP BTU/LBM-R		VISCOSITY LBM/HR-FT		THERM COND BTU/HR-FT-F	
TEMP-R	LIQUID	LIQUID	VAPOR	LIQUID	VAPOR
1240.0000	0.1824	0.4976	0.0554	23.3071	0.0103
1245.0000	0.1824	0.4948	0.0555	23.2582	0.0104
1250.0000	0.1824	0.4922	0.0556	23.2093	0.0104
1255.0000	0.1823	0.4895	0.0557	23.1605	0.0104
1260.0000	0.1823	0.4869	0.0558	23.1119	0.0104
1265.0000	0.1823	0.4843	0.0559	23.0633	0.0104
1270.0000	0.1822	0.4817	0.0561	23.0149	0.0105
1275.0000	0.1822	0.4792	0.0562	22.9665	0.0105
1280.0000	0.1822	0.4767	0.0563	22.9182	0.0105
1285.0000	0.1822	0.4743	0.0564	22.8701	0.0105
1290.0000	0.1821	0.4719	0.0565	22.8220	0.0105
1295.0000	0.1821	0.4695	0.0566	22.7741	0.0106
1300.0000	0.1821	0.4671	0.0567	22.7262	0.0106
1305.0000	0.1821	0.4648	0.0568	22.6784	0.0106
1310.0000	0.1820	0.4625	0.0569	22.6307	0.0106
1315.0000	0.1820	0.4602	0.0570	22.5831	0.0106
1320.0000	0.1820	0.4579	0.0571	22.5356	0.0107
1325.0000	0.1820	0.4557	0.0573	22.4882	0.0107
1330.0000	0.1820	0.4535	0.0574	22.4409	0.0107
1335.0000	0.1819	0.4514	0.0575	22.3937	0.0107
1340.0000	0.1819	0.4492	0.0576	22.3465	0.0107
1345.0000	0.1819	0.4471	0.0577	22.2995	0.0108
1350.0000	0.1819	0.4450	0.0578	22.2525	0.0108
1355.0000	0.1819	0.4430	0.0579	22.2056	0.0108
1360.0000	0.1819	0.4409	0.0580	22.1588	0.0108
1365.0000	0.1819	0.4389	0.0581	22.1120	0.0108
1370.0000	0.1819	0.4369	0.0582	22.0654	0.0109
1375.0000	0.1819	0.4350	0.0583	22.0188	0.0109
1380.0000	0.1819	0.4330	0.0584	21.9723	0.0109
1385.0000	0.1819	0.4311	0.0585	21.9259	0.0109
1390.0000	0.1819	0.4292	0.0586	21.8796	0.0109
1395.0000	0.1819	0.4273	0.0587	21.8333	0.0110
1400.0000	0.1819	0.4255	0.0589	21.7871	0.0110
1405.0000	0.1819	0.4237	0.0590	21.7410	0.0110
1410.0000	0.1819	0.4218	0.0591	21.6950	0.0110
1415.0000	0.1819	0.4201	0.0592	21.6490	0.0110
1420.0000	0.1819	0.4183	0.0593	21.6031	0.0111
1425.0000	0.1819	0.4165	0.0594	21.5573	0.0111
1430.0000	0.1819	0.4148	0.0595	21.5116	0.0111
1435.0000	0.1819	0.4131	0.0596	21.4659	0.0111
1440.0000	0.1819	0.4114	0.0597	21.4203	0.0111
1445.0000	0.1819	0.4097	0.0598	21.3747	0.0112
1450.0000	0.1819	0.4081	0.0599	21.3293	0.0112
1455.0000	0.1819	0.4064	0.0600	21.2839	0.0112
1460.0000	0.1819	0.4048	0.0601	21.2385	0.0112
1465.0000	0.1820	0.4032	0.0602	21.1933	0.0112
1470.0000	0.1820	0.4016	0.0603	21.1481	0.0112
1475.0000	0.1820	0.4000	0.0604	21.1029	0.0113
1480.0000	0.1820	0.3985	0.0605	21.0578	0.0113
1485.0000	0.1820	0.3969	0.0606	21.0128	0.0113
1490.0000	0.1820	0.3954	0.0607	20.9679	0.0113
1495.0000	0.1821	0.3939	0.0608	20.9230	0.0113
1500.0000	0.1821	0.3924	0.0609	20.8782	0.0114
1505.0000	0.1821	0.3910	0.0610	20.8334	0.0114
1510.0000	0.1821	0.3895	0.0611	20.7887	0.0114

HEAT CAP BTU/LBM-R		VISCOSITY LBM/HR-FT		THERM COND BTU/HR-FT-F	
TEMP-R	LIQUID	LIQUID	VAPOR	LIQUID	VAPOR
1515.0000	0.1822	0.3881	0.0612	20.7441	0.0114
1520.0000	0.1822	0.3866	0.0613	20.6995	0.0114
1525.0000	0.1822	0.3852	0.0614	20.6550	0.0115
1530.0000	0.1822	0.3838	0.0615	20.6105	0.0115
1535.0000	0.1823	0.3824	0.0616	20.5661	0.0115
1540.0000	0.1823	0.3811	0.0617	20.5217	0.0115
1545.0000	0.1823	0.3797	0.0618	20.4774	0.0115
1550.0000	0.1824	0.3784	0.0619	20.4332	0.0115
1555.0000	0.1824	0.3770	0.0620	20.3890	0.0116
1560.0000	0.1825	0.3757	0.0621	20.3449	0.0116
1565.0000	0.1825	0.3744	0.0622	20.3008	0.0116
1570.0000	0.1825	0.3731	0.0623	20.2568	0.0116
1575.0000	0.1826	0.3718	0.0624	20.2128	0.0116
1580.0000	0.1826	0.3706	0.0625	20.1689	0.0117
1585.0000	0.1827	0.3693	0.0626	20.1250	0.0117
1590.0000	0.1827	0.3681	0.0627	20.0812	0.0117
1595.0000	0.1827	0.3668	0.0628	20.0374	0.0117
1600.0000	0.1828	0.3656	0.0629	19.9937	0.0117
1605.0000	0.1828	0.3644	0.0630	19.9500	0.0118
1610.0000	0.1829	0.3632	0.0631	19.9064	0.0118
1615.0000	0.1829	0.3620	0.0632	19.8629	0.0118
1620.0000	0.1830	0.3609	0.0633	19.8194	0.0118
1625.0000	0.1830	0.3597	0.0634	19.7759	0.0118
1630.0000	0.1831	0.3586	0.0635	19.7325	0.0118
1635.0000	0.1831	0.3574	0.0636	19.6891	0.0119
1640.0000	0.1832	0.3563	0.0637	19.6458	0.0119
1645.0000	0.1833	0.3552	0.0638	19.6025	0.0119
1650.0000	0.1833	0.3541	0.0639	19.5593	0.0119
1655.0000	0.1834	0.3530	0.0640	19.5161	0.0119
1660.0000	0.1834	0.3519	0.0641	19.4729	0.0120
1665.0000	0.1835	0.3508	0.0642	19.4298	0.0120
1670.0000	0.1836	0.3497	0.0643	19.3868	0.0120
1675.0000	0.1836	0.3487	0.0644	19.3438	0.0120
1680.0000	0.1837	0.3476	0.0645	19.3008	0.0120
1685.0000	0.1837	0.3466	0.0646	19.2579	0.0120
1690.0000	0.1838	0.3455	0.0647	19.2150	0.0121
1695.0000	0.1839	0.3445	0.0648	19.1722	0.0121
1700.0000	0.1839	0.3435	0.0649	19.1294	0.0121
1705.0000	0.1840	0.3425	0.0649	19.0867	0.0121
1710.0000	0.1841	0.3415	0.0650	19.0439	0.0121
1715.0000	0.1842	0.3405	0.0651	19.0013	0.0121
1720.0000	0.1842	0.3395	0.0652	18.9586	0.0122
1725.0000	0.1843	0.3386	0.0653	18.9161	0.0122
1730.0000	0.1844	0.3376	0.0654	18.8735	0.0122
1735.0000	0.1845	0.3366	0.0655	18.8310	0.0122
1740.0000	0.1845	0.3357	0.0656	18.7886	0.0122
1745.0000	0.1846	0.3348	0.0657	18.7461	0.0123
1750.0000	0.1847	0.3338	0.0658	18.7037	0.0123
1755.0000	0.1848	0.3329	0.0659	18.6614	0.0123
1760.0000	0.1849	0.3320	0.0660	18.6191	0.0123
1765.0000	0.1850	0.3311	0.0661	18.5768	0.0123
1770.0000	0.1850	0.3302	0.0662	18.5346	0.0123
1775.0000	0.1851	0.3293	0.0663	18.4924	0.0124
1780.0000	0.1852	0.3284	0.0664	18.4502	0.0124
1785.0000	0.1853	0.3275	0.0665	18.4081	0.0124

TEMP-R	LIQUID	LIQUID	VAPOR	LIQUID	VAPOR
1790.0000	0.1854	0.3267	0.0665	18.3660	0.0124
1795.0000	0.1855	0.3258	0.0666	18.3239	0.0124
1800.0000	0.1856	0.3250	0.0667	18.2819	0.0124
1805.0000	0.1857	0.3241	0.0668	18.2399	0.0125
1810.0000	0.1858	0.3233	0.0669	18.1980	0.0125
1815.0000	0.1859	0.3224	0.0670	18.1561	0.0125
1820.0000	0.1860	0.3216	0.0671	18.1142	0.0125
1825.0000	0.1861	0.3208	0.0672	18.0723	0.0125
1830.0000	0.1862	0.3200	0.0673	18.0305	0.0125
1835.0000	0.1863	0.3192	0.0674	17.9887	0.0126
1840.0000	0.1864	0.3184	0.0675	17.9470	0.0126
1845.0000	0.1865	0.3176	0.0676	17.9053	0.0126
1850.0000	0.1866	0.3168	0.0677	17.8636	0.0126
1855.0000	0.1867	0.3160	0.0677	17.8220	0.0126
1860.0000	0.1868	0.3152	0.0678	17.7803	0.0127
1865.0000	0.1869	0.3144	0.0679	17.7388	0.0127
1870.0000	0.1870	0.3137	0.0680	17.6972	0.0127
1875.0000	0.1871	0.3129	0.0681	17.6557	0.0127
1880.0000	0.1872	0.3122	0.0682	17.6142	0.0127
1885.0000	0.1873	0.3114	0.0683	17.5727	0.0127
1890.0000	0.1874	0.3107	0.0684	17.5313	0.0128
1895.0000	0.1875	0.3099	0.0685	17.4899	0.0128
1900.0000	0.1877	0.3092	0.0686	17.4485	0.0128
1905.0000	0.1878	0.3085	0.0687	17.4072	0.0128
1910.0000	0.1879	0.3078	0.0687	17.3659	0.0128
1915.0000	0.1880	0.3071	0.0688	17.3246	0.0128
1920.0000	0.1881	0.3063	0.0689	17.2834	0.0129
1925.0000	0.1883	0.3056	0.0690	17.2422	0.0129
1930.0000	0.1884	0.3049	0.0691	17.2010	0.0129
1935.0000	0.1885	0.3043	0.0692	17.1598	0.0129
1940.0000	0.1886	0.3036	0.0693	17.1187	0.0129
1945.0000	0.1888	0.3029	0.0694	17.0776	0.0129
1950.0000	0.1889	0.3022	0.0695	17.0365	0.0130
1955.0000	0.1890	0.3015	0.0695	16.9955	0.0130
1960.0000	0.1891	0.3009	0.0696	16.9544	0.0130
1965.0000	0.1893	0.3002	0.0697	16.9134	0.0130
1970.0000	0.1894	0.2995	0.0698	16.8725	0.0130
1975.0000	0.1895	0.2989	0.0699	16.8315	0.0130
1980.0000	0.1897	0.2982	0.0700	16.7906	0.0131
1985.0000	0.1898	0.2976	0.0701	16.7497	0.0131
1990.0000	0.1899	0.2970	0.0702	16.7089	0.0131
1995.0000	0.1901	0.2963	0.0703	16.6680	0.0131
2000.0000	0.1902	0.2957	0.0703	16.6272	0.0131
2005.0000	0.1904	0.2951	0.0704	16.5864	0.0131
2010.0000	0.1905	0.2945	0.0705	16.5457	0.0132
2015.0000	0.1906	0.2938	0.0706	16.5050	0.0132
2020.0000	0.1908	0.2932	0.0707	16.4642	0.0132
2025.0000	0.1909	0.2926	0.0708	16.4236	0.0132
2030.0000	0.1911	0.2920	0.0709	16.3829	0.0132
2035.0000	0.1912	0.2914	0.0710	16.3423	0.0132
2040.0000	0.1914	0.2908	0.0710	16.3017	0.0132
2045.0000	0.1915	0.2902	0.0711	16.2611	0.0133
2050.0000	0.1917	0.2896	0.0712	16.2205	0.0133
2055.0000	0.1918	0.2891	0.0713	16.1800	0.0133
2060.0000	0.1920	0.2885	0.0714	16.1395	0.0133

HEAT CAP BTU/LBM-R		VISCOSITY LBM/HR-FT		THERM COND BTU/HR-FT-F	
TEMP-R	LIQUID	LIQUID	VAPOR	LIQUID	VAPOR
2065.0000	0.1921	0.2879	0.0715	16.0990	0.0133
2070.0000	0.1923	0.2873	0.0716	16.0585	0.0133
2075.0000	0.1924	0.2868	0.0716	16.0181	0.0134
2080.0000	0.1926	0.2862	0.0717	15.9777	0.0134
2085.0000	0.1928	0.2856	0.0718	15.9373	0.0134
2090.0000	0.1929	0.2851	0.0719	15.8969	0.0134
2095.0000	0.1931	0.2845	0.0720	15.8565	0.0134
2100.0000	0.1932	0.2840	0.0721	15.8162	0.0134
2105.0000	0.1934	0.2834	0.0722	15.7759	0.0135
2110.0000	0.1936	0.2829	0.0723	15.7356	0.0135
2115.0000	0.1937	0.2824	0.0723	15.6954	0.0135
2120.0000	0.1939	0.2818	0.0724	15.6551	0.0135
2125.0000	0.1941	0.2813	0.0725	15.6149	0.0135
2130.0000	0.1942	0.2808	0.0726	15.5747	0.0135
2135.0000	0.1944	0.2802	0.0727	15.5345	0.0136
2140.0000	0.1946	0.2797	0.0728	15.4944	0.0136
2145.0000	0.1947	0.2792	0.0728	15.4542	0.0136
2150.0000	0.1949	0.2787	0.0729	15.4141	0.0136
2155.0000	0.1951	0.2782	0.0730	15.3740	0.0136
2160.0000	0.1953	0.2777	0.0731	15.3340	0.0136
2165.0000	0.1954	0.2772	0.0732	15.2939	0.0136
2170.0000	0.1956	0.2767	0.0733	15.2539	0.0137
2175.0000	0.1958	0.2762	0.0734	15.2139	0.0137
2180.0000	0.1960	0.2757	0.0734	15.1739	0.0137
2185.0000	0.1962	0.2752	0.0735	15.1339	0.0137
2190.0000	0.1964	0.2747	0.0736	15.0939	0.0137
2195.0000	0.1965	0.2742	0.0737	15.0540	0.0137
2200.0000	0.1967	0.2737	0.0738	15.0141	0.0138
2205.0000	0.1969	0.2733	0.0739	14.9742	0.0138
2210.0000	0.1971	0.2728	0.0739	14.9343	0.0138
2215.0000	0.1973	0.2723	0.0740	14.8945	0.0138
2220.0000	0.1975	0.2718	0.0741	14.8546	0.0138
2225.0000	0.1977	0.2714	0.0742	14.8148	0.0138
2230.0000	0.1979	0.2709	0.0743	14.7750	0.0139
2235.0000	0.1980	0.2704	0.0744	14.7352	0.0139
2240.0000	0.1982	0.2700	0.0744	14.6955	0.0139
2245.0000	0.1984	0.2695	0.0745	14.6557	0.0139
2250.0000	0.1986	0.2691	0.0746	14.6160	0.0139
2255.0000	0.1988	0.2686	0.0747	14.5763	0.0139
2260.0000	0.1990	0.2682	0.0748	14.5366	0.0139
2265.0000	0.1992	0.2677	0.0749	14.4969	0.0140
2270.0000	0.1994	0.2673	0.0749	14.4573	0.0140
2275.0000	0.1996	0.2669	0.0750	14.4176	0.0140
2280.0000	0.1998	0.2664	0.0751	14.3780	0.0140
2285.0000	0.2000	0.2660	0.0752	14.3384	0.0140
2290.0000	0.2002	0.2656	0.0753	14.2988	0.0140
2295.0000	0.2005	0.2651	0.0754	14.2593	0.0141
2300.0000	0.2007	0.2647	0.0754	14.2197	0.0141
2305.0000	0.2009	0.2643	0.0755	14.1802	0.0141
2310.0000	0.2011	0.2639	0.0756	14.1407	0.0141
2315.0000	0.2013	0.2634	0.0757	14.1012	0.0141
2320.0000	0.2015	0.2630	0.0758	14.0617	0.0141
2325.0000	0.2017	0.2626	0.0758	14.0222	0.0141
2330.0000	0.2019	0.2622	0.0759	13.9828	0.0142
2335.0000	0.2022	0.2618	0.0760	13.9433	0.0142

HEAT CAP BTU/LBM-R		VISCOSITY LBM/HR-FT		THERM COND BTU/HR-FT-F	
TEMP-R	LIQUID	LIQUID	VAPOR	LIQUID	VAPOR
2340.0000	0.2024	0.2614	0.0761	13.9039	0.0142
2345.0000	0.2026	0.2610	0.0762	13.8645	0.0142
2350.0000	0.2028	0.2606	0.0762	13.8251	0.0142
2355.0000	0.2030	0.2602	0.0763	13.7857	0.0142
2360.0000	0.2033	0.2598	0.0764	13.7464	0.0143
2365.0000	0.2035	0.2594	0.0765	13.7070	0.0143
2370.0000	0.2037	0.2590	0.0766	13.6677	0.0143
2375.0000	0.2039	0.2586	0.0767	13.6284	0.0143
2380.0000	0.2042	0.2582	0.0767	13.5891	0.0143
2385.0000	0.2044	0.2578	0.0768	13.5498	0.0143
2390.0000	0.2046	0.2574	0.0769	13.5106	0.0143
2395.0000	0.2048	0.2571	0.0770	13.4713	0.0144
2400.0000	0.2051	0.2567	0.0771	13.4321	0.0144
2405.0000	0.2053	0.2563	0.0771	13.3929	0.0144
2410.0000	0.2055	0.2559	0.0772	13.3537	0.0144
2415.0000	0.2058	0.2556	0.0773	13.3145	0.0144
2420.0000	0.2060	0.2552	0.0774	13.2753	0.0144
2425.0000	0.2062	0.2548	0.0775	13.2362	0.0144
2430.0000	0.2065	0.2544	0.0775	13.1970	0.0145
2435.0000	0.2067	0.2541	0.0776	13.1579	0.0145
2440.0000	0.2070	0.2537	0.0777	13.1188	0.0145
2445.0000	0.2072	0.2534	0.0778	13.0797	0.0145
2450.0000	0.2075	0.2530	0.0779	13.0406	0.0145
2455.0000	0.2077	0.2526	0.0779	13.0015	0.0145
2460.0000	0.2079	0.2523	0.0780	12.9624	0.0145
2465.0000	0.2082	0.2519	0.0781	12.9234	0.0146
2470.0000	0.2084	0.2516	0.0782	12.8844	0.0146
2475.0000	0.2087	0.2512	0.0782	12.8454	0.0146
2480.0000	0.2089	0.2509	0.0783	12.8063	0.0146
2485.0000	0.2092	0.2505	0.0784	12.7673	0.0146
2490.0000	0.2094	0.2502	0.0785	12.7284	0.0146
2495.0000	0.2097	0.2499	0.0786	12.6894	0.0147
2500.0000	0.2099	0.2495	0.0786	12.6505	0.0147

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Appendix F

SONIC VELOCITY AND RELATED PROPERTIES OF
SATURATED POTASSIUM VAPOR

Values of the equilibrium sonic velocity are tabulated. GAMM-0 is the zero-pressure specific heat ratio. The instability in GAMM-0 is of numerical origin.

TEMP-R	EQ SONIC VEL-FT/SEC	CP/CV	GAMM-O	CP-BTU/LBM-R
960.00000	1387.54971	1.58479	1.58058	0.14559
970.00000	1393.06581	1.58161	1.57708	0.14663
980.00000	1398.51728	1.57843	1.57357	0.14769
990.00000	1403.90685	1.57528	1.57006	0.14879
1000.00000	1409.25806	1.57220	1.56661	0.14992
1010.00000	1414.50968	1.56905	1.56308	0.15107
1020.00000	1419.72869	1.56599	1.55961	0.15225
1030.00000	1424.89647	1.56297	1.55616	0.15346
1040.00000	1430.01523	1.55999	1.55275	0.15469
1050.00000	1435.11860	1.55714	1.54943	0.15596
1060.00000	1440.11579	1.55420	1.54601	0.15724
1070.00000	1445.10217	1.55139	1.54270	0.15856
1080.00000	1450.04925	1.54864	1.53943	0.15990
1090.00000	1454.95867	1.54596	1.53621	0.16126
1100.00000	1459.83257	1.54335	1.53303	0.16265
1110.00000	1464.67336	1.54081	1.52991	0.16406
1120.00000	1469.43337	1.53825	1.52674	0.16549
1130.00000	1474.20947	1.53586	1.52372	0.16695
1140.00000	1479.01393	1.53366	1.52088	0.16843
1150.00000	1483.67980	1.53132	1.51786	0.16992
1160.00000	1488.37525	1.52918	1.51502	0.17144
1170.00000	1493.04757	1.52713	1.51225	0.17298
1180.00000	1497.62753	1.52502	1.50940	0.17453
1190.00000	1502.25190	1.52314	1.50675	0.17610
1200.00000	1506.85564	1.52135	1.50417	0.17769
1210.00000	1511.44027	1.51965	1.50165	0.17930
1220.00000	1515.92099	1.51787	1.49903	0.18091
1230.00000	1520.46603	1.51635	1.49665	0.18255
1240.00000	1524.89603	1.51474	1.49414	0.18420
1250.00000	1529.50426	1.51360	1.49208	0.18587
1260.00000	1533.89711	1.51216	1.48970	0.18754
1270.00000	1538.37574	1.51101	1.48758	0.18923
1280.00000	1542.72581	1.50974	1.48531	0.19093
1290.00000	1547.17242	1.50877	1.48332	0.19264
1300.00000	1551.47985	1.50766	1.48116	0.19435
1310.00000	1556.02981	1.50714	1.47956	0.19610
1320.00000	1560.30230	1.50619	1.47752	0.19783
1330.00000	1564.55119	1.50532	1.47552	0.19956
1340.00000	1568.92887	1.50482	1.47386	0.20132
1350.00000	1573.13709	1.50411	1.47197	0.20307
1360.00000	1577.48847	1.50378	1.47044	0.20484
1370.00000	1581.83013	1.50356	1.46897	0.20662
1380.00000	1586.16327	1.50342	1.46757	0.20840
1390.00000	1590.29853	1.50303	1.46589	0.21016
1400.00000	1594.40902	1.50270	1.46424	0.21193
1410.00000	1598.70260	1.50282	1.46302	0.21372
1420.00000	1602.77315	1.50264	1.46146	0.21549
1430.00000	1606.81899	1.50251	1.45995	0.21726
1440.00000	1611.30485	1.50332	1.45931	0.21908
1450.00000	1615.31891	1.50335	1.45789	0.22084
1460.00000	1619.30702	1.50344	1.45651	0.22261
1470.00000	1623.27082	1.50359	1.45515	0.22437
1480.00000	1627.75134	1.50480	1.45480	0.22619
1490.00000	1631.68266	1.50509	1.45354	0.22794
1500.00000	1635.88129	1.50598	1.45283	0.22972
1510.00000	1639.77086	1.50639	1.45164	0.23147
1520.00000	1643.63375	1.50686	1.45048	0.23321
1530.00000	1647.46905	1.50738	1.44934	0.23493
1540.00000	1651.61679	1.50856	1.44882	0.23669
1550.00000	1655.40902	1.50919	1.44775	0.23841
1560.00000	1659.17357	1.50986	1.44669	0.24011

TEMP-R	EQ SONIC VEL-FT/SEC	CP/CV	GAMM-O	CP-BTU/LBM-R
1570.00000	1663.66660	1.51195	1.44697	0.24189
1580.00000	1667.00784	1.51204	1.44532	0.24353
1590.00000	1671.10537	1.51360	1.44504	0.24524
1600.00000	1674.78170	1.51447	1.44410	0.24690
1610.00000	1678.86340	1.51617	1.44392	0.24860
1620.00000	1682.49538	1.51714	1.44303	0.25023
1630.00000	1686.56413	1.51899	1.44295	0.25190
1640.00000	1690.15206	1.52005	1.44211	0.25351
1650.00000	1694.20875	1.52204	1.44213	0.25515
1660.00000	1697.75276	1.52318	1.44133	0.25672
1670.00000	1701.26652	1.52435	1.44054	0.25828
1680.00000	1704.74937	1.52555	1.43976	0.25981
1690.00000	1708.77013	1.52778	1.43994	0.26139
1700.00000	1712.79716	1.53010	1.44019	0.26296
1710.00000	1716.22400	1.53142	1.43948	0.26444
1720.00000	1719.61968	1.53275	1.43878	0.26590
1730.00000	1723.63133	1.53525	1.43916	0.26741
1740.00000	1726.98416	1.53665	1.43849	0.26882
1750.00000	1730.30616	1.53805	1.43782	0.27021
1760.00000	1733.59653	1.53947	1.43715	0.27158
1770.00000	1737.58733	1.54219	1.43769	0.27301
1780.00000	1741.59167	1.54499	1.43830	0.27442
1790.00000	1744.83154	1.54650	1.43769	0.27572
1800.00000	1748.03998	1.54801	1.43708	0.27698
1810.00000	1751.21701	1.54952	1.43647	0.27822
1820.00000	1754.36139	1.55103	1.43585	0.27943
1830.00000	1758.35155	1.55408	1.43666	0.28072
1840.00000	1761.45731	1.55562	1.43606	0.28188
1850.00000	1764.57772	1.55723	1.43554	0.28301
1860.00000	1767.62160	1.55875	1.43493	0.28410
1870.00000	1771.61848	1.56200	1.43591	0.28529
1880.00000	1774.62604	1.56354	1.43532	0.28632
1890.00000	1778.64565	1.56690	1.43641	0.28746
1900.00000	1781.61873	1.56846	1.43585	0.28843
1910.00000	1785.66619	1.57195	1.43705	0.28951
1920.00000	1787.47029	1.57152	1.43468	0.29028
1930.00000	1790.34881	1.57302	1.43408	0.29115
1940.00000	1794.39422	1.57659	1.43539	0.29214
1950.00000	1797.24130	1.57809	1.43480	0.29295
1960.00000	1801.32474	1.58178	1.43623	0.29388
1970.00000	1804.14283	1.58328	1.43566	0.29462
1980.00000	1808.26998	1.58709	1.43721	0.29549
1990.00000	1809.68599	1.58617	1.43449	0.29599
2000.00000	1813.82238	1.59004	1.43610	0.29680
2010.00000	1816.52532	1.59151	1.43548	0.29740
2020.00000	1819.22466	1.59289	1.43488	0.29796
2030.00000	1821.89280	1.59424	1.43427	0.29848
2040.00000	1826.09737	1.59828	1.43609	0.29917
2050.00000	1828.74327	1.59961	1.43549	0.29962
2060.00000	1831.35916	1.60089	1.43487	0.30004
2070.00000	1835.63765	1.60509	1.43688	0.30064
2080.00000	1838.23409	1.60635	1.43627	0.30098
2090.00000	1840.80038	1.60757	1.43565	0.30130
2100.00000	1843.33685	1.60873	1.43500	0.30157
2110.00000	1847.71476	1.61312	1.43724	0.30204
2120.00000	1848.31635	1.61092	1.43363	0.30201
2130.00000	1852.72576	1.61536	1.43595	0.30242
2140.00000	1855.18620	1.61641	1.43528	0.30255
2150.00000	1857.61631	1.61740	1.43457	0.30265
2160.00000	1862.12868	1.62202	1.43710	0.30298
2170.00000	1864.54651	1.62299	1.43641	0.30301
2180.00000	1866.93427	1.62391	1.43570	0.30301

TEMP-R	EQ SONIC VEL-FT/SEC	CP/CV	GAMM-O	CP-BTU/LBM-R
2190.00000	1869.29124	1.62478	1.43496	0.30298
2200.00000	1871.61767	1.62560	1.43419	0.30291
2210.00000	1876.28848	1.63048	1.43703	0.30311
2220.00000	1878.60528	1.63128	1.43628	0.30298
2230.00000	1880.89067	1.63203	1.43550	0.30283
2240.00000	1883.14519	1.63272	1.43469	0.30264
2250.00000	1887.97393	1.63789	1.43782	0.30274
2260.00000	1890.22054	1.63857	1.43703	0.30250
2270.00000	1892.43537	1.63921	1.43621	0.30224
2280.00000	1894.61764	1.63980	1.43535	0.30195
2290.00000	1896.76641	1.64035	1.43447	0.30163
2300.00000	1898.88262	1.64085	1.43355	0.30128
2310.00000	1900.96410	1.64130	1.43260	0.30092
2320.00000	1906.04807	1.64695	1.43619	0.30089
2330.00000	1908.12173	1.64742	1.43526	0.30049
2340.00000	1910.16059	1.64785	1.43429	0.30006
2350.00000	1912.16305	1.64825	1.43328	0.29962
2360.00000	1917.44379	1.65432	1.43721	0.29955
2370.00000	1919.43824	1.65475	1.43622	0.29908
2380.00000	1921.39496	1.65516	1.43520	0.29859
2390.00000	1923.31356	1.65553	1.43414	0.29809
2400.00000	1925.19212	1.65588	1.43304	0.29758
2410.00000	1927.03035	1.65621	1.43190	0.29705
2420.00000	1932.58203	1.66297	1.43630	0.29696
2430.00000	1934.40937	1.66338	1.43518	0.29642
2440.00000	1936.19228	1.66377	1.43401	0.29588
2450.00000	1937.93072	1.66416	1.43281	0.29534
2460.00000	1939.62324	1.66453	1.43156	0.29479
2470.00000	1941.26985	1.66491	1.43027	0.29424
2480.00000	1947.11061	1.67256	1.43518	0.29417
2490.00000	1948.73844	1.67307	1.43390	0.29363
2500.00000	1950.31551	1.67359	1.43258	0.29310

Fig. F.1. Comparison of frozen and equilibrium sonic velocities for saturated potassium vapor.

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Appendix G

MOLLIER CHARTS FOR CESIUM AND POTASSIUM

Fig. C.1. Mollier diagram for cesium.

Fig. G.2. Mollier dia-
gram for potassium.

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