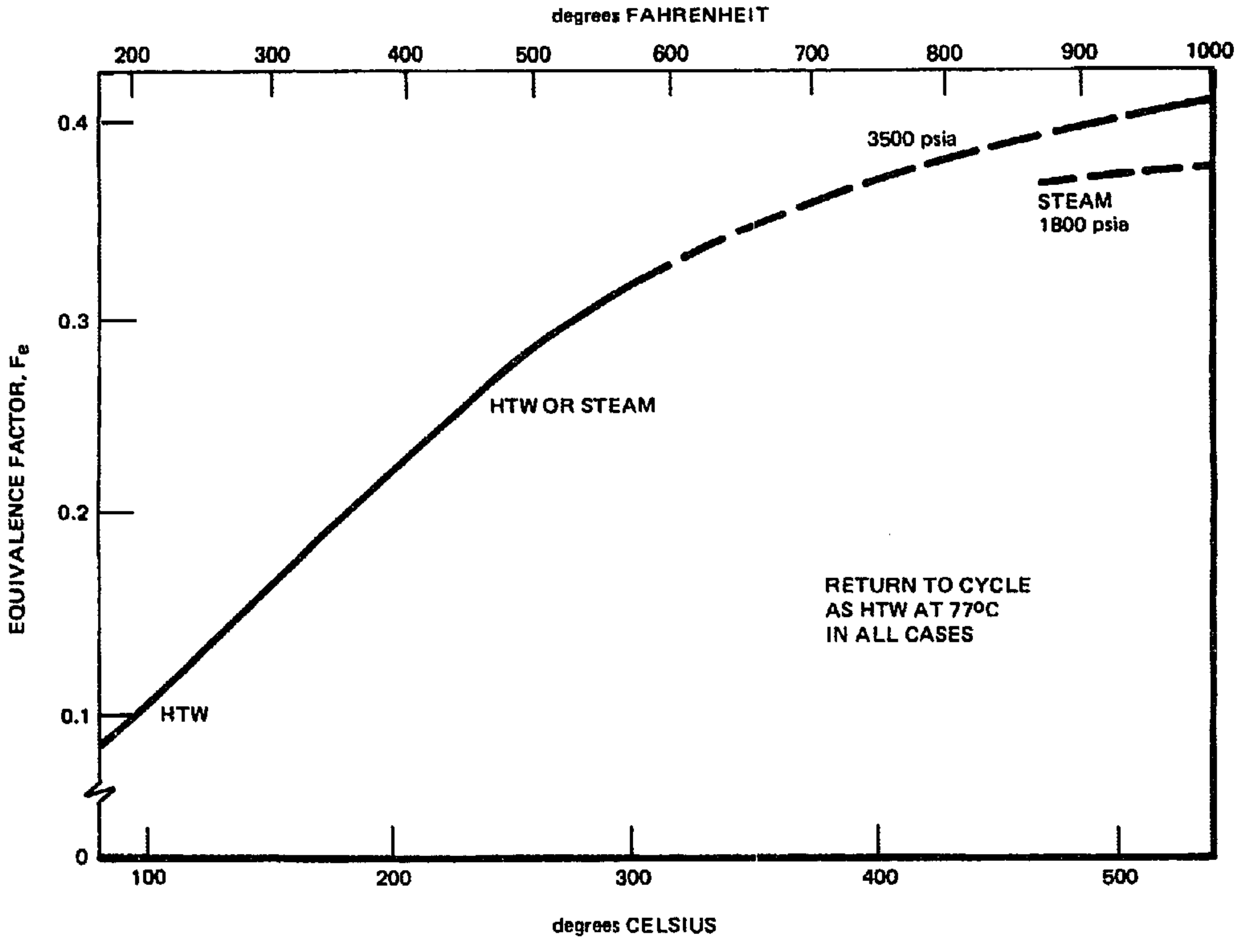


Figure A-1. EQUIVALENCE FACTOR (MW_e/MW_t)FOR COAL-FIRED POWER PLANT

Again, using data from the previous studies, updated to 1/79$ and a 1995 date of initial operation for a plant, \$51-53/MWh$_e$ is the busbar cost of base load electricity (CF = 0.80) with LWR and coal plants respectively. About 25 percent is added for transmission capital and operating (losses) costs to be more comparable to dispersed generation, giving \$65/MWh$_e$.

The allocation of costs to electricity and to heat is somewhat arbitrary. Recovery of all costs by the revenues received from sale of both products is the important criterion. Within reason the amount of thermal energy extracted can be increased, with a drop in electric output proportional to the equivalence factor.

At 0.80 capacity factor the 512 MW$_e$ plant, producing only electricity must recover revenues of \$298 M. At base load rates it will receive only \$233 M. The shortfall is \$65 M.

If 300 MW$_t$ are extracted, the electric output is reduced to 454 MW$_e$. Selling the heat at \$30/MWh$_t$ and electricity at \$65/MWh$_e$ gives an annual revenue of \$270 M. The shortfall is reduced to \$28 M. If 600 MW$_t$ are extracted, electric output is 396 MW$_e$ and the total revenues would be \$307 M, i.e. no shortfall. With more than 600 MW$_t$ extracted, both heat and electricity could be priced cheaper than the separate product alternatives.

The amount of thermal extraction that is permissible with a conventional steam turbine design is a function of the HTW temperature extracted; the equivalence factor is also a function of temperature. Extracting low temperature water, eg 100°C, for a low equivalence factor, will run into a limit, in that a point is reached at which the remaining steam flow to the condenser is too low for proper operation. Higher HTW temperatures reduce the flow less (higher ΔT) so can permit higher extraction rates in MW$_t$. However, the equivalence factor goes up, so the unit cost of heat is greater. Unconventional design for utilities, such as a backpressure turbine, permits the extreme in which no waste heat is generated, 20-30 percent of the output is electricity and 70-80 percent is useful heat. For the postulated CF = 0.80 this would give 314 MW$_e$ and 1024 MW$_t$ for example, with an annual revenue of \$371 M; both electricity and heat could be priced 20 percent lower than competing sources.

The above gives perspective but is oversimplified. The market for HTW at one temperature was not examined, nor were the problems and costs, advantages and disadvantages, of supplying different temperatures and/or steam. There is some conversion cost if HTW is extracted and is then converted to steam by industrial customers. On the other hand, with a high capacity factor load pattern, permitting a backpressure turbine configuration, there can be considerable capital cost savings, in that the generator and turbine need be sized only for 314 MW$_e$ instead of 512 MW$_e$, and no cooling towers (or ponds) are required. Further, the changes in system configuration and in costs for low capacity factor loads for both heat and electricity were not examined, particularly when both load types are varying independently.

Some of these cases will be treated briefly to see if they improve or deteriorate the cost benefit shown. First, a tool, the joint-products cost diagram, will be described to simplify the comparisons.

A.5.2.1 Joint Product Cost Diagrams

The revenue required to recover annual costs must be distributed over a quantity of electricity, Q_e, produced during the year and a quantity of heat, Q_h, produced over the year. The unit cost of electricity and heat, C_e and C_t, are related by the linear equation:

$$Q_eC_e + Q_tC_t = \text{Annual Fixed Costs} + \text{Annual Variable Costs} \qquad \text{(A-1)}$$

For any value of C_t there is a value of C_e that satisfies the equation and vice versa.

When there are competing alternatives for heat and electricity production, the relation of their unit costs to this linear tradeoff are easily shown by a diagram such as Figure A-2.

The unit costs of heat and electricity, C_t and C_e, are shown as abscissa and ordinate scales. Line 1 shows the linear equation for the last case described above, producing 371 MW_e and 1024 MW_t, at a capacity factor of 0.80.

The competing cost of base load generation plus transmission was found as 65 $\$/MW_e$. This is shown as a horizontal dotted line. The competing costs of heat from oil-fired and coal-fired boilers are shown as vertical dotted lines at \$31 and $\$33/MWh_t$. Any combination along line 1, between the points of intersection with the dotted lines represents a reduced annual cost, i.e., is favorable to the HTR-Multiplex.

Lines 2, 3, and 4 show the other cases discussed, for which the thermal extraction was 0 MW_t, 300 MW_t, and 600 MW_t extraction.

A.5.2.2 Backpressure Turbine Cost

Operating at high capacity factor with a backpressure turbine (i.e. all the "waste heat" is used and there is no portion of the steam flow that is condensed and rejected in a cooling tower) means that such cooling towers, and part of the turbine and electrical equipment are superfluous. They and their cost can be eliminated in the original design. There is a penalty in flexibility in that a backpressure turbine cogeneration system has no way to vary the ratio of electric and thermal outputs. Both can be reduced together by reducing the steam flow input to the turbine.

Where the heat as HTW or steam is required by an industry on a three-shift basis year-round this penalty is not important, and the reduced cost of both heat and electricity will be desired.

The Turbine Island cost in Table A-4 was $\$398/kW_e$. As a minimum, the capital investment could be reduced by the difference in capacity, 198 MW at this unit cost, or \$79 M. Actually, since cooling towers are completely eliminated and the very large diameter low pressure turbine stages eliminated are most costly than average for the turbine, the savings are greater. Estimating a 45 percent reduction in Turbine Island cost, for a 39 percent reduction in output gives the further reduction in cost illustrated by dash line 5 in Figure A-2.

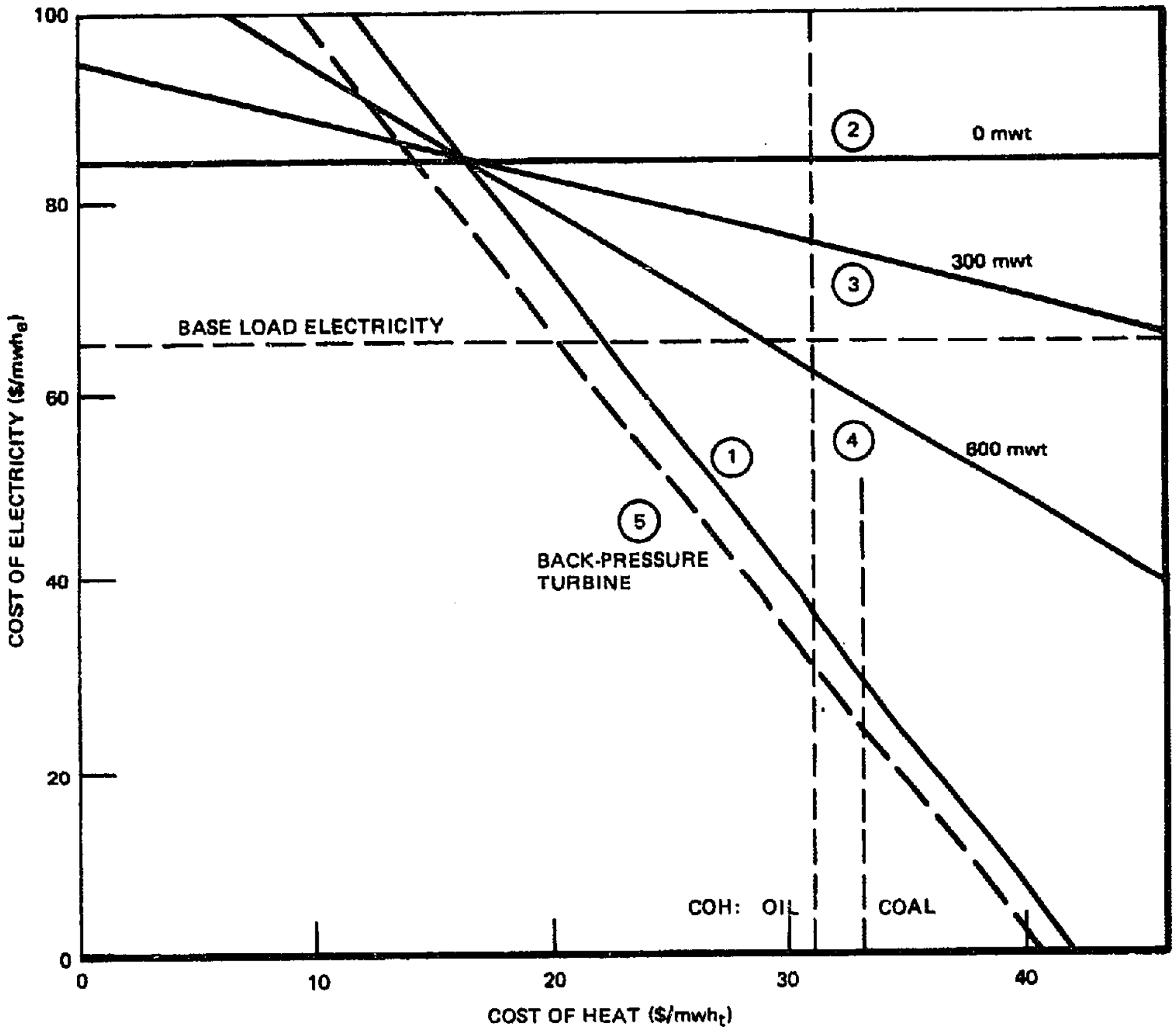


Figure A-2. JOINT PRODUCTS COST DIAGRAM

storage, e.g. for space heating in the lower part of the temperature range, aquifer storage appears promising but is not yet proven.

A.6.1 Cavern Storage

Excavated caverns in hard rock were examined in the above references using the analysis by J. Dooley, et al (1977). He found the direct costs of a 29,000 m^3 cavity to be \$172/m^3, plus an added \$15 M for the shaft and tunneling required. One shaft can serve several caverns of this size, or a larger diameter cavern if the rock quality and technology permit. The shaft size is in part related to the energy charging and discharging rate derived rather than the MWh of energy stored. Costs are somewhat dependent on the depth of shaft which in turn depends on the storage pressure required.

For simplicity in this analysis, three caverns per shaft are assumed and no attempt is made to separate power-dependent and energy-dependent costs. Allocating shaft costs gives a total of \$350/m^3 direct costs in 1/79\$. This is much higher than the figure sometimes quoted for excavation, i.e. \$1-1.50/ft^3 (\$35-52/m^3). The excavation cost item per se is a small part of the total direct or installed cost; Dooley, et al used \$1.30/ft^3 for this item.

This form of storage containment can be used for HTW storage or for gas storage, such as the endout and return gas mixtures of the TCP. For these latter, a steel liner may be unnecessary, and there is less concern with temperature fluctuations and more concern with pressure fluctuations. To assure public safety and a long life for the cavern and shaft, the precautions taken in cavern design cannot be greatly reduced, so the cost for gas storage is estimated as at best 20 percent less than for HTW storage. Three hundred fifty and \$280/m^3 will be used for HTW and gas storage respectively.

HTW at 177 °C, with a return temperature of 81 °C has an enthalpy difference and a density at the upper temperature such that 0.1 MWh$_t$/m^3 is stored (or 10 m^3/MWh). There must be an equivalent storage at the return temperature, as water is pumped from one to the other during charge and discharge. Alternatively one may depend on a thermocline, or stable temperature gradient region with hot water floating on the cold water. Separate return water storage in atmospheric pressure tanks above ground adds only \$29/m^3, so will be assumed. The total or investment cost in 1/79\$ is \$8000/MWh stored.

For comparison, at 40 bars pressure and ambient temperature, the volume required per MWh of energy difference between the hydrogen-rich sendout gas and the methane-rich return gas is 41.2 m^3 for the former and 14.5 m^3 for the latter. Two separate caverns are required, with transfer of gas into one being matched by transfer out of the other. If a pressure swing between 20 bars and 100 bars is assumed, the mass stored varies from half that at 40 bars to 2.5 times, a maximum storage of three times that indicated by the above specific volumes. The 18.57 m^3/MWh have an investment cost in 1/79\$ of \$11,000/MWh stored. This assumes intercooling of the compressed gas to near ambient temperature; if no intercooling is done, the cost nearly doubles.

For ready comparison with pipeline-packing storage we assume storage of 5814 MHW$_p$, 3.23 hours at 1800 MW$_p$. Allocated over the annual output, this adds \$0.94/MWh$_p$ to the cost. This is about half the cost of pipeline packing, and has the further advantage that the amount stored can be increased as much as desired, at proportionately greater cost.

Both thermal storage of HTW and storage of the TCP gases will require pumps/compressors, and a power cost dependent on the cycle of use. The cavern depth should be such that the maximum pressure does not greatly exceed the overburden pressure. For 100 bars this is about 560 m. HTW will be stored at a pressure that flashing to steam does not occur anywhere in the shaft. HTW at 40 bars at the top of the shaft would be at 66 bars at 305 m which is compatible with overburden pressures. The pump for thermal energy storage must be capable of lifting the HTW, i.e., about 26 bars; the compressors in each gas storage cavern must be capable of the maximum 80-bar pressure change.

The compressor power rating depends not only on the pressure rise needed but also on the mass flow pumped. A cycle of use must be assumed. A maximum charging rate is the full rated capacity of the TCP, i.e. the full 1800 MW$_p$ can be charged into storage for some part of the day. The greatest horsepower need is at the end of the charging cycle, when the hydrogen-rich gas must be boosted from 40 bars to 100 bars in the storage cavern, and the methane-rich gas must be boosted from 20 bars in its cavern to the 40-bar line pressure.

Assuming this maximum charging rate, the required compressor capability at the end of the charging cycle (when the full pressure rise from 40-bar line pressure to 100-bar storage is simultaneous with a pressure rise from 20 bars in the methane cavern to 40-bar line pressure) is 92 MW$_e$. This adds \$0.68/MW$_p$ to product cost.

If the compressor complement is considered as being four units capable of 20-bar pressure rise at the full mass flow rate: 1800 MW$_p$ or 775 Mg/hr, and each is equipped with an intercooler, considerable flexibility is gained if compressor technology permits each unit to be used with either gas, and over a range of pressure rises and flow capacities that do not exceed its power rating. Aircraft engine compressor technology, wherein compressor blade angle is adjusted to optimize operation over about an 8:1 variation of inlet pressure with altitude, may be relevant. Compressors may be put in series for maximum pressure rise, and in parallel for greater discharge flow rates to meet high load peaks. During part of each cycle, compressed gas is being expanded by throttling, as very high discharge rates are achievable without expenditure of power. During parts of the cycle when one gas is being compressed and the other expanded, the excavation may be by reverse flow through a compressor, reducing the required line power.

Without such regeneration, the maximum power required is for isothermal compression form 20 to 100 bars of each gas on a daily cycle. For the assumed 3.23 hours of storage (5814 MWh$_p$) approximately 268 MWh per day are required for compression. The annual cost of compressor power of \$7 M/yr adds \$0.50/MWh$_p$ to the cost of TCP product.

The three cost components, $\$0.94/\text{MWh}_p$ for storage volume, $\$0.56/\text{MWh}_p$ for compressor power, and $\$0.68/\text{MWh}_p$ for compressor capital costs, total $\$2.18/\text{MWh}_p$.

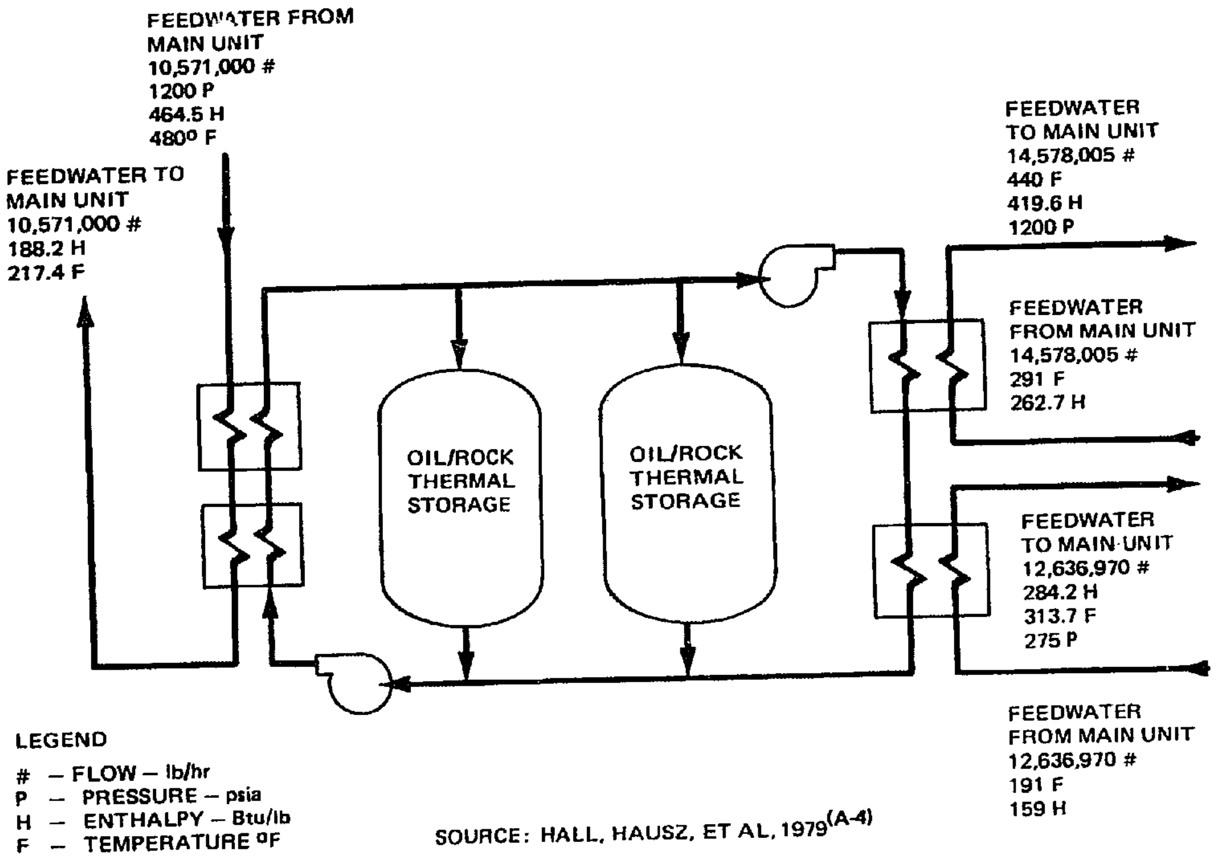
Each of these may be scaled up or down from the exemplar values assumed. If twice as much storage volume is required and is used daily, the first two components double, but compressor costs do not change. If the added volume is not used daily but is a precaution to buffer weekly load variations, the power cost should not be doubled. If storage charging is never done at more than half the 1800 MW_p pipeline capacity and discharging requirements are not excessive, the compressor costs would be halved.

What about the HTW storage? To store the same amount assumed above, 5814 MW_t, for ready comparison, at $\$8000/\text{MWh}$, the $\$46.5$ M cost allocated over the entire HTR-Multiplex output adds $\$0.68/\text{MWh}_p$. Pumps are required to recover the HTW, but none to charge storage. The head or pressure difference to be overcome corresponds to 1000 feet of cavern depth or about 27 bars. The energy required to discharge the cavern is 44 Mwh/day or $\$0.77$ M/yr or an added cost of $\$0.06/\text{MWh}_p$. The required pump capital cost is similarly small, adding $\$0.03/\text{MWh}_p$. Total of the cost components for 5814 MWH_t of storage is $\$0.78/\text{MWh}_p$, about a third the cost of pipeline gas storage in caverns. The amount of storage and the power and pump costs can similarly be scaled upward and downward for cases other than this example.

A.6.2 Dual-Media Storage

Other forms of thermal energy storage are available. Cavern storage requires suitable geologic conditions which are not available at all sites. Aboveground thermal energy storage in hot rocks in tanks, with the pore volume filled with a heat transfer hot oil such as Caloria HT43 were found by Hall, Hausz, et al(A-4) and Hausz, Berkowitz, and Hare(A-2) to be more economically viable than pressurized HTW containment. A heat exchanger from the extracted HTW to the dual storage media is required. While the media and tanks may be rated in MWH_t stored, the costs of the heat exchanger are power-dependent, rated in maximum MW_t in charging or discharging storage.

Typical of dual-media storage systems is the configuration shown in Figure A-3 from Hall, Hausz, et al. In this example, during charging the heat exchangers on the left transfer heat from 10.57 million pounds per hour of HTW to the heat transfer oil. The HTW temperature drops from 480 °F to 217 °F, raising the oil temperature from 204 ° to 440 °F. Hot oil entering the top of storage tanks heats the rocks and comes out, cold, at the bottom of the tanks to be recirculated. A thermocline, horizontal boundary between hot and cold rock, moved downward, reaching the bottom when the tank is fully charged. In discharging storage, the flow is reversed, with hot oil from the top passing through the heat exchanger on the right, heating water from 191 °F to 420 °F. The cold oil is returned to the bottom of the tanks and the thermocline moves upward. In practice, since the heat exchangers are a major cost, the same heat exchangers are used for charging and discharging through the use of necessary pipes and valves.



SOURCE: HALL, HAUSZ, ET AL, 1979(A-4)

Figure A-3. OIL/ROCK THERMAL STORAGE FOR FEEDWATER HEATING

The cost of this thermal storage subsystem, which stores 6848 MWh_t in six cylindrical tanks 120 feet in diameter and 40 feet high, was estimated at 45 M$ direct cost in mid-1976 dollars. Of this the power-dependent components, heat exchangers, pumps, piping, feedwater heaters, and instrumentation, were 85 percent and the energy-dependent tanks, rocks, and oil were only 15 percent. Power-dependent components were rated to discharge storage in six hours, i.e. 1140 MW_t.

Scaling this system to 5814 MWh_t to be comparable to the other storage examples, to the input temperature range 177 °C to 81 °C and to 1/79$ gives a total investment cost of $148 M. This is a capital cost of $25,4000/MWh stored. Allocated to the annual output of the HTR-Multiplex it adds 2.18 MWh_p, very comparable to cavern storage of the TCP gases.

While it is almost three times the cost of cavern storage of HTW, it should be noted that the number of hours of storage can be increased at low cost. For eight times as much storage, with the same heat exchanger rating (i.e. 48 hours of storage), the cost only doubles; for cavern storage of HTW and TCP gases the cost would be $5.60 and $9.00/$MWh_p$ respectively.

A.7 THE MULTIPLEX WITH MISMATCHED LOADS

With the component costs developed in the foregoing parts of this section, it is now possible to compare several alternatives for meeting low capacity factor and badly mismatched demand patterns for electricity and heat.

The daily demand pattern for electricity is highly peaked in the late afternoon in the summer months, over much of the country where air conditioning loads are high. In winter, as use of electric heating increases, there are daily peaks almost as high, often dual peaks in early morning and early evening. For most weekdays throughout the year, the utility system load swings daily from a peak to a minimum of 40 to 60 percent of the peak in off-peak hours. Base load plants take care of the continuous load, leaving mid-range and peaking plants to swing from zero or near zero to their peak output.

While some industrial heat demand is base load, round the clock all year, other plants operate one or two shifts. Their process heat needs may concentrate in a six- to eight-hour period; winter space heating needs similarly are greatest during the working shift. Individual plants may have capacity factors as low as 0.20 for heat, although others at 0.40 and higher, plus those that are continuously operated plus diversity effects will give a higher system capacity factor, perhaps comparable to the 50 to 70 percent found in system electricity demand.

A.7.1 Capacity Factor Effects

The importance of high capacity factor operation of the HTR-multiplex system has been illustrated earlier. Conventional electricity generating plants are similarly affected. A planning tool used by utility engineers as a ready comparison of alternative plant types, called the screening curve, is illustrated in Figure A-4. The total annual costs per megawatt, both fixed and variable, are plotted as a function of capacity factor.

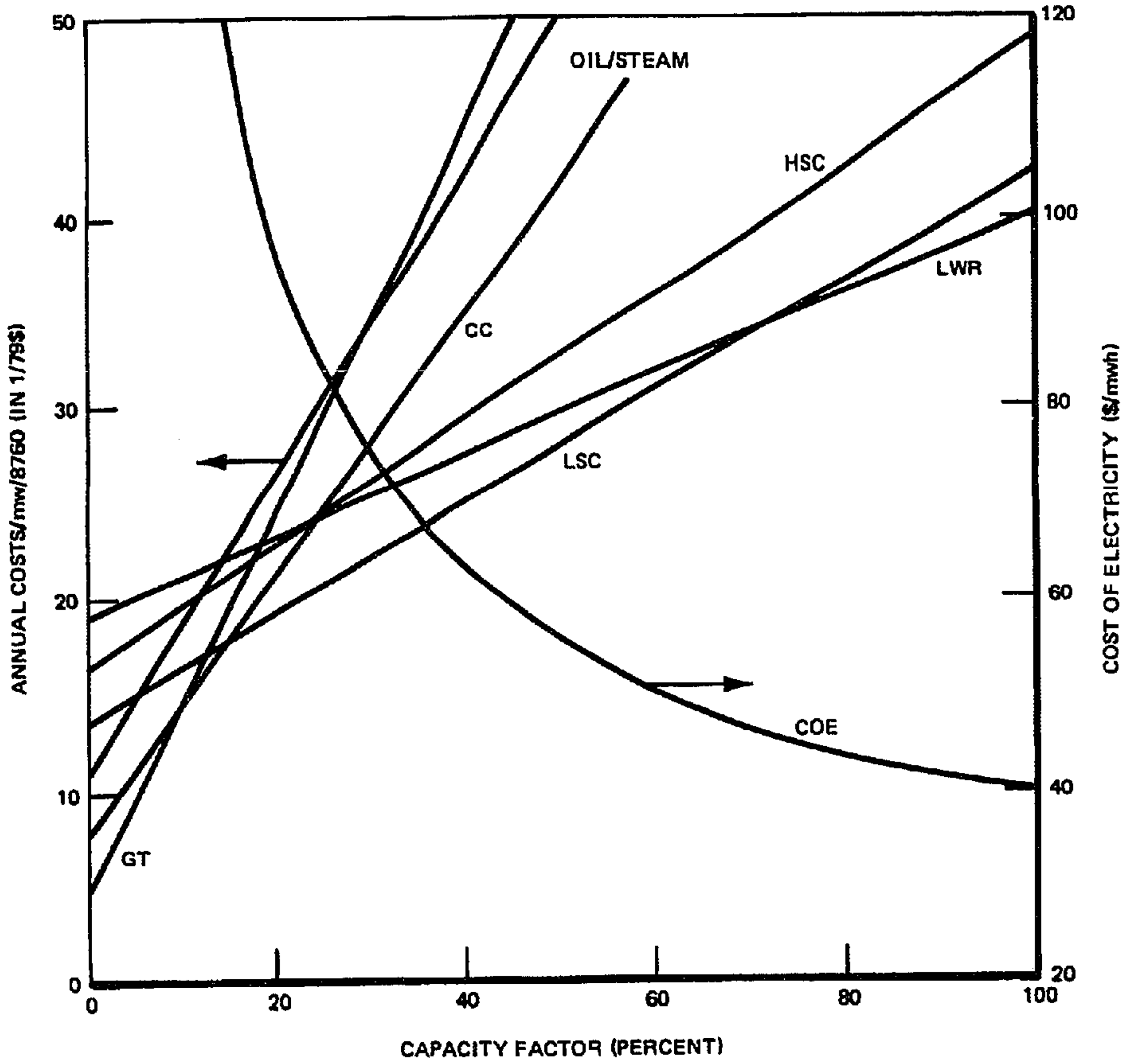


Figure A-4. SCREENING CURVES FOR CONVENTIONAL GENERATION PLANTS

By definition, the fixed costs are independent of capacity factor and the variable costs, fuel and O&M, are proportional to CF. At zero CF there are only the fixed annual costs, related to the capital cost of the plant. At CF = 1.00, the costs of fuel, etc. for 8760 hours per year are included. A straight line connects the two points. Thus a low capital cost plant, such as a gas turbine (GT) or a combined-cycle gas turbine plus steam turbine (CC) burning distillate oil, has a low intercept at CF = 0, and a steep slope. At the other extreme, a high capital cost plant option such as a light water reactor (LWR), with much lower cost fuel, has a high intercept at CF = 0 and a small slope. Coal-fired plants using high sulfur coal (HSC) or low sulfur coal (LSC) are intermediate in annual fixed costs and in fuel related costs (including scrubbing).

As a result, there is for each capacity factor one of the plants that has the lowest annual total costs. Utilities must plan for load demands that are exceeded only once a year for a short time, essentially CF = 0, to base load levels that are exceeded all year and are met by base load plants used at their maximum availability, say CF = 0.80. For minimum total cost of operation, a mix of plants is needed that roughly matches the load versus time duration curve of that utility. The screening curve is an aid in approximating this distribution, although more sophisticated computer simulations are needed to include reliability requirements, diversity, uncertainties in the demand forecast, etc.

It can be seen that at high capacity factors the LWR is most cost effective, while at the lowest capacity factors the gas turbine is most effective, despite the high cost of oil. Both high and low sulfur coal plants (HSC and LSC) are shown. In some areas of the country, low sulfur coal would not be available without very great transportation costs which would greatly increase the slope, making it inferior to HSC for base load. Oil-fired steam plants are still the dominant generation source in the Northeast and the Far West, for historic and environmental reasons. It currently has no region in the range of capacity factors where it is superior.

From this group of curves, the effective cost of electricity from a single plant is found by dividing the annual costs by the capacity factor. Using the lowest cost plant for each part of the range, the curve COE, using the scale on the right, gives the cost of electricity for this mix of plants. As is well known, peaking power is more costly than base load power. In this figure, from \$44/MWh$_e$ at CF = 0.80, it increases to \$95/MWh$_e$ at CF = 0.20 and to \$120/MWh$_e$ at CF = 0.15. In comparing the costs of low capacity factor operation of HTR-TCP systems, these data on conventional systems must be borne in mind.

The reader is warned again that the methodology consistently used herein of leveling fuel and O&M costs over a 30-year period starting in 1995 leads to unit costs of electricity and heat that are about double what he may be used to, thinking in terms of current costs of fuels.

Since the HTR-Multiplex-dispersed generation systems analyzed include 100 km of transmission, to be comparable the transmission costs of the conventional systems must be included. Figure A-5 repeats the COE curve of Figure A-4 as a dashed line and adds an estimated transmission cost for the curve labeled CONV. + TR. The cost of transmission includes both capital costs and the cost chargeable to the resistive

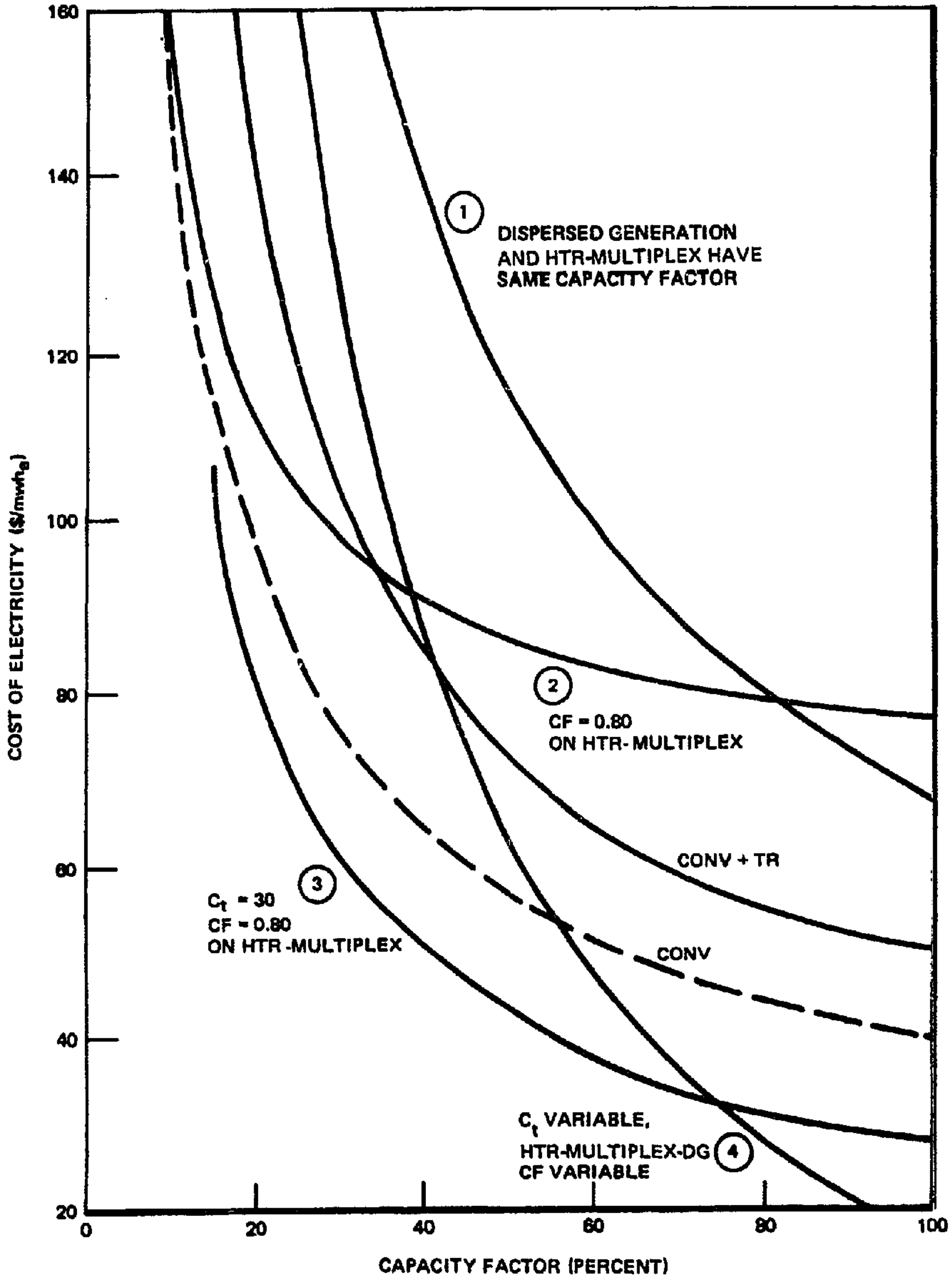


Figure A-5. COST OF ELECTRICITY FOR HRT MULTIPLEX VERSUS CONVENTIONAL GENERATION losses. Both of these components are capacity factor dependent, just as generation costs are. If transmission line capabilities are increased to carry peak loads, the incremental capital costs are distributed over the small increment of annual energy output during peak hours; if the transmission capabilities are not increased, the I^2R losses go up as the square of the energy carried.

Curve number 1 describes the cost of electricity for a dispersed electric generation system driven by an HTR-Multiplex as analyzed in Section A.5.1. This is for the extreme case in which the HTR-Multiplex has the same capacity factor as the dispersed generation. That is, no diversity of loads or storage is assumed. The COE rises dramatically at low capacity factors, and for no value of CF is superior to the conventional reference curve CONV. + TR.

Curve number 2 is the other extreme in which diversity is assumed, the HTR-Multiplex operates a capacity factor of 0.80, and only the dispersed generating plant is varied in capacity factor. Since only a part of the system varies in capacity factor, the curve crosses curve 1 at CF = 0.80 and has a considerably lower COE at capacity factors below this. In fact it is superior to the CONV. + TR. curve for CFs below 0.35.

Storage by pipeline packing would aid in achieving the high capacity factor operation of the HTR-Multiplex, but at an increase in fixed charges. The two points labeled STORAGE, at CF = 0.4 and 1.0, give an estimate of the amount curve 2 would be raised by the 3+ hours of pipeline packing storage.

In contrast to the limited regions of superiority for HTR-multiplex-dispersed generation of electricity alone, curves number 3 and 4 consider the joint production of heat and power. For curve number 3 the assumption of HTR-Multiplex operation at CF = 0.80 was retained and 1024 MW_t + 314 MW_e were generated for some fraction of the time, given by the capacity factor, as analyzed in Section 3.1.3.2. No storage was assumed. As both electricity and heat are generated, a value was assigned to the cost of heat, so the corresponding cost of electricity could be compared to the reference. A value of \$30/MWh_t is below the cost of alternative sources, oil-fired boilers at CF = 0.80 \$33/MWh_t) and coal-fired boilers at CF = 0.80 (\$31/MWh_t) from Section 3.1.2.2.

The resulting curve 3 is superior not only to the reference CONV. + TR. but also the cost of conventional generation alone, the dashed curve. It would be lower still in cost of electricity if the cost of heat assigned varied with capacity factor.

If the assumption of high capacity factor for the HTR-Multiplex is not made, curve number 4 results. As with curve 1, all components, HTR-Multiplex and the dispersed generation plant have the capacity factor indicated by the horizontal scale. Both heat and electricity are made at the same capacity factor. simultaneously. Instead of assigning a single value to heat to find the cost of electricity, the value of heat varies with capacity factor as indicated for coal-fired boilers in the table on page 3-15. The resulting cost of electricity is very low, below 20 \$/MWh_e at CF = 1.00 which is usually unattainable. It is superior to the reference CONV. + TR. for all capacity factors down to 0.40, and superior to the CONV. cost of generation alone down to 0.55.

It is clear that there are many combinations of variables that can be explored: assumptions of different HTR-Multiplex capacity factor, dispersed generation capacity factors for heat and for electricity separately and for TCP and/or thermal energy storage to mitigate low capacity factor penalties and to give flexibility. Only a few of these have been explored.

The information contained in Figure A-5 can also be displayed in the joint-products cost diagram format as shown in Figure A-6, with some additional perspectives. Each capacity factor of the CONV. + TR. conventional electric power reference value in Figure A-5 can be paired with a capacity factor for local coal-fired boiler (HSC) heat supply, to define a point giving the conventional, separate product, cost of electricity, and cost of heat for that product pair. For equal capacity factors from 1.0 to 0.3, such points are shown as bullets in Figure A-6. An HTR-Multiplex-dispersed generation system configuration can be compared with these competitive target values.

The straight lines repeat the data used in curve 4 of Figure A-5, i.e. the HTR-multiplex and the dispersed generation plant are both assumed to operate at the indicated capacity factor. The arrows from the conventional target values to the corresponding joint-product cost line indicate the bounds of the range over which the HTR-Multiplex-DG system has both a lower cost of electricity and heat than conventional separate product generation. At a capacity factor of 0.8 it can be seen that if the COE is set at the target value, heat can be sold for $8/$MWh_e$ less than the target value. If the COH is set at the target value, electricity can be sold for $26/$MWh_e$ below the target value. When capacity factors of 0.4 or less are considered, the superiority of the HTR-Multiplex-DG system has dispersed.

This figure particularly shows that the triangle defined by the target value point; the two arrows, and the segment of the joint-product cost line fully define the results for any case studied. Other target value points can be considered, in which the capacity factor of heat output and electric output are different. This combination is not feasible with a backpressure turbine configuration, but with an extraction turbine system the relative electric and heat output may be varied, within design constraints. The impact of storage, either of the TCP gases or of thermal energy storage, can be explored. Either a sketch, or the three points of the triangle in tabular form permit visualization of the merits or penalties of such changes.

A.7.2 Effects of TCP Gas Storage

The application of cavern storage of TCP gases, as described in Section A.6.1 will be considered, using the basic economic data derived for a dispersed generation plant rated 512 MW_e or 314 MW_e and 1024 MW_t when jointly producing both heat and power. To serve this plant, with a conversion efficiency of 0.375, requires HTR-multiplex input of 1365 MW_p, not the full 1800 MW output of the data presented in Table A-3. It is convenient to normalize by making the output of the TCP match the input needs of the dispersed generation plant so that a single system is analyzed rather than a diversity of loads. Neglecting economies-of-scale effects the HTR-Multiplex can be scaled down, the heat and power output could be

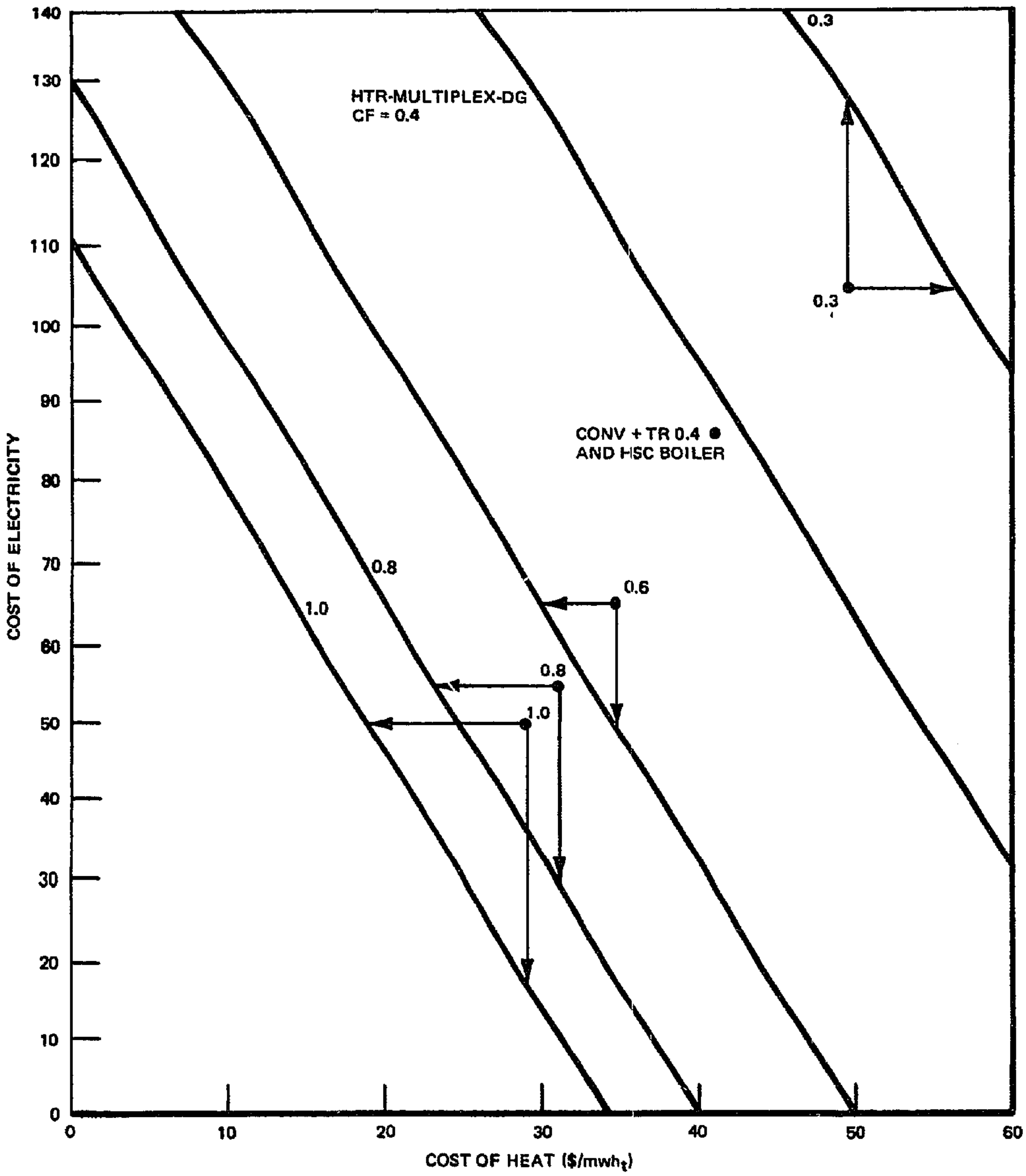


Figure A-6. JOINT-PRODUCTS COST DIAGRAM FORMAT scaled up, or an arbitrary unit level could be chosen. Keeping the plant size as 512 MW_e, the principal annual cost items before storage is added are:

HTR	\$136.8 M/yr
TCP (100 km)	35.4
Fuel (at CF = 1.0)	94.0
Methanator	13.5
Turbine Island	37.9
	223.6 + 94 × CF

Equation A-1 becomes:

$$314 \times 8760 \times CF \times C_e + 1024 \times 8760 \times CF \times C_t = (223.6 + 94 \times CF) \times 10^6$$

or

$$2.75\ C_e + 8.97\ C_t = 223.6/CF + 94 \qquad \text{(A-2)}$$

This equation, for various values of CF gives the lines displayed in Figure A-6. If cavern storage of TCP gases is added sufficient to store all the TCP output for X hours and to discharge it over (24 − X) hours, the methanator and Turbine Island will operate at a reduced capacity factor and higher output level, while the HTR and TCP continue to operate at their maximum availability, which we take to be CF = 0.8. The capital costs of methanator and Turbine Island must be increased in the ratio $(1 - X/24)^{-1}$ for their increased rating, and the storage capital costs must be added.

In Section A.6.1 a unit cost of \$2.18/MWh_p was found for a specific configuration of cavern storage: 5814 MWh stored, for discharge over six hours. The cost contained three components: 0.94 for capital cost of caverns and shafts, 0.56 for compressor power, and 0.68 for capital cost of compressors. For a daily cycle the first two are proportional to the MWh stored and the last one is proportional to the rage of discharge required. The annual costs for 5814 MWh stored was \$27.50 M.

For a case in which all TCP gas product is stored for 18 hours and discharged over 6 hours, 24570 MWh storage are required, at a cost of \$116.22 M. The methanator/Turbine Island will have four times the output, adding \$154.2 M. With the HTR-Multiplex operating at CF = 0.80 the Turbine Island operates at CF = 0.2 the equation becomes:

$$\frac{4}{4} \times (2.75\ C_e + 8.97\ C_t) = 617.5 + 94 \qquad \text{(A-3)}$$

Here the term 4/4 indicates that four times the power is output for one-fourth the time. The resulting joint product diagram is shown as Figure A-7.

The conventional target values for CF = 0.2 are \$140/MWh_e and \$63/MWh_t. For this case, the HTR-Multiplex-DG-Storage gives \$36.40/MWh_t at \$140/MWh_e and \$53.24/MWh_e at \$63/MWh_t. This can also be expressed as a 25 percent cost reduction in both the cost of electricity and heat.

Table A-5
TCP-GAS CAVERN STORAGE BENEFITS ($/MWh)

Capacity Factor	Target C_{tr}	Prices C_{er}	Point 1 C_t	Point 2 C_e	Margin: Percent
0.2	63.0	140	36.4	53.2	25
0.3	49.5	104	33.7	52.5	20
0.4	41.8	85	32.0	51.4	14
0.6	34.8	65	28.4	44.0	11
0.8	31.2	55	24.8	34.0	9

For other higher values of the CF of both heat and electricity, less storage and less added Turbine Island capacity are needed. The results are summarized in Table A-5.

The top line in the table repeats the data of Figure A-7. Points 1 and 2, as shown in that figure, have the coordinate listed plus one of the conventional reference or target values given by the columns C_{tr} and C_{er}. There is a significant margin for the HTR-Multiplex with storage over the target values at all capacity factors. It is most dramatic for capacity factors below 0.4, for which there was a shortfall rather than a benefit in Figure A-7, but is a cost reduction over the target values of up to 25 percent with storage.

The amount of storage is very small for CF = 0.6 and there is no storage needed for CF = 0.8. These, therefore, show little change.

A.7.3 Effects of Thermal Energy Storage

Thermal energy storage can similarly be applied to permit the capacity factor of the heat demand to differ markedly from the demand for electricity. Either cavern storage of HTW or dual-media storage of the heat in oil/rock filled tanks may be used. As previously derived, they differ in their cost components related to the amount of energy stored and to the maximum rate of charging or discharging storage. For a rate corresponding to storage discharge over six hours, cavern storage is lowest in cost; for a rate corresponding to discharge of storage over more than forty hours, dual-media storage is probably cheaper. In some cases a combination of the two may be warranted. Cavern storage requires suitable geologic conditions of course which are not available everywhere, so both will be treated.

For a case in which the HTR-Multiplex and the Turbine Island are all operated at CF = 0.8, but the continuously generated heat is stored for some fraction of the day to meet a daily load pattern of lower capacity factor, say CF = 0.2, the cost equation is:

$$2.75\,C_e + 8.97\,C_t = 373.5 + 39.0 \qquad \text{(A-4)}$$

In this equation the last number represents the annual cost increment for 18432 MWh of HTW thermal energy storage in caverns, rated for discharge over six hours. Figure A-8 portrays the joint product costs relative to the target or reference values of \$63/MWh$_t$ at CF = 0.2 and \$55/MWh$_e$ at CF = 0.8. Note that one could sell heat at 27 percent less than the target price and give all the electricity away free!!

The same case for dual-media storage differs only in the cost of storage increasing in the ratio 2.18/0.78 for six hours discharge. Even this more costly storage permits selling heat at 15 percent less than the target price and giveaway electricity!

As in Table A-5, Table A-6 gives the corresponding data on other capacity factors for the heat demand. Only above some capacity factor between 0.3 and 0.4 does the target price for heat become low enough to make the cost of electricity positive.

A result such as negative or very low cost of electricity can of course be translated into savings on both electricity and heat costs, feasibility of several hundred km of pipeline transmission, etc. The columns labeled "%" give the margin if equal savings on heat and electricity are wanted.

A.7.4 Peaking Electricity Generation

Using thermal energy storage, one of the options is to not extract thermal energy during the peak hours of the day, but to generate the full 512 MW$_e$ that the Turbine Island is capable of, if the cost savings of a backpressure turbine have not been taken. Assume that the Turbine Island generates 314 MW$_e$ (and 1024 MW$_t$) for 18 hours and 512 MW$_e$ for six hours this is the equivalent of adding 198 MW$_e$ of peak power generation at CF = 0.2 to a 314 MW$_e$ baseload plant at CF = 0.8. Naturally it is desired that this capability be usable during the peak electric demand hours, usually the daytime to early evening. Since peak and baseload heat demand are likely to be present during those same hours, some storage is necessary. Two

Table A-6
THERMAL ENERGY STORAGE BENEFITS

Capacity Factor	Target Price		Cavern Storage			Dual Media Storage		
	COH	COE	COH	COE	%	COH	COE	%
			①	②		①	②	
0.2	63.0	55	29.1	<0	42	36.9	<0	34
0.3	49.5	55	28.4	<0	30	32.0	<0	27
0.4	41.8	55	27.5	8.4	23	29.4	14.6	21
0.6	34.8	55	26.1	26.7	15	28.8	35.5	13
0.8	31.2	55	24.8	34.0	15	24.8	34.0	15

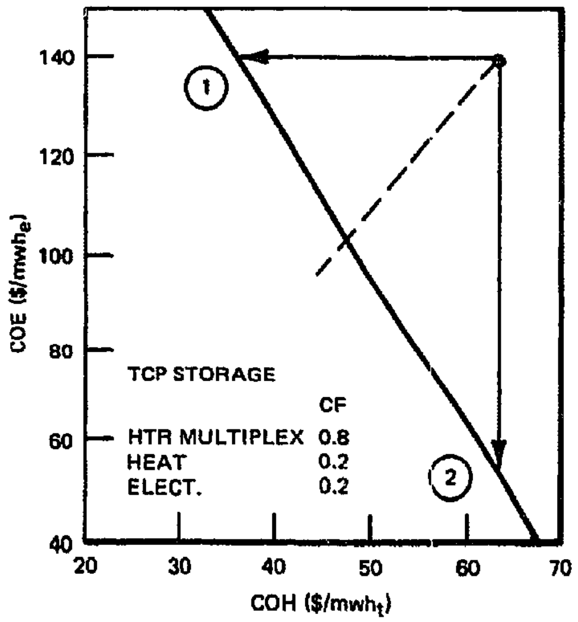


Figure A-7. TCP STORAGE EXAMPLE

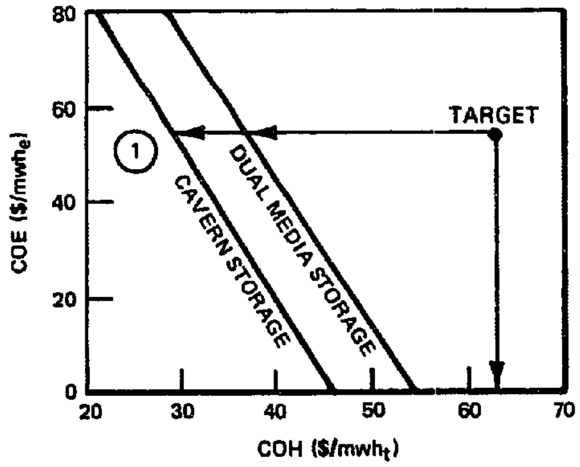


Figure A-8. TES STORAGE EXAMPLE

cases will be examined, one is for baseload heat demand at CF = 0.8; the other is for a heat demand at CF = 0.2 during the same hours as the peak electric demand. More thermal energy storage is required for the latter case of course.

The joint product cost equation becomes modified in several ways for these cases. First, heat is being extracted for only 18 hours a day; second, the electric output can be divided into two parts that have different target values: \$55/MWh$_e$ for base load, and \$140/MWh$_e$ for peaking at CF = 0.2

$$2.75\left(C_{eb} + 0.25 \times \frac{198}{314} C_{ep}\right) + 8.97 \times 0.75\, C_t = 373.5 + C_s/CF \qquad \text{(A-5)}$$

Here, C_{eb} has the target value of 55 and C_{ep} of 140. The amount of peaking electricity produced is less than the baseload electricity by the factors for one-quarter of the day and the relative ratings 198/314. C_s is the annual cost of storage to be included for each case.

For the first case, one-quarter of the heat extracted must be stored during 18 hours to maintain the base load heat output during six hours with no heat extraction. This, as dual-media storage, makes the total on the right 400.75. While, as before, combinations of heat and power costs can be plotted, the third variable, the target value of peaking power makes it more complex. A guide estimate of viability is to insert the target values \$31.8/MWh$_t$ for base load head and 55 and \$140/MWh$_e$ for base load and peaking electricity into the left side of the equation to find its margin over the right. As the left adds up to 421.82 there is a margin of 21.05 or 5.2 percent. This is less margin that that for baseload operation of both heat and electricity, with no peaking power.

For the second case, all of the heat extracted is stored for 18 hours, so the cost increment for storage is quadrupled. The target value for heat is \$63/MWh$_t$. The difference between revenue (left side) and costs (right side) is 153.25 or 24 percent. This is much larger than above, but is still lower than the corresponding case without peaking power.

While peaking power generation does not appear attractive in these examples, the benefits depend in part on the equivalence factor F_e, which in turn depends on the temperature of HTW extracted. If higher temperature HTW is extracted to serve customers requiring such temperatures, larger quantity of peaking electricity obtained per MWh$_t$ foregone may alter the conclusions.

A.8 SUMMARY AND CONCLUSIONS

Using a self-consistent economic methodology, recommended by EPRI, to compare High Temperature Reactors (helium cooled PBRs) in combination with a thermochemical pipeline (TCP) of the reformer/methanator type with alternative fuels and conversion systems at user sites showed some areas of inferiority and some of clear superiority of the HTR-Multiplex concept.

A cost estimate of \$616 M in January 1979 dollars was derived for a 3000 MT$_t$ HTR capable of producing 1800 MW of a hydrogen-rich syngas at 85 percent efficiency and, in addition, a net output of 353 MW of electricity. This leads to an estimate of a cost of syngas at the pipeline input of \$22.16/MWh of product. Upper and lower limits of 19.5 and 26.0 were estimated. To this is added about 17 percent for each 100 km of pipeline and 6 percent for terminal equiment (methanators). The system is capital intensive, fuel cost constitutes only 29 percent of the total, so use at a high capacity factor is necessary to be competitive with many alternatives.

For the inflation and fuel escalation scenarios assumed, the HTR-Multiplex was found to be competitive with industrial heat-only systems, with oil- or coal-fired boilers down to capacity factors of 60 percent; against electrode boilers it is competitive to below 40 percent.

While the diversity of demand pattens of different small users will give the HTR-Multiplex a better capacity factor than the average of that of the consumers, it is not maintained that for supplying heat loads alone the HTR-Multiplex without storage is competitive below 60 percent capacity factor.

Dispersed electric generation using steam derived from methanators was estimated to cost 83 $/MWh$_e$ at 80 percent capacity factor (CF = 0.8). At the same capacity factor, dispersed high sulfur coal-fired plants would cost only $74/MWh$_e$. At lower capacity factors of the whole HTR-Multiplex-dispersed generation system, the disparity is even greater. If through diversity the HTR-Multiplex is at CF = 0.80 but the dispersed generation plant must match a low capacity factor demand, the HTR-Multiplex becomes marginally competitive for one shift load demands because coal-fired boilers have higher capital costs than the methanators.

When combined heat and power generation by the HTR-Multiplex is considered, and compared against the costs of separate production of steam or high temperature water (HTW) and electricity, there is a significant competitive advantage at high capacity factors. In an example, with 314 MW$_e$ and 1024 MW$_t$ at 177° (350°F) produced, there is a 17 percent saving over separate production of heat and utility-delivered electricity. This is increased by 2.5 percent if a backpressure turbine is used for capital cost reduction. At lower capacity factors of all components, i.e. the HTR, TCP, dispersed heat and power generation (Multiplex), and the conventional utility electricity and dispersed heat production, the Multiplex system is competitive down to 40 percent capacity factor, without storage.

Storage was examined, both of the sendout and return gas mixtures of the TCP, and of thermal energy storage (TES), of the heat extracted from the dispersed generation Turbine Island. For storage of the gas mixtures, storage in underground excavated caverns and storage by pipeline packing (varying the pipeline pressure cyclically) were considered. Where geology is suitable the former is least expensive. For thermal energy storage, HTW in similar underground excavated caverns, and storage in tanks filled with rocks and oil (dual-media storage) were considered. Again, the former is least expensive where the site geology is suitable.

With cavern storage of TCP gases, it was found possible to operate the HTR-Multiplex at CF = 0.8, and store enough gas during 18 hours per day to operate the dispersed generation heat and power plant for one 6-hour shift, i.e. CF = 0.2. The cost savings over the separate production of heat and peaking electric power at this same capacity factor was 25 percent. For all capacity factors from 0.2 to 1.0, with suitable amounts of storage, there was a competitive advantage to the Multiplex system.

Thermal energy storage is appropriate when the electricity demand pattern differs radically from the heat demand pattern in capacity factor and/or the time of peak use. Using cavern storage to store heat for 18 hours and discharge over 6 hours, while baseload electricity at CF = 0.8 is being produced, gives a swing of 42 percent over the conventional sources, baseload electricity delivered from the utility and local coal-fired heat generation at CF = 0.2. Use of dual-media storage is only slightly less attractive, with a 34 percent cost advantage.

It should be noted that local cogeneration with a coal-fired plant with or without storage is not among the systems with which the HTR-Multiplex is compared. Some of the conservation benefits and economic benefits would apply almost equally to such systems and reduce the margin of superiority of the HTR-Multiplex system. On the other hand, the environmental problems and hazards of coal plants in urban industrial areas, the penalties of small sizes in coal plants which must meet environmental standards, and the balance of payments aspects of systems using oil or natural gas have not been quantified as benefits favoring the Multiplex.

It appears that the HTR-Multiplex system of electricity and heat generation at dispersed sites in a network within 100-200 km of a clean, safe HTR has a substantial margin of benefit over more conventional sources and should be pursued.

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APPENDIX B

NUCLEAR PROCESS HEAT APPLICATIONS

B.1 PROCESS HEAT DEVELOPMENTS

B.1.1 Introduction

Technological developments in fossil-fuel process heat applications in the 650° to 950°C temperature range are surveyed and cataloged in this section. Advancing technology for fossil-fuel process heat that has direct relevance to nuclear process heat applications is identified. Energy-intensive industries, such as chemicals, metals, and petroleum, are examined. Areas where nuclear process heat might be substituted in these processes are described.

DRAFT SUMMARY

This study examined nuclear process heat application in the 650°C-950°C range. Advancing process technologies and competing heat-generation technologies were studied for their potential effects on nuclear process heat applications. The study focused on large volume chemical processes and synthetic fuels, principally shale oil and hydrogen production by thermochemical means.

We conclude that the most promising processes for nuclear heat application are ammonia, methanol, ethylene, shale oil and hydrogen manufacture. The calcining of lime and pyrolysis of biomass may have potential for using nuclear heat. The summary table below indicates the maximum process temperatures and an estimate of the percentages of fossil fuel displaceable by nuclear heat.

Most existing industrial processes studied are not undergoing significant technological changes. There is a trend toward switching many of these processes from natural gas and petroleum to coal. Generally this would make it more difficult to apply nuclear heat to these processes, and also removes the incentive to switch from scarce oil and gas to the more abundant uranium resources.

There have been studies that considered the use of nuclear heat in steelmaking; if these concepts are developed, a large market for nuclear process heat would exist.

Developments in thermal energy generating devices that use fossil fuels are expected in the area of emissions reduction. Fluidized bed combustion of coal is an emerging technology that may show much promise for generating steam. Fuel cells generate heat as well as electricity; this technology is very well suited to cogeneration and could make substantial inroads in this area in the future. Heat pumps will probably find applications outside the industrial sector; they currently are not envisioned for high-temperature (650°C-950°C) industrial process use.

Hydrogen can be produced from nuclear heat and water. There are several promising thermochemical water-splitting cycles requiring heat only, hybrid cycles requiring both heat and power, and advanced electrolysis cycles using nuclear generated electricity. A major difficulty with heat-only cycles is the inability of nuclear reactors to provide iso-thermal heat, which is desirable for cycle efficiency.

NUCLEAR PROCESS HEAT APPLICATIONS

DRAFT SUMMARY

Application	Maximum Temperature, °C	Percentage of Fossil Fuel Displaced by Nuclear Heat[1]
Chemical Manufacture		
Sulfuric Acid	400	0
Lime	1350	10
Ammonia	850	10
Oxygen and Nitrogen[2]	25	0
Chlorine and Sodium Hydroxide[2]	100	0
Methanol	800	40
Metals		
Ferrous	1100	undeterminated[3]
Non-Ferrous	600	0
Petroleum and Related		
Ethylene	870	40
Coke	1100	0
Heavy Oils	120	0
Synthetic Fuels		
Oil Shale		
Aboveground	900	15
In Situ	800	15
Hydrogen		
From Methane	900	40
From Heavy Oil	870	40
By Thermochemical or Electrolytic Methods	950	100
Tar Sands	90	0
Biomass	980	25
Paper Products	150	0
Stone, Clay and Glass Products		
Cement	1500	0
Glass	1500	0
Food Industry	150	0

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1. Includes fuel and feedstock.
2. Most energy is requires as power, not heat.
3. Value not determined because of wide variety of manufacturing processes.

We believe that the pure thermochemical cycles (i.e., heat requirements only) are less developed than the hybrid or electrolytic cycles, which more closely resemble contemporary technology.

B.1.2 Processes and Developments

Chemicals

The largest volume chemicals were surveyed to determine the temperature and heat requirements of their production processes and the technological developments anticipated for them.

Sulfuric Acid

The first step in the manufacture of sulfuric acid is to produce SO_3. This is accomplished in the Contact process by heating sulfur or sulfide ore in the presence of oxygen and platinum or vanadium catalyst at about 400 °C. Higher temperatures increase the reaction rate, but result in less desirable equilibrium conditions. The SO_3 is carefully dissolved in water to react and form sulfuric acid. The largest manufacturing units produce 1500 tons of acid per day.(B-1)

Sulfuric acid is the leading volume chemical produced in the United States; in 1977, 69 billion pounds were produced.(B-2) No technological developments are anticipated that would significantly change the process operating conditions.

Lime

Lime is most often produced by thermally decomposing limestone in a kiln, where the following reaction occurs:

$$CaCO_3 \rightarrow CaO + CO_2$$

Kilns operate at temperatures from 900 °C to 1350 °C; in the United States they are fired with natural gas or oil. It is imperative to rapidly remove the carbon dioxide that is generated. The thermal requirements vary between 3 and 10 million Btu per ton of lime. Large plants have capacities of 600 tons per day.(B-3)

U.S. production of lime in 1977 was 19 million tons.(B-2)

Ammonia

There are three steps in ammonia manufacture: 1) generation of hydrogen and nitrogen gas, 2) purification of the gas, and 3) formation of ammonia by the following exothermic reaction:

$$N_2 + 3H_2 \rightarrow 2NH_3$$

A flow schematic of a hydrocarbon-based process is shown in Figure B-1. Gas exits the primary reformer at 750° to 850 °C. In the secondary reformer air is introduced as a source of nitrogen; the oxygen present is combusted with the hydrocarbon to provide heat for the endothermic reforming reaction.(B-4)

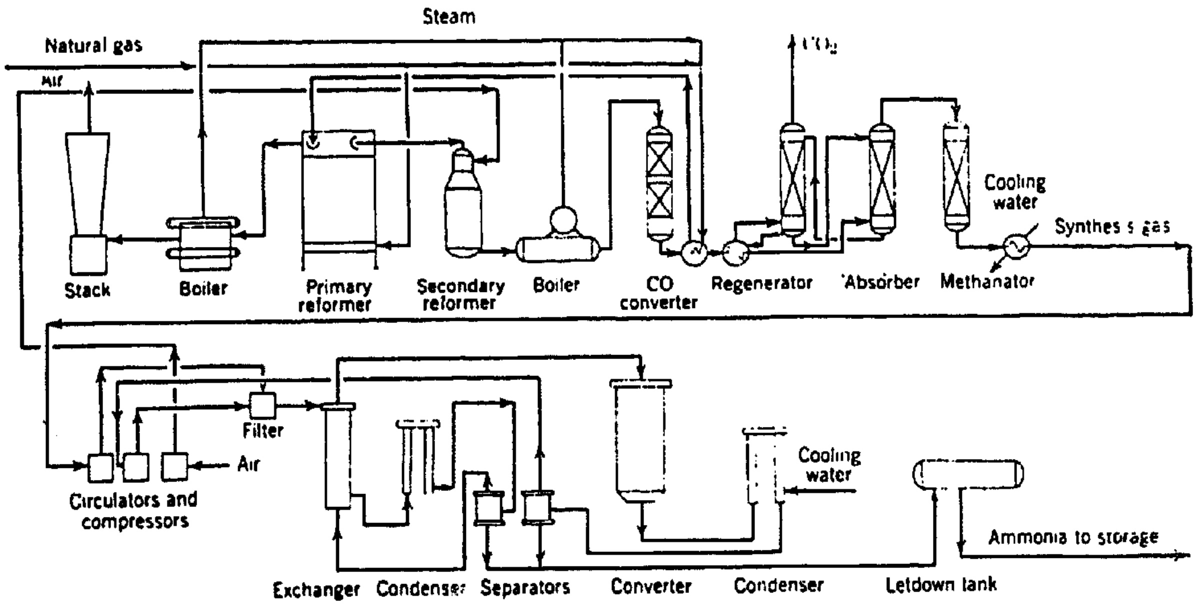

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Figure B-1. FLOW DIAGRAM OF A TYPICAL AMMONIA-SYNTHESIS PROCESS (CHEMICO PROCESS)

If natural gas is the hydrocarbon feed, manufacture of 1 ton of ammonia requires about 35,000 SCF of gas; 56% is used as the chemical feedstock and 44% is used as fuel.(B-5)

The optimal size for an ammonia plant is about 1500 tons of product per day.(B-4) Anhydrous ammonia production in the United States was 11 million tons in 1977.(B-2)

A major development under study for ammonia manufacture is the utilization of coal as the feedstock. Coal gasification is being investigated as a way of generating the synthesis gas. The Tennessee Valley Authority is leading this work.

Oxygen and Nitrogen

These gases are principally produced by separating them from air. Most energy requirements are for mechanical power to effect refrigeration; thermal power requirements are minimal.

Chlorine and Sodium Hydroxide

These chemicals are principally manufactured by electrolytically decomposing saturated brine (25% NaCl). The overall reaction can be simplified as follows:

$$NaCl + H_2O \rightarrow NaOH + 1/2CL_2 + 1/2H_2$$

In a diaphragm-type cell, the operating temperature is 80° to 100°C. A diagram of energy flow for chlorine production is presented in Figure B-2.

Sodium hydroxide is produced by evaporating the remaining cell liquor. An energy flow diagram for sodium hydroxide production is shown in Figure B-3.

Typical chlorine-caustic plants have capacities from 300 to 1500 tons of chlorine per day.(B-6) Eleven million tons were produced in the United States in 1977.(B-2)

Methanol

This alcohol is produced by catalytically reacting a synthesis gas of carbon monoxide and hydrogen as follows:

$$CO + 2H_2 \rightarrow CH_3OH$$

Carbon dioxide also participates in the reaction. Synthesis gas is often produced by steam-reforming light hydrocarbons, where the exit gas temperature typically reaches 800°C. When natural gas is used as a feedstock, the resulting synthesis gas is carbon deficient; this is corrected by adding carbon dioxide to the process. Care must be exercised in the reforming step to ensure the production of the proper mixture of carbon monoxide and hydrogen.(B-7)

Large methanol plants produce 5000 tons per day. U.S. production of this chemical in 1977 was 3 million tons.(B-2)

Metals

The primary metals industry uses about 20% of all industrial energy. Within this classification, energy usage breaks down as follows: ferrous metals about 70%, nonferrous metals about 30%.

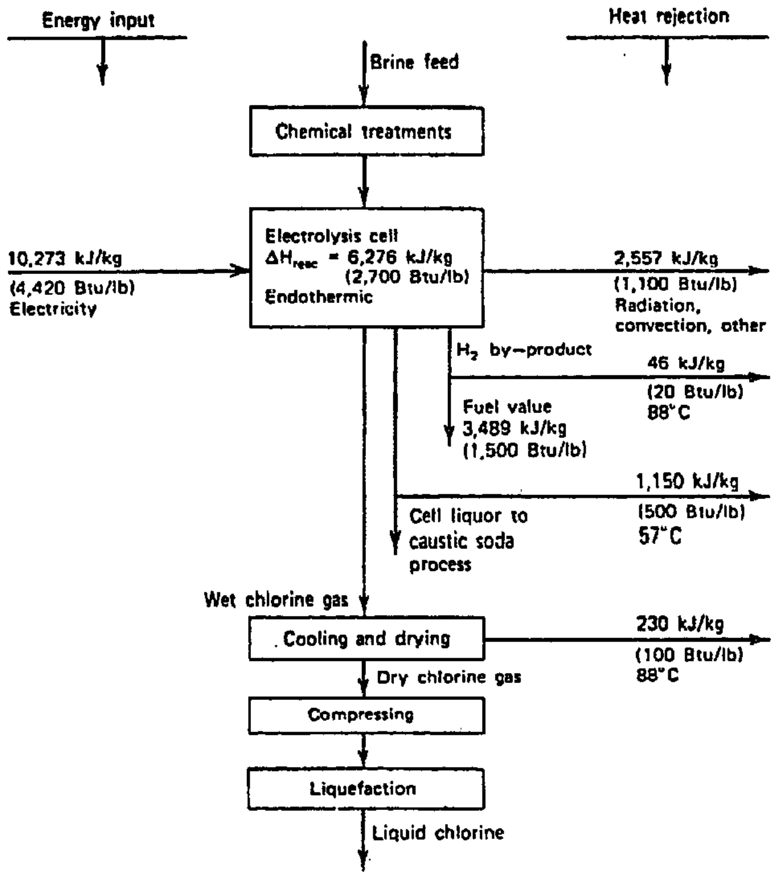

Figure B-2. ENERGY DIAGRAM OF CHLORINE PRODUCTION[B-6] (INEFFICIENCY IN ELECTRIC POWER IS EXCLUDED; ACTUAL HEAT VALUE IS ABOUT THREE TIMES 4,420 BTU LB)

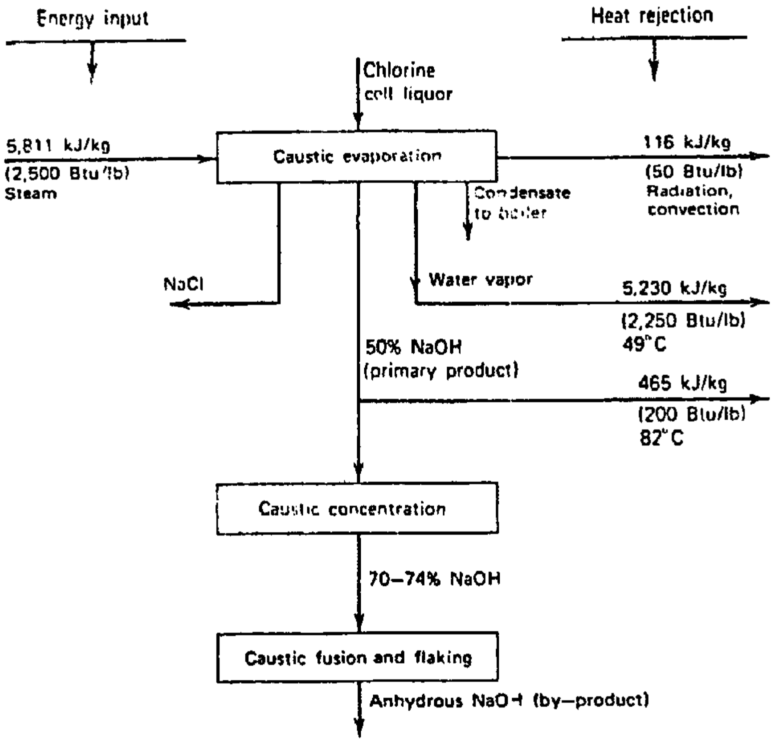

Figure B-3. ENERGY DIAGRAM OF CAUSTIC SODA PRODUCTION(B-6) (WEIGHT OF PRODUCTS AS 100% CAUSTIC SODA. ENERGY VALUES ARE IN TERMS OF ENERGY PER UNIT WEIGHT OF CAUSTIC SODA AS 100% CAUSTIC SODA.)

Primary Ferrous Metals

Operations under this category include —

- Ironmaking
- Steelmaking
- Working and heat treating of steels
- Steel mill coating and finishing
- Ferrous foundries.

Of the energy used in blast furnaces, about 30% is used in ironmaking, 12% in steelmaking, 33% in working and heat treating of steels, and 25% for all other purposes.

In ironmaking, coke (made from metallurgical coal) is used both as a heat source and as a reducing agent for the iron ore. Hydrocarbons (oil or natural gas) can be used as an additional source of reducing material. In the upgrading or beneficiation of the iron ore, pellets are formed and hardened by heating to a temperature over 1100°C. The hot blast is generally produced by heating the blast stove to a higher temperature with an auxiliary fuel. Thus, the major energy-consuming steps in blast furnace operation are (per ton of pig iron) —

- Energy equivalent of coke: 15 million Btu
- Pelletizing: 0.6 million Btu
- Hydrocarbon injection: 0.5 to 2.5 million Btu
- Hot blast: 2.25 million Btu
- Auxiliary fuel: 3 million Btu.

Trends in iron production that will affect these consumption figures are —

- Use of oxygen with blast air to increase hydrocarbon injection
- Improved beneficiation of iron ore
- Use of reformed natural gas for injection
- Recovery of iron fines from the stack gases to alleviate pollution. These fines are sintered (by heating) and recycled.(B-8)

As an alternative to the blast furnace, another established process that uses process heat in the 650° to 950°C range is direct reduction. This is performed in a shaft furnace with refining of the resultant sponge iron in an electric furnace. In the direct reduction process, iron ore is chemically reduced to sponge iron. Since this process does not remove the impurities or gangue constituents (mainly silica and alumina), low-gangue ores are favored for direct-reduction processes. Almost all direct reduction processes use a gas mixture of carbon monoxide and hydrogen at temperatures in the range of 800° to 980°C as a reductant.

Iron ore can be reduced to sponge iron with hydrogen alone, but sintering and reoxidation problems must be overcome before commercialization can occur. Assuming these problems can be worked out, a hydrogen-reduction system is another alternative to the blast furnace. Most of the process heat for this system would be converted to electricity to produce hydrogen by the electrolysis of water.(B-9)

Steel from any of these processes is generally cast into ingots prior to being worked into a finished shape. These ingots are brought to working temperatures in large, fuel-fired soaking pits. The ingots are then rolled into semifinished shapes, reheated in a fuel-fired or induction furnace, and then formed into finished products. A number of heating operations require temperatures in the 650° to 950°C range as shown in Table B-1.

Primary Nonferrous Metals[a]

In the primary nonferrous metals group, the principal energy users are aluminum, zinc, copper, and lead. Aluminum consumes over 70% of the total energy used by the nonferrous metals industry. Bauxite mining does not include any significant upgrading or beneficiation of the aluminum ore, except for rotary kiln drying at 590°C. Natural gas consumption of these kilns is about 1000 SCF per ton of ore. Although less than 15% of the bauxite being consumed in the United States is from domestic sources, bauxite shipment is preferred to refined alumina shipment due to import duties on the latter.

The total annual consumption in the United States of aluminum is currently around 10 million tons and is projected to triple by the year 2000. The Bayer Process, virtually the sole process for producing refined alumina, demands from 10 to 15 million Btu of thermal energy per ton of alumina.

Gas has generally been the preferred fuel. The thermal energy is applied at 1100°C to calcinate alumina trihydrate crystal in a rotary kiln. Approximately 2274 kWhr of thermal energy are required for this step. From this figure and the above demand projection, a total high-temperature thermal energy requirement for the U.S. aluminum industry can be projected as —

$$\frac{2274\ \text{kWhr/ton} \times 30 \times 10^6\ \text{tons/yr}}{365\ \text{days/yr} \times 24\ \text{hr/day}} = 8 \times 10^6\ \text{kWth / or } 8000\ \text{MWth}$$

Table B-1
HEAT AND TEMPERATURE REQUIREMENTS OF SELECTED IRON AND STEEL MILL HEATING OPERATIONS[a]

Operation and Furnace Type	Air Preheat, °C	Heat Requirements, 10^6 Btu/ton
Ingot Heating, Soaking Pit, 870°C Hot Steel; 7%-10% Cold Steel	480 - 650	1.0
Slab Heating, Pusher Type, 5-Zone, 3-Shift Operation	340	21.5
Billet and Bar Reheat, Rotary, Scale-Free	760 - 820	5.4

Since this is a total national figure, the projected market for process heat in this area (all nonferrous metals, since aluminum represents 70% of the energy) is apparently rather small.

Petroleum and Coal Products

This group consumes about 12% of total industrial energy.(B-10) Most of the energy is required as heat at temperatures from 400° to 500°C(B-11) Higher temperatures are required for certain processes, such as reforming (described in the section on oil shale) and the production of ethylene, an important chemical feedstock.

Ethylene

Nearly any hydrocarbon can be used to manufacture ethylene; in the United States, ethane and liquefiable petroleum gases (LPG) are the major feedstocks. Heating of the feedstock cracks it into ethylene and many other products. The process is typically conducted at 700° to 870°C and atmospheric pressure. Fired tubular heaters are employed for the reaction. The yield of ethylene is enhanced by high temperatures and short residence times.(B-12)

Economical plant size is from 200,000 to 400,000 tons per year.(B-12) U.S. production in 1977 was 12 million tons.(B-2)

Coke

Coke is produced by heating low sulphur, low ash, caking coal in the absence of air.

Coking occurs at 1100°C. Most of the heat energy comes from the off-gases generated in the process.

U.S. coke production in 1977 was 54 million tons.(B-2)

Thermal Processing of Oil Wells

Heavy oil cannot be recovered from wells by conventional techniques. Heated water or steam is injected to fluidize the oil and provide reservoir drive. Light gas oil has been used in Venezuela to dilute heavy crude to facilitate production. This recovered crude must still be heated to 120°C for easy pumping.(B-13)

Synthetic Fuels

Oil Shale

Aboveground Processing

Conventional aboveground processes to produce synthetic crude oil from oil shale involve a variety of retorting approaches. These processes include the Paraho Process, TOSCO II Process, Union Oil Process, and others. These processes typically operate at atmospheric pressure with maximum temperatures up to 900°C. Some of these processes have been tested in large-scale equipment and are reportedly ready for commercial-scale use. These processes would probably only be applicable to the Western U.S. Eocene shales.

The IGT HYTORT Process* can be used to produce either SNG or synthetic crude oil from either Western Eocene or Eastern Devonian shales. This process uses gradual, controlled heating in hydrogen at elevated pressures to achieve higher shale organic carbon conversions than conventional processes. This process also makes it possible to recover economically the organic matter (kerogen) in Eastern Devonian shales, which have in the past been largely ignored. The Devonian shale resource is estimated to exceed 400 billion barrels of shale oil.(B-14)

To produce an acceptable synthetic crude oil, all raw shale oils require upgrading by hydrotreating and/or hydrocracking. The HYTORT Process of Figure B-4 also requires makeup hydrogen (produced by catalytic steam-reforming of the effluent gaseous hydrocarbons when oil is the desired product) for the hydroretorting part of the process. However, it produces a better raw oil so that the required hydrotreating is less severe than for oils from other processes.

In-Situ Processing

As with aboveground processing, there are a number of processes which have reached advanced stages of development and which involve retorting oil shale *in-situ*. Examples of these processes are the Occidental Petroleum Corporation modified *in-situ* process, Dow Chemical true *in-situ* process, the Laramie Energy Technology Center true *in-situ* process, and others.

IGT has also made process-type calculations to show how the HYTORT process could be employed to hydroretort oil shales *in-situ*. Figure B-5 shows one such process configuration.

Hydrogen Production

As mentioned above, hydrogen is required for hydrotreating all raw shale oils. There are two major processes for hydrogen production which would be applicable here.

For heavy oil feeds (such as raw shale oil, or when SNG is the desired product and the HYTORT process is used) the Shell or Texaco processes would be applicable. The process reportedly can also be used to produce hydrogen from coal. Figure B-6 is a simplified diagram of the Texaco partial oxidation process. To produce high purity hydrogen, the raw product gas would have to be shifted and scrubbed.

The process is noncatalytic and can operate at elevated pressures in the range of 2000 psig or higher. Oil, steam and oxygen are charged to a generator containing a specially designed burner. Reactor temperatures are typically in the range of 1100° to 1500°F. The oxygen required by the process ranges from about 250 to 275 ft^3 per 1000 ft^3 of hydrogen produced. Reactants are usually preheated to temperatures in the range from 150° to 870°C. The raw product gas contains primarily H_2 + CO plus unreacted steam and soot, which can be recycled. Soot recycle reportedly increases the gas yield without increasing the specific oxygen consumption.

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*IGT offers HYTORT research and development, engineering, and technical services relating to the HYTORT process.

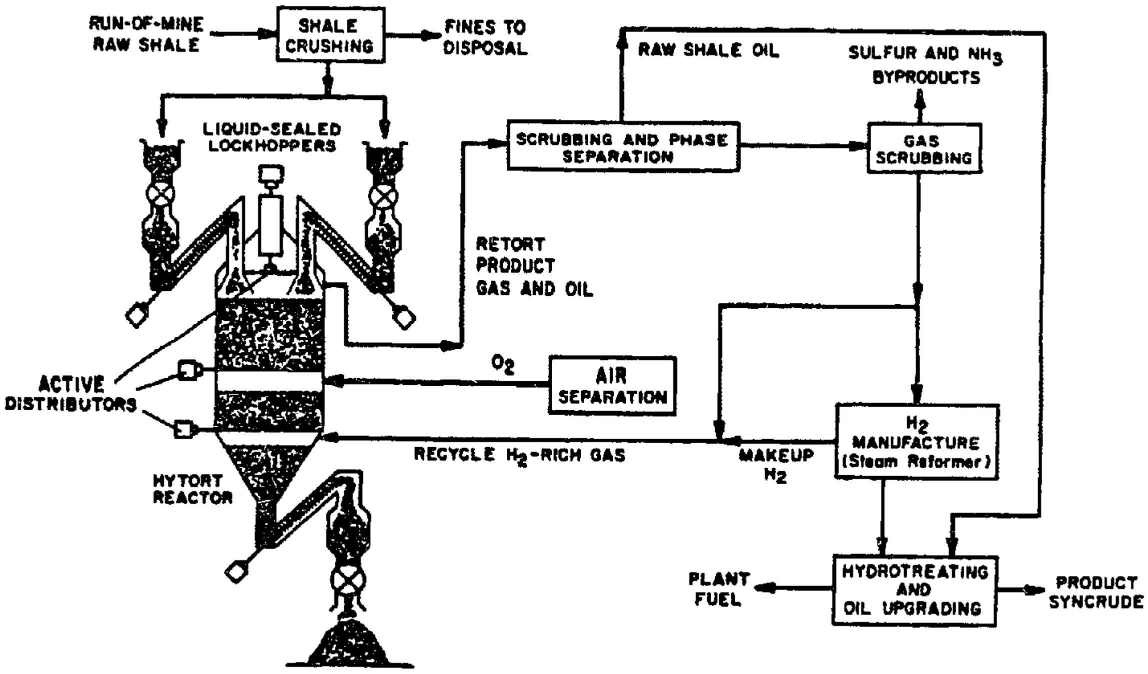

Figure B-4. HYTORT COMMERCIAL PLANT CONCEPT

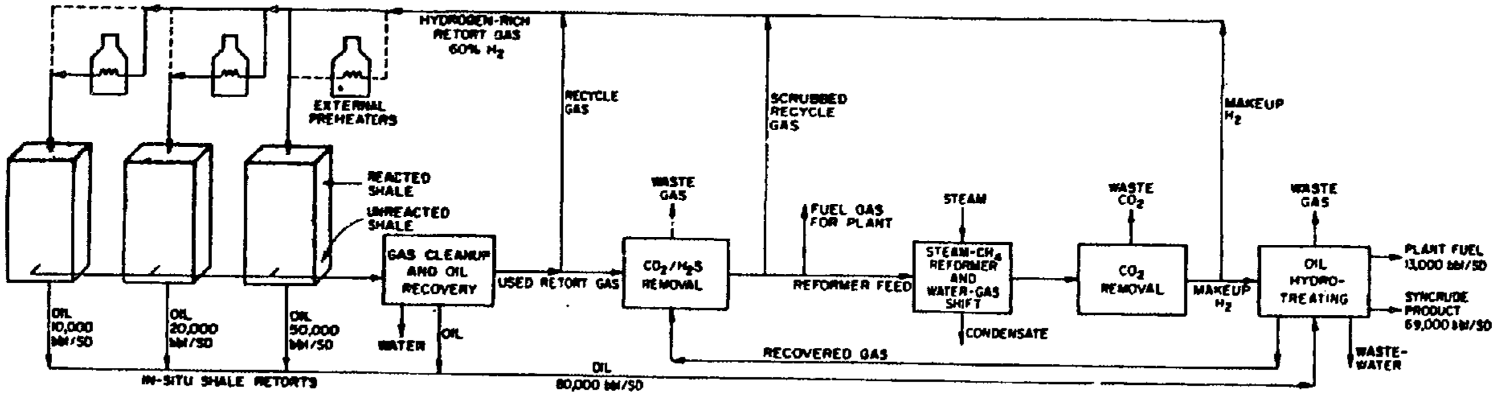

Figure B-5. IGT *in situ* OIL SHALE HYDRORETORTING PROCESS[(B-15)]

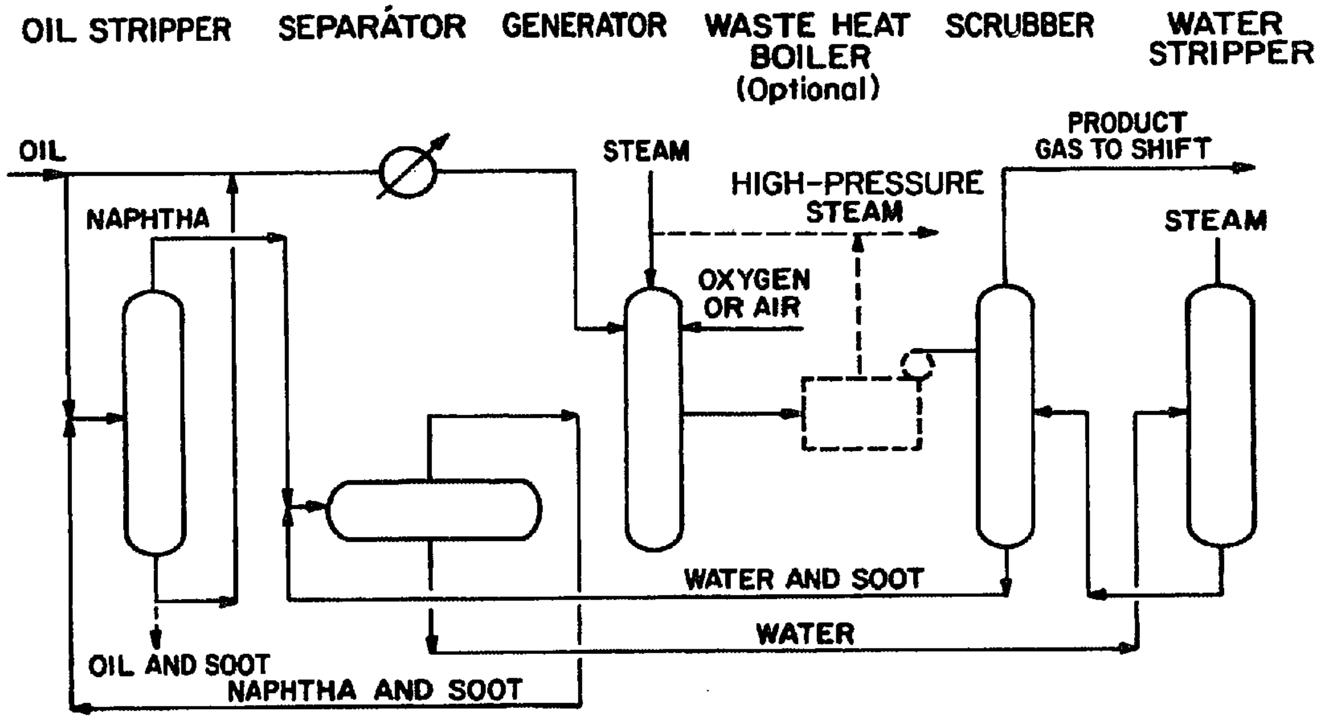

Figure B-6. TEXACO PARTIAL OXIDATION PROCESS

The raw product gas goes first through a waste heat boiler and then to a water scrubber. The soot is extracted from the quench water/carbon slurry with naphtha and recycled to the gas generator oil feed system. The raw synthesis gas is then shifted and scrubbed to remove acid gases; this results in a high purity hydrogen product. Acid gases can be sent to a Claus sulfur recovery unit to produce elemental sulfur as a by-product.

For light oils and gaseous hydrocarbons, catalytic steam reforming is the preferred route. The primary step in the production of hydrogen from methane is the endothermic reaction:

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

Natural gas is passed through a sulfur guard drum that contains activated carbon or zinc oxide to reduce any catalyst-poisoning sulfur compounds to a level of less than 5 ppm. The sulfur-free natural gas is combined with superheated steam and feed to the reforming furnace. The steam-to-methane ratios are usually 3:1 or 4:1 by volume. The steam/natural gas mixture is then passed through a vertical array of type 25-20 chrome-nickel steel tubes that contain a nickel catalyst.

Reforming has been performed at pressures of up to 600 psi, but 300 psi is typical. If a high-pressure hydrogen product is desired, it is preferable to reform the natural gas at a high pressure. However, high pressures require high temperatures to ensure a favorable methane conversion. A more efficient recovery of heat from the product gas is attained at the higher temperature. Gas typically exits from the reforming furnace at temperatures of from 820° to 900°C. The equilibrium constant for the steam-methane reforming reaction increases by a factor of 3 as the temperature is increased from 800°C to 850°C, so temperature is an important process operating parameter. Heat is transferred to the reformer furnace tubes from the radiant section of the furnace, and large amounts of heat are available for steam generation in the convection section.

The reformed gas is composed of hydrogen, water, carbon monoxide, carbon dioxide, and residual methane. It is cooled to about 370°C and is passed through a shift reactor in which excess steam reacts with carbon monoxide to form carbon dioxide and more hydrogen via the exothermic reaction —

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

An iron oxide-chromium oxide catalyst is used. The equilibrium constant for this reaction increases with a decrease in temperature, and at 370°C, a satisfactory conversion of carbon monoxide to hydrogen cannot be achieved. A second shift reactor is used that operates at lower temperatures.

The shifted gas stream enters a carbon-dioxide-removal system. In choosing a carbon-dioxide-removal system for the process, the desired purity of the product and the inlet pressure of the absorber feed-gas must be considered. Many absorption processes are commercially available (e.g., monoethanolamine, Selexol, hot-potassium-carbonate, etc.).(B-5)

Typical data for catalytic steam reforming of natural gas are presented in Table B-2.

Table B-2
EFFICIENCY, PROCESS, AND UTILITY REQUIREMENTS OF STEAM REFORMING NATURAL GAS FOR THE PRODUCTION OF 97% — PURE HYDROGEN
(Per 1000 SCF of Hydrogen)(B-5)

	Amount
Process Feed, lb of methane	11.9
Fuel, lb of methane	7.9
Electric Power, kWhr	0.4
Cooling Water, gal	400
Boiler Feedwater, gal	10
Condensate Returned, gal	6

$$\begin{array}{lcr} \text{Input } 19.8 \text{ lb of methane} \times 23{,}880 \text{ Btu/lb} & = & 472{,}800 \text{ Btu} \\ 0.4 \text{ kWhr Electricity} \times 3414 \text{ Btu/kWhr} & = & \underline{1{,}400 \text{ Btu}} \\ \text{Total} & & 474{,}200 \text{ Btu} \end{array}$$

$$\text{Output} = 1000 \text{ SCF of Hydrogen} \times 325 \text{ Btu/SCF} = 325{,}000 \text{ Btu}$$

$$\text{Efficiency} = \frac{\text{Output}}{\text{Input}} = \frac{325{,}000}{474{,}200} \times 100 = 68.5\%$$

Tar Sands

Tar sand is bitumen-impregnated sand that is more widespread than is generally known. Deposits occur in France, Germany, Romania, Nigeria, Canada, and the United States. Canada leads in commercial development; over 400,000 barrels per day of synthetic crude oil capacity are expected to be onstream in the early 1980's. Ultimate Canadian production is projected to total 25 billion barrels.(B-16)

The bitumen is recovered by extracting with warm water (80° to 90°C) at a pH of 8.0 to 8.5. The bitumen is of a poor quality (low hydrogen to carbon ratio and high sulfur content) and must be upgraded to produce a marketable synthetic crude oil. This is done by coking or hydrocracking. The H-Oil® process can be utilized in the hydrocracking step. Presumably, hydrogen can be manufactured for this process by reforming some of the oil product.(B-16)

Biomass

Pyrolysis "cracks" the chemical bonds of biomass using large amounts of heat at a lower temperature and/or pressure than gasification. A variety of products is formed, depending upon the reactor operating conditions and the chemical composition of the feedstock. The main challenge to R&D is to design a pyrolysis system which minmizes the undesirable products and maximizes the formation of usable gases and oils. Many pyrolysis systems solve this problem by using small amounts of oxygen or air as inputs. The oxygen or air is brought into the system's bed to generate the required heat by burning a small portion of the fixed carbon. The heat then migrates through the balance of the volatiles in an atmosphere devoid of further oxygen. In the pure pyrolytic systems, no air or oxygen is used directly, but the heat must still be produced, either by burning a portion of the output fuel or by adding electrical heat. Fuel oil is also burned.

Pure pyrolysis produces many products, such as chars, oils, and gases, of which some can create disposal problems. Pyrolysis with air or oxygen allows more control over the products. Low air-to-feed ratios favor the production of oils and chars. Oils are useful as a scrubbing medium, and as fuel oil. More air or oxygen favors gas production; thus, the air-to-feed ratio depends upon the needs of the user. Use of oxygen instead of air produces a higher Btu product, at the expense of greater complexity in the process deisgn and hence greater cost.

Commercial-scale pyrolysis of municipal refuse (municipal solid wastes and sewage) has been successfully achieved.* Data on system reliability and downtime are being compiled. One commercial system (50 tons per day) has been tested on wood, coal, manure, kelp, and MSW on a continuous-flow basis. See Figure B-7. The presence of glass and metals in municipal refuse creates special handling and separation requirements at the front end. Ferrous metals can be separated out, while nonferrous metals and glass can be fed into the reactors. Temperatures are kept below 870°C to avoid melting glass or aluminum. Slagging can be utilized, and/or tars and oils can be recycled into the reactor chamber. If wood is the feedstock, somewhat lower reactor temperatures can be used (590° to 760°C), in order to produce briquette material which has a ready market. Less gas and more tars and oils are produced at these temperatures, however.

Because pyrolysis proceeds readily at temperatures between 650° and 980°C, and it uses heat exchangers with no mass transfer, it seems well-matched to high-temperature nuclear process heat. Further research should concentrate on optimum feedstocks for such a system. One concern with municipal solid wastes, for example, would be whether such high ash materials could by pyrolyzed at those temperatures (i.e., slagging occurs at 1600°C). Another concern would be whether to use a number of parallel modular units or one or two large reactors.

Paper and Allied Products

In the manufacture of paper, wood chips are pulped by cooking under pressure in a digester, which breaks up the cellulose. The pulp is further refined into numerous paper products. Waste liquor and wood by-products from the process are used as fuel, as well as natural gas, oil, and coal.(B-17)

Maximum process temperature for the manufacture of writing paper is about 180°C, for newsprint it is about 150°C.(B-11)

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*For example, the "Purox" System, Tonawanda, New York. Also, Pyro-Sol, Inc., Redwood City, California.

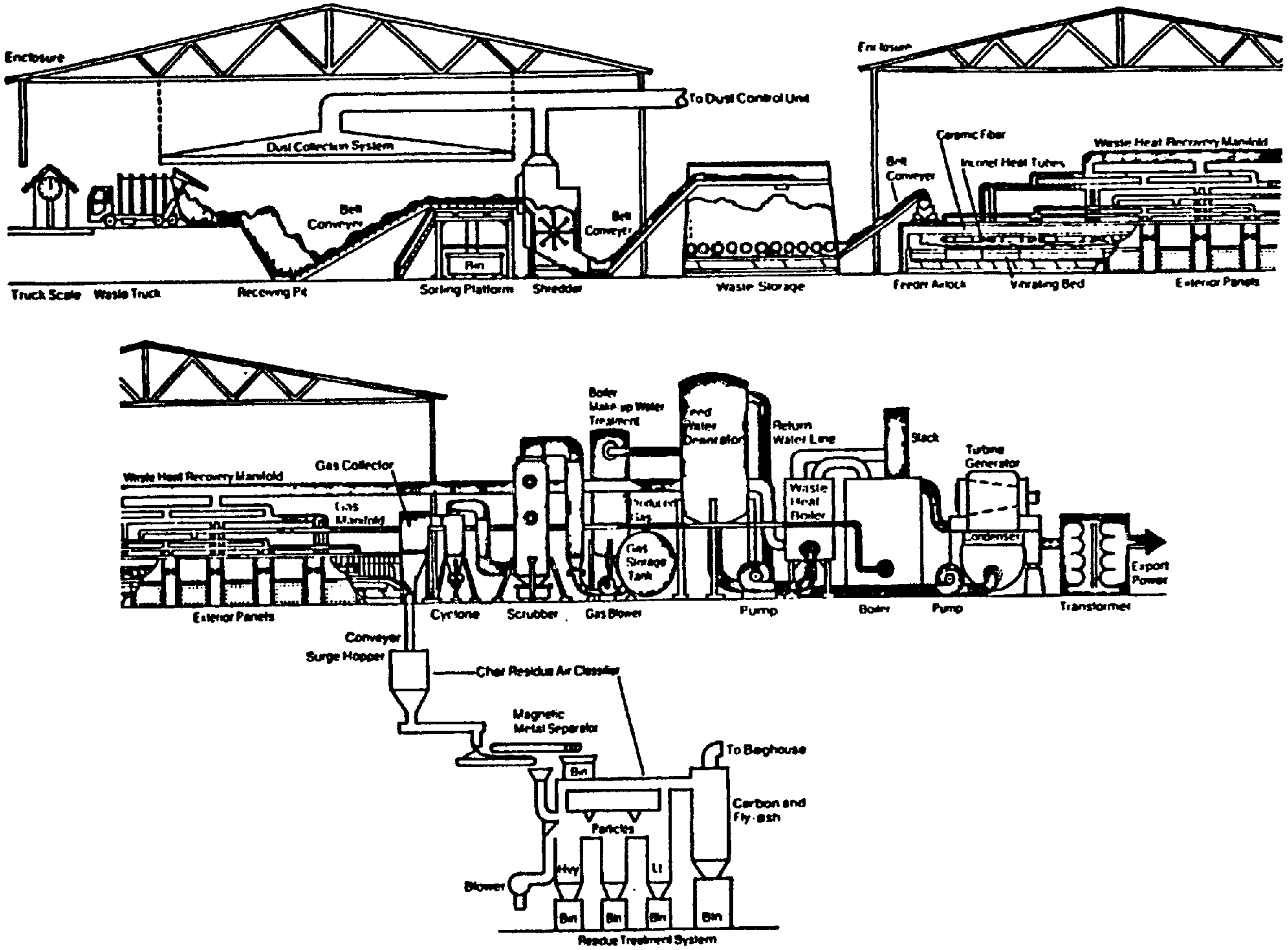

Figure B-7. BIOMASS PYROLYSIS SCHEMATIC (SOURCE - PYRO-SOL, INC.)

Stone, Clay, and Glass Products

Cement

This product is manufactured by combining limestone, clay, and other minerals and heating them to about 1500 °C. The resulting klinkers are ground into cement by either a wet or a dry process.(B-8)

The average energy requirement for the heating step in Portland cement production is 3100 Btu per pound; the most efficient plants require 1300 Btu per pound. Natural gas and coal are the principal fuels; there is a trend toward using pulverized coal.(B-18)

Typical plants in the United States have capacities from 1000 to 1500 tons per day; some Japanese plants have a capacity of 10,000 tons per day.(B-18)

U.S. production of Portland cement in 1975 was 67 million tons.(B-8)

Glass

Glass is made by mixing sand and other constituents and heating them to a melted condition (1500 °C). The process is shown in Figure B-8. About 800 Btu per pound of glass is required in manufacture.(B-19)

Clean fuels are required because fuel gases contact the product; natural gas and oil are used to melt the mixture. Electric melters are also used.

Food and Kindred Industries

Food processing plants are numerous and dispersed throughout the country. Most process heat requirements are for steam at temperatures up to 150 °C.(B-11)

B.1.3 Thermal Energy Generating Devices

Boilers, Furnaces, and Heaters

Most industrial heat (based on fossil fuels) is generated by these devices. Possible future developments are discussed in this section.

Gas

Gas boilers are clean and efficient devices for generating steam. They have required design modifications to control NO_x emissions because of their inherently high flame temperatures. Successful techniques include low excess air, staged firing, and flue gas recirculation. While these modifications add to the cost of gas boilers, they do not make them technologically unfeasible. Industrial gas boilers should have no problems in developing techniques to meet future clean air requirements.

Oil

Oil-fired industrial boilers generally burn the heavier residual (No. 6 or bunker "C") oil. Sulfur dioxide emission limitations are requiring these boilers to utilize low-sulfur resid, a commodity in short supply. Alternatively, the more expensive lighter distillate oils can be burned without significant boiler modification. These boilers may also require design modifications to meet NO_x standards.

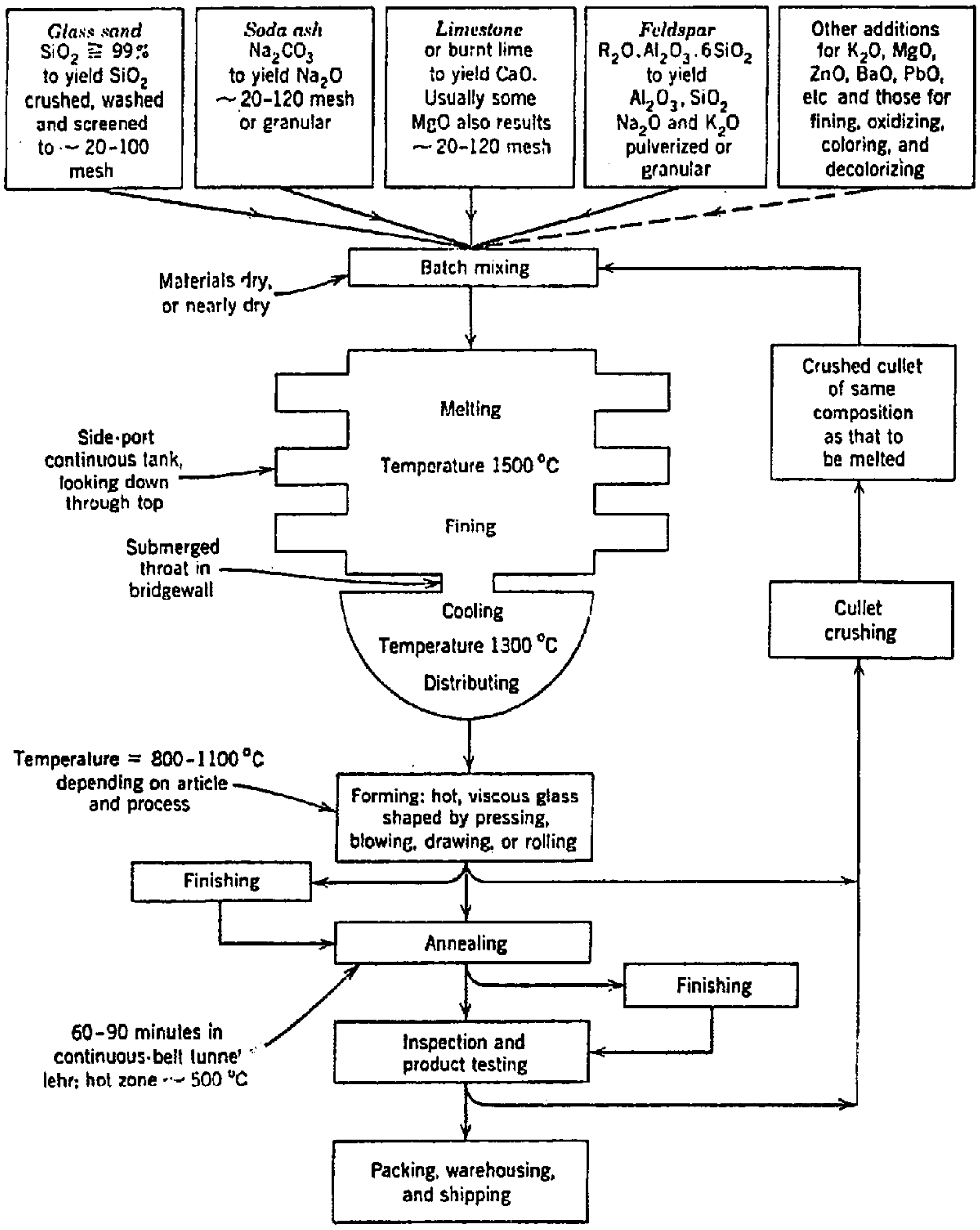

Figure B-8. GLASS PRODUCTION (TEMPERATURES ARE FOR COMMON GLASS; TEMPERATURE FOR SPECIAL GLASS IS HIGHER)(B-19)

Coal

Significant developments are proceeding to make coal boilers cleaner and more efficient. An impc tant new development is fluidized-bed combustion (FBC), which allows for more controllable combusti and attendant efficiency gains and emissions reduction. Sulfur dioxide is absorbed *in situ* by limesto which is admixed with the coal. FBC boilers have recently been made available in sizes from 2500 to 50,0 pounds of steam per hour at pressures up to 300 psig.(B-20)

Large utility boilers using high-sulfur coal meet SO_2 emission standards by flue gas desulfurizatic The technique based on lime/limestone absorption is rather well developed; new techniques are unc development that would reduce the sulfur to a saleable by-product (free sulfur or sulfuric acid).

A controversy exists regarding the role of SO_2 on morbidity effects and acid rain. Environmen considerations could significantly impede the future burning of coal or cause very expensive and comp coal boiler modifications to be developed.

Fuel Cells

Fuel cells convert chemical energy into electrical energy directly. They avoid heat-to-work (Carn efficiency limitations, hence they can be very efficient devices. Heat is generated as a by-product.

The phosphoric acid fuel cell is being demonstrated for utility use in New York City. It operates temperatures from 70° to 175°C.(B-21)

Great interest is being shown in molten carbonate fuel cells because of their potential for lower el tric generating costs. These cells operate from 590° to 700°C. Large quantities of heat are evolv cogeneration of steam or other use of heat is required.(B-22)

Heat Pumps

Heat pumps have been commercialized for the residential sector. Industrial applications are c rently under development. In one concept, a low temperature heat source (100°C) is used in an open-cy vapor-compression steam heat-pump system to generate 320°C process heat.(B-23) Westinghouse is currer marketing an industrial heat pump that generates heat at 100°C from a 40°C waste heat stream.

B.1.4 Nuclear Process Heat Substitution

Industrial processes that appear capable of utilizing nuclear heat are examined in this section conceptual ways of matching to a HTR.

Ammonia

In the United States, ammonia is most often manufactured by reforming light hydrocarbons, pı cipally methane. It is common practice in the production of ammonia to obtain the required nitrogen combusting air with methane to deplete the process stream of oxygen. The heat generated by this react cannot be replaced by nuclear generated heat due to this profess constraint. If nuclear process heat is s plied to the reforming step, at least 10% less natural gas is required to produce a ton of ammonia (

saving is at least 23%). Large ammonia plants of 1500 tons per day capacity could utilize an outside thermal source of about 60 MW.

Ammonia could also be made by directly combining nitrogen and hydrogen. For a 1500 ton per day plant using this conceptual process, about 100 million SCF of hydrogen are required.(B-5) Using the higher-heating value of hydrogen, this is equivalent to about 400 MW of thermal energy. Assuming hydrogen can be produced thermochemically at an efficiency of 50%, an 800-MWth heat source is required.

Methanol

This chemical is also manufactured by reforming natural gas. Assuming 25% of the natural gas consumed is required for the reforming step and can be replaced by an outside heat source, about 800 MWth of heat is required for a large, 5000 ton per day plant.

Lime

In the production of this chemical, a high-temperature gas (900° to 1350°C) must pass over limestone. If a helium stream from an HTR intermediate heat exchanger is employed, a carbon dioxide removal process is required, as carbon dioxide impedes the calcination process. Alternatively, another heat exchanger could be interposed to heat air that could pass over the limestone. Either option may not provide adequate temperatures and full utilization of the nuclear heat. Conceivably, a 40-MWth heat source could be utilized by a large 600 ton per day lime plant.

Oil Shale

A flow diagram showing the IGT HYTORT Process *(in situ)* coupled with an HTGR helium loop is presented in Figure B-9. The major heat requirements and mass flow for a 69,000 bbl per day syncrude plant are as follows:

Process Heat Requirements	**Heat Flow Rate (10^9 Btu/hr)**	**Mass Flow Rate (10^6 lb/hr)**
External Preheaters	2.00	3.0
Reformer Tubes	1.10	0.1
Process Steam (555,000 lb/hr saturated at 150 psi)	0.66	0.6
CO_2 Stripper Steam (477,000 lb/hr saturated at 20 psi)	0.56	0.5
Electric Power (30,000 kWe or about 90,000 kWth)	0.31	—
TOTAL	4.63 (1400 MWth)	

Here, high-temperature heat would be "skimmed off the top" of the primary HTGR helium loop to drive the steam-methane reformer and to replace the fossil-fuel-fired external preheaters.(B-15) The oil hydrotreating step is not included.

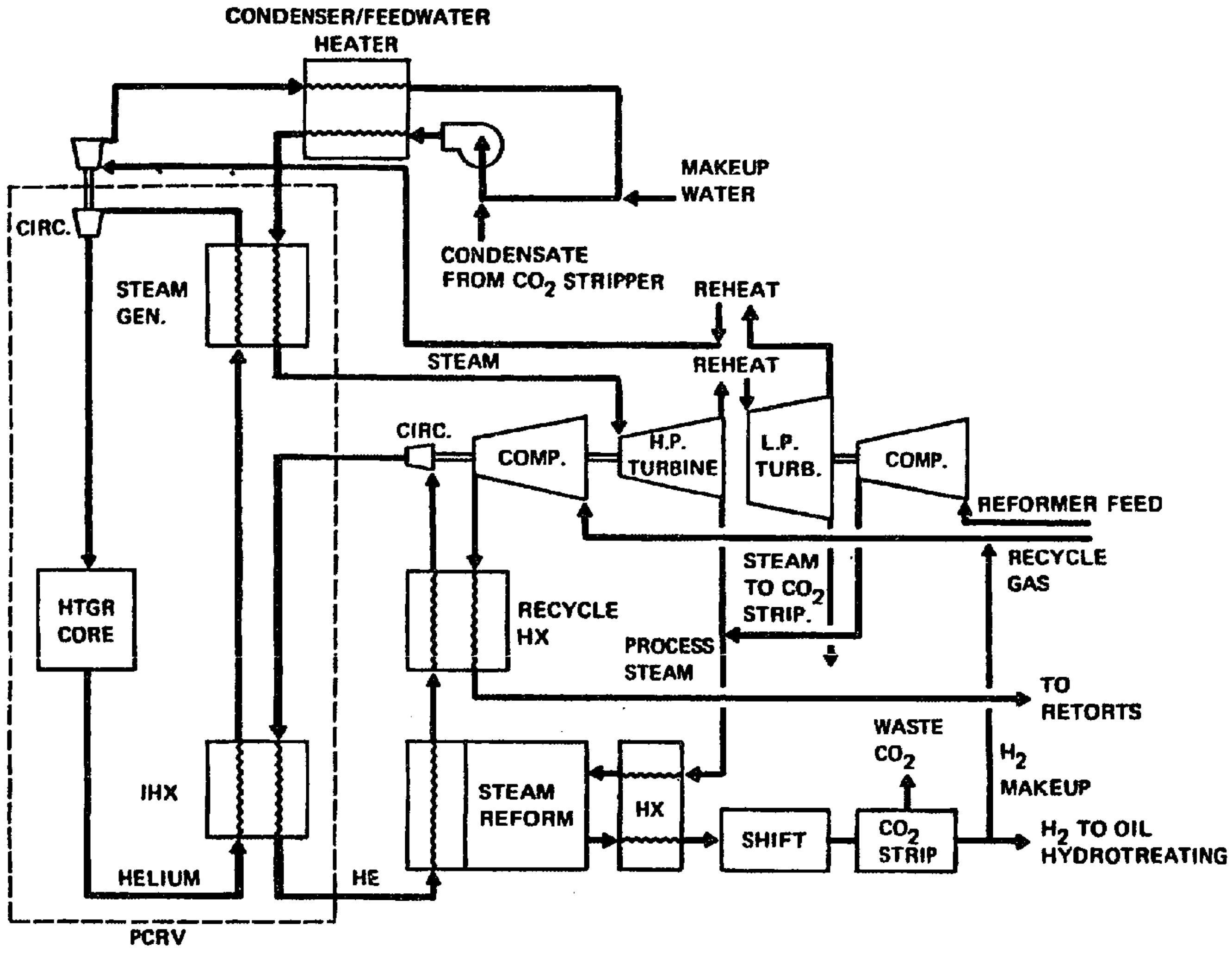

Figure B-9. SIMPLIFIED FLOW DIAGRAM OF HEAT-TRANSFER, POWER, AND PROCESS STREAMS COUPLING HTGR TO IGT HYDRORETORTING PROCESS[15]

Biomass

As an example of the number of commercial biomass plants (50 tons per day) that could be served by a HTR, considered a 300 MWth heat source. One dry ton of wastes produces approximately 16 million Btu of low-Btu gas energy. The heat requirement to pyrolyze that ton of wastes is one-third of its output, or 5.3 million Btu. The amount of wastes that could be processed in a day is about 4600 tons.

Assuming about a 20% moisture content, 110 50-ton per day pyrolysis units would be required. Clearly, if units that "small" are to be used, a formidable heat exchange design problem must be addressed. This would seem to be a major R&D requirement.

B.2 THERMOCHEMICAL HYDROGEN FROM NUCLEAR PROCESS HEAT

B.2.1 Introduction

A thermochemical water-splitting process is a sequence of chemical reactions in which every species except water is recycled. Ideally, the net inputs are only water and thermal energy. The net outputs are hydrogen, oxygen, and degraded heat. Thermochemical water-splitting processes offer a closed-cycle, nonmaterial-polluting route to fuel synthesis. They are environmentally compatible because the only by-product is oxygen and because combustion of the product hydrogen re-creates the raw material, water.

In the long-term, thermochemical water-splitting processes offer a technology for transforming heat from any moderate- or high-temperature source into chemical energy by using a perpetually available resource. For the near term, hydrogen from a heat source, such as a nuclear reactor, can be used to supplement fossil-fuel sources, such as natural gas (blending), petroleum (hydrotreating), and especially coal, shale kerogens, or sand bitumens (hydrogenation to liquids or gases). If hydrogen and oxygen can be produced by a water-splitting process at low cost, they would assume increased importance as industrial commodities as well as fuel sources.

Ideally, water can be split into hydrogen and oxygen by supplying the enthalpy of reaction with a combination of thermal energy (for entropy requirements) and work energy (for free-energy requirements). The present technology for water-splitting is electrolysis, in which work energy (electricity) in excess of the reaction enthalpy is supplied to produce hydrogen and oxygen from a water-electrolyte solution. A heat-to-work transformation is needed to generate the required electrical energy from primary thermal energy (fossil, nuclear, or solar). The efficiency of this transformation is restricted by thermodynamic limitations and by practical constraints in operating power plants.

To eliminate the heat-to-work cycle, water can be decomposed in a single step by heating it to very high temperatures (2500° to 4000°C) and separating the gaseous products. The materials required for containment and separation limit practical applications at these temperatures. The same thermal decomposition and separation can be accomplished through multiple chemical reaction steps operating at lower temperatures, i.e., through a thermochemical water-splitting process. According to the second law of thermodynamics, quantities of heat in excess of the reaction free energy plus entropy requirements of water-splitting must be supplied. The chemical reactions are driven principally by thermal energy, with a small requirement for mechanical or electrical energy.(B-24)

B.2.2 Classes of Thermochemical Cycles

Nearly all thermochemical water-splitting cycles can be placed into one of the five general classes below.

- **Class 1: Metal-Metal Oxide Cycles**

  $M + H_2O \rightarrow MO + H_2$

  $MO \rightarrow M + 1/2O_2$

- **Class 2: Metal Oxide-Metal Hydroxide Cycles**

  $M + 2H_2O \rightarrow M(OH)_2 + H_2$

  $M(OH)_2 \rightarrow MO + H_2O$

  $MO \rightarrow M + 1/2O_2$

- **Class 3: Metal Oxide-Metal Sulfate Cycles**

  $MO + SO_2 + H_2O \rightarrow MSO_4 + H_2$

  $MSO_4 \rightarrow MO + SO_3$

  $SO_3 \rightarrow SO_2 + 1/2O_2$

- **Class 4: Metal-Metal Halide Cycles**

  $M + 2HX \rightarrow MX_2 + H_2$

  $MX_2 \rightarrow M + X_2$

  $X_2 + H_2O \rightarrow 2HX + 1/2O_2$

- **Class 5: Metal Oxide-metal Halide Cycles**

  $M + H_2O \rightarrow MO + H_2$

  $MO + 2HX \rightarrow MX_2 + H_2O$

  $MX_2 \rightarrow M + X_2$

  $X_2 + H_2O \rightarrow 2HX + 1/2O_2$

M is considered to be a zero-valent metal, or the lower valence state of an oxide or halide pair. Other types of cycles are known, such as hybrid thermoelectrochemical processes that incorporate electrical energy for electrolysis in one of the steps.(B-24)

Many combinations of cycles within these classes can be conceived and a plethora of cycles has been published. It is expected that the simpler concepts will be easier to develop into commercial processes.

Cycle B-1, represented by Reactions 1-4, is one of IGT's most developed thermochemical water-splitting cycles:

$$3FeCl_2 + 4H_2O \rightarrow Fe_3O_4 + 6HCl + H_2 \tag{1}$$

$$Fe_3O_4 + 8HCl \rightarrow 2FeCl_3 + FeCl_2 + 4H_2O \tag{2}$$

$$2FeCl_3 \rightarrow 2FeCl_2 + Cl_2 \tag{3}$$

$$Cl_2 + H_2O \rightarrow 2HCl + 1/2\,O_2 \tag{4}$$

A conceptual flowsheet for this cycle has been prepared; Figure 10 shows a schematic of the flowsheet and Table B-3 provides a summary of molar flow rates based on the production of 1 gram-mole of hydrogen.(B-25)

Many separations are required for Cycle B-2 Reactions 1, 2, and 4 form gaseous product streams (S-1, S-7, and S-14) whose primary components are H_2O and HCl, which must be separated for recycling. Areas of major difficulty include the HCl/H_2O separation and the large steam requirement for the $FeCl_2$ hydrolysis (Reaction 1).(B-26) Overall thermal to hydrogen energy conversion efficiency is estimated as 18%, based on a nonoptimized flowsheet.

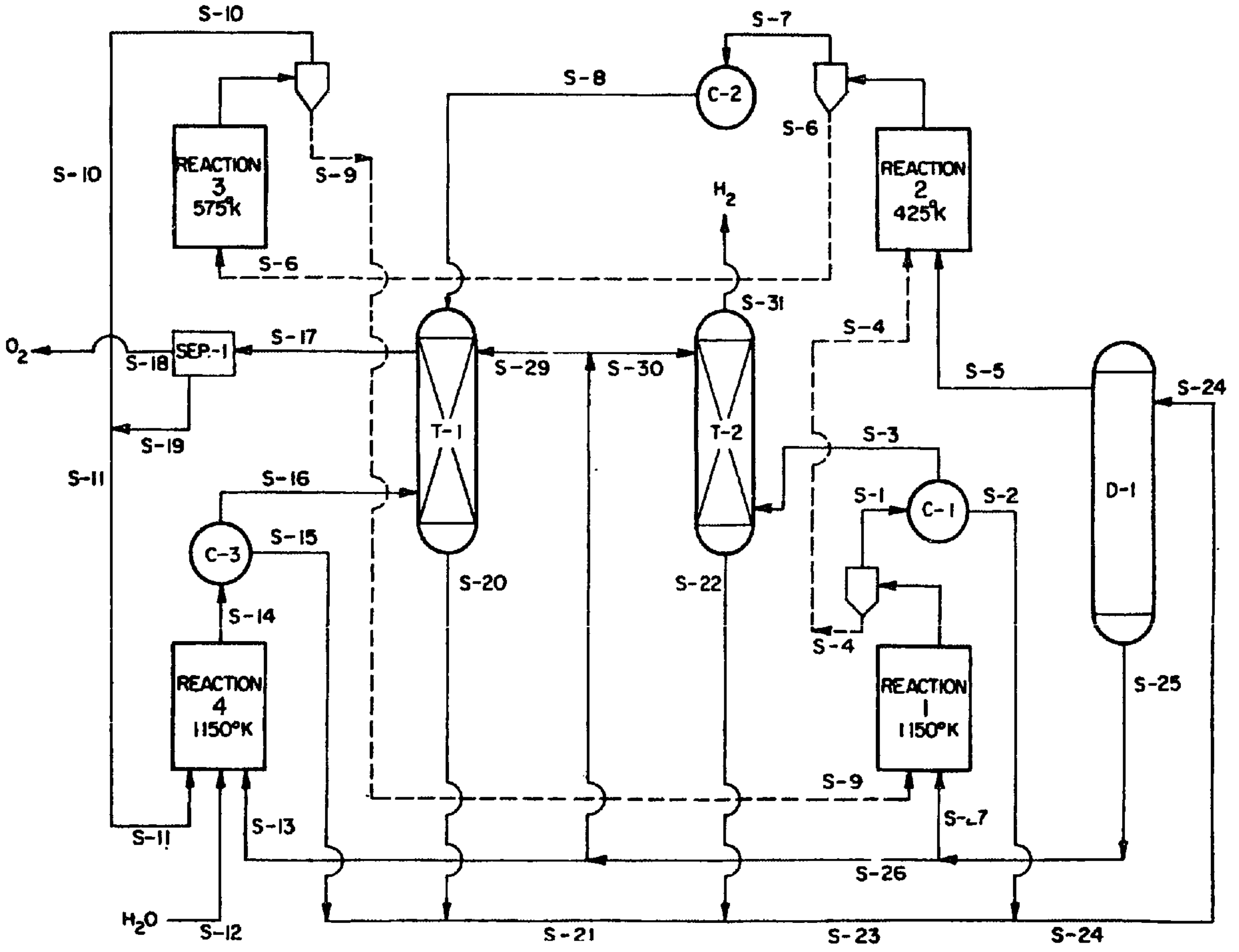

Figure B-10. SCHEMATIC DIAGRAM FOR CYCLE B-1

Table B-3
MOLAR FLOW RATES FOR THE FLOWSHEET SHOWN IN FIGURE 10

Stream	Comosition, mol		Temperature at Source, °C
S-1	1 H_2	(g)	877
	6.94 Hcl	(g)	
	3.54 H_2O	(g)	
S-2	3.54 H_2O	(ℓ)	25
	1.23 Hcl	(aq)	
S-3	1 H_2	(g)	25
	5.71 Hcl	(g)	
S-4	1 Fe_3O_4	(s)	877
S-5	9.04 HCl	(g)	60
	0.48 H_2O	(g)	
S-6	2 $FeCl_3$	(s)	152
	1 $FeCL_2$	(s)	
S-7	4.48 H_2O	(g)	152
	1.04 HCl	(g)	
S-8	4.48 H_2O	(ℓ)	25
	1.04 Hcl	(aq)	
S-9	3 $FeCl_2$	(s)	302
S-10	1 Cl_2	(g)	302
S-11	1.08 Cl_2	(g)	302
S-12	1 H_2O	(ℓ)	25
S-13	3.27 H_2O	(ℓ)	108
	0.41 HCl	(aq)	
S-14	0.50 O_2	(g)	877
	0.068 Cl_2	(g)	
	2.421 HCl	(g)	
	3.27 H_2O	(g)	
S-15	3.27 H_2O	(ℓ)	25
	1.14 HCl	(aq)	
S-16	0.50 O_2	(g)	25
	0.068 Cl_2	(g)	
	1.27 HCl	(g)	

Stream	Composition, mol		Temperature at Source, °C
S-17	0.50 O_2	(g)	25
	0.068 Cl_2	(g)	25
S-18	0.50 O_2	(g)	25
S-19	0.068 Cl_2	(g)	25
S-20	7.79 H_2O	(ℓ)	25
	2.72 HCl	(aq)	
S-21	11.0 H_2O	(ℓ)	25
	3.86 HCl	(aq)	25
S-22	25.49 H_2O	(ℓ)	25
	8.89 Hcl	(aq)	
S-23	36.56 H_2O	(ℓ)	25
	12.76 HCl	(aq)	
S-24	40.09 H_2O	(ℓ)	25
	13.99 HCl	(aq)	
S-25	39.62 H_2O	(ℓ)	108
	4.96 HCl	(aq)	
S-26	32.08 H_2O	(ℓ)	108
	4.01 HCl	(aq)	
S-27	7.54 H_2O	(ℓ)	108
	0.94 HCl	(aq)	
S-28	28.81 H_2O	(ℓ)	108
	3.60 HCl	(aq)	
S-29	3.32 H_2O	(ℓ)	25
	0.41 HCl	(aq)	
S-30	25.49 H_2O	(ℓ)	25
	3.19 HCl	(aq)	
S-31	1 H_2	(g)	25

A more promising cycle (designated H-5) studied at IGT is a hybrid copper oxide/copper sulfate thermochemical water-splitting cycle:

$$CuO(s) + SO_2(g) + 6H_2O(\ell) \rightarrow CuSO_4 \cdot 5H_2O(s) + H_2(g) \tag{5}$$

$$CuSO_4 \cdot 5H_2O(s) \rightarrow CuSO_4(s) + 5H_2O(g) \tag{6}$$

$$CuSO_4(s) \rightarrow CuO(s) + SO_3(g) \tag{7}$$

$$SO_3(g) \rightarrow SO_2(g) + 1/2\ O_2(g) \tag{8}$$

The hydrogen-producing step (Reaction 5) proceeds by electrolysis; the remaining reactions require thermal inputs. A conceptual flow sheet has been developed and is presented in Figure 11; a mass balance for the H-5 cycle (per gram-mole of hydrogen) is shown in Table B-4.(B-27)

Primary heat source requirements are summarized in Table B-5. Thermal-to-hydrogen energy conversion efficiency is estimated as about 40%; electricity for the electrolysis step is assumed to be generated from high temperatures at 50% efficiency.

B.2.3 HTR Heat Source Utilization

IGT Cycles B-1 and H-5 do not mate well with high-temperature nuclear fission heat sources. The temperature of the secondary helium loop is high enough for the B-1 cycle, but the capacity for the transfer of heat to isothermal chemical reactions at high temperature is too low. For the H-5 cycle, heat is required at about 1030°C, which is a fundamental mismatch with an HTR core exit temperature of 950°C.

The difficulty in matching these two processes to an HTR points out the basic mismatch between thermochemical hydrogen production cycles and the sensible heat supplied by a nuclear coolant stream. The most efficient thermochemical processes are those that 1) have the fewest number of chemical reactions and 2) require the bulk of their primary heat inputs at the highest temperatures. The primary heat used to drive high-temperature chemical reactions will be relatively isothermal. Nuclear fission heat sources, on the other hand, supply sensible heat, which means that the temperature of the coolant stream constantly cascades downwards as heat is removed. Therefore, only a small fraction of the heat supplied by a coolant stream is available at the highest temperature levels. The supply of heat at several (falling) temperature levels calls for a cycle with several different isothermal heat requirements, rather than 1 or 2 at high temperature. Such cycles are inherently less efficient and more complex than cycles with fewer reaction sequences. The higher the temperature and the more nearly isothermal the heat source, the higher the efficiency and the lower the complexity of the thermochemical cycle capable of being driven by that heat source. Solar and nuclear fusion heat sources currently being developed appear better suited for thermochemical cycles because they should be capable of supplying high-temperature heat isothermally.

There are, however, a number of thermochemical cycles with attractive efficiencies capable of mating with an HTR. A sulfuric acid cycle (Westinghouse) and a sulfur-bromine cycle (Joint European Community) are the most developed of these cycles.

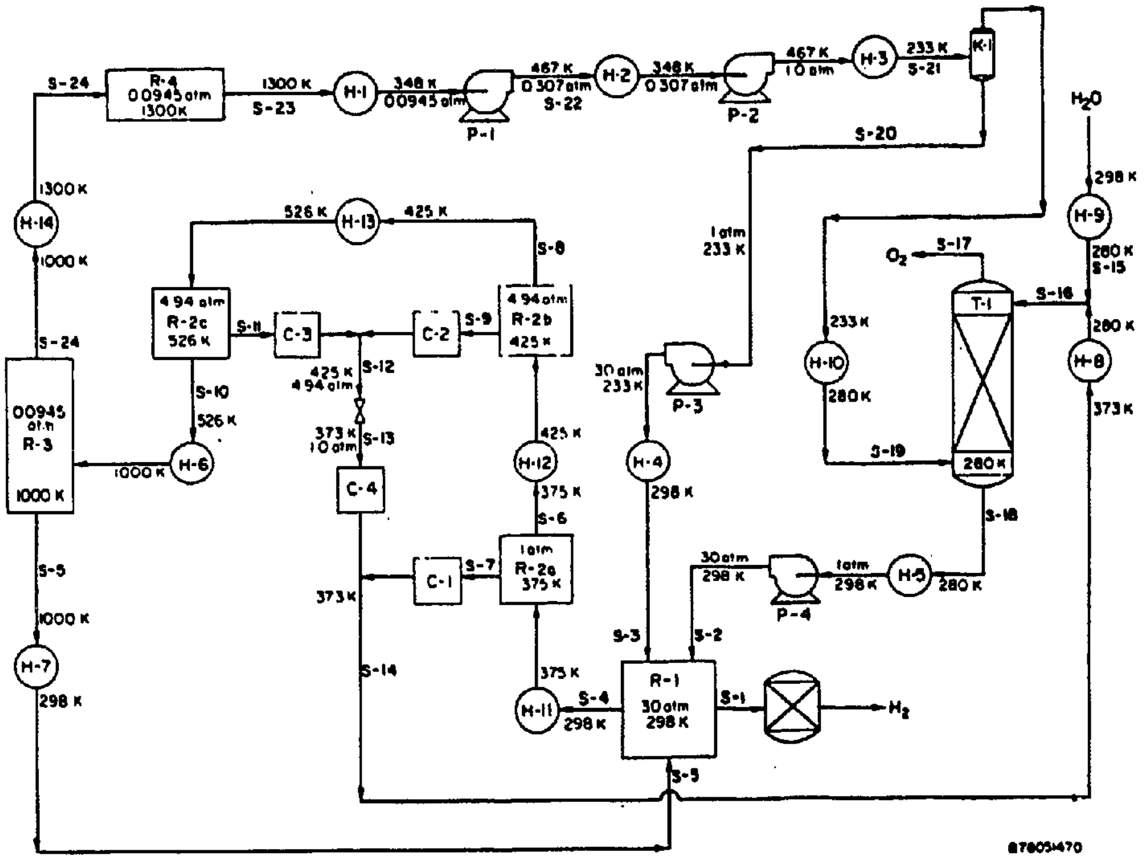

Figure B-11. FLOW SHEET FOR CYCLE H-5

Table B-4
MASS BALANCE FOR CYCLE H-5[14]

	Inlet Streams		Outlet Streams			
Unit	**Stream No.**	**Composition, mol**	**Stream No.**	**Composition, mol**	**Pressure, atm**	**Temperature, °C**
R-1	S-3	0.022 SO_3 (in SO_2)	S-1	1.000 $H_2(g)$	30.0	25
	S-3	0.874 SO_2 (ℓ)	S-4	1.002 $CuSO_4 \cdot 5H_2O(s)$		
	S-2	0.126 SO_2 (aq)				
	S-2	6.108 $H_2O(\ell)$				
	S-5	1.022 CuO(s)				
R-2a	S-4	1.022 $CuSO_4 \cdot 5H_2O(s)$	S-6	1.022 $CuSO_4 \cdot 3H_2O(s)$	1.0	102
			S-7	2.043 $H_2O(g)$		
R-2b	S-6	1.022 $CuSO_4 \cdot 3H_2O(s)$	S-8	1.022 $CuSO_4 \cdot H_2O(s)$	4.94	152
			S-9	2.043 $H_2O(g)$		
R-2c	S-8	1.022 $CuSO_4 \cdot H_2O(s)$	S-10	1.022 $CuSO_4(s)$	4.94	253
			S-11	1.022 $H_2O(g)$		
R-3	S-10	1.022 $CuSO_4(s)$	S-5	1.022 CuO(s)	0.0945	700
			S-24	0.233 $SO_3(g)$		
			S-24	0.789 $So_2(g)$		
			S-24	0.394 $O_2(g)$		
R-4	S-24	0.233 $SO_3(g)$	S-23	0.022 $SO_3(g)$	0.945	1030
	S-24	0.789 $SO_2(g)$	S-23	1.000 $SO_2(g)$		
	S-24	0.394 $O_2(g)$	S-23	0.500 $O_2(g)$		
K-1	S-21	0.022 $SO_3(g,s)$	S-20	0.021 SO_3 (in SO_2)	1.0	−40
	S-21	1.000 $So_2(g,\ell)$	S-20	0.874 $SO_2(\ell)$		
	S-21	0.500 $O_2(g)$	S-19	0.001 $So_3(g)$		
			S-19	0.126 $SO_2(g)$		
			S-19	0.500 $O_2(g)$		
T-1	S-19	0.001 $SO_3(g)$	S-17	0.500 $O_2(g)$	1.0	7
	S-19	0.126 $SO_2(g)$	S-18	0.001 $SO_3(aq)$		
	S-19	0.500 $O_2(g)$	S-18	0.126 $SO_2(aq)$		
	S-16	6.108 $H_2O(\ell)$	S-18	6.108 $H_2O(\ell)$		

p

Table B-5
PRIMARY HEAT SOURCE REQUIREMENTS FOR CYCLE H-5
(Basis: 1 gram-mole of hydrogen or 271.1 Btu)

Unit	Q, Btu	Temperature, °C
R-4	19.3	1030
H-14	5.0	960 - 1030
Work	265.1	1030
R-265	100.3	152
R-2c	72.5	253
R-3	274.1	700
R-6	18.5	500 - 700
Total	754.8	

$$\text{Efficiency} = \frac{271.1}{754.8} = 0.36 \text{ or } 40\%$$

- **Westinghouse Sulfuric Acid Cycle**

$SO_2 + 2H_2O \rightarrow H_2SO_4 + H_2$ (electrochemical)

$H_2SO_4 \rightarrow H_2O + SO_2 + 1/2\ O_2$

- **Euratom Sulfur-Bromine Cycle**

$SO_2 + Br_2 + 2H_2O \rightarrow H_2SO_4 + 2HBr$

$H_2SO_4 \rightarrow H_2O + SO_2 + 1/2\ O_2$

$2HBr \rightarrow Br_2 + H_2$ (electrochemical)

B.2.4 Electrolytic Hydrogen Production

IGT has studied the electrolytic production of hydrogen by high-temperature nuclear reactors.(B-27) The conceptual facility integrated a high-temperature gas-cooled reactor with direct-current acyclic generators and solid polymer electrolyte electrolyzers. All subsystems are close-coupled and optimally interfaced for hydrogen production alone (i.e., without separate production of electrical power). The HTGR coolant exit temperature was 980°C.

It was found that thermal-to-hydrogen energy conversion efficiency for this advanced facility was 43%. The hydrogen production cost was estimated to be about one-half that of contemporary electrolysis. This concept appears competitive with other water-splitting methods based on nuclear-fission.

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APPENDIX C
NUCLEAR FUEL CYCLE

C.1 FUEL CYCLE SUMMARY

The nuclear fuel cycles studied were all based on the conversion of thorium to fissionable uranium-233 (U-233) using fuel enriched in U-235 or U-233 as a feed material. The high neutron efficiency of U-233 makes U-233 particularly well suited for use in graphite moderated reactors. The high conversion efficiency of the U-Th cycle helps to minimize the annual fuel requirements and fuel costs of the Pebble Bed Reactor (PBR).

Four basic fuel cycles are of current worldwide interest. Two of these cycles are "stow away" cycles in which the fuel elements are stored without immediate reprocessing once they achieve their final burnup. Feed material for these cycles is a mixture of either high enriched U-235 (HEU) or medium enriched U-235 (MEU) particles (as oxide or carbide) mixed with enough Th-232 particles to provide the desired feed enrichment. The MEU feed option is of interest solely because the uranium feed material is not useful for use in weapons manufacture. The two remaining fuel cycles involve reprocessing and recycling of the spent fuel. The feed enrichment is achieved with either U-233 or U-235 makeup (HEU). With recycle the uranium ore requirements are reduced relative to the "stow away" cycles, however the front-end of the fuel cycle requires remote handling because of the activity of the U-234 decay products. In this report the fuel cycles with no reprocessing will be referred to as Once-Through. This is not to be confused with the "Once-Through Then Out" or OTTO fuel management scheme in which the fuel achieves its discharge exposure in one pass through the core. The fuel cycles with reprocessing and recycling will be referred to as Recycle. All of the fuel cycles considered here are of the OTTO type.

For this study, two designs have been selected for each fuel cycle to show the range of expected fuel parameters which might occur. The lower bound represents fuel designs which approximate the characteristics of current Thorium High Temperature Reactor (THTR) fuels; the upper bound represents an improvement in one of the design characteristics of the current THTR fuels, usually the heavy metal loading per ball. Neither the upper nor lower bound are intended to show maximum or minimum characteristics but rather characteristics which might reasonably be expected.

The Once-Through MEU fuel designs are characterized by high fissile burnups (100,000 MWD/MT) with medium enriched uranium-235 used both as first core and reload core fuel. The designs were developed to minimize proliferation risks while still maintaining a high burnup, and THTR and AVR fuel characteristics.

The Once-Through HEU fuel designs are representative of extended parametric research on Once-Through HEU fuel cycles in PBR's in Germany. The designs are characterized by high fissile burnups (100,000 MWd/MT) with high enriched uranium-235 used both as first core and reload core fuel. The designs were developed as introductory fuel designs, to be used before recycle occurred.

With fuel recycle, the PBR can recover its bred U-233 and increase the amount of U-233 in the core. The net results are fuels with a higher conversion ratio and lower uranium-235 makeup fuel requirements. The fuel designs studied represent the equilibrium conditions that would occur after many U-235 fueled-cycles had created an inventory of U-233. Once the inventory has reached equilibrium no net gain or loss of U-233 occurs, but the makeup fuel requirements are again U-235.

If U-233 is available, the above HEU recycle designs might be initally fueled by U-233 rather than by U-235, and U-233 might also be substituted as makeup fuel. With a fully enriched U-233 core, the PBR can reach its highest conversion ratio. The Recycle U-233 designs present this high enriched uranium-233 fuel concept.

A summary of the fuel design parameters and results are shown in Table C-1.

Once-Through Versus Fuel Recycle

Both the Once-Through and the Recycle cases (Figure C-1) have distinctive advantages. The relative worths of these fuel cycles will depend on such issues as uranium conservation, fuel cycle economics, availability of backend fuel cycle services and the nation's perception of weapons proliferation.

The Once-Through fuel cycles are characterized by high uranium utilization without fuel reprocessing and refabrication. The fuel cycle is simple, requiring fewer and less expensive manufacturing steps than with Recycle, and relatively inexpensive when uranium is moderately priced, making it appealing as a means of introducing the PBR. Additionally, because the fuel is optimized for a high burnup, the Once-Through cycles generally produce a smaller volume of fuel wastes than do the Recycle cases.*

The Recycle cases are characterized by a higher uranium utilization due to fuel reprocessing and refabrication. The fuel cycle however is more complex than the Once-Through case because of the need for fuel reprocessing and remote fabrication facilities, but this additional complexity can reduce fuel costs when uranium prices increase. The use of these last two technologies involves proliferation concerns.

A summary of characteristics which distinguish the Once-Through and the Recycle cases is shown in Table C-2.

HEU vs MEU Fuels

The original fuels developed for Pebble Bed Reactors and those which have been investigated most fully have been high enriched uranium (HEU) fuels. To reduce proliferation potential, the use of medium enriched uranium (MEU) as a feed material has been suggested for Pebble Bed Reactors. These fuels have fissile fuel enrichments of less than 20% U-235 or 15% U-233. The dilutant in both cases would be the fertile element U-238, making the fissile portion of the fuel chemically inseparable and unuseable in weapons manufacture. The MEU feed material would be mixed with thorium to provide the desired feed enrichment.

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*The mass of fission products produced is the same for both cycles, but the Recycle case uses more thorium per unit of energy.

Table C-1
FUEL INFORMATION SUMMARY

	Once-Through MEU Fuels		Once-Through HEU Fuels		Recycle HEU Fuels		Recycle U-233 Fuels	
	THTR	Advanced	THTR	Advanced	THTR	Advanced	THTR*	Advanced
Fuel Recycle	No	No	No	No	Yes	Yes	Yes	Yes
U-235 (and U-233) Enrichment	20%	20%	93%	93%	93%	93%	99%	99%
Burnup (MWd_o/Kg)	100	100	100	100	23.7	32.0	35.0	35.0
Conversion Ratio	0.64	0.58	0.594	0.665	0.77	0.85	0.86	0.91
Fuel Residence Time ($days_o$)	1223	872	1217	1070	382	687	507	676
Ball Loading (gm/ball)	11.24	8.07	11.24	20.00	11.24	20.0	15.0	20.0
Carbon/Heavy Metal Ratio	325	458	325	180	325	180	240	180
Capacity Factor	80%	80%	80%	80%	80%	80%	80%	80%
One-Cycle Fuel Cycle Cost (1979 ¢/MBtu)	110	102	105	98	128	94	67	62
Equilibrium Annual Uranium Use (ST U_3O_8/GW_{Th}-year)	60	58	54	53	21	14	NA (55 kg U-233)	NA (44 kg U-233)

NA = Not applicable, design uses U-233 for makeup requirements.

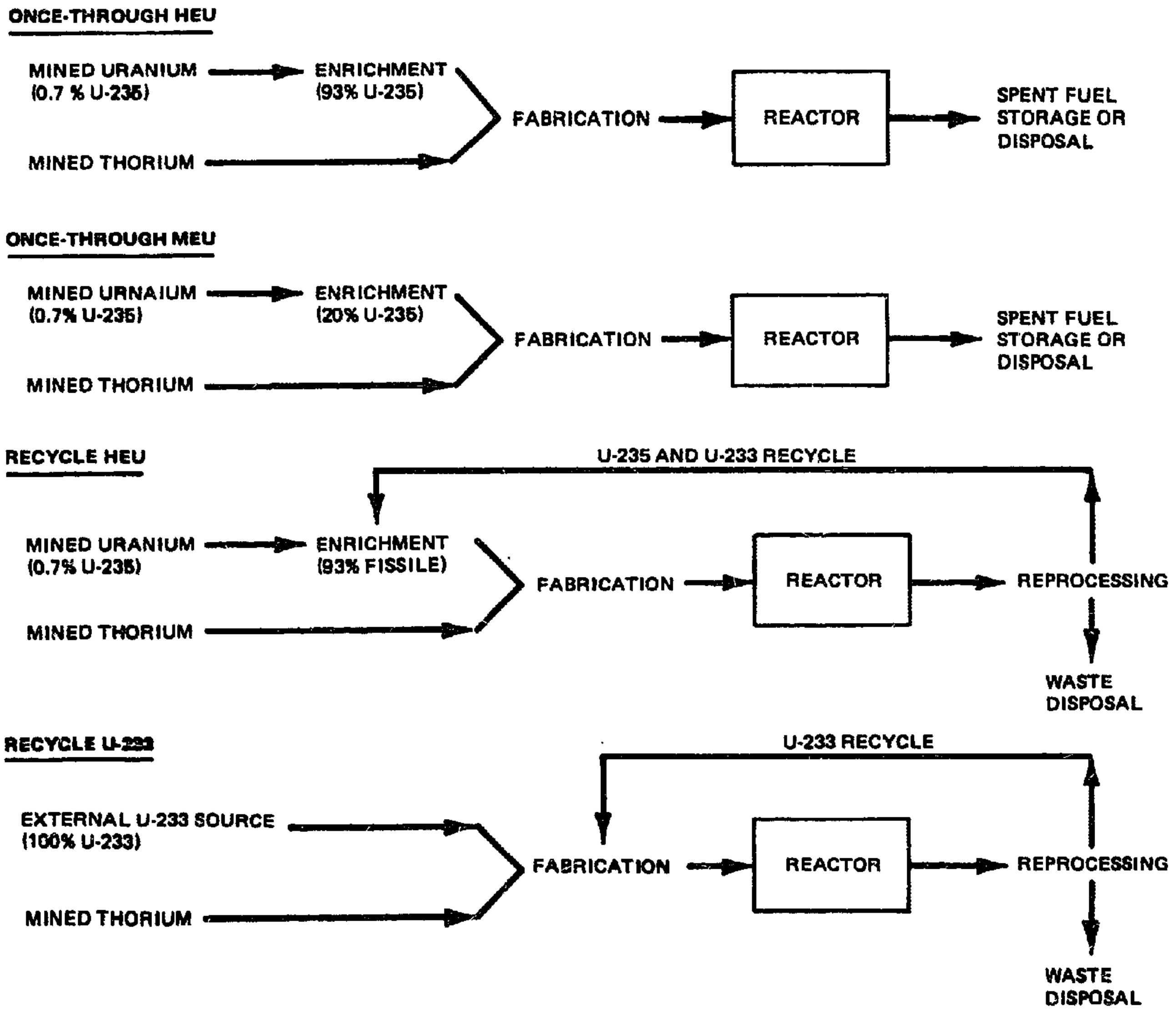

Figure C-1. PEBBLE BED REACTOR FUEL CYCLES

Table C-2
FUEL CYCLE COMPARISON OF ONCE-THROUGH WITH RECYCLE*
(Values Relative to Once-Through for Each Category)

	Once-Through	Fuel Recycle
Fuel and Fuel Service Use (MT/MT)		
Uranium	1.0	0.4
Thorium	1.0	2.0
Conversion	1.0	0.4
Fabrication	1.0	2.5
Enrichment	1.0	0.4
Reprocessing	NA	—
Waste Disposal	1.0	2.2
Fuel Transportation	1.0	3.2
Fuel Service Costs ($/kgHM)		
Fabrication	1.0	1.4
Reprocessing, Waste and Transportation	1.0	1.5
Fuel Cycle Cost (¢/MBtu) (Conditions of Tables C-22 and C-23)		
Cost	1.0	1.2

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*Based on equilibrium HEU Once-Through THTR and Recycle HEU THTR designs of Table C-4.

A summary of characteristics which distinguish MEU fuels from HEU fuels is shown in Table C-3. The penalty for using MEU Once-Through fuels over HEU Once-Through fuels is small.

C.2 FUEL CHARACTERISTICS

For this study, two designs have been selected for each of the four fuel cycles described in above to show the range of expected fuel parameters which might occur. Results are shown in Table C-4.

The lower bound represents fuel designs which approximate the characteristics of current THTR fuels; the upper bound, an improvement in one of the design characteristics of the current THTR fuels, usually the heavy metal loading per ball. Neither the upper nor lower bound are intended to show maximum or minimum characteristics but rather characteristics which might reasonably be expected.

Each of these fuel designs were also developed to meet different fuel cost and uranium ore conservation objectives. With the Once-Through fuel designs, an objective was to reduce cycle costs. This was accomplished by making the fuel usable for a longer period (high fuel burnup), thus distributing the unit fuel service costs over a larger heat output base. With the Recycle fuel design, an objective was to reduce uranium requirements, because of high uranium prices or low uranium availability. This was accomplished by lowering the fuel burnup (shortening fuel life) to remove the fission products more often plus reoptimizing the fuel design. Thus the selection of the fuel burnup resulted from the design objective.

Table C-3
COMPARISON OF ONCE-THROUGH CYCLES WITH MEU AND HEU FEED*
(Values Relative to MEU for each Category)

	MEU	HEU
Fuel and Fuel Service Use — MT		
Uranium	1.0	0.9
Thorium	1.0	1.9
Conversion	1.0	0.9
Fabrication	1.0	1.2
Enrichment	1.0	1.0
Reprocessing	NA	NA
Waste Disposal	1.0	
Fuel Transportation	1.0	1.0
Fuel Service Costs — $/kg HM		
Fabrication	1.0	1.0
Reprocessing, Waste and Transportation	1.0	1.0
Fuel Cycle Cost — ¢/MBtu (Conditions of Tables C-22 and C-23		
Cost	1.0	0.95

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*Based on equilibrium MEU Once-Through THTR and HEU Once-Through THTR designs of Table C-4.

The Once-Through MEU fuel designs of Table C-4 are characterized by high fissile burnups (100,000 MWd/MT) with medium enriched uranium-235 used both as first core and reload core fuel. The designs were developed to minimize proliferation risks while still maintaining a high burnup, and THTR and AVR fuel characteristics.(C-1) A summary of mass and isotopic fuel balances for the Once-Through MEU fuel designs are shown in Table C-5.

The Once-Through HEU fuel designs of Table C-4 are representative of extended parametric research on Once-Through HEU fuel cycles in PBRs in Germany(C-1) and are more fully evaluated than the above MEU fuel designs. The designs are characterized by high fissile burnups (100,000 MWd/MT) with high enriched uranium-235 used both as first core and reload core fuel. The designs were developed as introductory fuel designs, to be used before recycle occurred. A summary of mass and isotopic fuel balances for the Once-Through HEU fuel designs are shown in Table C-6.

Table C-4
SUMMARY OF FUEL CHARACTERISTICS

	Once-Through MEU Fuels		Once-Through HEU Fuels		Recycle HEU Fuels		Recycle U-233 Fuels	
	THTR	**Advanced**	**THTR**	**Advanced**	**THTR**	**Advanced**	**THTR***	**Advanced**
Makeup Fuel Isotope	U-235	U-235	U-235	U-235	U-235	U-235	U-233	U-233
Fuel Recycle	No	No	No	No	Yes	Yes	Yes	Yes
Uranium Enrichment	20%	20%	93%	93%	93%	93%	99%	99%
Fresh Fuel/Total Fuel Load Fraction	100%	100%	100%	100%	71.6%	71.6%	0%	0%
Burnup (MWd_o/Kg)	100	100	100	100	23.7	32.0	35.0	35.0
Maximum Power Density (kW/ball)	2.2	2.6	2.9	2.4	3.6	3.4	1.59	1.53
Conversion Ratio	0.64	0.58	0.594	0.665	0.77	0.85	0.86	0.91
Fuel Residence Time ($days_o$)	1223	872	1217	1070	382	687	507	676
Ball Loading (gm/ball)	11.24	8.07	11.24	20.00	11.24	20.0	15.0	20.0
Carbon/Heavy Metal Ratio	325	458	325	180	325	180	240	180
Capacity Factor	80%	80%	80%	80%	80%	80%	80%	80%
Cycle Identification Number	3225	M020	M093	18005	325/32	180/32	6D	6B
	Reference 2		Reference 2		Reference 1		Developed at ARSD	

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*Moderately Advanced THTR (Higher ball loading)

Table C-5
FUEL BALANCES FOR ONCE-THROUGH MEU FUEL CYCLES
(1000 $\mathrm{MW_{th}}$)

	THTR Series		Advanced Fuel	
	Load	**Discharge**	**Load**	**Discharge**
FUEL MASS BALANCES ($\mathrm{kg/gW_{Th}d}$)				
HEAVY METAL	9.999	8.839	10.017	8.878
Th-232	5.932	5.511	6.076	5.671
Uranium	4.067	3.328	3.941	3.207
U-233	—	0.161	—	0.133
U-234	—	0.031	—	0.033
U-235	0.805	0.107	0.780	0.056
U-236	—	0.115	—	0.114
U-238	3.262	2.914	3.161	2.871
Pu-239	—	0.039	—	0.021

	THTR Series			Advanced Fuel		
	1st Core	**Reload**	**30 Year**	**1st Core**	**Reload**	**30 Year**
FUEL REQUIREMENTS (1,000 $\mathrm{MW_{Th}}$; 80% CF; 0.2% tails)						
Uranium (ST U_3O_8)	250	60	1990	170	58	1850
Thorium (MT Th)	7.1	1.7	56	5.4	1.8	58
Enrichment (MT)	230	54	1800	160	54	1700
Fabrication (MT)	12	2.9	96	9	2.9	93
Spent Fuel (MT) Storage	12	2.9	96	9	2.9	93

Table C-6
FUEL BALANCES FOR ONCE-THROUGH HEU FUEL DESIGNS
(1000 MW_{Th})

	THTR Series		Advanced Fuel	
	Load	Discharge	Load	Discharge
FUEL MASS BALANCES (kg/gW_{Th}-d)				
HEAVY METAL	9.988	8.895	9.921	8.841
Th-232	9.209	8.467	9.155	8.357
Uranium	0.779	0.428	0.766	0.484
U-233	—	0.187	—	0.229
U-234	—	0.056	—	0.054
U-235	0.724	0.034	0.712	0.059
U-236	—	0.107	—	0.104
U-238	0.055	0.044	0.054	0.038
Pu-239	—	>0.1	—	>0.1

	THTR Series			Advanced Fuel		
	1st Core	Reload	30 Year	1st Core	Reload	30 Year
FUEL REQUIREMENTS (1,000 MW_{Th}; 80% CF; 0.2% tails)						
Uranium (ST U_3O_8)	230	54	1800	200	53	1740
Thorium (MT Th)	11	2.7	90	10	2.7	90
Enrichment (MT)	230	54	1800	200	53	1740
Fabrication (MT)	12	2.9	96	11	2.9	95
Spent Fuel (MT) Storage	12	2.9	96	11	2.9	95

With fuel recycle, the PBR can recover its bred U-233 and increase the amount of U-233 in the core. The net result are fuels with a higher conversion ratio and lower uranium-235 makeup fuel requirements.

The fuel designs of Table C-4 represent the equilibrium conditions that would occur after many U-235 fueled cycles had created an inventory of U-233. Once the inventory has reached equilibrium no net gain or loss of U-233 occurs, but the makeup fuel requirements are again U-235. A summary of the isotopic and mass balances for the Recycle HEU fuel designs(C-2) are shown in Table C-7.

If U-233 is available, the above HEU recycle designs might be initially fueled by U-233 rather than by U-235, and U-233 might also be substituted as makeup fuel. The Recycle U-233 designs of Table C-4 represent this high enriched uranium-233 fuel concept. With a fully enriched U-233 core, the PBR can reach its highest conversion ratio. A summary of the isotopic and mass balances for the Recycle U-233 designs are shown in Table C-8.

C.3 FUEL REQUIREMENTS

U_3O_8 Requirements

The uranium (U_3O_8) requirements for the PBR will depend both on the fuel design (and fuel cycle choice) as well as on the operating tails assay of the enrichment plants. The fuel design principally determines how efficient the fuel will be burned in the reactor, while the enrichment tails assay determines how much new uranium (99.3% U-238 and 0.7% U-235) are needed to produce the fuel product (either 93% or 20% U-235). But for any tails assay, the relative usage of uranium (in %) between any two fuel designs will remain constant.

Table C-9 shows the first core, reload and 30 year uranium requirements for the eight fuel designs described above based on today's enrichment plant operating assay of 0.2% U-235.

Uranium-233 Requirements

All of the PBR fuel designs described above produce U-233. However, only the Recycle U-233 fuel designs require an external source of U-233. The U-233 fuel requirements for the Recycle U-233 designs are shown in Table C-10.

Thorium Requirements

All of the Pebble Bed Reactor fuel designs studied utilize thorium to breed uranium-233, thereby minimizing fissile fuel requirements. In addition to this role, thorium also dilutes the fissile enrichment of the fuel to useful levels of about 7 or 8%. For the HEU fuel designs which are 93% enriched uranium (U-235 or U-233), the amount of thorium needed to dilute the uranium fuel is larger than for 20% MEU fuels. The amount of thorium used in PBR's also varies with the flow rates of fuel through the reactor. For the Recycle designs selected this flow rate is about three times that of the Once-Through designs; with

Table C-7
FUEL BALANCES FOR RECYCLE HEU FUEL DESIGNS
(1000 MW_{Th})

	THTR Series			Advanced Fuel		
	Load	Load				
	Fresh	Recycle	Discharge	Fresh	Recycle	Discharge
FUEL MASS BALANCES (kg/gW_{Th}-d)						
HEAVY METAL	22.701	9.02	30.626	22.461	8.923	30.291
Th-232	22.394	7.470	29.146	22.255	7.404	28.782
Uranium	0.307	1.550	1.480	0.206	1.519	1.509
U-233	—	0.660	0.655	—	0.742	0.749
U-234	—	0.218	0.220	—	0.190	0.203
U-235	0.285	0.146	0.148	0.191	0.181	0.174
U-236	—	0.372	0.322	—	0.296	0.279
U-239	0.022	0.154	0.135	0.014	0.110	0.104
Pu-239	—	—	>0.1	—	—	>0.1

	THTR Series			Advanced Fuel		
	1st Core	Reload	30 Year	1st Core	Reload	30 Year
FUEL REQUIREMENTS (1,000 MW_{Th}; 80% CF; 0.2% tails)						
Uranium (ST U_3O_8)	76	21	685	190	14	595
Thorium (MT Th)	11.2	6.5	200	20.0	6.5	210
Enrichment (MT)	90	21	700	190	14	595
Fabrication (MT)	11.9	9.3	282	21.1	9.2	288
Reprocessing (MT)	11.9	9.3	282	21.1	9.2	288

thorium use correspondingly about three times as large, assuming no thorium recycle.* This result depends primarily on the discharge burnup which was selected to minimize uranium ore requirements on the recycle cases. The thorium requirements for the eight fuel designs of Section 10.2 are shown in Table C-11.

C.4 FUEL AVAILABILITY

Domestic U_3O_8 Availability

The U.S. Department of Energy (DOE) and mining industry estimates of domestic uranium resources are based on both known reserves and expected but undiscovered resources in and around existing mining districts. Exploration activities throughout the geologically attractive areas since 1960 have

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*The discharged thorium from a reprocessing plant is slightly radioactive. At this time it appears cost effective to use new thorium when fabricating new and recycle fuels to avoid remote handling problems. If waste costs should rise, thorium recycle may occur.

Table C-8
FUEL BALANCES FOR RECYCLE U-233 FUEL DESIGNS
(1000 MW_{th})

	Advanced THTR		Advanced Fuel	
	Load	Discharge	Load	Discharge
FUEL MASS BALANCES ($kg/gW_{Th}d$)				
HEAVY METAL	28.86	27.80	28.84	27.76
Th-232	27.97	27.09	27.96	27.01
Uranium	0.89	0.71	0.88	0.75
U-233	0.86	0.67	0.85	0.71
U-234	—	—	—	—
U-235	0.03	0.04	0.03	0.04
U-236	—	—	—	—
U-238	—	—	—	—
Pu-239	—	—	—	—

	THTR Series			Advanced Fuel		
	1st Core	Reload	30 Year	1st Core	Reload	30 Year
FUEL REQUIREMENTS (1,000 MW_{Th}; 80% CF; 0.2% tails)						
U-233 (Kg)	341	55	1940	482	44	1760
Thorium (MT Th)	13.9	8.2	250	18.5	8.2	260
Enrichment (MT)	0	0	0	0	0	0
Fabrication (MT)	14.3	8.4	260	18.9	8.4	260
Reprocessing (MT)	14.3	8.4	260	18.9	8.4	260

Table C-9
URANIUM-235 REQUIREMENTS — SHORT TONS U_3O_8
(1000 MW_{th}, 80% Capacity Factor, 0.2% Tails)*

Fuel Cycle	U_3O_8 Requirements — Short Tons					
	THTR Series			Advanced Fuel		
	1st Core	Reload	30 Years	1st Core	Reload	30 Years
Once-Through MEU	250	60	1990	170	58	1850
Once-Through HEU	230	54	1800	200	53	1740
Recycle HEU	76	21	685	190	14	595
Recycle U-233**	0	0	0	0	0	0

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*To convert to another tails assay use the following table:

Tails Assay (%)	Pounds Natural Uranium (HM) per Pound of New Fuel (HM)		Uranium Saved by Changing Tails Assay — %
	20% Enriched	93% Enriched	
0.10	32.5	151.9	16%
0.15	35.4	165.4	9%
0.20 (Reference)	38.7	181.4	0%
0.25	42.8	201.0	(11%)
0.30	47.9	225.3	(24%)

**U-233 must be supplied annually from an external source.

not located any new districts, in spite of the fact that exploration effort since 1960 is four times greater than all cumulative exploration through 1960.

The 1978 DOE U.S. uranium resource estimate is divided into four information categories: 'Reserves', and 'Probable', 'Possible', and 'Speculative' Resources, as shown in Table C-12. The four information categories represent ores in differing states of discovery. The 'Reserves' and 'Probable' categories are defined by the largest information base and the 'Possible' and 'Speculative' categories by progressively less physical evidence.

The known 'Reserves' category represents ore resources for which definitive drilling and mineralogical statistics have been determined and which are now being developed. To many, the 'Reserves' are comparable to those ores which a mining company would be willing to sell today for future delivery.

The 'Probable' resource category is defined as ores surrounding existing mines which have not been as extensively drilled as the 'Reserves' category ores. The 'Probable' resource estimates are found by mapping local ore trends and interpolating data where necessary.

The 'Possible' resource category represents ores which may be found when a mining district is fully explored and those ores in mining districts not currently producing uranium. This estimate is found by

Table C-10
URANIUM-233 REQUIREMENTS - Kg
(1000 MW_{Th}, 80% Capacity Factors)

	Uranium-233 Requirements — MT					
	THTR Series			Advanced Fuel		
Fuel Cycle	**1st Core**	**Reload**	**30 Year**	**1st Core**	**Reload**	**30 Year**
Once-Through MEU	0	0	0	0	0	0
Once-Through HEU	0	0	0	0	0	0
Recycle HEU	SGR	SGR	SGR	SGR	SGR	SGR
Recycle U-233	341	55	1940	482	44	1760

SGR = Self Generated recycle; no net external requirements

Table C-11
THORIUM REQUIREMENTS — MT
(1000 MW_{th}, 80% Capacity Factor)

	Thorium Requirements — MT					
	THTR Series			Advanced Fuel		
Fuel Cycle	**1st Core**	**Reload**	**30 Year**	**1st Core**	**Reload**	**30 Year**
Once-Through MEU	7.1	1.7	56	5.4	1.8	58
Once-Through HEU	11.0	2.7	90	10.0	2.7	90
Recycle HEU*	11.2	6.5	200	20.0	6.5	210
Recycle U-233*	13.9	8.2	250	18.5	8.2	260

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*No thorium recycle.

Table C-12
1978 DOE U.S. URANIUM RESOURCE ESTIMATE
(Tons U_3O_8)

		Estimated Resources			
Foward Cost ($/lb U_3O_8)	**Known Reserves**	**Probable**	**Possible**	**Speculative**	**Total**
Less Than 15	360,000	560,000	485,000	165,000	1,570,000
15 to 30	330,000	505,000	635,000	250,000	1,720,000
Up to 30	690,000	1,065,000	1,120,000	415,000	3,290,000
30 to 50	185,000	385,000	350,000	155,000	1,075,000
Up to 50	875,000	1,450,000	1,470,000	570,000	4,365,000
By-Product	140,000	—	—	—	140,000
TOTAL	1,015,000	1,450,000	1,470,000	570,000	4,505,000

multiplying the 'Reserves' and 'Probable' categories by the ratio of the total surface area of known mining districts to the surface area currently well explored in these mining districts.

The 'Speculative' resource category is defined as ores which might be discovered if additional mining districts are located. These additional districts are expected by probabilistic techniques but have yet to be found.

DOE believes there is a reasonable prospect that something in the range of 4.5 million tons of U_3O_8 will ultimately be recovered. In contrast, many in industry believe the significance of the DOE estimates lies in the supporting geological information that has been accumulated. They believe that because of the importance of being conservative in projecting national energy supplies, that probabilities should be placed on actually finding and developing the above ores. When such a procedure is used, U.S. resources have been estimated to be less than half of an estimate using the above DOE technique. DOE has also recognized the uncertainty of any geological estimating procedure and the importance of conservatism. For planning purposes, DOE considers 2.5 million tons consisting of ore 'Reserves' plus 'Probable' resources, plus byproduct sources, as a prudent planning basis.

For this assessment two U.S. uranium resource bases will be used. The first, a prudent planning base of 2.5 million tons. The second, the full DOE uranium estimate of 4.5 million tons (Table C-13).

Uranium-223

All of the fuel designs utilize uranium-233 (U-233) as a fuel. However, only four of the designs, the two Recycle HEU and the two Recycle U-233, require U-233 as a fresh fuel.

With the two Recycle HEU fuel designs, the design assumption was that the PBR would produce its own U-233 by starting up on U-235 and as U-233 is produced, converting to the equilibrium cycle shown in Table C-7. The equilibrium design then assumes that U-235 will be used as the purchasable fuel and no net U-233 purchases will occur.

With the two Recycle U-233 fuel designs, the design assumption is different. The Recycle U-233 designs assume that an external source of U-233 will be available as a replacement for U-235 as the makeup fuel.

Table C-13
U.S. URANIUM RESOURCE ESTIMATE

	Basis	Estimate
PRUDENT PLANNING BASE	Reserves Plus Probable	2.5 million tons U_3O_8
FULL DOE ESTIMATE	Reserves, Probable, Possible and Speculative Ores	4.5 million tons U_3O_8

This U-233 fuel might be supplied from any one of the following sources:

- spent PBR fuel
- LMFBRs
- fusion hybrids, or
- accelerator breeders

One of these sources must be developed for U-233 to become available as a makeup fuel for PBRs.

Thorium

It is important to recognize that the nation and the world have abundant resources of both fertile thorium and U-238. It is the fissile fuels that are scarce. The availability of thorium is thus not seen to be a problem.

Fuel Cycle Service Requirements

Fuel cycles services are those manufacturing processes by which raw fuel such as natural uranium (U_3O_8) or uranium-233 (U-233) are converted to Pebble Bed Reactor (PBR) fuel and later recovered or disposed. For the PBR these manufacturing services are conversion, enrichment, fabrication, reprocessing, waste disposal and transportation. Each of these manufacturing processes are not limited by resource availability as are uranium-235 and uranium-233.

The usage and type of fuel process selected is a compromise between high resource utilization (fuel conservation), fuel cycle cost, technical advancement and political acceptability of fuel reprocessing and high enriched fuels. For example, removing and reprocessing the fuel from a PBR more often will reduce uranium fuel requirements but at the expense of increased fuel service requirements.

The fuel service requirements for the PBR fuel cycles are shown in Tables C-14 through C-19. As shown in these tables, the high burnup fuel designs, such as the Once-Through MEU and HEU fuels, require the least fuel cycle services while fuel designs with lower burnups but higher resource utilization, such as the Recycle HEU and U-233 fuel designs, require more fuel services.

C.5 FUEL CYCLE COSTS

Economic Ground Rules

In comparing the cost of heat from alternate heat sources, it is important to look at the time discounted costs for fuel over some operating period. Ideally this operating period should be the life of the unit. For this analysis, the operating period over which costs are discounted will be thirty (30) years.

Table C-14
CONVERSION REQUIREMENTS — MT HM
(1000 MW_{th}, 80% Capacity Factor)

	THTR Series			Advanced Fuel		
Fuel Cycle	**1st Core**	**Reload**	**30 Year**	**1st Core**	**Reload**	**30 Year**
Once-Through MEU	190	46	1520	130	45	1440
Once-Through HEU	180	42	1400	150	41	1340
Recycle HEU	58	16	530	146	11	460
Recycle U-233	0	0	0	0	0	0

Table C-15
FABRICATION REQUIREMENTS — MT HM
1000 MW_{th}, 80% Capacity Factor)

	THTR Series			Advanced Fuel		
Fuel Cycle	**1st Core**	**Reload**	**30 Year**	**1st Core**	**Reload**	**30 Year**
Once-Through MEU	12	2.9	96	9	2.9	93
Once-Through HEU	12	2.9	96	11	2.9	95
Recycle HEU	11.9	9.3	282	21.1	9.2	288
Recycle U-233	14.3	8.4	260	18.9	8.4	260

Table C-16
ENRICHMENT REQUIREMENTS

	Enrichment Requirements — MT					
	THTR Series			Advanced Fuel		
Fuel Cycle	**1st Core**	**Reload**	**30 Year**	**1st Core**	**Reload**	**30 Year**
Once-Through MEU	230	54	1800	160	53	1700
Once-Through HEU	230	54	1800	200	53	1740
Recycle HEU	90	21	700	190	14	595
Recycle U-233	0	0	0	0	0	0

Table C-17
REPROCESSING REQUIREMENTS — MT HM
(1000 MW_{th}, 80% Capacity Factor)

Fuel Cycle	THTR Series 1st Core	Reload	30 Year	Advanced Fuel 1st Core	Reload	30 Year
Once-Through MEU	NA	NA	NA	NA	NA	NA
Once-Through HEU	NA	NA	NA	NA	NA	NA
Recycle HEU	11.9	9.3	282	21.1	9.2	288
Recycle U-233	14.3	8.4	260	18.9	8.4	260

NA: Not applicable, no reprocessing for this fuel cycle.

Table C-18
WASTE DISPOSAL REQUIREMENTS — MT HM*
(1000 MW_{th}, 80% Capacity Factor)

Fuel Cycle	THTR Series 1st Core	Reload	30 Year	Advanced Fuel 1st Core	Reload	30 Year
Once-Through MEU	12	2.9	96	9	2.9	93
Once-Through HEU	12	2.9	96	11	2.7	95
Recycle HEU	11	6.5	200	20	6.5	210
Recycle U-233	14	8.2	250	19	8.2	260

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*Carbon fixation not included.

Table C-19
EQUILIBRIUM TRANSPORTATION REQUIREMENTS — MT HM/Yr

Fuel Cycle	THTR Series Fresh Fuel	Spent Fuel	Waste	Advanced Fuel Fresh Fuel	Spent Fuel	Waste
Once-Through MEU	2.9	—	2.9	2.9	—	2.9
Once-Through HEU	2.9	—	2.9	2.9	—	2.9
Recycle HEU	9.3	9.3	6.5	9.2	9.2	6.5
Recycle U-233	8.4	8.4	8.2	8.4	8.4	8.2

These discounted costs should also include the interest charges on fuel inventories. For this analysis, the financial assumptions used to calculate fuel cycle costs are those typical of a U.S. investor owned utility operating in an environment of a constant 6% inflation, as described below:

Table C-20
FINANCIAL ASSUMPTIONS

		Constant $ Assumptions %	Reference Assumptions %
1.	Rate of Inflation	0	6
2.	Cost of debt money for zero inflation	2.75	2.75
3.	Cost of debt money with inflation (1 × 2), e.g., [(1.06 × 1.0275) − 1] × (100) = 8.92%	2.75	8.92
4.	Cost of equity money for zero inflation	5.5	5.5
5.	Cost of equity money with inflation (1 × 4), e.g., [(1.06 × 1.055) − 1] × (100) = 11.83%	5.50	11.83
6.	Assumed debt/equity ratio for utility industry capitalization	55/45	55/45
7.	Assumed federal plus state tax as percent of total earnings	50	50
8.	Return on debt (cost for use of debt portion of money), (3 × 6); e.g., (0.55 × 8.92) = 4.91%	1.52	4.91
9.	(a) Return on equity (cost for use of equity portion of money), (5 × 6) e.g., (0.45 × 11.83%) = 5.32%, and; (b) Federal tax plus state tax (for 50% total tax on earnings)	2.48	5.32
10.	(a) Weighted-average interest rate (net-of-taxes), (8 + 9)	4.00	10.24
	(b) Discount rate for utility decision making process (net of taxes), (8 + 9)	4.00	10.24
11.	Level annual revenue requirement excluding depreciation (effective interest rate plus taxes) (10a + 9b)	6.48	15.56

An example of how to calculate a fuel cycle cost for fabrication is shown in Table C-21 and more fully in C-3.

Table C-21
EXAMPLE OF A PBR FUEL COST CALCULATION FOR FABRICATION SERVICES

Assumptions

- \$420/kg HM Fabrication Cost
- Payment for fabrication service is one year before fuel use
- Fuel lifetime of 4 years (1170 full power days, 80% capacity factor)
- Fuel burnup of 100,000 kWd_o/kg
- 10.24% discount rate
- 15.56% interest rate
- 6% inflation

Fuel Cycle Cost Calculation

- Service Cost

$$420 \frac{\$}{\text{kgHM}} \times \frac{1}{100{,}000} \frac{\text{kg}}{\text{kWd}_o} \times \frac{d_o}{24\text{ hr}} \times \frac{\text{Kwhr}}{0.003413\text{ MBtu}} = 5.1¢/\text{MBtu}$$

- Value at time of fuel loading (with one year of inflation and interest)

$$5.1 \frac{¢}{\text{MBtu}} \times \frac{1.1556}{1.06} = 5.6\ ¢/\text{MBtu (in first year of use dollars)}$$

- Depreciation and Inventory Costs (¢/MBtu)

Year	Book Value at Beginning of Year	Depreciation at End of Year	Interest at End of Year	Total Annual Cost
0	5.6	0.4	0.871	2.271
1	4.2	1.4	0.653	2.053
2	2.8	1.4	0.436	1.836
3	1.4	1.4	0.218	1.618
		1/4 × Initial Book Value	15.56 × Book Value	

- Fuel Cycle Cost (Levelized by Amount of Heat Produced)
  Cost = 6.2 ¢/MBtu

Fuel Service Costs

The annual operating cost of a Pebble Bed Reactor is made up of six components which occur during the fuel cycle at different times. These components are:

- Conversion
- Enrichment
- Fabrication
- Reprocessing
- Waste Transportation
- Waste Disposal

For this study, the "Draft NASAP Provisional Data Base"(C-4) has been the basis for all fuel service cost estimates other than for fabrication. The fabrication cost estimates for PBR fuel are those of the fuel design reports from which the design of this report have been extracted.(C-1, C-2) To convert the above cost estimates to 1979 $ from 1978 and 1977 dollar estimates, an escalation rate of 10% per year has been used. Table C-22 summarizes the fuel service costs used in this report.

Fissile and Fertile Fuel Costs

The fuel materials for the Pebble Bed Reactor are predominantly uranium-235 and uranium-233 with thorium added to allow breeding of additional uranium-233. The uranium-235 and thorium fuel

Table C-22
SUMMARY OF FUEL SERVICE COSTS
(1979 $)

Fuel Services	Price
Conversion	$4.4/kg U
Enrichment	$110/SWU
Fabrication	Per Design Reference
Reprocessing*	Cost ($/kg) = 0.371* (c/HM) + 256 [Semi-remote AGNS Type Facility]
Spent Fuel Shipping*	Cost ($/kg) = 0.567* (c/HM) + 4.8
Waste Shipping	$33/kg HM
Waste Shipping	$66/kg HM
Permanent Spent Fuel Storage	$260/kg HM

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*c/HM = carbon to heavy metal ratio.

Table C-23
SUMMARY OF FUEL COSTS
(1979 $)

Fuel	Price
Natural Uranium	$75/lb U_3O_8
Thorium	$42/kg HM
Uranium-233	Indifference value to Recycle HEU fuel design ($40/gm fissile for $75/lb U_3O_8)

materials are obtained by mining domestic resources while the uranium-233 must be produced in the Pebble Bed Reactor or by another device. The prices assumed in this report for these fuel materials are shown in Table C-23 and described below.

Uranium-235 Prices

How uranium prices will rise as U.S. and worldwide resources are depleted, is as difficult to predict as is the exent of the resources themselves. Today, spot market prices are between $40 and $50/lb U_3O_8 (0.7% U-235, 99.3% U-238). For the future, the success of explorations, the cost of developing and operating new mines and mills, and other factors such as environmental protection will all affect the price of uranium.

For this assessment, it is assumed that uranium prices will rise as higher grade resources are depleted and lower grade resources are developed. Prices are additionally assumed to rise linearly from $40/lb U_3O_8 (1979 $) today to $110/lb U_3O_8 (1979 $) when the resource base is completely depleted, as is shown in Table 10.5.3-2. In the year 2000 time period, about 1.1 million tons of U_3O_8 should be mined for electric utility purposes, implying a uranium price of about $75 per pound U_3O_8 based on prudent planning assumptions.

The above resource estimates implicitly assumes that there is a fixed uranium resource base and once it has been mined that lower and lower grade (and higher) cost resources will not become productive. For uranium this appears to be the case. In the west, high grade uranium deposits were formed by river water precipitating through carbonacious fields leaving behind previously dissolved uranium. This process does not appear to produce large low or medium grade deposits which are necessary for uranium prices to

Table C-24
U.S. URANIUM PRICE ESTIMATES
($/lb U_3O_8 — 1979 $)

Cumulative Use (Millions of Tons)	Price — $/lb Prudent Planning Base	Price — $/lb Full DOE Estimate	Price ($/lb)	Cumulative Demand (Millions of Tons) Prudent Planning Base	Cumulative Demand (Millions of Tons) Full DOE Estimate
0	40	40	40	0	0
0.5	54	48	50	0.36	0.64
1.0	68	56	60	0.71	1.29
1.5	82	63	70	1.07	1.93
2.0	96	71	80	1.43	2.57
2.5	110	79	90	1.79	3.21
3.0	—	87	100	2.14	3.86
3.5	-	94	110	2.50	4.50
4.0	—	102			
4.5	—	110			
5.0	—	—			

continue to be a function solely of ore grade. And eastern ores, such as the Tennessee shales* and the Conway granites**, appear to have too low an ore grade to be environmentally mined.

Uranium-233 Prices

Uranium-233 for PBRs must be manufactured and is not available as a minable element. The principle means of making U-233 is in fission reactors such as in the PBR. Other means such as using fusion-hybrid reactors, accelerator, or LMFBR's may eventually be used but are not available today.

For this report, the method of costing (and crediting) U-233 will be the classical indifference approach, where the indifference values are determined by equating the cost of heat from U-235 fuel balls to the cost of heat from U-233 balls in "Recycle HEU" fuel designs.

Thorium Prices

The requirements for thorium in a Pebble Bed Reactor are small in comparison to expected but yet delineate domestic resources. For this study the price of thorium will be based on the "Draft NASAP Provisional Data Base"(C-4) The price of thorium is assumed to remain constant at $42/kg HM (1979 $).

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*The Tennessee — or Chattanooga — shales occupy a large part of eastern Tennessee, bedded in three layers beneath 150 feet of limestone overburden. Only the upper two layers are normally considered recoverable. The top layer is about seven feet thick and contains about 70 ppm U_3O_8 while the next layer is approximately eight feet thick and contains uranium in concentrations 25 to 60 ppm. The top layer is estimated to contain nearly 5 million tons of U_3O_8, while the bottom may contain up to 8 million tons.

**The Conway granites contain 20 to 40 ppm of uranium, and are thought to aggregate 25 million tons of U_3O_8, plus by product thorium, at costs greater than $220 per pound, or more, plus environmental effects comparable to or worse than that involved in mining the Chattanooga shales.

Nuclear Heat Costs

Each of the fuel designs of section 10.1 were developed to meet different fuel cost, uranium conservation and fuel recycle objectives. The objective of the Once-Through fuel designs was principally to reduce fuel cycle costs without fuel recycle, while the objective of the Recycle fuel designs was to reduce uranium usage through fuel recycle. Thus the Once-Through and Recycle fuel designs are usable to show the flexibility of the PBR to changing uranium availability and costs, and less of the difference between a Once-Through and a Recycle fuel design. Tables C-25 through C-28 show the nuclear heat cost results for these objectives.

The Once-Through designs are characterized by relatively high charges for uranium of from 50 to 55% of the total heat cost (for THTR fuels), Tables C-25 and C-26. These high uranium charges are balanced by low fuel service charges, combining to give a low fuel cost even for moderately priced uranium ($75/lb U_3O_8).

Table C-25
ONE-CYCLE FUEL CYCLE COSTS
FOR
ONCE-THROUGH MEU FUELS — ¢/MBtu
(1979 $)

	Unit Cost (1979 $)	Fuel Cost — ¢/MBtu	
		THTR Series	Advanced Fuel
Uranium	75 $/lb U_3O_8 42 $/kg Th 4.4 $/kgu conversion	59.0	53.2
Enrichment	110 $/SWU	38.1	34.3
Fabrication	420 and 520 $/kg respectively	7.8	9.0
Reprocessing	—	—	—
Spent Fuel Transportation	190 and 250 $/kg respectively	2.0	2.9
Waste Disposal	260 $/kg	2.8	2.9
Bred Fuel Credits	—	—	—
TOTAL		110	102

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*Cost includes indirect interest charges. (Discount rate = 10.24%; Interest rate = 15.56%).

Table C-26
ONE-CYCLE FUEL CYCLE COSTS FOR ONCE-THROUGH HEU FUELS — ¢/MBtu
(1979 $)

	Unit Cost (1979 $)	Fuel Cost — ¢/MBtu	
		THTR Series	Advanced Fuel
Uranium	85 $/lb U_3O_8 42 $/kg Th 4.4 $/kgU conversion	53.9	51.6
Enrichment	110 $/SWU	38.1	36.4
Fabrication	420-310 $/kg	7.8	5.6
Reprocessing	—	—	
Spent Fuel Transportation	190 and 110 $/kg respectively	2.0	1.2
Waste Transportation and Disposal	260 $/kg	2.8	2.8
Bred Fuel Credits	—	—	—
TOTAL		105	98

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*Cost includes indirect interest charges. (Discount rate = 10.24%; Interest rate = 15.56%.)

Of the two Once-Through fuel designs, the HEU design showed lower fuel costs than the non-proliferation MEU fuel design. However, the cost penalty of using the MEU fuel designs is modest (an increase in costs of about 4%).

The Recycle fuel designs have lower uranium charges, ranging from 30 to 35% of the fuel cost, than do the Once-Through designs. These low uranium charges are balanced by higher fuel service charges due to lower fuel exposures available to distribute fuel service costs, making the THTR Recycle HEU fuel design 20% more expensive than the THTR Once-Through HEU fuel design. However, at higher uranium prices this trend would reverse.

Of the two Recycle fuel design concepts, the Recycle U-233 fuel design provides the lowest fuel cycle cost. This low fuel cost however is subject to the cost and availability of U-233. No significant quantities of U-233 currently exist, thus it must be manufactured for this concept to be useful. The cost of producing U-233 is uncertain at this time.

Tables C-29 and C-30 provide a comparison of each of the four reference fuel cycles based on their one-cycle and thirty year levelized fuel costs.

Table C-27
ONE-CYCLE FUEL CYCLE COSTS FOR RECYCLE HEU FUELS — ¢/MBtu

	Unit Cost (1979 $)	Fuel Cost — ¢/MBtu	
		THTR Series	Advanced Fuel
Uranium	50-100 $/lb U_3O_8 42 $/kg Th 4.4 $/kgU conversion	26.4	17.2
Enrichment	110 $/SWU	16.9	10.7
Fabrication	600 and 440 $/kg respectively	39.1	22.8
Reprocessing	380 and 320 $/kg respectively	18.5	11.3
Spent Fuel Transportation	190 and 110 $/kg respectively	9.3	3.9
Waste Transportation and Disposal	100 $/kg	4.9	3.5
Bred Fuel Credits	40 $/gm	13.1	24.9
		128	94

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*Cost includes indirect interest charges. (Discount rate = 10.24%; Interest rate = 15.56%.)

Table C-28
ONE-CYCLE FUEL CYCLE COSTS FOR RECYCLE U-233 FUELS — ¢/MBtu
(1979 $)

	Unit Cost (1979 $)	Fuel Cost* — ¢/MBtu	
		THTR Series	Advanced Fuel
Thorium	42 $/lb Th	1.7	1.8
Fabrication	530 and 440 $/kg respectively	24.1	20.8
Reprocessing	345 and 320 $/kg respectively	11.3	10.4
Spent Fuel Transportation	100 $/kg	4.6	3.6
Waste Transportation and Disposal	100 $/kg	3.3	3.2
U-233 Costs*	$40/gm	22.2	22.2
TOTAL		67	62

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*Cost includes indirect interest charges. (Discount rate = 10.24%; Interest rate = 15.56%.)

Table C-29
SUMMARY OF ONE-CYCLE NUCLEAR HEAT COSTS — ¢/MBtu
(1979 $)

	Nuclear Heat Cost — ¢/MBtu	
Fuel Cycle	THTR Series	Advanced Fuels
Once-Through MEU	110	102
Once-Through HEU	105	98
Recycle HEU	128	94
Recycle U-233	67	62

Table C-30
SUMMARY OF 30-YEAR LEVELIZED NUCLEAR HEAT COSTS — ¢/MBtu
(1979 $, 6% per year escalation)

	Nuclear Heat Cost — ¢/MBtu	
Fuel Cycle	**THTR Series**	**Advanced Fuels**
Once-Through MEU	210	190
Once-Through HEU	200	180
Recycle HEU	240	180
Recycle U-233	130	120

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*Conversion factor between one cycle cost and 30 year costs with inflation = 1.87

C.6 RESEARCH AND DEVELOPMENT

Reactor Research and Development

The research and development of gas reactors has proceeded in several countries, such as in the United States, Federal Republic of Germany, England, France and Switzerland, with each nation investigating slightly different design variations.(C-5) The Pebble Bed Reactor used in this report has been developed principally in the Federal Republic of Germany.

In Germany the Pebble Bed gas reactor development program led to the construction and operating in 1967 of the AVR reactor. This first Pebble Bed Reactor demonstrated the use of spherical graphite fuel elements and an on-line refueling system. With continued successful operation of AVR, a second Pebble Bed Reactor was commissioned and construction began in 1971. This second reactor Thorium High Temperature Reactor (THTR), forms the basis for which the reactor characteristics of this report have been founded.

There are three reactor research and development issues of important in importing the FRG Pebble Bed Reactor for use in the United States. First, there will be continuing R&D necessary to improve the near-commercial FRG technology must be converted to a process heat technology. And, third, the FRG technology must adapt to U.S. licensing and safety practices. Each of these issues however, do not appear to prohibit the use of PBR's in the U.S.(C-5, C-6)

Fuel Research and Development

The fuel research and development program in the Federal Republic of Germany is presently focusing on the HHT/PNP-target fuel elements with three variants of the THTR fuel element being assessed. The three variants differ from the THTR fuel element only in the particle coatings and include both the BISO and TRISO particle types. The near-term objectives of the fuel development program include the specification of methods for analyzing the SIC coatings for the TRISO elements, the reduction and removal of waste coming from the particle manufacturer, the continuation of failure fraction — irradiation tests, and the specification of modes for the analysis of all reference particle variants.

Testing of fuel elements is also continuing in the AVR. During the past ten years fuel elements of up to 20 g of heavy metal per ball have been irradiated in this facility. These tests have indicated the feasibility of going to high heavy metal loadings. However, the tests for the 20 g/ball elements were not entirely satisfactory to date, and these tests will be continued.

REFERENCES FOR APPENDIX C

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C-4. "Revised NASAP Provisional Data Base", February 6, 1979.

C-5. NUS Corporation. "Gas Reactor International Cooperative Program Interim Report, 'Construction and Operating Experience of Selected European Gas — Cooled Reactors' ", COO-4057-8. U.S. Department of Energy, September, 1978.

C-6. NUS Corporation. "Gas Reactor International Cooperative Program Interim Report, 'Safety and Licensing Evaluation of German Pebble Bed Reactor Concepts' ", COO-4057-7. U.S. Department of Energy. September, 1978.

APPENDIX D

APPLICATION OF HTR HEAT SOURCE TO H-COAL LIQUID REFINERY FOR MOTOR FUEL PRODUCTION

This document presents the results of an evaluation of the application of a high temperature gas-cooled nuclear reactor (HTR) as a heat source for a coal liquid refinery. These results are very preliminary and are to be used as a guideline for study purposes only.

A review of the two intermediate heat transfer schemes presented by GE indicates that both are probably feasible, but would require a significant amount of process engineering to determine what metallurgical problems might be encountered and how they might be resolved. Also, the high helium temperatures available from the HTR would cause excessive temperature gradients in most of the H-Coal liquid upgrading process applications, as all the refinery processes, except for hydrogen production, and catalytic reforming, are operated at about 400°C or below. To avoid possible thermal shock or coke deposition in process heat exchangers most of the low temperature process heat utilized in the coal liquid refinery will first have to be converted to an intermediate heat source (steam or hot oil) at temperatures and pressure not too far in excess of those for process streams. The catalytic reformer and hydrogen production facility process heat can probably be derived directly from the hot helium gas being used much as a flue gas.

By the proper positioning of steam generation and superheating sections as well as consideration of allowable approach temperatures for the various steam sections, almost all of the coal liquid refinery steam requirements can be met by usage of the sensible heat available in the high temperature helium. Refinery power can also be generated when de-superheating or lowering the pressure of high pressure, superheated steam.

Specific comments are provided below:

1. A schematic of the coal liquid refinery scheme selected as the most economically attractive H-Coal liquid processing alternative in our study for the DOE shown in Figure D-1. The sensible heat available from the hot helium might be used directly for process heating of both the catalytic reformer and hydrogen plant. High pressure, superheated steam would be utilized for the high pressure hydrotreating sections, and low/medium pressure steam would be used for fractionation and in the low temperature first stage of the coal liquid naphtha hydrotreater.

   Although only a steam reformer was used in our study for the DOE, a partial oxidation plant would be incorporated in the HTR scheme as shown in the attached schematic. The partial oxidation plant shown would utilize a combination of hot helium as well as steam. Because of the heat received from the HTR all of the heavy distillate (see Table D-1 — refinery material balances) would now have to be diverted from refinery fuel to the partial oxidation plant for hydrogen production, for it is of too poor quality to be sold directly as a product. With the hydrogen production gained from the partial oxidation plant, only a small portion of the total refinery hydrogen requirements would have to be met by the steam reformer. The hydrogen produced in the partial oxidation unit will replace that produced first from C4-C6, and second from fuel gas in the steam reformer.

2. The estimated total amount of heat which would be recovered from the HTR for use in the coal liquid refinery is about $9{,}500 \times 10^6$ KCAL per day. Thirteen percent of this amount reflects the addition of power generation facilities in the refinery offsites. In our previously mentioned DOE study, the purchase of electric power had been assumed, but working on the assumption that the maximum utilization of heat available from the HTR is desired, the addition of power generating facilities seems a realistic option. In this evaluation, no allowance has been made for variations in heat transfer efficiencies which might be expected due to the differences in the heating mediums involved.

3. As mentioned previously, since the low quality heavy distillate would now be the primary source of hydrogen, the quantity of hydrogen produced from fuel gas would be relatively small (about 12% of the total). In terms of most effective utilization of the hot circulating helium as a heat source, it may be desirable to include both the steam reformer and partial oxidation hydrogen production facilities within the reactor containment boundary.

   The amounts of feed material and product hydrogen expected for the two hydrogen sources is:

	MT/D Hydrogen (metric tons/day)	
	Partial Oxidation	**Steam Reformer**
Fuel Gas	—	42
Heavy Distillate (Feedstock to Produce H_2)	623	—
Hydrogen Produced	138 (57,000,000) (SCF/D)	18 (7,500,000) (SCF/D)

4. The nominal projected coal liquid refinery on stream time is 330 days per year.
5. The original estimated capital costs for the steam reformer, and offsites facilities for steam and power generation which are affected by the high temperature plant are:

Existing Refinery
(Base Case)

Units Being Replaced	**Estimated Erected Cost**	
Steam Reformer	\$28,000,000	(156 MT/D H_2)
Steam Plant Power Generation	\$40,000,000	

Installation of comparable facilities within the HTR containment area may be more expensive because of possible exotic metallurgical requirements, and because of the need for both a steam reformer and a partial oxidation unit. Incorporation of high temperature helium as a heat source may result in engineering considerations outside the realm of normal practice.

Aside from cost variance due to required design modifications, the cost of steam and power generation facilities within the context of a HTR heat source cannot be compared directly because in the HTR scheme all of the required refinery power would be internally generated. In our previous study some power was generated in the offsites in pressure reduction turbines, but most of the refinery power requirement was purchased.

In the HTR refinery scheme the estimated cost of the revised steam reformer and the added partial oxidation unit — without allowance for engineering modifications is:

Refinery with HTR Heat Source (New Case)

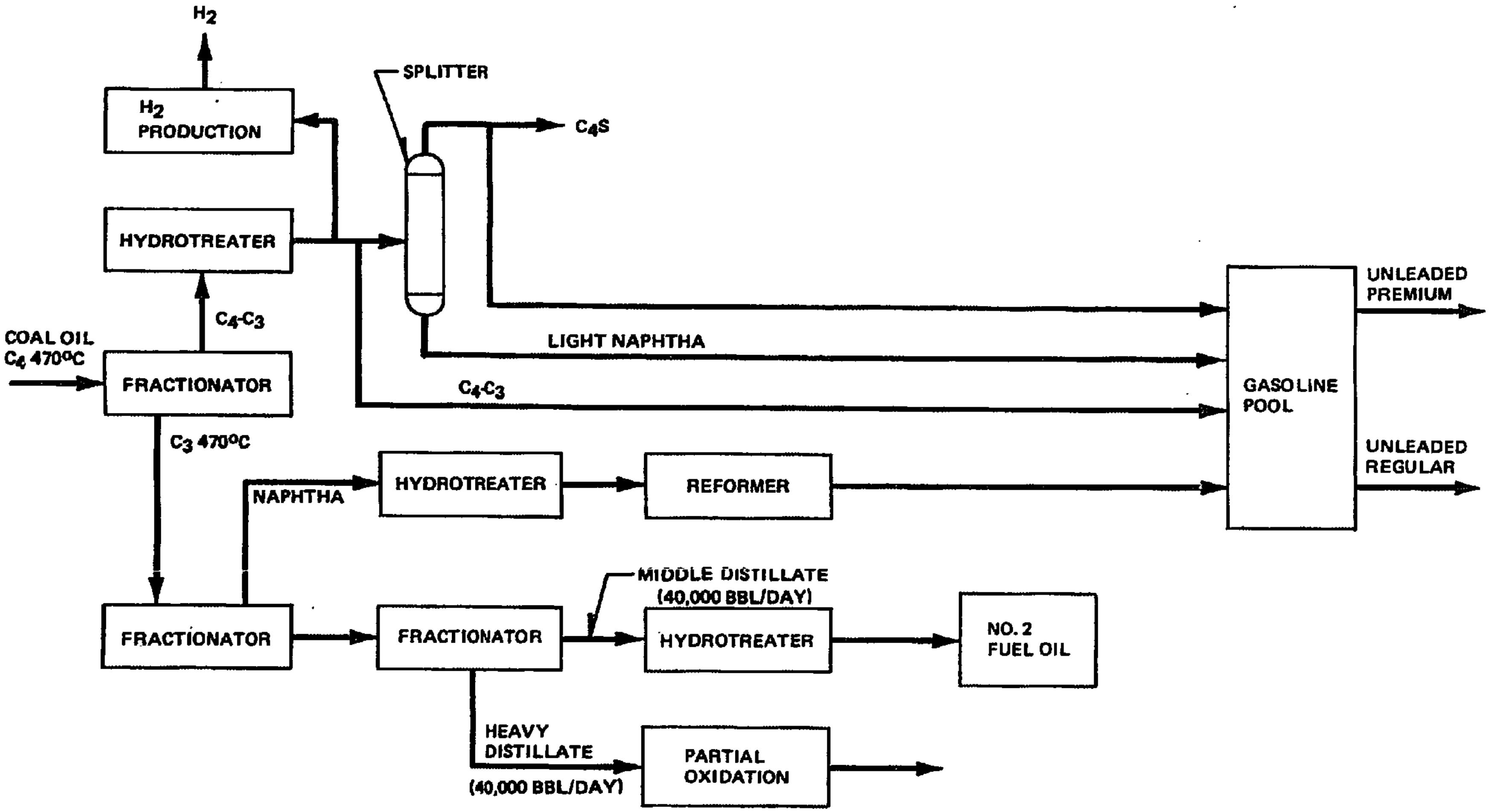
Unit	**Estimated Erected Cost**	
Steam Reformer	\$5,000,000	(18 MT/D H_2)
Partial Oxidation	\$50,000,000	(130 MT/D H_2)

6. The reduction in refinery feed material (H-Coal liquid) for nuclear-heat source compared with fossil heat is 3220 BPCD and is provided in Table D-2. This comparison is based upon producing a fixed quantity of gasoline and No. 2 fuel oil.
7. The change in refinery product and waste stream for nuclear heat source compared with fossil heat at a fixed refinery charge rate are presented in Table D-3. The results in this table can be compared directly with those of the base case in Table D-1. Included in the three tables are refinery chargestock and product material balances, nature and quantity of refinery streams converted to utilities, as well as process waste streams. Waste stream handling is not anticipated to be a problem in the coal liquid refinery. Since the coal liquid is low in sulfur, SO_x flue gas emissions are no problem even if the heavy coal liquid distillate is used as a fuel oil. Treatment for the recovery of NH_3, Ash, and small quantities of acidic or basic nitrogenous compounds would be essentially unchanged. The waste gas streams listed in the above tables are primarily CO_2.

It must again be cautioned that a large scale engineering involvement would be required to properly define all the revised design parameters implied by this study. Some of the areas to be considered are:

1. Minimization of long pipe runs where movement of high temperature material is involved.
2. Maintenance of acceptable pressure differentials across and shell and tube heat exchangers.
3. Proper consideration of approach temperatures and temperatures gradients in any heat exchangers.
4. Full examination of metallurgical requirements.
5. Consideration of possible problems associated with catastrophic failure in any integrated refinery/HTR equipment.
6. Effects of either refinery or HTR downtime and reliability upon the overall operating requirements of each component.

The properties of coal liquid and H-Coal distillate used for this study are shown in Tables D-4 and D-5, respectively.



FUEL FOR PROCESS HEAT = 7% OF FEEDSTOCK (942 MT/DAY)

FEED FOR H_2 PRODUCTION = 5% OF FEEDSTOCK (665 MT/DAY)

Figure D-1. SCHEMATIC OF COAL OIL REFINERY. HYDROTREATING CASES WITH LOW TEMPERATURE FRACTIONATION

Table D-1
FOSSIL-FIRE HEAT SOURCE — BASE CASE
UNIVERSAL OIL PRODUCTS CO. STANDARD OPTIMIZATION REPORTS, P51.0

DOE Task NO. 9 — **17:42 September 11, 1979**
Coal Oil Refinery Hydrotreating Case
Gasoline/Distillate = 1.0

Material Balance Summary

Charge	Sp. Gr.	BBL/D	LV%	M. Tons/D	Wt %	M. Tons/Yr
H-Coal Feedstock	.87330	100000.0	100.000	13868.73	100.000	5062086
•				.01	.000	5
Total Charge		100000.0	100.000	13868.74	100.000	5062091
Products	**Sp. Gr.**	**BBL/D**	**LV%**	**M. Tons/D**	**Wt %**	**M. Tons/Yr**
Liquified Pet. Gas	.53422	1859.2	1.859	157.73	1.137	57572
Unleaded Regular	.74872	32025.1	32.025	3807.88	27.457	1389875
Unleaded Premium	.78920	13744.7	13.745	1722.64	12.421	628764
No. 2 Fuel Oil	.92325	45815.2	45.815	6717.43	48.436	2451854
Sulfur				16.58	.120	6051
Total Products Sold		93444.2	93.444	12422.26	89.570	4534125
Streams Converted to Utilities	**Sp. Gr.**	**BBL/D**	**LV%**	**M. Tons/D**	**Wt %**	**M. Tons/Yr**
345-471C Cut	1.06550	3713.2	3.713	628.32	4.530	229336
H-Coal Feedstock	.87330	1243.8	1.244	172.50	1.244	62963
C6-471C H-Coal Cut	.91450	974.9	.975	141.59	1.021	51681
Total Streams Converted		5932.0	5.932	942.41	6.795	343980
Streams Not Utilized	**Sp. Gr.**	**BBL/D**	**LV%**	**M. Tons/D**	**Wt %**	**M. Tons/Yr**
Water				230.03	1.659	83960
Ammonia				47.40	.342	17301
Waste Gas				226.63	1.634	82721
•				.01	.000	5
Total Not Utilized				504.07	3.635	183985
Total Products Made		99376.2	99.376	13868.74	100.000	5062091

Table D-2
HTR NUCLEAR HEAT SOURCE REDUCED H-COAL FEED CASE
UNIVERSAL OIL PRODUCTS CO. STANDARD OPTIMIZATION REPORTS, P52.0

DOE Task No. 9 **10:03 September 20, 1979**
Coal Oil Refinery Hydrotreating Case
Gasoline/Distillate = 1.0

Material Balance Summary

Charge	Sp. Gr.	BBL/D	LV%	M. Tons/D	Wt %	M. Tons/Yr
H-Coal Feedstock	.87330	96779.2	100.000	13422.04	99.999	4899045
•				.10	.001	38
Total Charge		96779.2	100.000	13422.14	100.000	4899083
Products	**Sp. Gr.**	**BBL/D**	**LV%**	**M. Tons/D**	**WT %**	**M. Tons/Yr**
Liquified Pet. Gas	.55762	3438.9	3.553	304.53	2.269	111152
Unleaded Regular	.74627	32025.0	33.091	3795.41	28.277	1385326
Unleaded Premium	.78920	13744.6	14.202	1322.63	12.836	628761
No. 2 Fuel Oil	.92417	45815.0	47.340	6724.04	50.097	2454275
Sulfur				17.57	.131	6414
Total Products Sold		95023.5	98.186	12564.19	93.608	4585929
Streams Converted to Utilities	**Sp. Gr.**	**BBL/D**	**LV%**	**M. Tons/D**	**Wt %**	**M. Tons/Yr**
Refinery Fuel Gas				75.30	.561	27483
Total Streams Converted				75.30	.561	27483
Streams Not Utilized	**Sp. Gr.**	**BBL/D**	**LV%**	**M. Tons/D**	**Wt %**	**M. Tons/Yr**
Water				227.04	1.692	82870
Ammonia				46.34	.345	10913
Waste Gas				25.03	.191	9355
Partial OX Off-Gas				483.55	3.603	176495
•				.10	.001	38
Total Not Utilized				782.66	5.831	285671
Total Products Made		95023.5	98.186	13422.14	100.000	4899083

Table D-3
HTR NUCLEAR HEAT SOURCE INCREASED REFINERY OUTPUT CASE
UNIVERSAL OIL PRODUCTS CO. STANDARD OPTIMIZATION REPORTS, P51.0

COE Task No. 4 11:31 September 20, 1979
Coal Oil Refinery Hydrotreating Case
Gasoline/Distillate = 1.0

Material Balance Summary

Charge	Sp. Gr.	BBL/D	LV%	M. Tons/D	Wt %	M. Tons/Yr
H-Coal Feedstock	.87330	100000.0	100.000	13868.73	99.999	5062096
•				.11	.001	39
Total Charge		100000.0	100.000	13868.84	100.000	5062125
Products	**Sp. Gr.**	**BBL/D**	**LV%**	**M. Tons/D**	**Wt %**	**M. Tons/Yr**
Liquified Pet. Gas	.55762	3553.3	3.553	314.66	2.269	114851
Unleaded Regular	.74627	33090.8	33.091	3921.73	28.277	1431430
Unleaded Premium	.78920	14202.1	14.202	1779.96	12.834	649686
No. 2 Fuel Oil	.92417	47339.7	47.340	6947.82	50.097	2535954
Sulfur				18.16	.131	8621
Total Products Sold		98185.9	98.186	12982.33	93.608	4738550
Streams Converted to Utilities	**Sp. Gr.**	**BBL/D**	**LV%**	**M. Tons/D**	**WT %**	**M. Tons/Yr**
Refinery Fuel Gas				77.80	.561	28398
Total Streams Converted				77.80	.561	28398
Streams Not Utilized	**Sp. Gr.**	**BBL/D**	**LV%**	**M. Tons/D**	**Wt %**	**M. Tons/Yr**
Water				234.60	1.692	85628
Ammonia				47.88	.345	17416
Waste Gas				26.48	.191	9666
Partial OX Off-Gas				499.64	3.603	182368
•				.11	.001	39
Total Not Utilized				808.71	5.831	295178
Total Products Made		98185.9	98.186	13868.84	100.000	5062125

Table D-4
COAL LIQUID PROPERTIES

	C_4- 471°C	C_6- 471°C	C_4/C_6	C_6/ 177°C	C_6/ 193°C	C_6/ 204°C	177/ 204°C	177/ 193°C
Yield, vol. %	100.0	86.13	13.87	30.84	38.9	41.25	10.41	8.10
Yield, wt. %	100.0	90.2	9.8	28.39	37.5	40.14	11.75	9.11
API	30.5	23.2	97.9	44.5	36.8	35.0	12.2	11.1
Sp. Gr.	0.8733	0.9145	0.617	0.804	0.8408	0.8498	0.9844	0.9823
Oxygen, wt. %	1.72	1.97	—	0.52	2.7	3.0	8.6	8.2
Nitrogen, wt. %	0.37	0.40	—	0.052	0.29	0.30	0.9	1.0
Sulfur, wt. %	0.15	0.15	0.1	0.2	0.2	0.2	0.2	0.22
CCR, wt. %	0.10	0.10	—	—	—	—	—	—
Carbon, wt. %	86.7	87.1	83.5	85.5	85.1	84.9	83.5	83.7
Hydrogen, wt. %	11.0	10.5	16.5	13.1	11.9	11.6	7.9	8.1
Bromine Number	41.7	43.1	26	11.0	11.0	12.0	11.0	11.0
Ash, ppm	67	63.0	—	1	1	1	1	
C_7 Insol.	0.10	0.11	—	—	—	—	—	—

Table D-5
H-COAL DISTILLATE PROPERTIES

	204/343°C	343°C+	204°C+	204/260°C	204/288°C	260°C+	288°C+
Yield, vol. %	40.89	3.99	44.88	21.88	32.58	23.0	12.3
Yield, wt.%	45.19	4.87	50.06	23.55	35.66	26.51	14.4
API	15.1	1.3	13.6	19	16.5	9.1	6.9
Sp. Gr.	0.9652	1.0655	0.9742	0.9402	0.9561	1.0067	1.0225
Oxygen, wt.%	1.0	1.3	1.0	1.0	1.0	1.05	1.1
Nitrogen, wt.%	0.4	1.3	0.5	0.38	0.39	0.55	0.8
Sulfur, wt.%	0.1	0.2	0.1	0.1	0.1	0.12	0.14
CCR, wt.%	<0.1	2.0	0.2	<0.1	<0.1	0.4	0.7
Carbon, wt.%	88.8	89.2	88.8	87.9	88.4	89.0	89.1
Hydrogen, wt.%	9.7	8.0	9.6	10.1	9.8	8.7	8.3
Bromine Number	50		50		about the same		
Ni + V, wt. ppm			(small — other metals also present)				
Ash, ppm	10	1200	120	1	2	220	340
C, Insol.	0	2.0	0.2	0	0	0.4	0.7