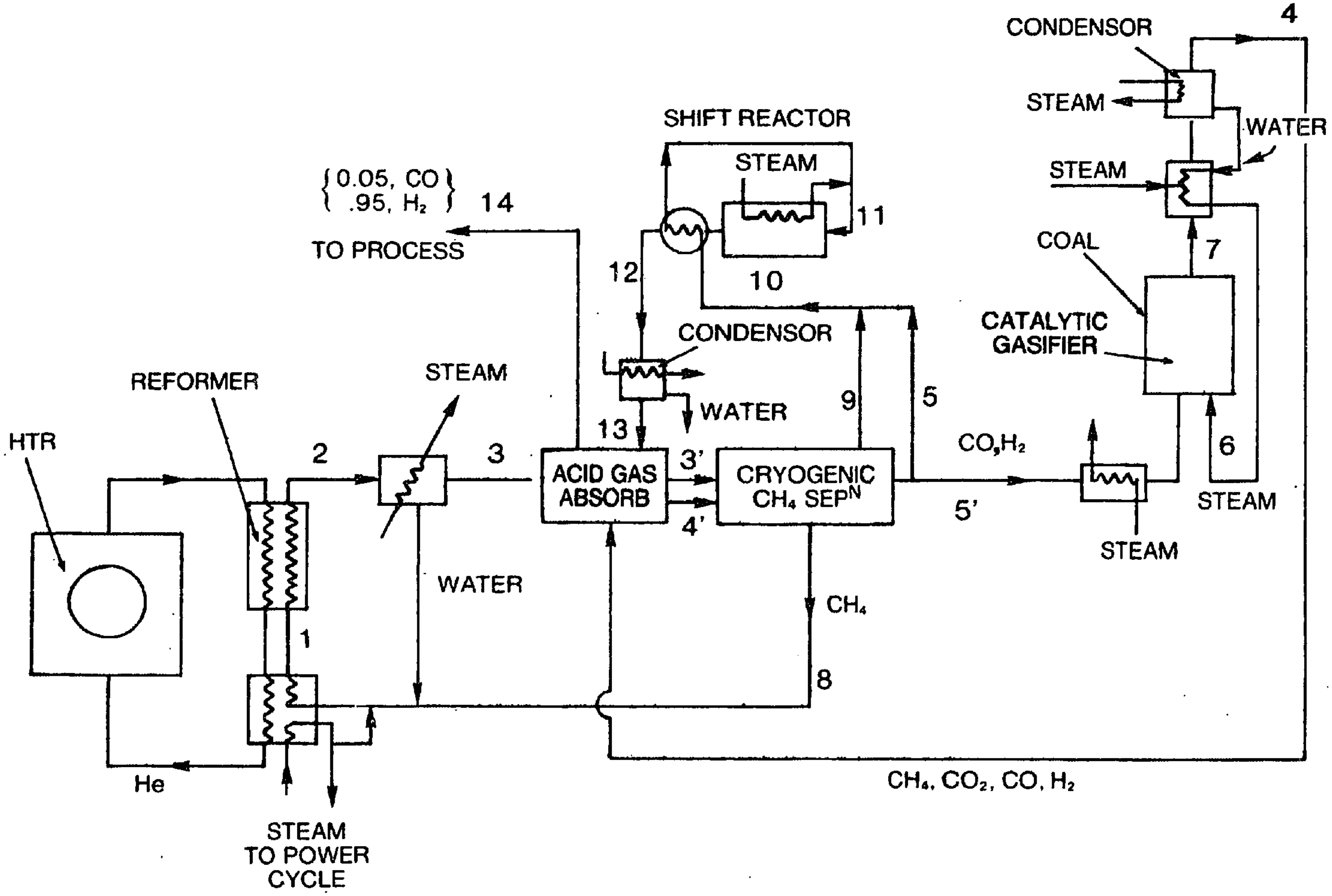

Figure 5-14. SCHEMATIC OF H_2 PRODUCTION FOR COAL LIQUEFACTION VIA HTR + COAL

Dry coal, CO and H_2 with some additional steam (to effect the reactions leading to methane) are injected into the gasifier (at pressures essentially equal to that of the nuclear heat source cooling loop). This produces methane, which after cleaning procedures, will be used to feed the nuclear reformer. The reforming products, CO and H_2, (stream 5) and the unconverted CO and H_2 in the effluent purified gasifier effluent (stream 9) are added together forming the "Syn gas" (stream 10). The syn gas is then conducted to a shift reactor in which the exothermic shift reaction

$$CO + H_2O \rightleftharpoons CO_2 + H_2$$

yields a H_2-rich product by addition of excess steam. Since the shift reaction is reversible it is necessary to supply sufficient steam to drive the equilibrium toward the H_2 product. Some of the reaction heat is removed by steam jacketing the shift reactor by low temperature steam from the power cycle. This heated steam is then used as the steam feed to the shift reaction. Excess steam is removed from the product CO_2/H_2 gases and the condensate can be used as part of the feed to the gasifier. The final CO and H_2 fractions are 0.05/0.95. The condenser cooling water can be interchanged with the power cycle as BFW, etc. CO_2 is removed by a conventional acid gas removal process yielding the desired hyrogen product. The flow streams of the various lines are listed in Tables 5-15 through 5-17, which correspond respectively to the combinations of reformer and gasifier temperatures exemplified in Tables 5-12 through 5-14 respectively.

5.3.2.3 SRC Processes

A flow process diagram for a 50,000* barrel/day plant using Illinois 6 coal is shown as Figure 5-15(5-11). Reference 11 gives flow rates of each component. Approximately, 1,640,050 lbs/hr of coal is feed into the system of which 130,820 lb/hr is feed to the power plant, 52,160 lb/hr to the fuel gas production unit and 228,500 lb/hr is consumed in the drying process. Net dry coal feed is 1,228,570 lb/hr to the process. In the SRC process, the reactor in which the actual liquefaction is carried out operates at approximately 1800 psi and at between 650-850°F. The hydrogenated coal liquid is separated into naphtha, fuel oil and solid

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*Based on fuel oil equivalent of $6.3x10^6$ Btu/bbl and 43,185 bbl/day residuum, 5383 bbl/day fuel oil and 1935 bbl/day naptha.

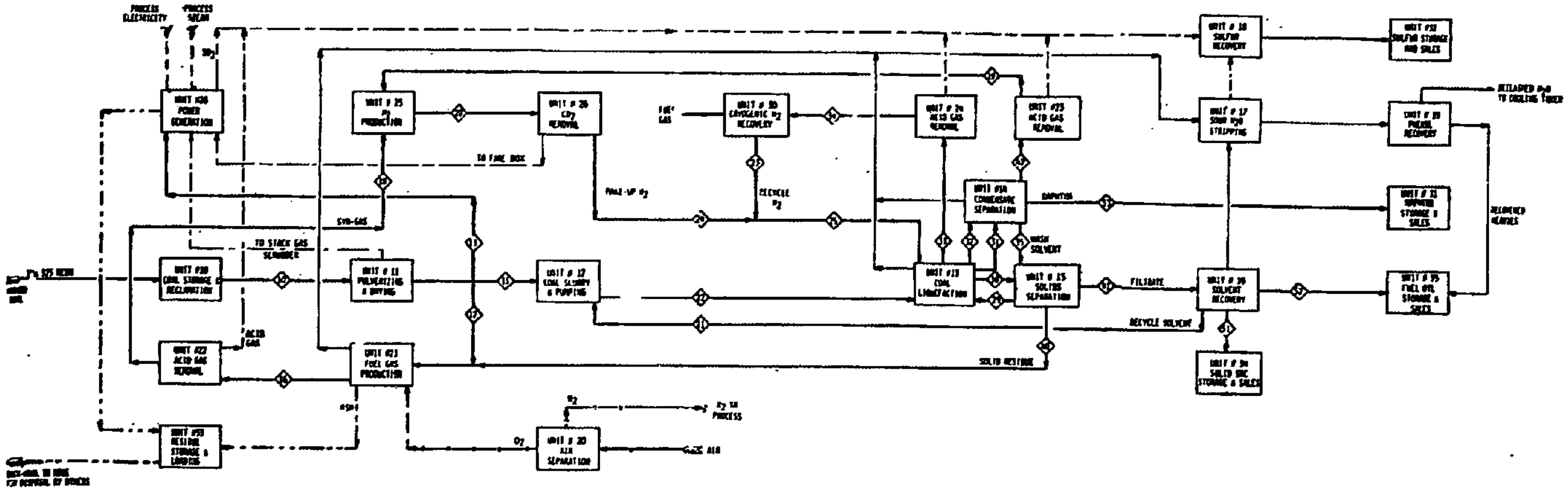

Figure 5-15. SRC SCHEMATIC (REF 11)

Table 5-15: Product Streams in ST moles/hr for 2000 ST/D Illinois #6 Coal Feed to Make 95/5 H_2/CO Syn Gas

Stream	5	9	10	11	12	13	14
CH_4	-	-	-	-	-	-	-
H_2O	-	-	-	3.360	0.919	0.312	-*
CO	1.187	1.784	2.971	2.971	0.531	0.531	0.531
CO_2	-	-	-	-	2.440	2.440	-
H_2	7.650	0.004	7.654	7.654	10.090	10.090	10.090
NPH	950°C						
Reformer	825°C						
Gasifier	700°C						

*Assumes additional drying step not shown in Figure 5-14.

Table 5-16: Product Streams in ST moles/hr
for 2000 ST/D Illinois #6 Coal
Feed to make 95/5 H_2/CO Syn Gas

Stream	5	9	10	11	12	13	14
CH_4	-	-	-	-	-		
H_2O	-	-	-	3.315	0.907	0.319	-*
CO	1.187	1.755	2.942	2.942	0.532	0.532	0.532
CO_2	-	-	-	-	2.408	2.408	-
H_2	7.635	0.068	7.703	7.703	10.107	10.107	10.107
NPH	950°C						
Reformer	825°C						
Gasifier	825°C						

*Assumes additional drying step not shown in Figure 5-14.

Table 5-17: Product Streams in ST moles/hr
for 2000 ST/D Illinois #6 Coal
Feed to make 95/5 H_2/CO Syn Gas

Stream	5	9	10	11	12	13	14
CH_4	-	-	-	-	-	-	-
H_2O	-	-	-	1.876	0.514	0.254	-*
CO	0.540	1.353	1.893	1.893	0.531	0.531	-
CO_2	-	-	-	-	1.361	1.361	-
H_2	8.719	0.013	8.732	8.732	10.094	10.094	10.094

NPH	825°C
Reformer	700°C
Gasifier	700°C

*Assumes additional drying step not shown in Figure 5-14.

SRC (in this case) products. An important step is the recovery of anthacene oil from the hydrogenated liquids and its recycle to dissolve fresh coal. It is very difficult to filter the residues, and these amount to 243,820 lb/hr in this case. The residues are subsequently gasified to hydrogen. The liquid air plant and fuel gas producer, which in combination, gasify the residues and additional coal. In this particular plant the power requirement for O_2 production is 22.4 MW_e and the fuel gas plant is 32 MW_e out of a total of 182 MW_e expended. Finally, the gross hydrogen (via synthesis gas) requirement for this plant is 8070.59 lb moles/hr.

5.3.2.4 Modified SRC Process

It is clear that HTR produced H_2 can be substituted for the syngas used in the "standard" SRC process described above. The modified liquefaction plant is shown in Figure 5-16 where certain key items of Figure 5-15 associated with H_2 manufacture have been deleted. These are:

(1) The oxygen plant

(2) The fuel gas production unit

(3) The acid gas clean-up for the above

(4) The much reduced H_2 production shift reactor dealing with light tops from the liquefaction reactor.

We must supply 8070.59 lb moles/hr of H_2 to the system. Coal feeds to the fuel gas unit and the power generation unit have been eliminated saving a total of 182,980 lb/hr of coal.

5.3.2.5 Comparison of SRC With and Without HTR

Given below is a comparison of the SRC process with and without the use of the HTR to supply thermal energy. The coal saved (including power reduction) in the HTR case amounts to approximately 10.3% of the coal feed to the standard plant.

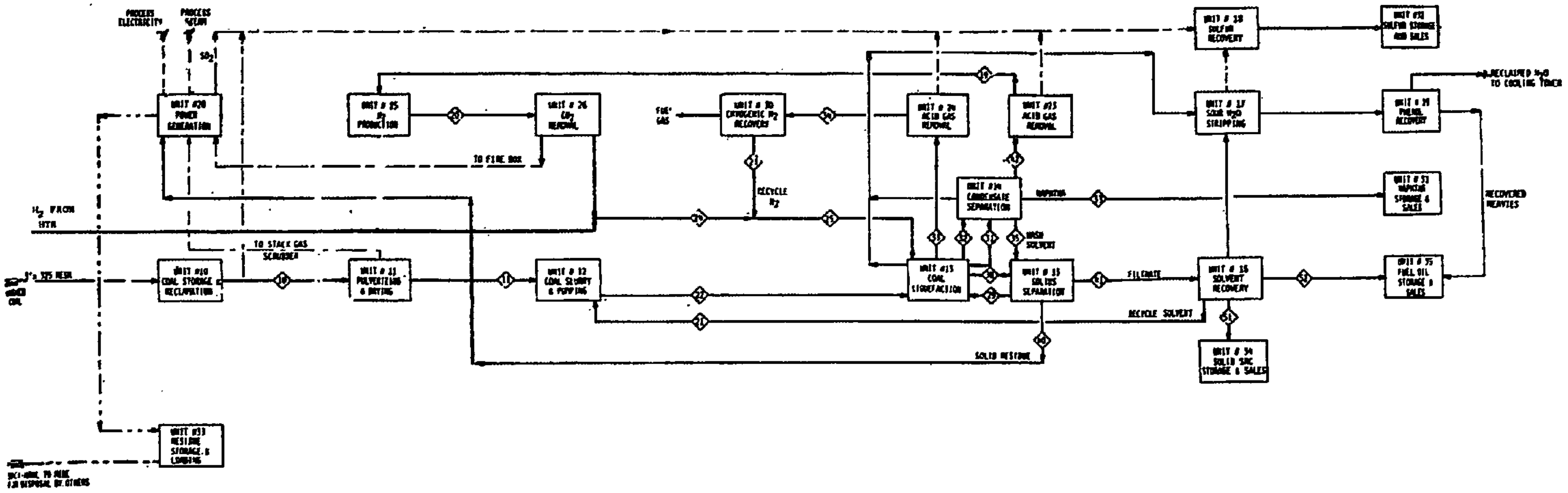

Figure 5-16. MODIFIED SRC SCHEMATIC

Basis: 50,000 bbl/day fuel oil equivalents (FOE)

	Standard	With HTR	Savings
(I) Credits			
As received coal: lb/hr	1,640,050	1,457,070	182,980
Power: MW_e	182	127.6	54.4
Power: MW_t*	-	-	163.2
(II) Debits			
H_2 from HTR (lb moles/hr)	-	8070.59	(8070.59)
As received coal: lb/hr for H_2 production	-	66,655	(66,655)
Nuclear heat/MW_t	-	86.2	(86.2)
(III) Net			
As received coal: lb/hr	-	-	116,325
Power: MW_t	-	-	77.0

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*$MW_t = 3.0 x MW_e$

5.3.2.6 H-Coal Process

Figure 5-17 shows a schematic flow diagram for the H-Coal Process(5-11) and is reproduced from the data presented there. The primary difference between the H-coal and the SRC processes is the use of a cobalt-molybdenum/zinc oxide catalyst in the liquefaction reactor itself which requires somewhat different operating conditions -- 2300 psi at 650-850°F. Because of the higher activity of the H-coal catalyst, there is a greater use of hydrogen and a greater yield of lighter products. The nominal 50,000 bbl/day products of this plant are 30,800 bbl/day residium, 10,788 bbl/day fuel oil, and 6578 bbl/day naphtha -i.e., approximately 2 times the fuel oil and 3 times naphtha production of the SRC case. The vital statistics of this case are:

(1) Gross as received coal feed: 1,689,350 lb/hr

(2) Coal to fuel gas and to power generation: 197,370 lb/hr

(3) Hydrogen equivalents to process, 13,281.31 lb mole/hr

(4) Power to oxygen plant: 28.3 MW_e

(5) Power to fuel gas plant: 51.2 MW_e

(6) Power to total plant: 238 MW_e

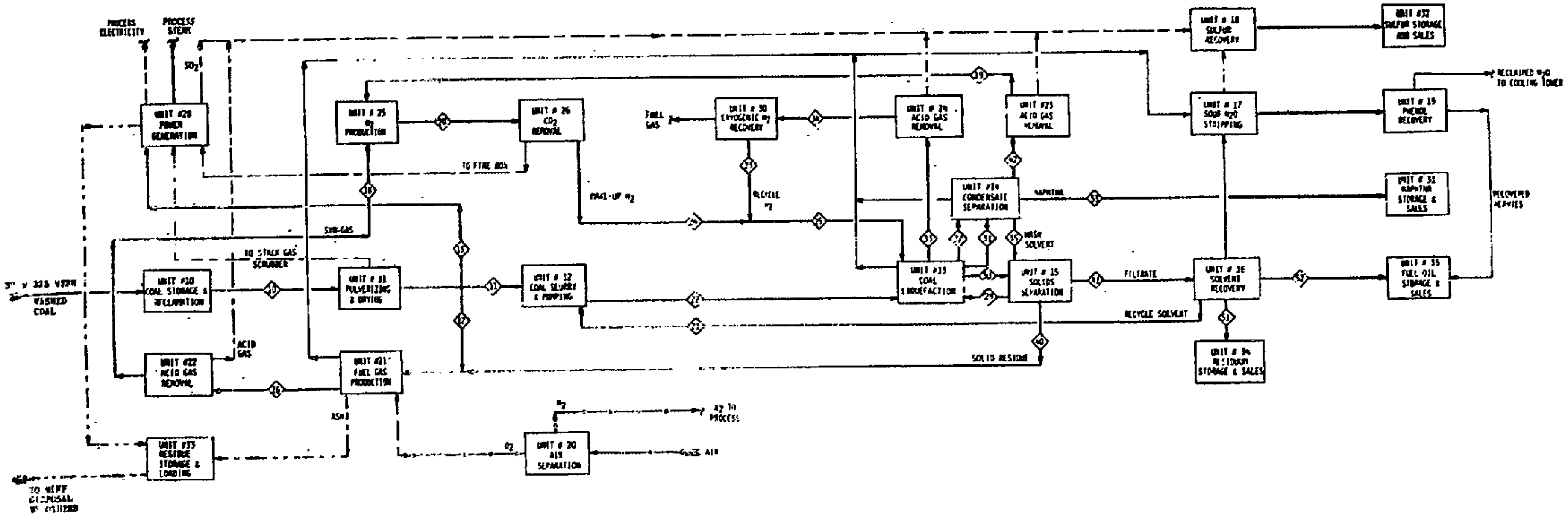

Figure 5-17. H-COAL SCHEMATIC (5-11)

5.3.2.7 Modified H-Coal Process

Figure 5-18 shows the proposed alternate plant design which follows the same philosophy as in the case of the SRC plant -- elimination of the air plant/O_2 enrichment, elimination of the fuel gas plant, and reduced diverted coal to the power generation unit offset in part by the otherwise normal consumption of solid residue from the solids or filtration plant.

5.3.2.8 Comparison of H-Coal with and without HTR

Given below is a brief comparison of the conventional H-Coal process with the proposed modified process employing a nuclear thermal source. The coal savings for the nuclear case (including power production) amounts to approximately 9.3% of the coal feed to the standard plant.

Basis: 50,000 bbl/day FOE

	Standard	With HTR	Savings
(I) Credits			
As received coal: lb/hr	1,689,350	1,491,980	197,370
Power: MW_e	238	158.7	79.3
Power: MW_t*	-	-	237.9
(II) Debits			
H_2 from HTR (lb moles/hr)	-	13,281.31	(13,281.31)
As received coal: lb/hr for H_2 production	-	109,690	(109,690)
Nuclear Heat, MW_t	-	141.9	(141.9)
(III) Net			
As received coal, lb/hr	-	-	87,680
Power: MW_t	-	-	96.0

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*$MW_t = 3 \times MW_e$

5.3.2.9 The Exxon Donor Solvent Process

Figure 5-19 shows the essential chemistry of the EDS Process. The naphthalene solvent is hydrogeneated to tetralin. Dissolution of coal in tetralin releases free hydrogen radicals in solution which in turn seek and attach to free radical fragments of coal. The hydrogenated fragments are then stabilized and can

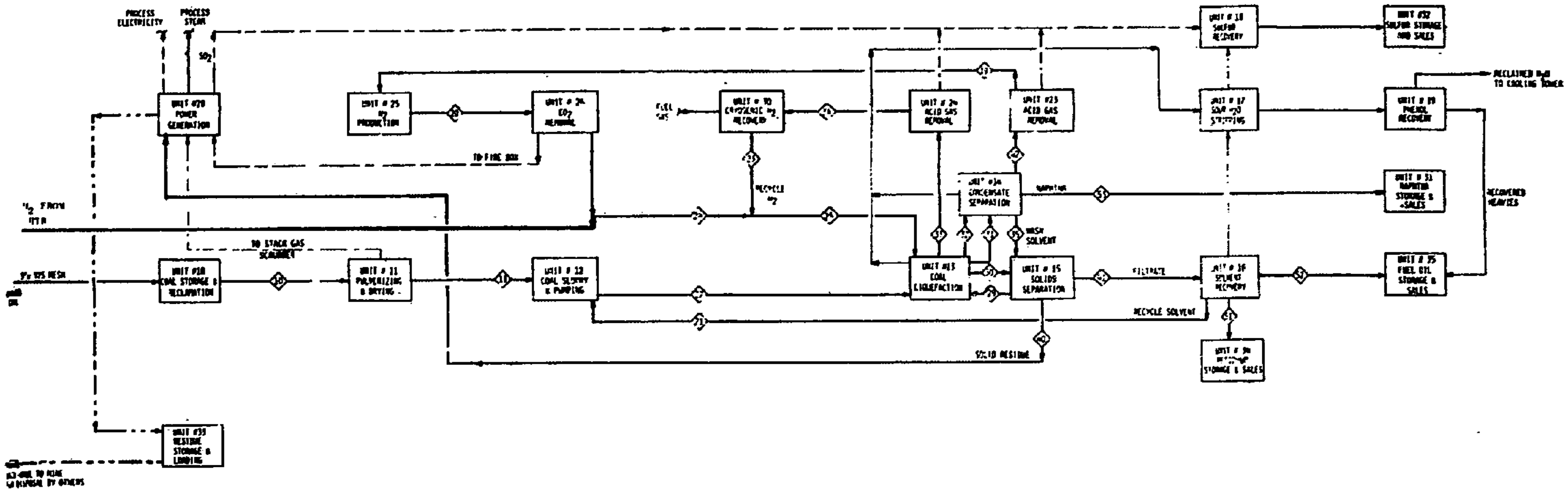

Figure 5-18. MODIFIED H-COAL SCHEMATIC

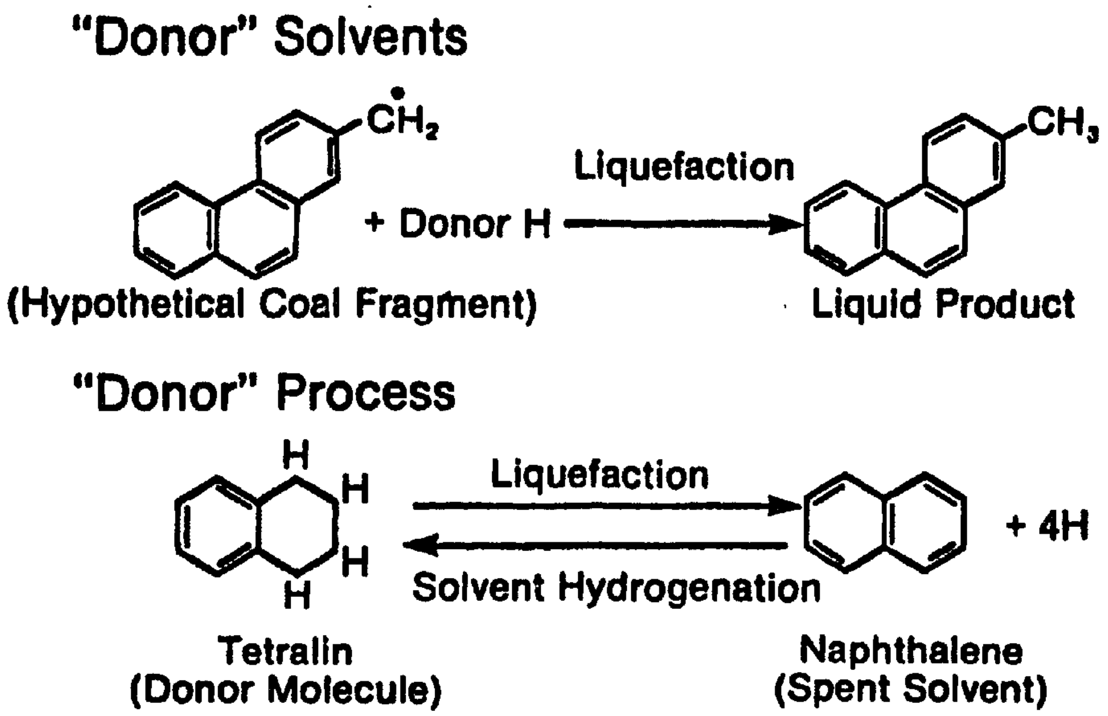

Figure 5-19. CHEMISTRY OF THE EXXON DONOR SOLVENT PROCESS

be removed as a liquid product. The overall flow sheet follows Figure 5-20 where coal is liquefied in the presence of tetralin, the products separated, the liquid coal solution removed and the spent solvent, naphthalene, returned to the hydrogenerator and recycled as tetralin back to the coal reactor. As with SRC and H-Coal the production of H_2 is a key process. Figure 5-21 shows the features of the EDS process excluding details which are available in Reference 5-13. Dried coal is slurried with solvent and pumped with H_2 into an oven and then into the liquifer vessel. Typical conditions are 1500 psi and 820°F. The liquefaction products are flashed, the light gases scrubbed of acid gases and hydrogen returned to the feed. The bottoms are reheated, flashed in a vacuum unit and the solid residue extracted.* The lighter liquids are then fractionated to produce a bottoms of oil fuel products and an overhead of spent solvent. The spent solvent is then heated and hydrogenated with fresh hydrogen at approximately 2200 psi and 550°F. The solvent then follows a path similar to the coal liquids leading to fuel gas manufacture, light low sulfur oils, heavier liquids and replenished solvent.

Case #311, Reference 5-13 (p. 68) gives the following process characteristics:

Coal feed, as received Illinois coal: 1,353,200 lb/hr
Hydrogen used, 33,359 lb moles/hr
Power consumption: 124.5 MW_e

The product distribution reflects the high H_2 consumption.

LPG	3654 bbl/day
Naphtha	14,457 bbl/day
Low sulfur fuel oil	17,050 bbl/day
Residuum	3180 bbl/day
Fuel gas	11,570 bbl(FOE)/day

In the EDS process, the Flexicoker adds significant fuel value apart from syngas to the process.

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*Exxon actually uses a "Flexicoker" system for the bottoms for a variety of products such as CO, H2 and pyrolysis products such as hydrocarbons.

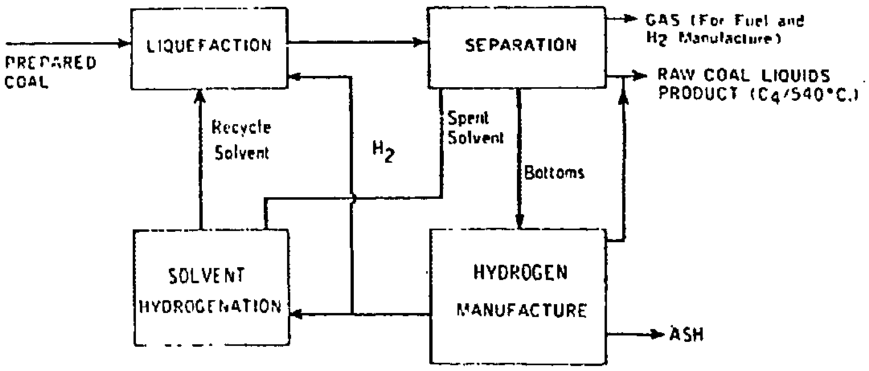

Figure 5-20. EDS PROCESS SIMPLIFIED BLOCK DIAGRAM (5-12)

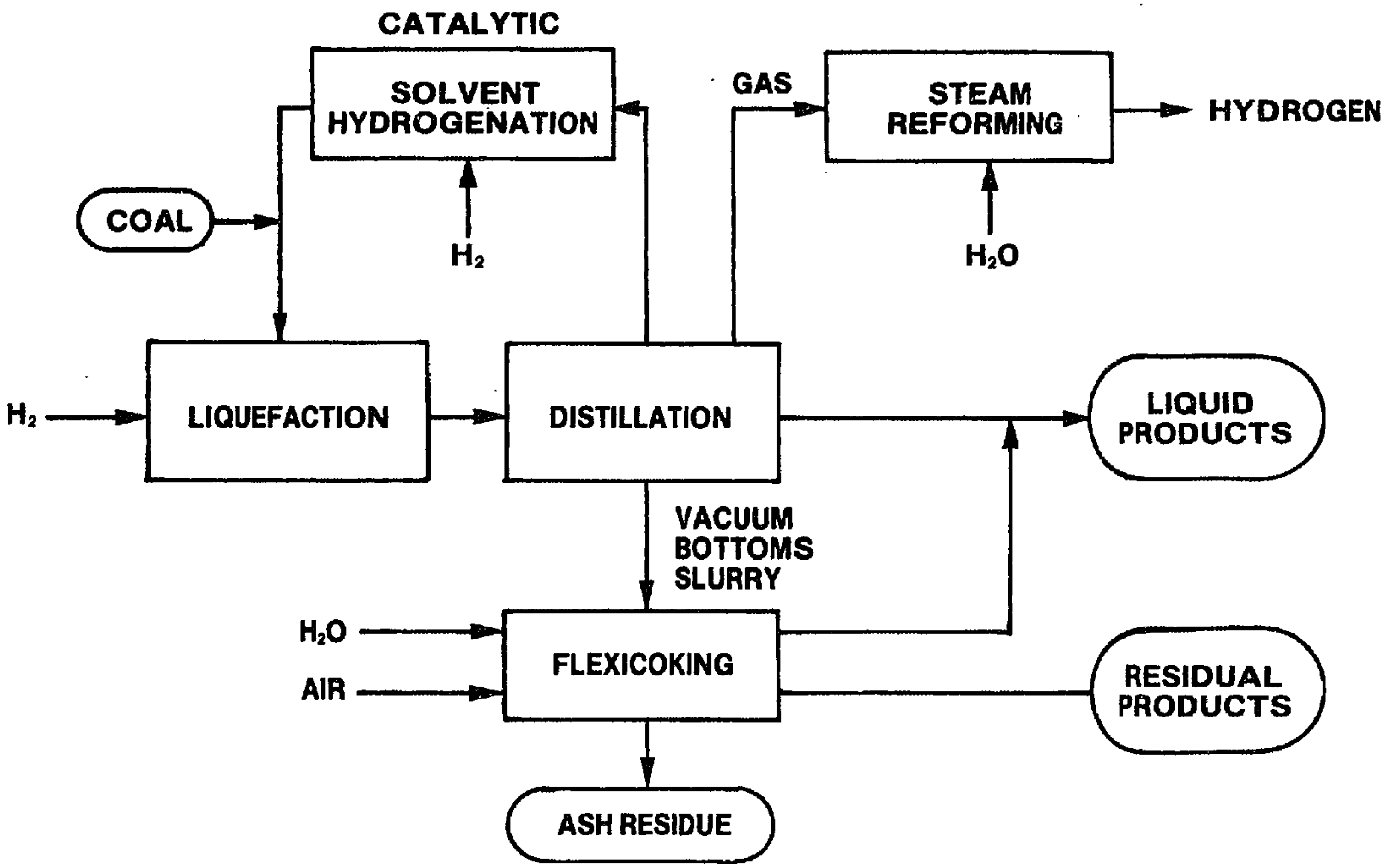

Figure 5-21. SCHEMATIC FLOW DIAGRAM OF EXXON DONOR SOLVENT COAL LIQUEFACTION PROCESS

5.3.2.10 Modified EDS Process

Exxon assumes H_2 is supplied "over the fence"* to the process. An HTR based gasification process could be used to supply this hydrogen and process heat requirements could be supplied by a nuclear/process interface. These changes would not show up directly in the flowsheet in Figure 5-21 since the details of the hydrogen production and the heat exchange to the process are not shown.

5.3.2.11 Comparison of EDS Process with and without HTR

Given below is a comparison of the conventional EDS process and nuclear based EDS process. Care must be taken in making the comparison since the increase in coal and power consumption results from the production of H_2 which is not included in the conventional case. The amount of coal saved is approximately 15% when both the savings in power production and in hydrogen manufacture are considered.

Basis: 50,000 bbl/day FOE

	Standard	With HTR	Savings
(I) Credits			
As received coal: lb/hr	1,353,200	1,353,200	0
Power: MW_e	124.5	124.5	0
Power: MW_t*	-	-	0
H_2 lb moles/hr	33,359	-	33,359
(II) Debits			
H_2 from HTR: lb moles/hr	-	33,359	(33,359)
As received coal for H_2 production	-	275,512	(275,512)
Nuclear heat, MW_t	-	356.4	(356.4)
(III) Net			
As received coal: lb/hr	-	-	(275,512)
Power: MW_t	-	-	(356.4)
H_2: lb moles/hr	-	-	33,359

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*This hydrogen may be purchased from another portion of one's own process with economic separation between the two portions of the overall plant.

5.3.2.12 Capital Costs and Annual Revenue Rates

Reference 5-11 quotes the cost of the SRC plant to be $\$1097 \times 10^6$ for investment and $\$1,510 \times 10^6$ for total capital*. In the H-Coal process these figures are $\$1,228\text{x}10^6$ and $\$1,690\text{x}10^6$ respectively. Exxon, (5-13) quotes prorated costs of $\$678\text{x}10^6$ for investment and $\$947\text{x}10^6$ for total capital. Note however 23% of their product is in the form of fuel gas which is otherwise consumed in the other two considered processes. If we assume a penalty free recycle of this gas the costs become $\$835\text{x}10^6$ and $\$1,116\text{x}10^6$ respectively.

Annual Savings

	Coal	Power	Total	% of Capital
1. SRC	$\$10.1\text{x}10^6$	$\$7.2\text{x}10^6$	$\$17.3\text{x}10^6$	6.9%
2. H-Coal	$\$10.3\text{x}10^6$	$\$8.9\text{x}10^6$	$\$19.2\text{x}10^6$	7.7%
3. EDS	$\$33.0\text{x}10^6$	$\$24.8\text{x}10^6$	$\$57.8\text{x}10^6$	23.1%

As a "ball-park" investment we shall assume capital requirements of $\$1.25\text{x}10^9$ and a revenue requirement return of 20% or $\$250\text{x}10^6$/year. As an estimate of the economic benefits we shall assume savings of coal at $25/ton and of power requirements at 3¢/kWhr.

For the EDS plant it is necessary to assume some net credit for the HTR hydrogen versus the over-the-fence cost of $3.20/1000 SCF assumed in the Exxon study(5-13). To generate the amount of H_2 required for the EDS plant would require 301,165 lb/hr of coal and 94.3 MW_e of power if we pro-rate from the SRC and H-Coal cases. Hence this amount of coal and power have been used as the basis of the savings in the above table.

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*These approximate numbers are assumed to be constant 1978 dollars.

Against these savings an important fact to be understood is the risk. An optimistic 90% availability is assumed for both the HTR and liquefaction plants. If these are independent the combined availability is 9% less than either one. The capital cost of each is roughly $\$2\times10^9$ (HTR) and $\$1.25\times10^9$ (liquefaction) so that a 9% loss of availability translates to

$$0.09 \times (0.2 \times 2\times10^9 + 0.2\times 1.25\times10^9)\ \$/\mathrm{yr}$$

expense which totals $\$58.5\times10^6$/year. This is of the order of the potential savings of even the high H_2 demand of the EDS process.

5.3.2.13 Coal Liquefaction: Summary and Conclusions

In this section, the application of HTR heat to coal liquefaction has been investigated. A key to successful coupling of the HTR and the various liquefaction processes is the production of hydrogen from coal using the catalytic gasification/-nuclear reformer concept developed previously. Here a process has been outlined for the production of 95% pure H_2 for use in liquefaction. This purity was chosen arbitarily and the optimum purity would have to be chosen by a more detailed design. The liquefaction processes themselves would require a minimum of modification in order to match up with an HTR. Heat exchange to the processes where needed could be supplied by steam generated from contact with the primary Helium coolant from the HTR. Again it should be noted that a significant portion of the process heat might be obtained from an LWR. However, again the temperatures required to produce the hydrogen in the gasifier could only be achieved by an HTR.

The analysis performed here indicates that 10 to 15% of the coal used in the conventional process could be substituted by nuclear heat. This is significantly less than in the analysis of the coal gasification processes. This is not surprising since the majority of the coal feed is converted to coal liquids and the generation of H_2 consumes a relatively small portion of the coal feed.

The cases chosen here produce products that range in quality from a boiler fuel up to a fairly select product of naphtha. The quality of the product is directly related to the amount of hydrogen added to the coal and the coal saved in the process is in turn related to the hydrogen consumption. Therefore, the amount of coal saved is greater for higher quality products and less for the lower quality products. Each of the processes considered here can operate over a range of process conditions to produce a range of products. Therefore, hydrogen requirements and coal savings can vary from process to process. Generally, the range of coal savings for any of the processes would fall within the 10-15% value calculated here.

Once the liquids are formed, they must be further refined if they are converted into gasoline. In order to accomplish this refining additional hydrogen must be added to the coal "crude" and, in general, the hydrogen for this step could also be produced by the catalytic gasification/nuclear reformer. Even more coal could be saved in this process, so that the overall coal savings in converting coal to gasoline would probably be of the order of 20%.

5.3.3 Ammonia Manufacture

5.3.3.1 Introduction

Ammonia is one of the major chemicals produced in the world today. It is the source of "fixed nitrogen" in the fertilizer industry which has as a basis various ammonia salts such as ammonium nitrate, phosphate, and sulfate. The most popular of these is ammonium nitrate in which the nitrate is obtained by oxidation of ammonia to nitric acid followed by neutralization by ammonia.

In the future, ammonia production will continue to increase to meet increasing world demand (particularly in developing countries) for these important ammonia based fertizilers. Currently, methane is the source of hydrogen for ammonia manufacture. As world supplies of natural gas diminish, coal will become a logical replacement for natural gas and the possibility exists for supplying with the HTR a portion of the energy required to convert the coal into hydrogen. In this section, a process for the production of high purity hydrogen from coal using supplemental HTR heat for use in ammonia manufacture is described, then a brief description of conventional ammonia manufacturing processes is given, a process is then described which would use the high purity nuclear/coal-hydrogen in the production of ammonia, and finally these two types of processes are compared.

5.3.3.2 Production of "Pure" Hydrogen

For many purposes hydrogen free of impurities is desirable. In the manufacture of ammonia, the presence of CO at concentrations greater than 10 ppm will poison and deactivate the Fe_2O_3 based catalyst. The nuclear/coal based process for "pure" hydrogen (shown in Figure 5-22) consists of proceeding as far as practical with the shift reactor and completing the removal of CO over a Ni based methanation catalyst which is tolerant to CO. The rationale for these steps is twofold. The shift reaction cannot be carried to completion since it is reversible and steam consumption increases excessively if the concentration of CO is reduced below 0.3% at the outlet of the reactor. The methanation step, carried out at 475°F/600 psig, is thermodynamically favorable to almost complete conversion of the residual CO to CH_4.

$$CO + 3H_2 = CH_4 + H_2O$$

The impurity of 0.3% methane in the final H_2 gas does not affect the ammonia production step since it may be recovered and recycled or burnt for a small additional heat source. The quantity of processed gas produced by this means is shown in Tables 5-18 through 5-20 which corresponds to conditions in Tables 5-12 through 5-14 and Tables 5-15 through 5-17, i.e., reformer at 825°C, gasifier at 700°C or 825°C and reformer and gasifier at 700°C respectively.

The steam consumption for "pure" and "impure" hydrogen is well demonstrated by comparing columns 11 and 12 (i.e., prior to shift and post shift) of Tables 5-15 and 5-18 for the prime case of a 950°C HTR/825°C reformer/700°C gasifier. These numbers are summarized in Table 5-21.

Note the absolute high levels of steam consumption, the latent heat requirement to generate them, and the high fractions of steam available in the shift reactor effluent. For "pure" hydrogen, recovery of the latent heat of the unreacted steam could supply approximately 153 MW out of the 176 MW total required. It must be emphasized that the temperature level of this heat recovery step must be matched such that a reasonable temperature difference exists across the heat exchanger. The presence of the steam power plant, as shown in Figure 5-23, allows the interchange of this heat at whatever temperature it appears and

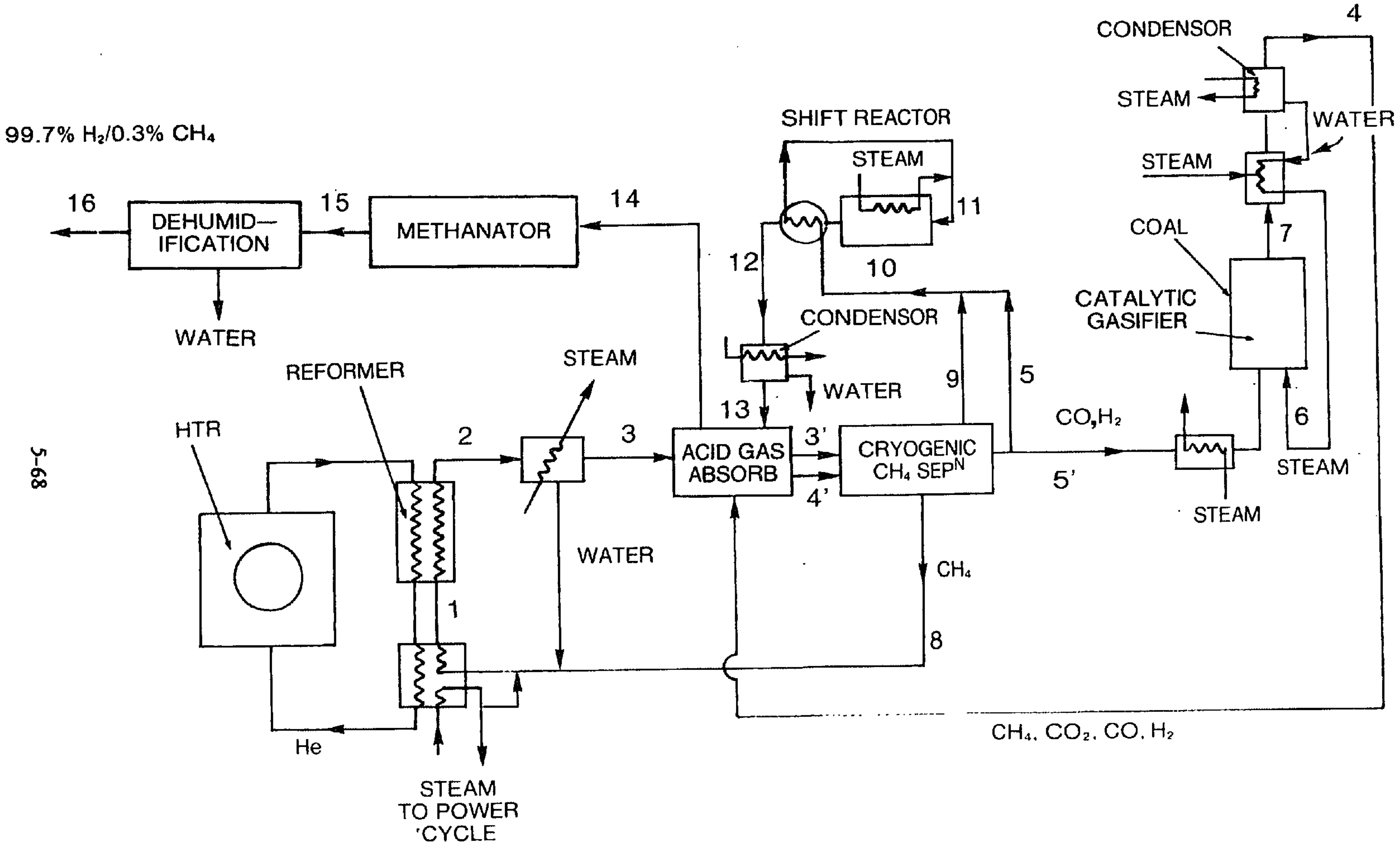

Figure 5-22. SCHEMATIC OF H_2 PRODUCTION FOR AMMONIA PRODUCTION & CHEMICALS VIA HTR + COAL

Table 5-18: Product Streams in ST moles/hr for 2000 ST/D Illinois #6 Coal to Make H_2

Stream	10	11	12	13	14	15	16
CH_4	-	-	-	-	-	0.031	0.031
H_2O	-	22.904	19.961	0.323	0.253	0.285	-
CO	2.971	2.971	0.031	0.031	0.031	-	-
CO_2	-	-	2.938	2.938	-	-	-
H_2	7.654	7.654	10.588	10.588	10.588	10.495	10.495

NPH	950°C
Reformer	825°C
Gasifier	700°C

Table 5-19: Product Streams in ST moles/hr for 2000 ST/D Illinois #6 Coal Feed to Make H_2

Stream	10	11	12	13	14	15	16
CH_4	-	-	-	-	-	0.031	0.031
H_2O	-	22.670	19.764	0.323	0.254	0.347	-
CO	2.942	2.942	0.031	0.031	0.031	-	-
CO_2	-	-	2.843	2.843	-	-	-
H_2	7.703	7.703	10.614	10.614	10.614	10.521	10.521

NPH	950°C
Reformer	825°C
Gasifier	825°C

Table 5-20: Product Streams in ST moles/hr for 2000 ST/D Illinois #6 Coal Feed to Make H_2

Stream	10	11	12	13	14	15	16
CH_4	-	-	-	-	-	0.035	0.035
H_2O	-	14.158	12.293	0.324	0.254	0.360	-
CO	1.893	1.893	0.035	0.035	0.035	-	-
CO_2	-	-	1.864	1.864	-	-	-
H_2	8.732	8.732	10.591	10.591	10.591	10.486	10.486

NPH	825°C
Reformer	700°C
Gasifier	700°C

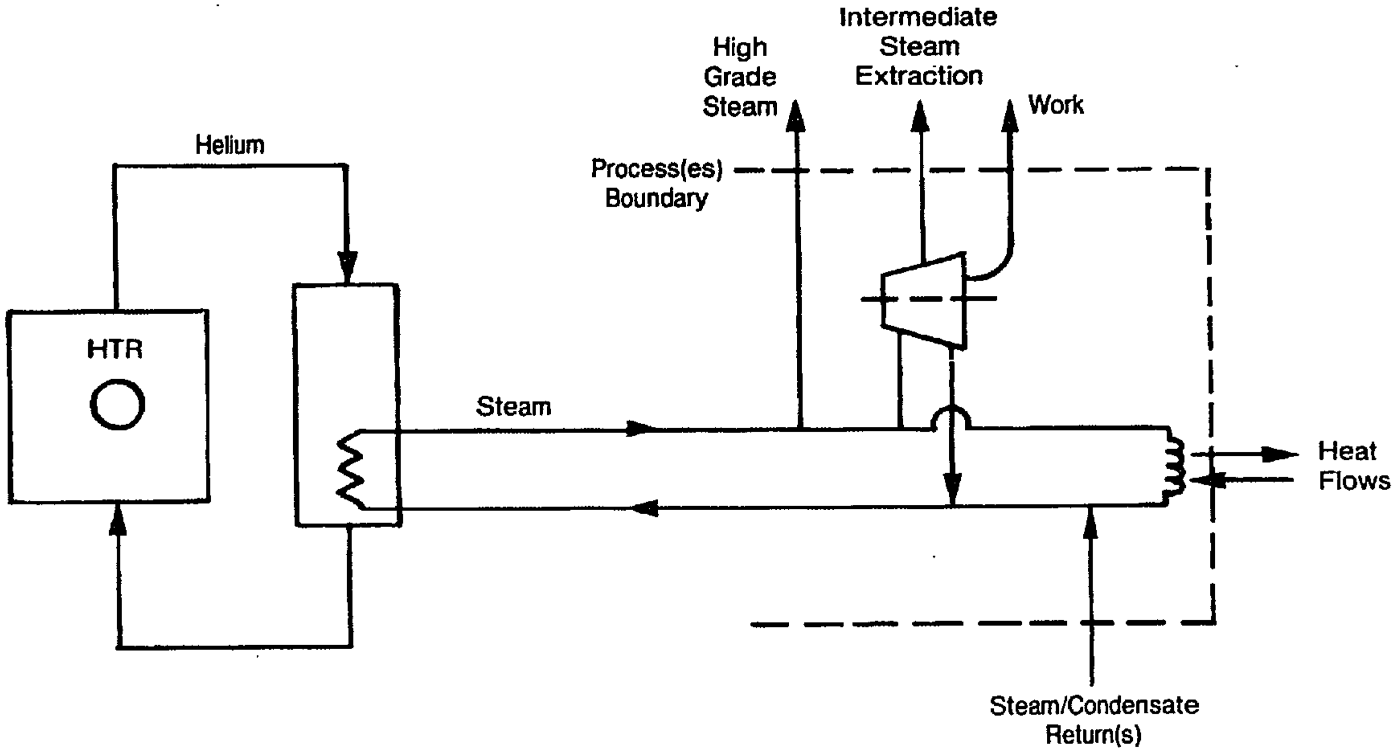

Figure 5-23. BOTTOMING STEAM CYCLE INTERACTIONS WITH PROCESS(ES)

recovery of the work available in it by the power plant*. Indeed the ability to exchange "low" grade heat in the form of steam or condensate to an HTR power cycle may be a significant advantage of the coupling of the HTR to chemical processes.

Table 5-21

STEAM CONSUMPTION IN THE SHIFT REACTOR

Product	Steam Pre-Shift ton moles/hr	Steam Post-Shift ton moles/hr	Latent Heat in MW	Potentially Recoverable Latent Heat in MW
5% CO	3.360	0.919	25.8	7.1
0.3% CO	22.904	19.961	176	153

5.3.3.3 Conventional Process for the Production of Ammonia

Figure 5-24(5-14) shows schematically a conventional ammonia production process. The process can be divided into 3 major sections: 1) hydrogen production, 2) raw gas treatment, and 3) ammonia synthesis. Natural gas is normally used as the source of hydrogen. Steam is added to the feed natural gas which is catalytically reformed to hydrogen and carbon monoxide in an externally fired furnace. Conversion of the methane in this unit is limited by control of steam addition and temperature in the furnace. The process gases are then fed to a secondary reformer along with a well controlled amount of air. The air serves a dual purpose. It supplies oxygen for the partial oxidation of a portion of the gas to supply heat for the endothermic reforming reaction which is carried out simultaneously and it supplies nitrogen which will subsequently be used in the ammonia synthesis reaction. The stoichiometry of the various components in these reactors must be carefully controlled such that hydrogen production is maximized and the proper N_2/H_2 ratio for the synthesis reaction is obtained in the outlet raw gas. The raw gas is treated to shift the carbon monoxide to hydrogen by the water gas

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*Chemical processes have not traditionally been designed for high efficiency. When the product is a saleable chemical the cost of energy may be considered a relatively small flow-through charge. Often the effluent heat may be exhausted to a cooling tower. As energy becomes more expensive, processes must be efficiently designed to provide the lowest cost products.

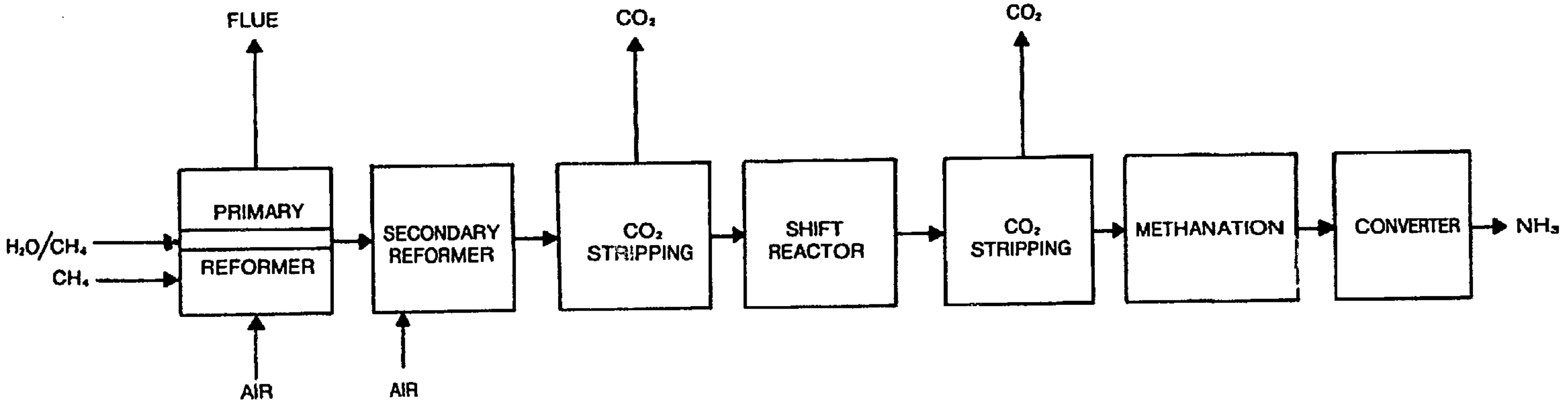

Figure 5-24. CONVENTIONAL AMMONIA FROM NATURAL GAS

shift reaction and then the carbon dioxide is removed by a conventional absorption unit. The residual carbon oxides are methanated to prevent poisoning of the synthesis catalyst. The pure N_2/H_2 mixture is then compressed to between 100 and 300 atm, mixed with recycled unconverted N_2/H_2, and preheated before being fed to the synthesis reactor. Iron oxide is employed as a catalyst in the synthesis reactor. Although unfavorable thermodynamically for high conversions, high temperatures are required to obtain sufficient reaction rates; thus, the preheat is required. High pressure is employed to help drive the synthesis reaction toward ammonia production. Despite the extreme pressures employed, conversion is limited to 15 to 20% per pass. This low conversion requires that the ammonia product be separated from the unconverted N_2/H_2 mixture and the reactant be recycled back to the reactor. The ammonia can be delivered as either a liquid or gaseous product.

5.3.3.4 Nuclear/Coal Based Ammonia Production Process

The key element of a nuclear based ammonia process is the production of hydrogen from coal using nuclear heat. This process has been outlined in Section 5.3.3.2. Once this pure hydrogen is obtained, it is fed to a modified synthesis process. The major difference between this process and the conventional process is that air could no longer be used as the source of nitrogen, since the oxygen present is no longer needed for the partial oxidation reaction. Therefore, as shown in Figure 5-25, an air separation plant is added to the flowsheet to produce a pure nitrogen product for use in the synthesis reaction. An enriched air or oxygen byproduct is available for sale, or if no market existed, is vented. Once the nitrogen is obtained, it is compressed to a pressure equivalent to that of the feed hydrogen and then mixed with it. This mixture is then compressed to reaction pressure and fed to a process essentially equivalent to existing synthesis processes.

The overall material and energy balance for the nuclear based ammonia process is given in Table 5-22 excluding the production of the hydrogen which is discussed in Section 5.3.3.2. Major energy consuming operations include the compression of the reactants to pressure and the refrigeration of the product stream to increase the recovery of ammonia from the recycle N_2/H_2 mixture.

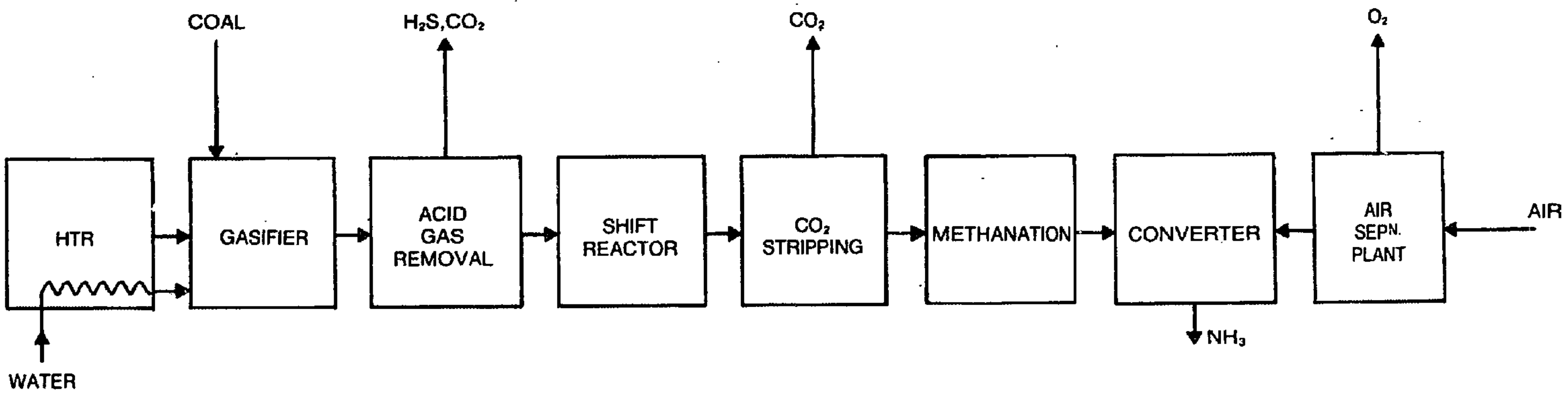

Figure 5-25. AMMONIA FROM COAL AND HRT HEAT

Table 5-22
OVERALL MATERIAL AND ENERGY BALANCE FOR NUCLEAR/COAL BASED AMMONIA PLANT

Ammonia Production		3000 ST/D
Coal Consumed		2100 ST/D
Electrical Power		
Oxygen Plant	4.4 MW_e	
N_2 compression	16.3 MW_e	
N_2/H_2 compression	21.5 MW_e	
Refrigeration	23.6 MW_e	
Total	65.8 MW_e	
Thermal Equivalent NH_3 Synthesis		197.4 MW_t
Thermal Requirement for H_2 Production		794 MW_t
Total Thermal Requirements		991.4 MW_t

5.3.3.5 Comparison of Nuclear Based and Conventional Ammonia Production Processes

The ammonia synthesis process in both the conventional and nuclear based processes is essentially identical. However, there is a major difference in the method uses to obtain the nitrogen and hydrogen reactants. The conventional process uses air and natural gas while the nuclear process uses coal, nuclear heat and an air separation plant. Although the processes were not evaluated in detail, it is obvious that the viability of the nuclear based process is dependent on the cost of nuclear heat, coal, and the nitrogen separation as compared to the cost of natural gas. Assuming that the nuclear based process can be designed as efficiently as the conventional process, the lower cost of both coal and nuclear heat make this route attractive as a means of producing ammonia. Again it should be noted that the key use of HTR heat is in the manufacture of hydrogen where the conversion in the reformer is directly related to the temperature of the gas coolant.

5.3.4 Steel Manufacture

5.3.4.1 Introduction

Two major processes are involved in the manufacture of steel. The first is the reduction of iron ore to form pig iron. The second is the treatment of this raw product in a Basic Oxygen Furnace (BOF) where the residual carbon is burned out of the pig iron and alloying components are added. The first step requires a reducing agent and a supply of thermal energy and, hence, an HTR could be utilized in conjunction with it. The second step is highly exothermic and direct application of the HTR is unneeded. However, pure oxygen is required in the BOF, and low grade steam generated from the HTR cycle could be used to drive the compressors in the oxygen plant refrigeration system.

In this portion of the report, a brief discussion of direct reduction steel manufacturing processes using gaseous reducing agents will be given. Then the application of a HTR synthesis gas production unit to produce the reducing gas for the iron ore treatment will be described.

5.3.4.2 Process Description of Direct Reduction Processes

There are many complex steps in the manufacture of steel. Today's convention is to use a coking coal blast furnace to produce the pig iron and a BOF or modifications thereof to produce the steel. However, there is a growing trend toward direct reduction of iron ore pellets which are then converted to steel in an electric furnace. This method is very compatible with the use of synthesis gas from the high temperature gasification of coal.

Over 1000 direct gaseous reduction (DR) processes have been patented but only a handful have become commercial. Principal among these are:

HYL	Static Bed
MIDREX	Moving Bed
ARMCO	Moving Bed
PUROFER	Moving Bed
HIB	Fluid Bed

There are also several processes using rotary kilns and solid fuels.

For the purpose of this discussion, we will consider the most fuel efficient gas reduction process, the MIDREX process shown in Figure 5-26(5-15), but any of the gaseous reducing processes could be used.

The efficiency of the gas utilization and the rate of reduction vary widely with temperature. For example, the per pass utilization of CO varies from 48% at 575°C to 30% at 900°C, while that of H_2 varies from 22% at 575°C to 37% at 900°C.

Reduction rates also vary with temperature and iron oxide can be reduced rapidly at greater than 550°C, then more slowly through the 600 to 800°C range until at 900°C reaching the same rate as at the lower temperature. However the metal iron produced at temperatures below 500°C is usually pyrophoric and must be heated to about 900°C to sinter and coalesce, thus reducing the surface area and stabilizing the reduced iron.

It should be noted that the 90 to 92% metallization is not limited by thermodynamics. It is controlled to prevent sintering and agglomerization of the charge.

5.3.4.3 Nuclear Based Steel Manufacture

HTR heat could be utilized in two ways in the manufacture of steel: 1) to supply energy for the production of synthesis gas from coal via the gasification scheme presented previously, and 2) to supply heat for the sintering of the product sponge iron to stabilize it.

The product composition given in Table 5-12 (950°C NPH/825°C reformer/-700°C gasifier) could serve the same duty as the reducing gas from the reformer in the MIDREX process as shown in Figure 5-26.

Using synthesis gas as shown in column 5 of Table 5-12 would serve the same purpose as the natural gas feed in Figure 5-26.

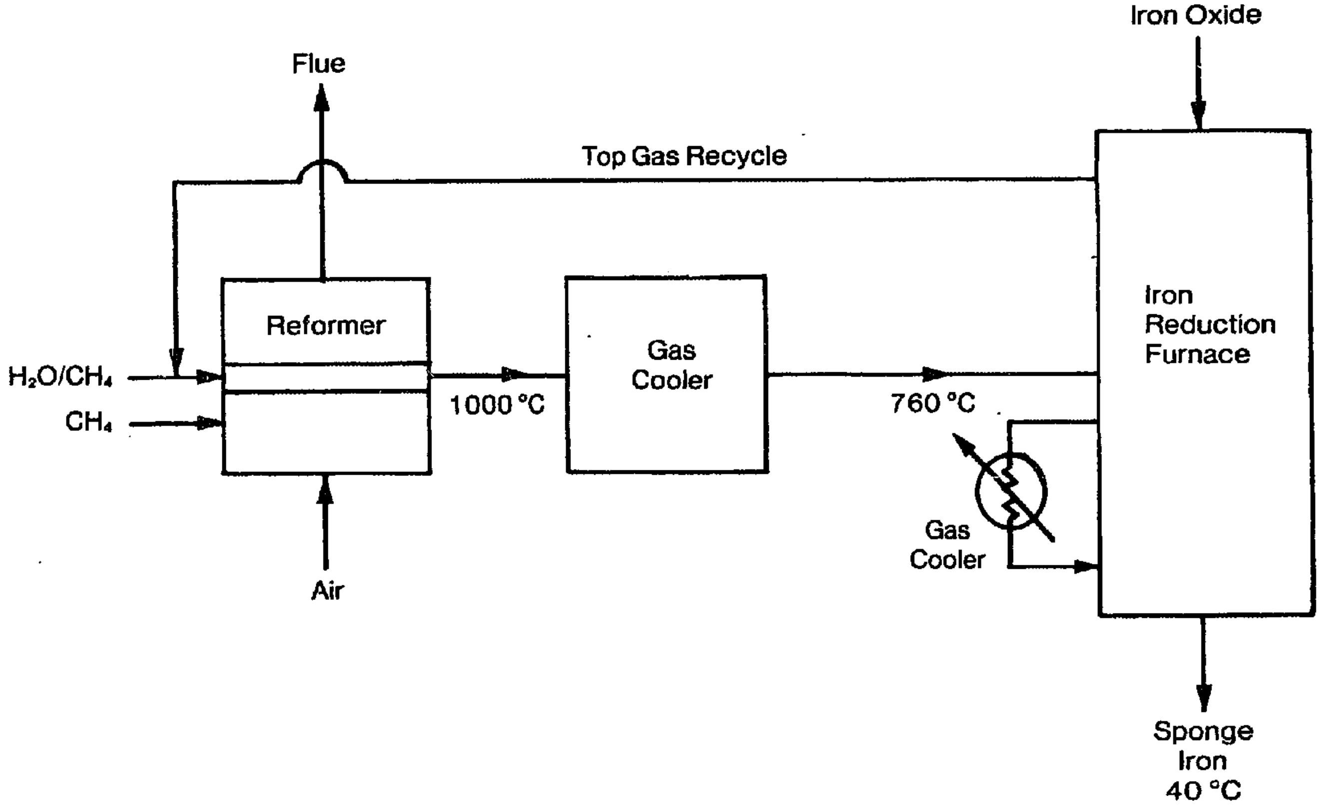

Figure 5-26. SCHEMATIC FLOW DIAGRAM FOR MIDREX DIRECT IRON REDUCTION PROCESS (5-15)

MIDREX and HYL have generated some energy use data from which capacity calculations may be made; for example, per short ton of product

	MIDREX	HYL	
Metallization	92%	90%	
Direct Fossil Energy	9.51×10^6 Btu	430 Nm^3	Nat. Gas
Electrical Energy	121 KWh	70 KWh	

Other data from MIDREX quote 1.1×10^7 Btu/ST for natural gas use. The MIDREX figures above total 9.54×10^6 Btu. At 1000 Btu/cu ft, the HYL figures would equal 1.4×10^7 Btu/ST for total energy.

Using a rounded average of 1.1×10^7 Btu/ST and the energy figure from Table 5-12, 6.254×10^{10} Btu per day leads to an estimate of 5682 STD or 2.1 million standard tons/year. This is the output of three large MIDREX units. However, DR plants of 2.7 million standard tons per year are in the planning state and Several larger conventional steel plants exist.

Calculations based on the gas production shown in Table 5-12, 2.971 ST moles/hr of CO and 7.654 ST moles/hr of H_2 show a stoichiometric yield of 396 tons of Fe/hr, 9493/day, 3.47 million tons/year. Thus the actual thermal efficiency based on the above data is about 55% of the stoichiometric value.

Since the process description is not complete, it is not clear as to where the excess thermal energy is rejected. Some of it may be used (by combusting off gas) to preheat the iron or carryout the sintering operation. Since, in theory, the HTR could provide both of these functions, the gas used in these functions could be recycled to the process thus increasing the amount of steel produced per unit of syn gas feed to the direct reduction furnace.

5.3.5 Other Chemicals and Process Heat Applications

5.3.5.1 Introduction

The use of HTR heat in coal liquefaction, ammonia production, and steel manufacture would result in a very large market for this type of reactor. These processes all are carried out in large units which would be compatable with a reasonable size HTR. Because of the attractiveness of these three processes to HTR applications, somewhat detailed analyses were carried out in this report. To complete the spectrum of possible HTR process heat applications, other chemical processes will be briefly considered in this section.

In theory, HTR heat could be used in any application that requires thermal energy. In reality, size limitations, siting difficulties, competition from more convenient fuels, and process design constraints might limit HTR applications in many chemical processes. Some of these constraints could be overcome by the use of the Thermochemical Pipeline (described in detail in Section 4 of this report). The other constraints will be ignored for this discussion.

Other applications of HTR heat can be divided into two general categories: 1) applications requiring the production of hydrogen or syn gas and 2) direct heat applications. In the first category, petroleum refining, methanol production, and gasoline via the Fischer-Tropsch route will be discussed briefly in the remainder of this section. In the second category, recovery of crude oil from oil shale, and heavy oil deposits and production of inorganic chemicals by heat treatment will be discussed.

5.3.5.2 Petroleum Refining

The purpose of refining is to separate the crude oil into a wide variety of distinct and usable products. This is shown schematically in Figure 5-27(5-16) where the crude is treated and separated into products ranging from the very "light" liquefied petroleum gas (LPG) to the "heavy" asphalts. The products contain many chemical compounds that are blended together to give the desired characteristics (boiling point range, vapor pressure, viscosity, octane rating) required for that particular product.

In the transformation of the crude into the final products, it is estimated(5-16) that 70% of the crude undergoes a chemical conversion of some type. These chemical conversions are processes such as cracking, isomerization, reforming, polymerization or alkylation. In order to carry out the transformation of the crude oil into useful products, approximately 10% of the net input to the

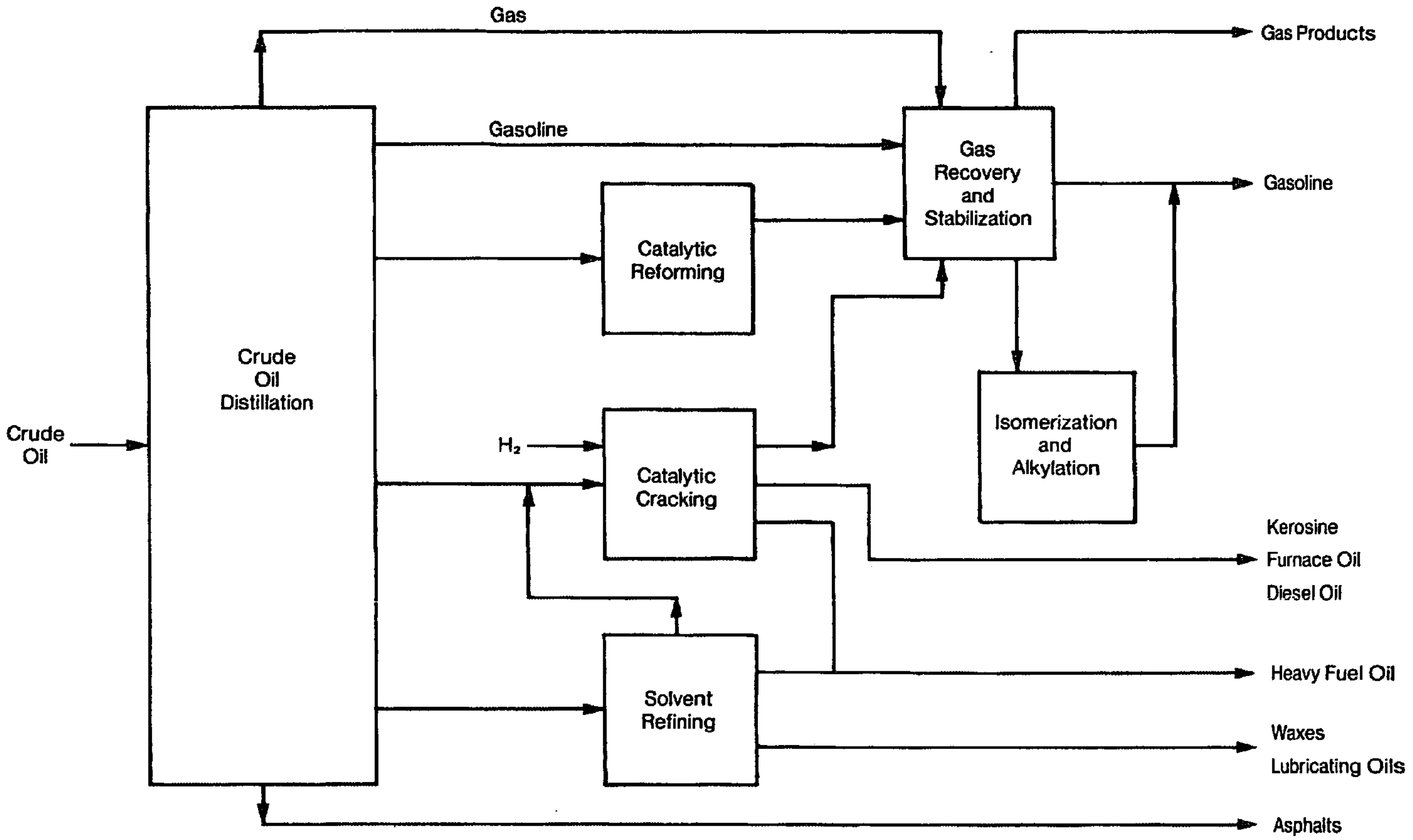

Figure 5-27. SCHEMATIC DIAGRAM OF OIL REFINERY (5-16)

refinery is consumed as a source of energy to drive the reactions. Much of this energy is consumed in the production of hydrogen which is added chemically to the crude to obtain modified chemical compounds which meet the desired product specifications.

The HTR could be used as a heat source for hydrogen production from crude or coal and thus reduce the amount of crude required for energy production. This operation would require the high temperature energy available from the He coolant and the lower temperature heat could be used to drive the chemical and separation processes in the refinery. Assuming that 10% of the crude is consumed for energy requirements, and that the heating value of crude is 6 x 10^6 Btu/bbl, a 50,000 bbl/d refinery would require approximately 370 MW of thermal energy. Obviously larger refineries would require proportionally larger amounts of energy such that a typical 150,000 bbl/d refinery would consume 1100 MW_{th} of HTR heat. The major problems associated with the application of HTR supplied energy to petroleum processing are the siting problems associated with the nuclear reactor close to a large source of fuel and personnel, the opposition associated with replacing the convenient petroleum energy source with an external source, and the need for high reliability and on stream time required for economical operation of the refinery.

5.3.5.3 Methanol

Another logical candidate industry for utilization of nuclear/coal based syn gas is methanol production. In existing methanol synthesis processes, as shown in Figure 5-28(5-16), the syn gas feed (2:1 H_2 to CO) is compressed to 300 atm prior to reaction in the methanol converter which operates at approximately 300°C. The catalyst employed is usually silver or copper promoted with oxides of zinc, chromium, manganese, or aluminum. The methanol synthesis reaction is exothermic and, hence, no heat is required in the reactor. Conversion is low (approximately 15% per pass) which requires large recycle rates. The recycle gas must be compressed to account for system pressure drop and the feed gas must be compressed to the reaction pressure.

The use of syn gas as the feed and the compression power requirements make the methanol synthesis process ideal for matching to the coal derived/nuclear supplemented syn gas production process discussed previously. The high grade

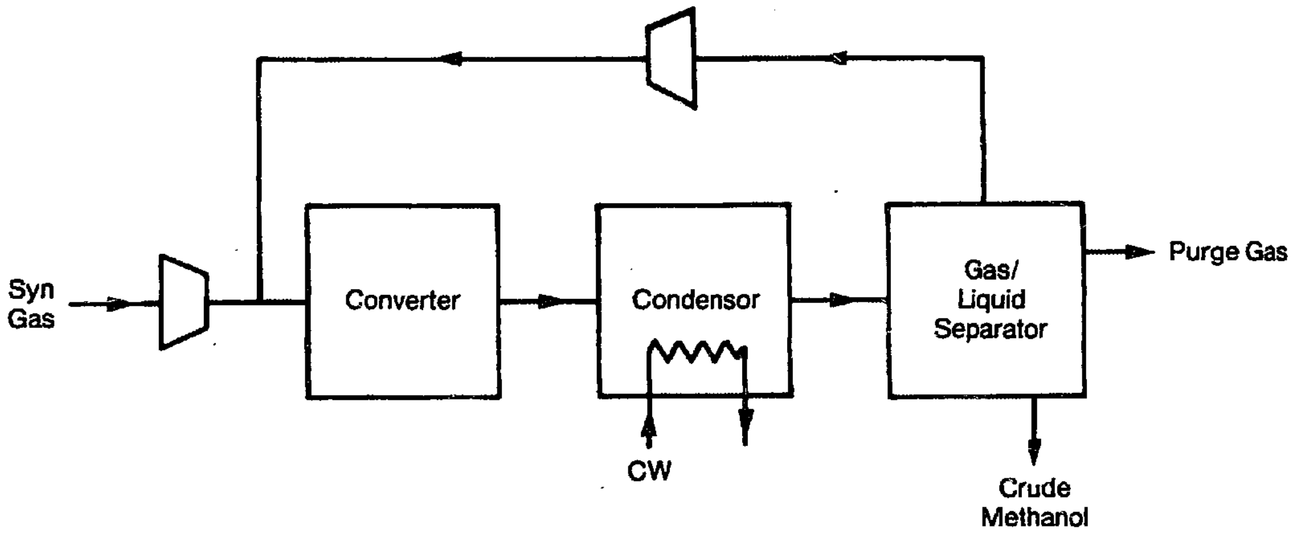

Figure 5-28. SCHEMATIC DIAGRAM OF METHANOL SYNTHESIS PLANT (5-16)

energy from the coolant loop could be used for synthesis gas production and the lower grade energy used for steam generation to drive the compressors. The syn gas product from the catalytic gasifier/reformer system could easily be shifted (via the water gas shift reaction) to produce the exact composition required for methanol synthesis.

The methanol synthesis industry is well suited for possible nulcear/coal applications since the plant could be located near the mine mouth and the product shipped by pipeline to its final use point or distribution center. The current feed of synthesis gas is supplied by the reforming of natural gas and, hence, the substitution with efficiently generated coal derived syn gas should become economically attractive in the future.

5.3.5.4 Fischer-Tropsch Route to Gasoline

One alternate scheme to direct coal liquefaction to produce liquid products is the complete gasification of coal to synthesis gas which is in turn liquefied by the Fischer-Tropsch synthesis. The Fischer-Tropsch synthesis involves the reaction of a 2:1 to 3:1 ratio of hydrogen to carbon monoxide (depending on the desired product mix) over an iron based catalyst to form the higher molecular weight liquid products. The products obtained range from methane to hard waxes. As shown in Figure 5-29, this distribution of products requires refinery-like processing to separate the products and selectively convert a portion of them to more desirable products. HTR heat could be used in a Fischer-Tropsch process to supply energy for the gasification to syn gas to be used as feed and for the production of hydrogen for treatment of a portion of the product stream. Lower grade energy from the bottoming cycle on the HTR cooling loop could be used to produce steam to run the compressors in the plant and to provide energy for the various separation processes.

An interesting option would be to gasify the coal at the mine mouth, transport the syn gas to a convenient industrial process heat user site, and then run the exothermic Fischer-Tropsch reaction to supply thermal energy for the industrial user and also to supply the liquid product. This would be an alternate embodiment of the half Thermochemical Pipeline (Section 4.0) producing liquid products instead of SNG.

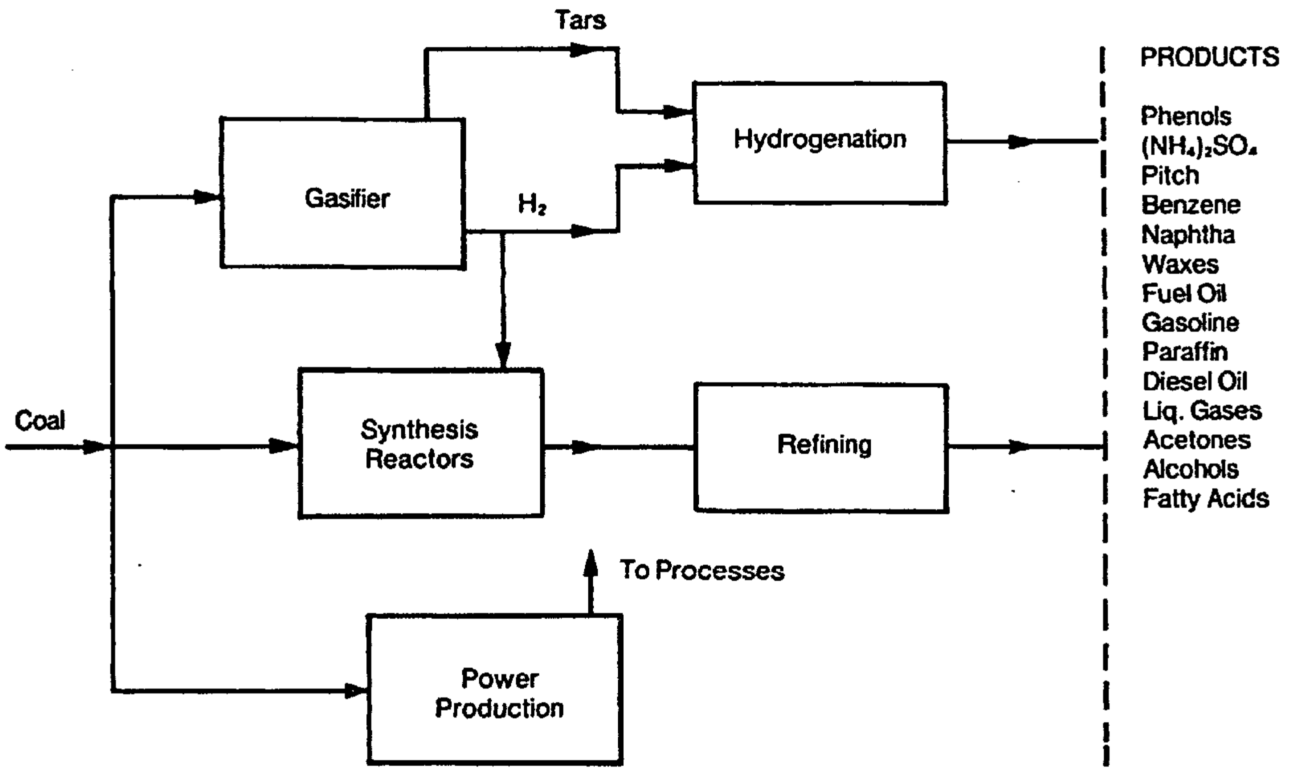

Figure 5-29. FISCHER TROPSCH ROUTE TO GASOLINE

In principal, there is no reason why the catalytic gasification/HTR/reformer process discussed earlier in this section could not be used to supply the syn gas feed to a Fischer-Tropsch process. The incentives for this application would again be economic (the cost of nuclear based Fischer-Tropsch liquid products vs. the cost of petroleum) and environmental (a reduction of approximately 40% of the coal consumption and coal source emissions from the gasification plant). The Fischer-Tropsch synthesis itself is commercial under unique economics with a plant operated by the South African Coal, Oil, and Gas Corp. Ltd. (Sasol) at Sasolburg, South Africa.

5.3.5.5 Direct Heat Applications

The direct heat applications of NPH have to be considered in the context of both total heat demand and of the temperature requirements for the process.

Various chemical industries produce sufficient quantities of product to be considered for direct NPH. Considered here are cement, lime, gypsum and glass manufacture in the inorganic chemical industry, shale oil and enhanced oil recovery in the fuel industry.

<u>Cement manufacture</u> is carried out on a large scale - 81 x 10^6 ST/yr in the U.S. in 1972(5-16). The total theoretical power requirement is about 1.5 x 10^{14} Btu/yr, which is equivalent to about 5000 MW_t of power at 100% capacity factor. However actual power usage has been about 20,000 MW_t. It is indeed unfortunate with respect to the current study that the temperature levels at which the key endothermic reactions occur seem to be beyond even optimistic HTR projections.

Table 5-23 is reproduced from Shreve and Brink(5-16).

At 900°C and above the fraction of heat required is about 57% of the total. Obviously nuclear heat could be used for the dehydration steps (F 600°C). These would amount to the majority of the remaining 43% additional heat. Thus NPH <u>could</u> substitute from 2150 MW_t to 8600 MW_t in the U.S. depending on the efficiency of processing steps. However, it is more probable that the dispersed

Table 5-23
REACTIONS OF CEMENT MANUFACTURE

Temperature, °C	Reaction	Heat Change
100	Evaporation of free water	Endothermic
500 and above	Evolution of combined water from clay	Endothermic
900 and above	Crystallization of amorphous dehydration products of clay	Exothermic
900 and above	Evolution of carbon dioxide from calcium carbonate	Endothermic
900-1200	Main reaction between lime and clay	Exothermic
1250-1280	Commencement of liquid formation	Endothermic
1280 and above	Further formation of liquid and completion of formation of cement (compounds)	Probably endothermic on balance

character of the cement industry would require that conventional energy recovery methods be seriously considered before the centralization of the industry around a nuclear island.

Lime Calcination: The industry is greater than 20x10^6 ST/yr in the U.S.(5-16). In theory, at 900°C the energy requirement for the calcining process

$$CaCO_3 = CaO + CO_2$$

(limestone) (lime)

is about 4.25 x 10^6 Btu/ton CaO. Thus, the industry wide requirement is about 2850 MW_{th}. Unfortunately, apart from the dispersed nature of the industry, the fact is that in practice calcining temperatures must be as high as 1200-1300°C. This is an inherent constraint if the center of a lime particle is to be raised to its thermodynamically constrained temperature of 900°C.

Again the prognosis for NPH in the lime industry is poor when one considers that it requires helium from an HTR at 1325-1425°C.

Gypsum manufacture is the dehydration of gypsum to its demihydrate, i.e.,

$$\mathrm{CaSO_4}\,\tilde{}\,2\mathrm{H_2O} = \mathrm{CaSO_4}\,\tilde{}\,\tfrac{1}{2}\mathrm{H_2O} + 1\tfrac{1}{2}\mathrm{H_2O}$$

The dehydrated or calcined gypsum is made into familiar products such as wall plaster. However it only requires heat at 120-190°C and is not a serious candidate for NPH from an HTR.

Glass manufacture is a major industry in the U.S. Sales volume in 1972 was about 15×10^6 ST using data in references 5-16 and 5-17. The energy content of glass is about 800 Btu/lb(5-17)* so that the industry consumed approximately 800 MW_t at 100% capacity factor in that year.

It is instructive to note the softening "point" of various glasses - 700°C for common soda lime glass, 820°C for pyrex to 1425°C for fused silica. Clearly glass making must be at temperatures in excess of these points. In fact the processes operate at between 1015°C and 1425°C for common glass at various process steps.

Again one should conclude that the glass industry will use electric heat or more efficiently use fossil heat in the future. It is not likely to become a customer for NPH.

Oil Shale Processing: Oil shale processing has been addressed only briefly in this study, relative to the effort on understanding coal conversion processes. Four HTR-Multiplex output energy forms match requirements of oil shale recovery and processing:

- Electrical power, for the mine and the surface processes.
- Heat from TCP methanators, to replace product used for fuel in the mine and surface processing.
- Hydrogen produced by steam reforming, electrolysis, or water splitting, to replace product used for fuel and feedstock in onsite steam reformers.

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*Fossil energy consumption efficiency is 10-30% and electric furnace efficiency is about 70% (5-17).

- Heat at temperatures not yet defined, for a syncrude refinery (probably collocated with the HTR-Multiplex), to replace product otherwise burned as fuel.

"Product" in the above refers to syncrude, diesel, naphtha, off gas or make gas, and other hydrocarbons produced from the oil shale. We have briefly considered both the HYTORT process for Eastern, lower-grade Devonian shales, and the TOSCO process for Western shales. The following discussion focuses on the TOSCO process.

Many energy-company holdings surround large tracts of federal land in Colorado, Utah and Wyoming; there are some holdings within the federal lands, and more leases will be sold. An HTR-Multiplex could be located more easily on federal land than on private land -- which would be no more than 15 or 20 miles from much of the oil shale recovery and processing now being undertaken or planned. The transmission of TCP energy and other energy forms is then easy and inexpensive. The establishment of a buffer zone around the nuclear plant should also be easy. The thermal capacity required in the HTR-Multiplex appears to be in the order of several thousand megawatts. For reliability, to accommodate growth most economically, and for other reasons, four 1.5-GW_t reactors might be built at one site rather than, say, two 3-GW_t reactors.

The underground mining and surface processing by the indirect-heat TOSCO II retorting and upgrading process is one of the most advanced of the many processes that have been proposed, and is believed capable of commercial operation. In situ and directly heated retorting processes will be similar in some respects, different in others. Input rates to the conventional process are 132,000 tons/day (10,374 MW_t) of oil shale and 500 MW_t (170 MW_e) of coal for an output of 100,000 bbl/day syncrude. If the electricity is provided from an HTR-Multiplex, the problems associated with handling and burning the coal, with stack gas cleanup, and ash and sludge would be avoided. In addition, more than 850 MW_t is generated by burning gas and oil derived from the shale oil to provide heat for the process.

Oil shale retorting yields a viscous, waxy, high-nitrogen, and moderate-level sulfur liquid product that is undesirable for transportation or storage. For this reason, development plants usually include an upgrading or hydrotreating process to treat shale oils before they are shipped to petroleum refineries. The upgrading involves heating, hydrogeneration, and possibly some cracking of the crude shale oil. The TOSCO II process includes some upgrading.

In order to improve the economics of transporting oil shale products to the ultimate consumers, the products should be refined into the final, most valuable form at a location as close as possible to the point of production of the syncrude and other plant products. Locating the refinery at the HTR-Multiplex site would satisfy this need. It would also make easily available to the refinery the high temperature output of the gas cooled reactor, for use in refinery processes in lieu of hydrocarbon fuels. A typical refinery uses about 10% of the input crude to produce the output products. It also produces significant air pollution emissions. Both the economics and the air quality of the refinery might be improved by the use of nuclear heat.

Use of nuclear heat could release for sale substantial amounts of hydrocarbon products which would otherwise be used in recovery and processing. Nuclear heat could also make unnecessary the development of facilities for transporting and handling coal and its waste products after combustion and scrubbing, or the installation of long electrical transmission lines, and could substantially reduce air pollution caused by combustion of oil shale products.

The extent to which various energy forms from the HTR-Multiplex might substitute for use of oil shale products can only be ascertained by further study of mining, retorting, upgrading, and refinery operations in some detail in order to provide a level of detail similar to that currently available on coal conversion processes.

Enhanced oil recovery (EOR) has several forms - these include steam injection, CO_2 injection, surfacant injection and in situ combustion of oil wells. The rationale is that only approximately 30% of the oil in the ground is recovered by normal oil field operation. To extract more requires some energy input into the crude, primarily to lower its viscosity so that it may flow through the porous rock structure which commonly constitutes the environmental medium in the well.

The application of the HTR in this case would be to supply the steam for njection into the well. In order to cover an entire oil field, the TCP concept might e employed to transport the high grade energy to distant portions of the field from a centrally located HTR.

5.4 SUMMARY

The use of an HTR for process heat requires that the reactor system be apportioned in some way for the simultaneous production of high temperature nuclear process heat (NPH) and for a lower temperature steam cycle. This is exemplified in Figure 5-30. It is here assumed that the primary helium is available at a core exit temperature of 950°C or 800°C, representative of the two prototypical graphite moderated HTRs under development. The return temperature of the helium is assumed to be 300°C as compatible with the multivarious requirements of core and core hydraulics design. It is assumed in this study that a natural split between the NPH and the steam cycle is a helium temperature of 575°C since the very best steam systems appear to be confined to temperatures below this(5-18). Thus the basic power cycle of a nuclear process heat HTR will be 58% (or 45%)* in NPH and 42% (or 55%)* in the steam "bottoming" cycle. Thus, as indicated in Figure 5-30, for 1000 MW_t in NPH the steam cycle must produce 733 MW_t (1222 MW_t)*.

The use of NPH does not imply the tapping of a single source temperature point. In fact any real application will involve several staging steps in which lower temperature steam heat as well as electrical energy from the steam cycle will be required. Furthermore, several applications of NPH will contain large recycles of steam, which are needed to effect chemical changes in vital processes. The consequence of these considerations is that for efficient use of NPH it is mandatory to return used (i.e., low availability) steam and condensate back to the steam bottoming cycle. The philosophy was previously shown in Figure 5-23. In that figure a typical leg of the steam bottoming cycle was shown in its interactions with a set of processes. The steam cycle is a source of low to high grade steam for the production of electrical work or for use in direct drive process steam turbines to drive compressors. A further fraction of the steam heat may be exchanged with

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*Figures in parenthesis indicate the lower helium exit temperature case.

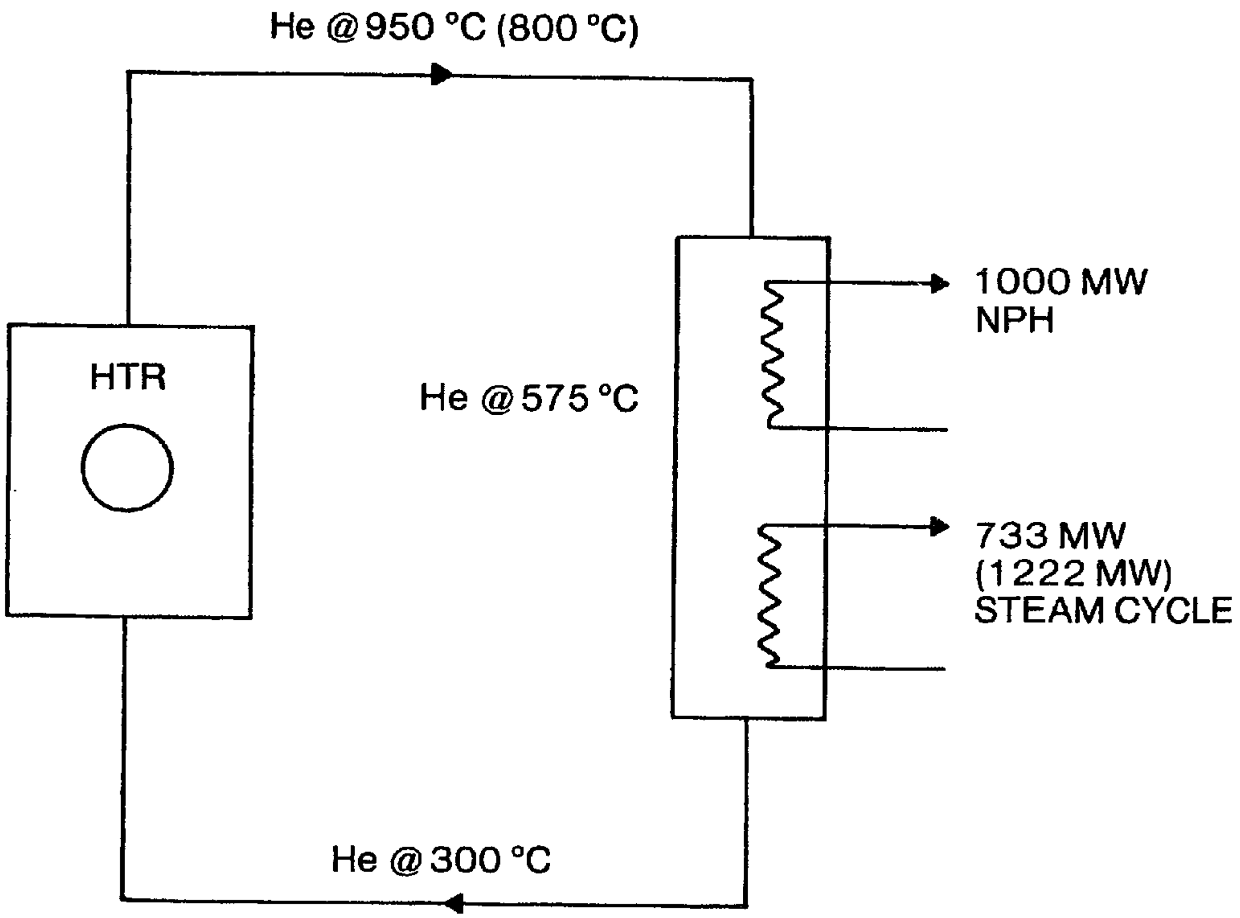

Figure 5-30. HTR WITH STEAM BOTTOMING CYCLE

the process with a significant return heat flow from the lower availability exhaust process steam flows. Finally hot condensate should be returned to the system as BFW.

The fact that chemical processes will be incorporated with a nuclear reactor has one interesting nucleonic interaction. The chemical plant will need maintenance-typically a steam reformer is shutdown every two years. While no allowance for refueling is necessary for a PBR the opportunity exists to match the burn-up (and hence enrichment) of a prismatic HTR with this shutdown cycle.

The key interface between the HTR and chemical manufacturing processes identified in this study has been the catalyzed fluid bed coal gasifier as exemplified by a development program at Exxon(5-6,-7). This gasifier uses a fluidized bed of coal catalyzed by a large addition of solid potassium carbonate. It can operate at temperatures down to 700°C. It is also unique among coal gasifiers in that its prime product is methane (SNG) rather than CO/H_2 (synthesis gas) as with conventional gasifiers. Apparently, in ways not understood, the addition of K_2CO_3 promotes the methanation reaction(s). The process is also unique in that the gasification step itself is almost athermal (i.e., requires no energy input) - although the process overall will consume considerable energy in preheat, catalyst recovery, recirculation H.P. and coal drying. The energy absorbtive step in this process will be in subsequent steam reforming of the methane to synthesis gas. This is the process which will absorb the high temperature NPH. This process operates best at the higher temperatures. For example the conversion of methane to synthesis gas using NPH at 950°C (825°C peak in the reformer) is about 65% with 3/1 steam recycle compared with only 50% at 800°C NPH (700°C peak in the reformer) with a steam recycle of 5/1. Furthermore, the steam recycle is a severe penalty for lower process temperatures in that for 1000 MW_t of NPH the higher temperature process uses 584 MW_t of steam heat and the lower temperature process uses 1222 MW_t.

It is evident from the above description that the catalytic coal gasification step is indirect and it is worthwhile to note the considerations that have dictated the need for this circuitious route. The primary criterion is that the transfer of heat in the range of 700-900°C into a coal gasification environment presents major materials problems. These problems are associated with the extremely corrosive and erosive atmosphere inside the gasifier. By adopting the indirect route there is no direct interface between coal and the nuclear helium circuit*. It is also to be noted that non-catalyzed coal gasification processes will operate at 900°C or higher(5-19) - which would present further corrosion difficulties as well as putting the process beyond the lower temperature HTR system.

The synthesis gas product is a key flexible resource. Its utility was indicated in Figure 5-12 in which a "Chemplex" was shown. There the symbiosis of the HTR and the catalytic gasification system was the basis of processes for industrial NPH, cogenerated electricity, SNG, hydrogen production, ammonia and fertilizer production, coal liquefaction to synthetic liquid fuels, refining crudes to gasoline, Fischer-Tropsch synthesis of methanol and gasoline and for highly fuel efficient steel production from high grade iron ore. Closely allied to the catalytic gasification/reformer/HTR system is the Thermochemical Pipeline (TCP) as described in Section 4.

The breakdown between NPH and the steam bottoming cycle has significant consequences with respect to the coupling to chemical processes. Table 5-24 gives a global breakdown of the energy flows for the major processes examined in this study. The ordinate of this matrix gives several entries as to the various MW_t required for each element. Thus for example, in the case of the TCP, 1003 MW_t of NPH is absorbed in the reformer unit. The total reactor (NPH + steam cycle) is based upon this size. This size of NPH demands that the steam cycle be 736 MW_t but the net steam flow requirement of this cycle is 177 MW_t**. Hence the exportable energy flow is 559 MW_t (or approximately 186 MW_e) and the total reactor will be 1739 MW_t. Likewise for the H-coal case, the NPH is 142 MW_t into the process and 559 MW_t for process steam. In this case the latter is controlling so that the corresponding high temperature section of the reactor is 762 MW_t for a total reactor of 1321 MW_t. This means that the process must export 620 MW_t of high temperature heat. This is highly undesirable but is indicative of the tight overall constraints put on the Chemplex with the HTR concept. In

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*The development of licensable alloys/ceramics for the nuclear reformer is itself not a minimal task.

**There are various shaft work processes to be considered so that the exergy flow them has been assessed at 3*MW_e. In effect the pessimistic assumption (5-20) that an electrical power plant operating at 33 1/3% efficiency using 1000°F steam has been made.

Table 5-24

Summary of Energy Flows Between HTR and Chemical Processes

	TCP*	Lurgi Nuclear	Catalytic Gas. 700C Gas. 825C Ref.	Catalytic Gas. 825C Gas. 825C Ref.	Catalytic Gas. 700C Gas. 700C Ref.	SRC	H Coal	EDS	NH_3 from Coal
MW_{th}									
Process Heat	1003	1563	1294	1324	1294	590	762	794	1354
Used	1003	846	1294	1324	1086	86	142	356	227
Exported	0	717	0	0	208	504	620	438	1127
Steam Cycle	736	1146	949	971	1581	433	559	582	993
Used	177	1146	756	820	1581	433	559	582	993
Exported	559	0	193	151	0	0	0	0	0
Total Reactor	1739	2709	2243	2295	2875	1023	1321	1376	2347
Nature of Product	MW_{th} or MW_e	SNG	Syn Gas	Syn Gas	Syn Gas	Boiler Fuel	Boiler Fuel + Fuel Oil	Fuel Oil + Naphtha	NH_3
Product									
MW_{th}	986	-	-	-	-	-	-	-	-
bbl/d	-	-	-	-	-	50,000	50,000	50,000	-
Btu/d	-	2.8×10^{11}	3.75×10^{11}	3.76×10^{11}	3.76×10^{11}	-	-	-	-
ST/D	-	-	-	-	-	-	-	-	3000
Total Consumed ST/D	-	12,000	12,000	12,000	12,000	18,300	19,200	19,500	2,100

*He @ 950°C

Figure 5-31, the example of the HTR coupled to a 579 MW_t TCP and a 30,000 bbl/day H-Coal refinery is shown. The ratio of the high temperature NPH (492 + 85 = 577 MW_t) and the steam cycle (87 + 336 = 423 MW_t) is correct with respect to the split in the helium cycle (Figure 5-30).

A Chemplex could be conceived using a larger number of components than the two in the above example. The key to the Chemplex system is in Table 5-24. The details in such numbers will change but their totality will determine the viable NPH processes with the HTR source.

Beyond the fuels industry, nuclear steel making and ammonia synthesis, the areas for NPH applications seem limited by either or both of temperature levels or the discretization of the demand pattern. Thus we have identified no other inorganic chemical process which could effectively use the NPH/HTR.

5.5 CONCLUSIONS

In the area of nuclear assisted coal processing, we have identified essentially one key coupling. This is the catalyzed fluidized bed coal gasifier as exemplified by the Exxon process(5-6,-7). Its advantages are that it operates at temperatures as low as 700°C and it does not require the direct transfer of heat into a corrosive atmosphere.

We have identified synthesis gas (CO/H_2) rather than SNG (CH_4) as the best candidate for export from the coal processing plant. This is because it represents the opportunity to add the high temperature NPH and to export this energy at a remote methanating plant and also export SNG at these points. Furthermore synthesis gas is a highly flexible resource allowing one to make hydrogen, ammonia, steel, petroleum and coal refinery products.

There is a tradeoff between the problem of developing higher temperature materials vs. the advantages accrued in the processes using the higher temperature heat. In this study, we have concentrated on the latter and find that there are compelling reasons to be attracted to the 950°C Helium reactor outlet temperature. When a T of 125°C is accepted across the reformer, the temperature levels favor the 950°C reactor outlet temperature for both the once through conversion

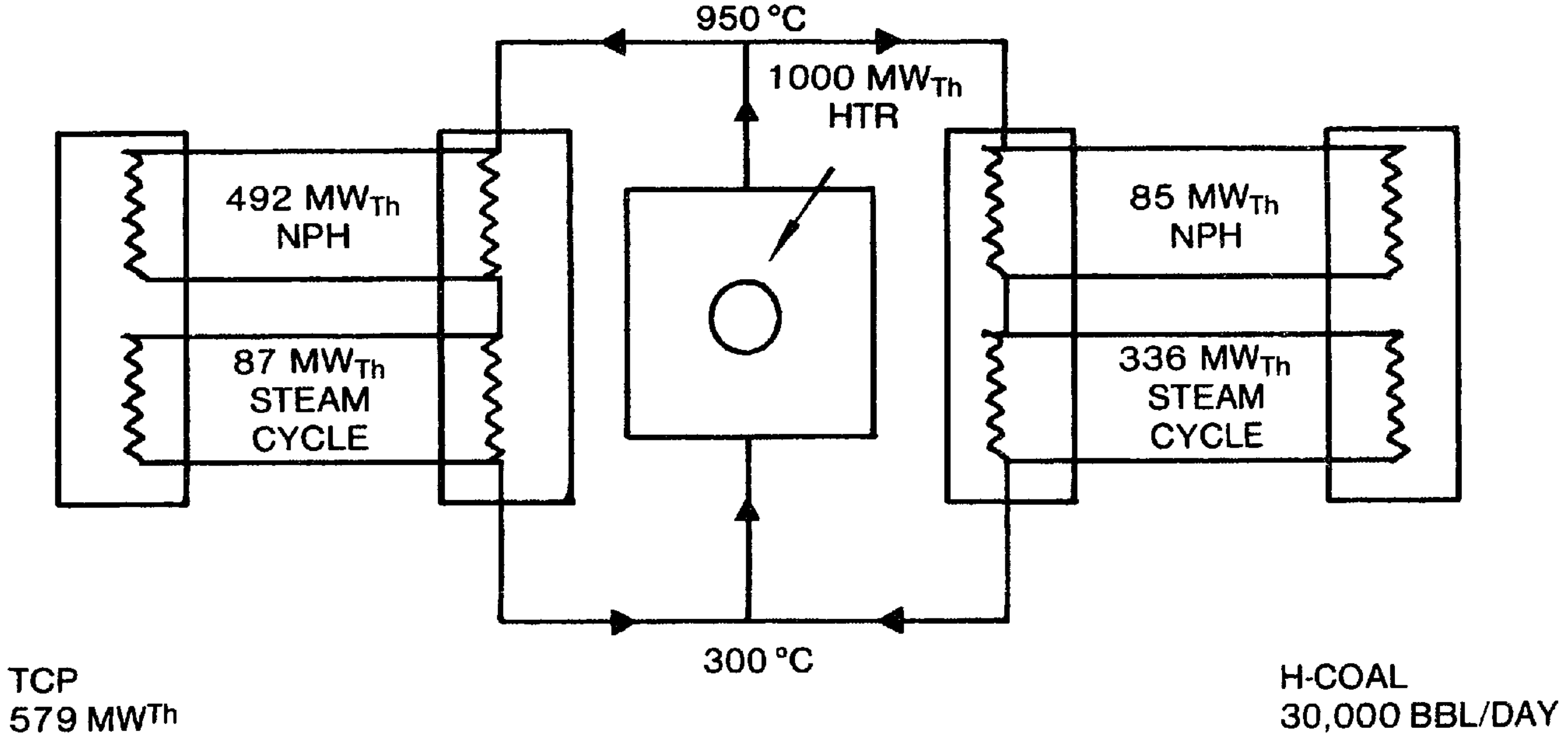

Figure 5-31. EXAMPLE OF COMPATIBLE HTR AND CHEMPLEX

efficiency (thus less inerts to pump or to recycle) and the quantities of steam raised to assist in the reactions. The steam utilization is a major factor in the methane based system.

Finally no direct heat applications of NPH have been identified. In the main the endothermic step has been beyond the reach of the presently envisaged HTRs. The use of NPH for direct preheat purposes is unlikely since there is more incentive to improve the existing processes by regenerative reheat than to centralize for a HTR source. Indeed, for several of the large energy users such as cement and lime manufacture, the location of the production facilities is optimized with respect to transportation. It is unlikely this pattern will change. For all of these reasons no further work was done on these processes.

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SECTION 6
HTR TECHNOLOGY

6.1 CURRENT STATUS

The high temperature reactor (HTR) uses helium as a coolant and graphite as the moderator, fuel cladding, and structural material. The temperature of the HTR is high only by comparison to the light water reactors. The HTR is capable of reactor outlet temperature around 950°C, whereas light water reactors have outlet temperatures around 290°C. Other utility fuels - coal, gas, oil -burn at temperatures on the order of 1650°C. There are two types of cores usually considered for the HTR, the pebble bed and the prismatic block. In the pebble bed concept, the fuel is contained in spheres which fill the circular cavity formed by the graphite radial and bottom reflectors. The spheres are loaded at the top of the bed and removed at the bottom. The refueling of the pebble bed is continuous and takes place while the reactor is at power. In the prismatic concept, the fuel is contained within large hexagonal graphite blocks, which are stacked into the core and arrayed to form a near-circular core. The active fuel blocks are surrounded by removable reflector blocks and then by permanent reflector. Refueling of the prismatic blocks occurs while the reactor is shut down, and a fuel handling machine loads and unloads the blocks from above. Generally, the core and the entire primary coolant circuit are enclosed in a prestressed concrete pressure vessel (PCRV). Many of the early HTR concepts used a single cavity in the PCRV to house the core and primary circuit, but most large conceptual designs utilize a multi-cavity PCRV, with separate cavities for the core and major components.

The HTR fuel element is a matrix of ceramic coated fuel particles in a graphite-coke-resin binder matrix. This matrix is made in the form of spheres for the pebble bed reactor and in the form of cylindrical fuel sticks for the prismatic design. The fuel spheres for the pebble bed reactor have a pyrolitic coating and are 6 cm in diameter. In the prismatic fuel design the fuel sticks, which are 1.6 cm diameter and 75 cm length, are loaded into 132 machines holes in a graphite block of hexagonal shape. These blocks are 79 cm long and 36 cm across flats; this is the block that is handled during refueling. This graphite block is called a fuel element and eight of these are stacked to give a fuel column. The graphite block has 72 machined coolant holes. Also the center block in the pattern of seven (hex-array) has two holes for the control rods.

Experience with the basic process of the fuel kernels and irradiation experience with this type fuel pertains to both the pebble bed and the prismatic type. Both of the designs have focused on the use of a fully enriched driver fuel kernel with thoria fertile kernels. These kernels are of different sizes so as to separate the kernels at the head end of the recovery process; the thoria fuel would go through a separate chemical recovery line. The thoria kernel is the larger kernel and is about 500 μm in diameter. Either reactor could use slightly enriched uranium with a single sized particle. The development of the process (ORNL, GA, Germany and England) is well along and is well suited to remote fabrication. In fact, ORNL and GA are developing the process equipment to be operated and maintained remotely. The basic fabrication process is to convert the fuel into a solution - usually a nitrate solution - and to load the solution onto a carrier, e.g., weak acid resin particles, and then heat the particles in a fluidized bed. Each particle is given an inner porous pyrolytic carbon coating and two high density carbon coatings. TRISO uses a third layer of high density silicon carbide coating.

These particles have been tested in capsules in the Dragon Project (British), at Peach Bottom and at Fort St. Vrain. Peach Bottom used a vented rod and the fission gases were collected and measured. To date overall fuel experience makes for an optimistic forecast. Statistical proof of large quantities of fuel operating at long exposures in a process heat reactor is years away. In Germany, the pebble bed reactor, AVR*, has operated for 10 years with helium temperatures up to 950°C. Many thousands of fuel pebbles have been tested in the AVR, with a total of 1.5 million individual fuel element movements through the core at full power operation. A large (300 MWe) German pebble bed reactor, THTR, is scheduled to start operation in 1982.

Both the prismatic and pebble bed reactors have top, bottom, and side reflectors in the form of graphite blocks. In the prismatic design, the top and bottom reflectors are formed by the hexagonal blocks identical to the fuel blocks; three blocks are used above and below the fuel blocks to form the top and bottom reflector. There are two types of side reflectors in the prismatic design. A row of replaceable side reflector blocks, similar to those used for the top and bottom

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*Arbeitsgemeinschaft Versuchsreaktor GmbH reflectors, completely surround the core and are handled the same as a fuel column when replaced. The second type of side reflector blocks are shaped to form the transition from a hexagonal block array to a cylindrical core configuration.

In the pebble bed reactor, the graphite reflector blocks form the boundaries of the pebble bed. The side reflectors are a cylindrical array of interlocking graphite blocks. The bottom reflector blocks are contoured to facilitate ball flow to discharge ports located in the bottom reflector. The top reflector blocks are suspended from the top head of the reactor vessel, permitting a small cavity above the pebble bed to accommodate variations in the bed level due to ball loading. The reflector blocks are not normally replaced in the pebble bed concept, however, the present design efforts in Germany are directed at accommodating some radiation damage to part of the side reflectors and provisions to remove the blocks if necessary. It is planned to be able to replace the side reflector blocks without greatly disturbing the pebble bed, but the entire core must be removed to replace the bottom reflector. The fuel cycle scheme for the advanced concepts is such that radiation damage to the bottom reflector is small and mostly restricted to the top part of the side reflectors.

The control rod systems for the prismatic and small pebble bed reactors basically operate by inserting rods containing boron carbide into holes through the graphite blocks, which in the pebble bed are in the reflector and are for fast scram to hot shutdown. In the large pebble bed concepts, fast scram to hot shutdown is achieved by moving the control rods to the top of the pebble bed. Long term cold shutdown requires inserting the control rods into the pebble bed. The THTR uses control rods with a smooth tapered contour on the leading end to force its way into the pebble bed. Prior to insertion, ammonia is injected into the coolant to reduce the friction between the control rod and the pebbles. The forces required to insert this type of rod into larger pebble beds appears to be too great, and a screw-type rod is being developed in Germany to reduce the forces during insertion. This type of control rod has a screw-type contour on the lower end, and the rod is rotated as it is inserted. Tests have shown that this type has reduced insertion forces, but there appears to be some problems with relocation of the fuel after repeated insertion of the control rods. In tests with a static pebble bed, repeated insertion of the control rod caused some of the fuel around the rod to have much higher burnup than normal and could lead to fuel failures. The present German shutdown concepts for pebble bed reactors have one system that provides fast and long-term shutdown and another system that provides backup long term shutdown. The first system is the rod type system described above. The backup system for long term shutdown is the insertion of small absorber balls (called KLAK) into the pebble bed. These balls are one-sixth the diameter of the fuel balls and fit within the interstices of the fuel balls. It is anticipated that the KLAK system would be manually operated in the event the rod system failed. The present German concept does not have a backup to the fast shutdown system.

Reactivity control in the prismatic design is also provided by two separate systems, control rods operated in pairs and the reserve shutdown system of absorber balls. The control rod pair are rods which are inserted into holes in the fuel block and operated by a control drive in a PCRV penetration. The control rod drives are electrically powered winches that raise and lower the control rods by means of flexible steel cables. Gravitational force acts to insert the control rods into the core during a trip. Each control rod is composed of articulated segments, and each segment consists of a metal container filled with boron carbide dispersed in a graphite matrix. A manually actuated reserve shutdown system utilizing boronated graphite pellets is provided for backup shutdown capability. The pellets, which are contained in hoppers located in the refueling penetrations, are released into a channel in the center column of each refueling region by an electrically actuated gate. The reserve shutdown system is sufficient by itself to achieve and maintain reactor shutdown from hot operating conditions to room temperature without the use of control rods.

The pebble bed reactor may also have an important safety characteristic - the ability to withstand a loss of coolant flow accident without fuel failure. This characteristic has been demonstrated in the AVR at Juelich, West Germany. The AVR coolant gas flow was interrupted by stopping the circulators at full power operation, and the control rods were deliberately prevented from shutting the reactor down. The reactor shut itself down without fuel failure to a power level in the kilowatt range. It appears possible that this characteristic can exist in large cores, particularly annular cores.

The pebble bed reactor operates on a continuous refueling basis, but the prismatic reactor must be shutdown for refueling. The refueling machine for the prismatic reactor sits over one of the penetrations in the top head of the reactor vessel, normally used for a control and orifice assembly, and replaces the fuel blocks in the vicinity of that penetration. The pebble bed reactor continuously removes fuel balls through discharge ports in the bottom reflector and loads balls onto the top of the pebble bed through ports in the top reflector. The early pebble bed designs, AVR and THTR, continuously recycle the balls through the core until target burnup is reached. For THTR, the balls normally make six passes through the core before being removed from the reactor.A small critical assembly below the core measures the burnup and then the ball handling system either recycles the ball or discharges it. The advanced pebble bed concepts utilize the OTTO (one through, then out) cycle, where the ball makes one pass through the core in about 3 years. The OTTO cycle greatly skews the flux and power to the top of the pebble bed where the fresh fuel is. An important advantage of the OTTO cycle is that, despite the 950°C helium temperature compared with 750°C for THTR, the fuel element center temperature in the OTTO cycle is not higher than that calculated for THTR.

The prestressed concrete pressure vessel (PCRV) is a key component of the high temperature gas reactor concept. The PCRV houses all of the major components of the reactor and primary heat transport circuit. The small HTGRs use a single large cavity to house the core and other major concepts. However, for large PCRV designs, the pod concept is used whereby the major components are located within separate smaller cavities. Each cavity is contained by a steel helium-tight liner. The liners are anchored to the concrete, cooled on the concrete side by coolant tubes, and protected inside by thermal barriers. The PCRV is constructed of high strength concrete reinforced by bonded reinforcement steel and prestressed by steel tendons. The PCRV liners and closures serve as primary containment for the reactor; the PCRV and secondary closures act as the secondary containment.

6.2 HTR IN PROCESS HEAT APPLICATIONS

The HTR is considered for process heat applications because of its high temperature capability, and the higher the achieveable helium outlet temperature, the greater the number of processes for which it is adaptable. There are two primary circuit concepts considered for process heat applications; one uses an intermediate heat exchanger (IHX) so that the primary coolant is not used in the process (indirect cycle), while the other includes the reformer and steam generator in the primary circuit (direct cycle). The use of an IHX in the primary circuit generally reduces the helium temperature available for the chemical process by about 50°C relative to the direct cycle. The reasons for using an IHX are both operationally and safety related. Safety considerations are affected since the IHX and the secondary helium system provides additional separation of the process gas circuit from the primary coolant boundary. This would be important for the thermochemical pipeline application where the process gas is transported off-site without intervening systems. In addition, using an IHX avoids bringing combustible substances (such as process gas) into the reactor containment. Locating the reformer outside of reactor containment permits easy accessibility to components requiring frequent maintenance (e.g., replacement of reformer catalyst.

Many of the process heat applications (see Section 5.0) require the use of a steam reformer to convert gaseous hydrocarbons into hydrogen, carbon monoxide, and/or carbon dioxide. Since a high temperature is desirable, it is preferable to include the reformer in the primary circuit. The duplex tube reformer mitigates some of the safety concerns by providing double wall separation between the primary helium and the process gas; it also can permit monitoring of the gap between the two tubes to determine leakage, if any. Since there is little use in the reformer for helium temperatures below about 570°C, a steam generator is used downstream of the reformer to utilize the heat unusable in the reformer. Figures 5-10, 5-11, 5-13 and 5-22 illustrate schematically some applications of the HTR using a reformer and steam generator. The schematic for the indirect cycle would be similar except that an intermediate heat exchanger would replace the reformer and steam generator in the primary circuit and a secondary loop containing the reformer and steam generator would be added.

A power plant for process heat applications would probably produce electricity for its own use, and, in most cases, would have electricity available for export. Table 6-1 shows the reactor power requirements for several major process heat applications: thermochemical pipeline, coal gasification, coal liquefaction, and ammonia production. Two coal gasification processes, the Lurgi process and the catalytic gasification process, are shown with several variations of the catalytic gasification process dependent on the temperatures in the reformer and

Table 6-1
REACTOR POWER REQUIREMENTS FOR PROCESS HEAT APPLICATIONS

	Thermo Chemical Pipeline		Coal Gasification				Coal Liquefaction			Ammonia From Coal
				Catalytic Gasification						
	$T_R = 825^oC$	$T_R = 700^oC$	Lurgi	$T_G = 700^oC$ $T_R = 825^oC$	$T_G = 825^oC$ $T_R = 825^oC$	$T_G = 700^oC$ $T_R = 700^oC$	SRC	H-Coal	EDS	
Thermal Power (MWt)										
Power to Reformer*	1003	1106	1563	1294	1324	1294	590	762	794	1354
Used as Process Heat	1003	1106	846	1294	1324.	1086	86	142	356	227
Unused in Process	0	0	717	0	0	208	504	620	438	1127
Power to Steam Generator**	736	812	1146	949	971	1581	433	559	582	993
Used as Steam in Process	177	139	1146	756	820	1581	433	559	582	993
Unused Steam in Process	559	673	0	193	151	0	0	0	0	0
Total Reactor Power	1739	1918	2709	2243	2295	2875	1023	1321	1376	2347
Product										
Nature of Product	MWt or MWe		SNG	Syngas	Syngas	Syngas	Boiler fuel	Boiler fuel and fuel oil	Fuel Oil and Naptha	Ammonia
Quantity Produced										
MWt	986	980	-	-	-	-	-	-	-	-
Bbl/D	-	-	-	-	-	-	50,000	50,000	50,000	-
BTU/D	-		$2.8 x 10^{11}$	$3.8 x 10^{11}$	$3.8 x 10^{11}$	$3.8 x 10^{11}$	-	-	-	-
ST/D	-	-	-	-	-	-	-	-	-	3000
Input										
Coal Consumed, ST/D	-	-	12,000	12,000	12,000	12,000	18,300	17,200	19,500	2100

Notes: T_R = Peak Temperature on Process Side in Reformer
T_G = Peak Temperature in Gasifier
* Heat Available at Temperatures Above 575°C (1067°F)
** Heat Available at Temperatures Below 575°C (1067°F)

gasifier. All of the coal gasification cases are sized to handle 12,000 tons of coal per day. Three processes are included for direct coal liquefaction, Solvent Refined Coal (SRC), H-Coal and the Exxon Donor Solvent (EDS), and are sized to produce 50,000 barrels of fuel per day. There is no relationship in the sizes used for the different applications. Table 6-1 shows the total power required for the reactor and the split between the power to the reformer and to the steam generator. The table is based on a 950°C reactor outlet temperature, a 300°C reactor inlet temperature, and a 575°C outlet temperature from the reformer. The 575°C reformer outlet is as low as practical for use in reforming, although higher temperatures could be used in the steam generator for superheating the steam.

Table 6-1 shows that the reformer power requirements determine the reactor size in the applications for thermochemical pipeline and catalytic gasification with a 825°C peak reformer temperature on the process side. In these applications, the processes cannot use all of the steam produced, so there is excess steam available for use in other processes or to produce exportable electricity. The remaining processes in Table 6-1 determine the reactor power on the basis of the steam requirements, and the processes do not utilize all of the high temperature heat. Although it would be possible to use this high temperature heat to provide some of the steam required and reduce the reactor power required, it would waste high temperature heat that could be used for other processes. As discussed in Section 5.4, an optimized plant would probably use more than one process or produce exportable electricity.

Other than the reformer and the high outlet temperature, there are no special requirements for the HTR in many of the process heat applications over what is needed for electricity generation using a steam cycle. However, there may be a few more reasons for the HTR in process steam applications to use an indirect cycle. It can be expected that there would be differences in the duty cycles, but it is not clear at this time that it would be an adverse difference. In fact, the duty cycle may be better for a process like the thermochemical pipeline than for electricity generation because it can be base loaded.

6.3 HTR DEVELOPMENT NEEDS

6.3.1 Summary Of The Nuclear Heat Supply System

Many of the development needs for the nuclear heat supply system are dependent on the type of core selected and the degree of information exchange with the Germans. The major development programs in the U.S. have been directed toward the prismatic core and steam cycle applications. Most of the pebble bed development has been in Germany. However, many of the development activities in both countries are generic to either reactor concept.

There is considerable experience with the fuel kernels of the HTGRs, but there has been no demonstration that the prismatic blocks can operate satisfactorily at the 950°C helium temperatures needed for the process heat application. The pebble bed fuel has operated with prototypic heavy metal loadings, power density, and burnups at 950°C in the AVR. Statistical proof of large quantities of fuel operating at long exposures and high temperatures would be needed for commercial operation. Development of higher heavy metal loadings would be needed for high converter/near breeder concepts. Long range commercialization would probably also require closing the fuel cycle by developing reprocessing of thorium/U-233 and fuel fabrication with recycled fuel.

Characterization of the graphite for long term exposure is also needed. The pebble bed concept has a somewhat greater need for graphite reflectors which can sustain the reactor lifetime fluence because it is more difficult to replace. In the prismatic design, the reflectors are replaced the same as the fuel. Development in the areas of seismic movement and vibration of the fuel and reflectors is also needed.

There is much design and development required in the areas of reactor shutdown and control, particularly for the large reactors. The development efforts for the pebble bed concepts might be more extensive because of the need to insert the control rods into the pebble bed for long term shutdown. Control rod temperatures are currently prohibiting the insertion of the rods into the pebble bed for fast shutdown. The present German schemes only provide functional diversity and redundancy for long term shutdown; it is not clear that this will be sufficient for U.S. applications. For the large pebble bed reactors, the forces to insert the rods into the pebble bed are quite large, so the Germans are developing turning rods which screw into the pebble bed. However, these results show that repeated insertion causes considerable displacement of the balls. This can cause some balls near the control rods to receive much higher exposures. Control rods are needed which can insert with low force without disrupting the bed. Absorber balls for emergency shutdown have been developed for both pebble bed and prismatic cores, but the development efforts are unique for each concept. In the pebble bed concept, the absorber balls (called KLAK) must flow between the larger fuel balls. There is more development needed in the insertion, performance, and removal of the balls.

The basic fuel handling technology is fairly well developed for both concepts (the prismatic in the U.S. and the pebble bed in Germany), but licensing of the pebble bed system in the U.S. must be demonstrated and may require some additional effort. The pebble bed concept may also need some additional development effort for the ball flow in large reactors.

There is considerable experience with PCRVs in the UK and FRG, and limited experience in the U.S. The major need for development is extending the concepts to the larger sizes and greater number of penetrations. Spanning the large cores may be a major design problem in developing the commercial size plants (around 3000 MWt). There is also development needs for the liners and hot ducts. The present concepts for the liner utilize a thermal barrier within the liner, so that the liner cannot be inspected. It may be necessary to develop inspectable liners. The hot ducts must be developed to contain the high temperatures needed for the process heat application.

6.3.2 Status Of Nuclear Heat Supply Components

The following sections describe the status of the reactor plant components, assess the concerns of the requirements for the components relative to the state-of-the-art, and recommend the development needs of the components. These sections summarize the development needs of the components that will have to be fulfilled to provide a timely introduction of the HTR plant into the U.S. market and to derive a design and manufacturing capability within the U.S. The components that are discussed in the following sections are listed in Table 6-2 under the

Table 6-2

Nuclear Heat Supply
- a. Steam Reformers
- b. Catalysts
- c. Steam Generators
- d. Intermediate Heat Exchangers
- e. High Temperature Reactor Helium Turbine
- f. Helium Circulators
- g. Hot Gas Ducts
- h. Reactivity Control System
    - Control Rod System
    - Nuclear Instrumentation

Auxiliary Systems
- a. Fuel Handling
- b. After Heat Removal
- c. Gas Purification
- d. Process Control & Instrumentation

Balance of Plant
- a. Chemical Heat Pipe
- b. Hydrogasification
- c. Steam Gasification
- d. Methanator

} Not discussed in this report

Materials

NSSS Base Technology
- Fuel & Fuel Cycle & Fuel Process
- Graphites
- Coolant & Fission Product Chemistry
- PCRV & Structural Technology
- Safety & Reliability
- Structural Materials
- Analytical Methods

} Not discussed in this report

heading of Nuclear Heat Supply along with other components that require consideration in a development program.

a. Steam Reformers

The steam reformer concept is being actively developed by the Federal Republic of Germany. They have accumulated a significant amount of experience in chemical and nuclear steam reformer design and experimentation. A great deal of experience with conventional steam reforming exists in the U.S. A background and summary of the FRG experience follows:

- Conventional (nonnuclear) steam reforming uses sulfur-free fuel combustion for heating primarily by radiation at a maximum flame temperature of approximately 1500°C. The reforming temperature is in the range of 750°C to 850°C with a pressure of 1 to 30 bar and with a steam/methane ratio of 2:1 to 5:1. A summary data comparison of nuclear and "conventional" steam reforming plants is shown in Table 6-3. The process is technically well developed today and is applied worldwide for the production of gases for ammonia and methanol synthesis as well as H_2 production for hydrocracking processes.
- The selection of reforming operating parameters is a complex evaluation of equilibria methane conversion factors and reaction kinetics influenced by the subsequent use of the product H_2 or H_2 + CO mixture. In general, for all applications which require a high operating pressure, a high reforming pressure is advantageous (compression energy can be saved by compressing the gas before the steam reforming process). A disadvantage of increasing pressure is that the unreformed methane content of the product gas will increase with increased pressure. As an example, for typical parameters of temperature and H_2O/CH_4 ratio, an increase in reforming pressure from 30 b to 40 b will decrease the CH_4 conversion by about 10%. Obviously optimization is required.
- At the operating regions of interest (i.e., temperature of 600 to 800°C, pressure of 20 b to 30 b, H_2O/CH_4 ratio of 2.5 to 3), the reforming reaction rates are found to be limited by heat flux to the reformer tube. Therefore, the thermal-heat transfer characteristics of the reformer are extremely important. Trade-off studies have been made of heat transfer

TABLE 6-3

COMPARISON OF CONVENTIONALLY HEATED (FLUE GASES) AND NUCLEAR HEATED (HELIUM) STEAM REFORMERS

Parameter	Conventional Plants	Nuclear Plants
Tube length	8....12 m	10 m
Internal diameter	100 ... 150mm	100 mm
Wall thickness	15 ... 20 mm	15 mm
Product gas removal	Outside reformer tube	Within reformer tube
Reforming pressure	1 .... 25 b	40 b
Reforming temperature	800 ... 850°C	800 ... 850°C
Heating side pressure	1 b	40 b
Heat transfer	Radiation	Convection
Space utilization	1 tube/m^2	45 tubes/m^2
Max. heating temperature	1400 ... 1500°C	950°C
Max. tube wall temp.	900°C	900°C
Max. pressure difference across tube wall	0 ... 25b	1 bar (hot part)
H_2O/CH_4 ratio	2/1 ... 5/1	2/1 ... 5/1
Mean Heat flux	60000 kcal/m^2h	60000 ... 70000 kcal/m^2h
Heat flux max/min	10/1	1.5/1
Rate of gas flow	50000 $Nm^3 H_2 + CO/m^2h$	50000 $Nm^3 H_2 + CO/m^2h$
Service life aim	100,000 h (60,000 h attained) today)*	100,000 h
Reformer tube materials	G-X40 CrNiNb 2524 (W.-No. 1.4855; IN 519) G-X45 NiCrCoWNb (IN 643) G-X45 NiCrCoWNb (IN 638)	To be determined
Product gas tube materials	Incoloy 800, Incoloy 807	

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*Individual tube life can be considerably shorter, but in conventional plants repair is relatively easy.

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coefficients and pressure drops on the process and helium sides versus gas velocities to provide size and helium pumping power requirements. It was also found that utilizing an inner gas return duct for hot product gas down to approximately 650°C can be used to transfer heat to the catalyst filling and leads to an approximate 20% increase of the heat transferred to the reformer tube.

- Stress analyses of reformer tubes in the reference were conducted for internal pressure, external buckling, thermal stress during startup and shutdown with a temperature transient of 1°C/min, and thermal stress during operation. Results indicated that the reformer tubes are capable of satisfying the conditions evaluated. Neither a fatigue analysis of the tubes nor an analysis of the carrier plate support for the tubes was provided. Based on this, it can be generally concluded that the design can be shown adequate if materials characteristics are known. Low and high cycle fatigue and creep/fatigue characteristics, however, must be demonstrated.

- Materials properties under actual operation conditions have been and are continuing to be evaluated, and this appears to be the greatest area of uncertainty. Creep properties in the 900°C range, hydrogen permeation through the reformer tube wall into the primary gas stream, and the corrosion effects of the reactor coolant on candidate tube materials and their possible control are all areas of concern. The problems and concerns of tritium permeation from a systems standpoint must be demonstrated and licensability must be determined. The design criteria and design code for use of properties must also be developed and accepted for the proposed materials at the temperatures of interest.

Data for steam reformer operation in a helium-heated loop simulating nuclear plant operation have been and are being established through operation of the EVA-I plant at KFA in Juelich, Germany.

Work to date shows that the helium-heated steam reforming of methane is basically possible and has established fundamental design parameters. However, the design of a large tube bundle has to be proven. This testing is planned in the EVA-II facility at KFA. That facility is scheduled for completion and checkout in late 1979. A 30-tube steam reformer test section is planned with 3 tubes each of Incoloy 800H, Incoloy 802; Manuarite 36A and the remainder of Inco-519.

In summary, it is apparent that considerable basic work has been done by the Federal Republic of Germany to determine the design criteria for steam reforming processes. Important parameters and relationships between temperature, pressure, H_2O/CH_4 ratio, and reformer heat transfer characteristics have been estalished. What remains to be done is the considerable effort to provide a steam reformer design that will satisfy the manufacturing, opreating, maintenance life, and safety requirements of a nuclear plant installation. Of immediate concern are the properties of the materials to be used as affected by the service conditions. In the longer term design problems of weld joint fatigue, tube vibration, insulation attachment, and flow-induced vibration, large-diameter expansion joint design, transient safety analysis, and the general area of maintainability must be resolved.

b. Catalysts

The incentive for developing an alternate catalyst for the nuclear process heat plant is to remove two limitations of conventional reformer catalyst and tube designs. First, the life of conventional catalyst is between two and eight years, which is much less than the 30-40 year design life of the entire plant. Since replacement of the catalyst results in costly plant shutdown (approx. $250,000/day plant downtime), a catalyst that can be reactivated in place would reduce operating costs. Secondly, the nature of the pellet catalysts used in conventional reformers requires the reformer tubes to be rather large (90 mm minimum). Because of the large tubes, the reformer is a low power density device, which in a nuclear process heat application represents a significant cost to the system. In order to reduce capital costs, it is therefore desirable to minimize the size of the reformer.

Performance of the large diameter reformers using pellet catalysts and heated with helium is limited by heat transfer rather than reaction diameters. As a result it is possible to make the steam reformer assembly more compact and to achieve significant size and weight reductions by reducing the steam reformer tube diameter to increase the heat transfer surface per unit of core volume. In addition when the tube diameter is reduced the tube wall thickness can also be reduced, which increases the thermal conductance. The effect of reformer tube diameter on tube bundle diameter, overall heat exchanger pressure vessel diameter, and on the total reformer/steam generator heat exchanger weight is shown in Figure 6-1.

Recent experiments at the Foster Wheeler Development Corporation (FWDC) in the U.S. and at the KFA in Germany have shown that it may be possible to design a catalyst for easy removal or even reactivation in place. Also these new geometric catalyst designs have the potential to significantly reduce the reformer size and increase its power density. The EVA test facility at the KFA in Juelich, West Germany is well suited for testing a full scale helium heated reformer tube and the Germans have indicated a willingness to test a reformer containing an alternate catalyst. The ultimate goals of a catalyst development program are:

1. To reduce the steam reformer tube size in order to reduce the assembly size, weight and cost.
2. To reduce the catalyst replacement time and cost.
3. To determine limits for reactivating the catalyst in place in order to increase operating life.
4. To determine feasibility of periodic inspection of steam reformer tubes.

c. Steam Generators

Information on helical-type steam generators is available from a number of gas-cooled reactors in France and Britain and from Fort St. Vrain in the U.S. It is considered that some testing, however, is necessary to confirm the feasibility of the proposed straight tube concept. Vibration and flow distribution testing is needed. The requirement for in-service inspection (at least of the superheater tubes and the supporting structure) will necessitate changes from previous design experience. In the available literature, no stress analyses for the stationary and transient operations are given. Hence, the thicker walled supporting structures should especially be analyzed. It is stated that ferritic steels are used in the helix bundle and that ferritic steel and Incoloy 800 are used in the hot part of the superheater. The Incoloy material needs further qualification for long-term applications in helium circuits, especially for a temperature of 800°C. This work has started as a part of the FRG national program for the PNP Project. Methods

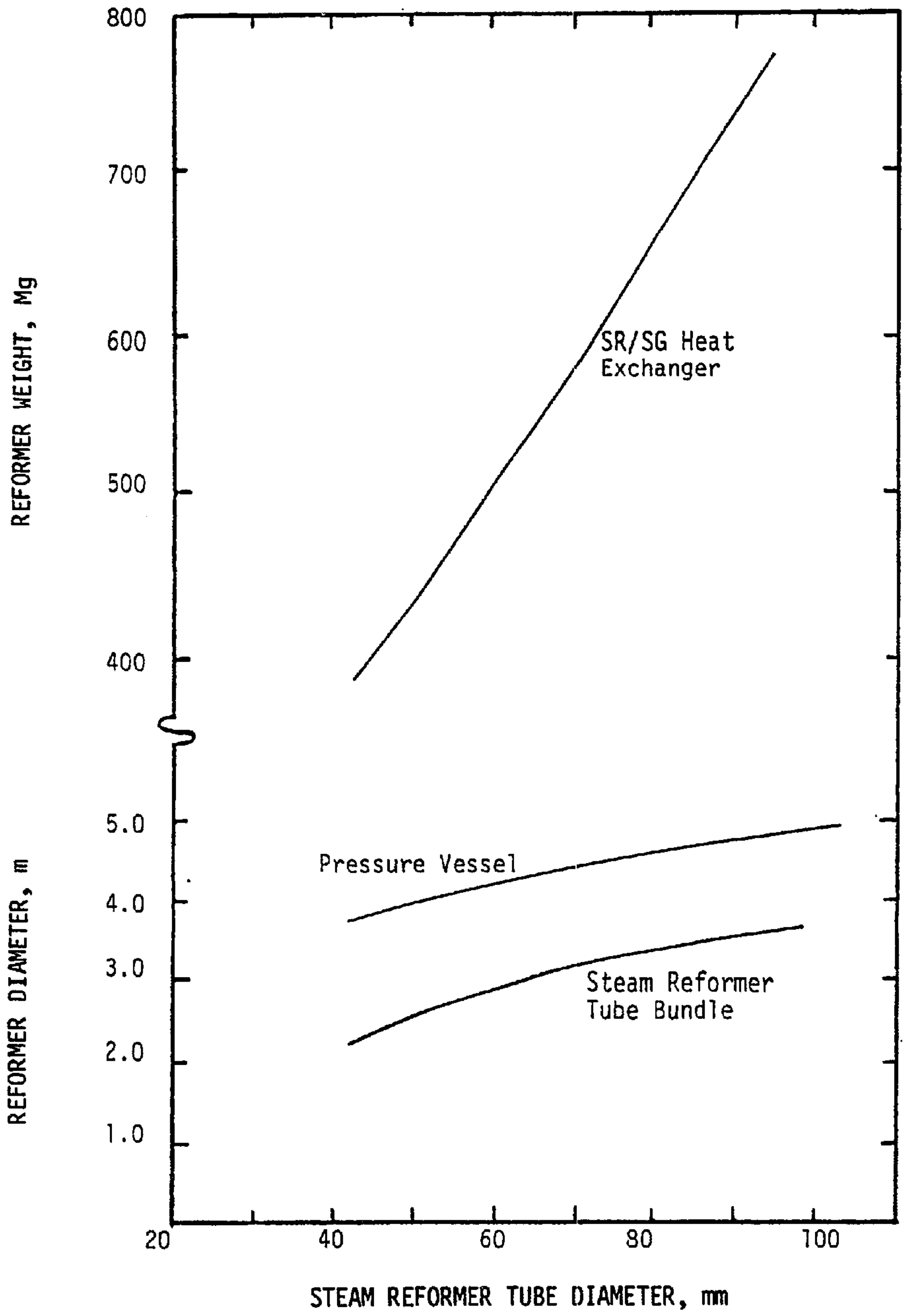

Figure 6-1. HEAT EXCHANGER DIMENSIONS AND WEIGHT VERSUS TUBE DIAMETER

for in-service inspection are available today for straight tubes using knowledge from the field of light water reactors. Helical bundles today can only be tested if their length is less than 20 m and if the number of turns is no more than two. The tubes in the designs are much longer and cannot be tested with current techniques. The future requirements of safety authorities in this field are unknown, and perhaps further change to straight tubes in the economizer and the evaporation may be required. This would cause more space to be needed for the steam generator within the multicavity vessel. It can be generally stated that the basic elements of the proposed steam generators are known; however, some confirmatory tests remain to be done.

d. Intermediate Heat Exchangers (IHX)

The processing plant used to convert coal to methane by steam gasification, unlike the hydrogasification plant, utilizes an intermediate circuit to separate the reactor plant from the rest of the gasification equipment. This intermediate loop uses helium as an energy transport medium, and therefore, requires a He-to-He heat exchanger between the secondary loop and the reactor primary helium system. The He-to-He heat exchanger is located within the PCRV.

Two alternate design concepts for the intermediate heat exchanger are being evaluated in parallel in the Federal Republic of Germany. One plant design consists of 24 exchanger units, where 4 units are provided for each of 6 primary circulating loops. These units are of helical tube construction. The other plant design consists of 12 exchanger units each of 8 U-tube modules with 2 units provided for each of 6 primary circulating loops. The primary advantage of the helical tube units is that they are of such size that all construction can be done in the fabrication shop and no field welding is required. The advantage of the U-tube-type IHX is that the tubes are more accessible for visual inspection, and therefore it is easier to find and isolate a failed tube.

A comparative evaluation has been made of the results of the helical and U-tube IHX that resulted in a U-tube configuration being rated superior to a helical IHX in all categories, which consisted of safety-related mechanical design, thermal hydraulic design, and cost aspects. The major advantages of the U-tube exchanger included ease of in-service inspection, ability to replace a module, leaky tube isolation, and cost differences resulting from less tube weight and significantly smaller tube sheets. One of the major inspection features was the ability to boroscope-inspect the U-tubes without opening the primary loop. This advantage is not available with the current FRG design, but could be incorporated with redesign of the secondary loop flow. An advantage retained in the FRG design is the ability to pressure or vacuum test for and then isolate leaks by plugging both ends of the faulty tube. Inspection models are being developed for the helical type exchanger, but are much more complex and the sensitivity is reduced.

A problem in most U-tube heat exchanger designs is the thermal stress and deflection due to restraining the ends of the U-tube considering the difference in thermal growth rates between the cold and hot legs. This has been greatly reduced in the FRG design through the use of a constant load hanger to support one end of the U-tube and by making the cold leg of the exchanger longer than the hot leg to reduce the differential expansion

The helical tube IHX can be completely assembled and tested in the fabrication plant and requires a minimum of field assembly and testing. In contrast, the U-tube IHX has a large number of joints to be made in the field. The current design shows seven flange joints for each module plus eight sliding seal joints for the total exchanger. These joints could become potential problems.

The problem of tube vibrations will appear in each of the tube designs and must be evaluated and the results substantiated by test. Since details of the designs (such as tube length between supports and the method of tube spacing/support) are not available, no evaluation can be made of this concern other than that it must be considered.

Similar to the situation for the steam reformer, the major effort remaining is to provide an intermediate heat exchanger design to be tested in sufficient size and power rating to demonstrate the operating, maintenance, life, and safety requirements of a nuclear plant installation. Efforts are underway in FRG to perform such large-scale tests. Of immediate concern are the screening and selection of candidate materials for IHX components at the high temperatures for the long life and special environment required and providing the materials properties required for the design.

e. High Temperature Reactor Helium Turbine

Since the high temperature reactor helium turbine (HHT) is not a component in the process nuclear heat plant that is being proposed, there will be no direct involvement in development of such a component. However, the selection and development of the HHT in Germany will have to be closely followed and understood, since the development needs are different relative to the requirements placed on the components for a process heat plant.

The selection of the HHT in Germany has made the high-temperature reactor program somewhat of a long term research and development activity. Even though there are many areas where development needs will allow common HHT and process heat plant programs, such as PCRV technology, graphite structures, and fuel handling systems, some development programs will be different. Therefore, care will have to be taken in working cooperatively with the FRG to separate the development needs of the process heat plant components from the development requirements that are unique to the HHT concept. For example, the liner concepts are different with different development needs; after-heat removal requirements will be somewhat different; even the fuel requirements may have to be different, due to the HHT requirements of extremely low reactor coolant contamination levels. Materials development for many of the HHT components will have to be somewhat different, since the severe conditions that must be considered in HHT are not the same as those for the process heat plant, particularly during transients.

f. Helium Circulators

The design of the helium circulators are in the advanced stages of a preliminary design in FRG with further work awaiting decisions on the specifications on the final plant and the establishment of final product requirements. Both of these items are necessary to establish the size and operating conditions. Of particular concern will be development of the off-design and transient operating conditions for both normal and faulted operation, which are used to establish inlet temperature and pressure variational conditions. FRG analysis indicates that 70 to 80°C overtemperature is the maximum attained during any mode of operation.

The general layout and design features of the proposed circulator are based on scaling up the design for the THTR circulators but without the THTR variable-speed motor. The THTR circulators, while not yet proven in reactor operation, have been built and tested in a simulated environment. Similar circulators using both fixed and variable-speed motors have been designed and operated as part of the United Kingdom CO_2-cooled Magnox and AGR reactor programs, at much smaller power levels, however. A summary of the THTR, Dragon and other helium circulators is shown in Table 6-4. In addition it is reported by the Germans that an electrically driven circulator of 11 MW has been used by Great Britain in their CO_2-cooled reactor work. Based on this experience and background, no "barrier" problems are anticipated in the development of the 3000 MW_t system circulators, providing that transient and off-design conditions do not significantly exceed the steady state conditions.

Similar to the circulator design for the THTR, the 3000 MW_t system unit should be designed as a module for ease of removal and replacement. Where cavity liner diameter is not of critical concern, as in the hydrogasification design, the circulator and closure assembly may be removed as a unit. Where space is more limiting, the unit may be removed by removing the closure head, unbolting the mounting plate from the liner flange using the extended bolts provided, and lifting out the unit. Note that it would be possible to separate the main blower assembly from the outlet diffuser for ease of assembly.

A separate test facility in FRG, which should be operating, is the HHV facility (High-Temperature Helium Test Plant) located at KFA. The test loop is designed to test out a closed-loop helium gas turbine. The test will also provide useful high-temperature component data and some materials information including operational experience with the auxiliary circulators, which are nearly equivalent to the circulators required for a 3000 MW_t plant.

g. Hot Gas Ducts

The major hot duct problems appear to be in the selection of materials for the environment and long life requirements. Particular emphasis must be placed on obtaining design experience with ceramic ducting components. It is considered that partial ceramic construction will be needed because of the severe thermal and

TABLE 6-4

SUMMARY OF GAS CIRCULATOR CHARACTERISTICS

Data	Dragon	Dragon Refer.	Peach Bottom	Fort St. Vrain	770 MW_e (Summit)	1160 MW_e (Fulton)	AVR	THTR
Medium	He	He	He	He	He	He	He	He
Number of Circulation	6	13	2	4	4	6	2	6
Driving Power, MW	0.07	1.26	-	3.88	10.8	10.7	0.128	2.5
Kind of Drive	electr.	electr.	electr.	steam	steam	steam	electr.	electr.
Mass flow, kg/s	1.8	32.4	27.7	110	235	235	8.5	50.7
Rotational Speed, RPM	1100-12000	6000	-	9550	6750	6750	4400	5600
Inlet pressure, bar	8.25	40.2	24	46.4	45.9	45.9	10.2	38
Pressure rise, bar	0.45	-	-	0.96	1.43	1.43	0.083	1.22
Inlet Temp, °C	350-420	325	343	395	-	330	175-330	250
Circulator design/stage	radial/1	radial/1	radial/1	axial/1	axial/1	axial/1	radial/1	radial/1
Circulator arrangement	horiz.	-	horiz.	upright	upright	upright	horiz.	horiz.
Regulation	speed	initial whirl throttle	-	speed	speed	speed	speed	bypass-frequency variation

environmental conditions and the poor high-temperature characteristics of most weldable and formable metallic alloys. In addition, testing of the selected materials and configuration must be done to evaluate the effects of flow and acoustically induced vibration, depressurization transients, and long-term effects on conductivity. Component testing is required to select the optimum design of expansion compensators, attachment devices (particularly to the core reflector), and flange connectors for ready removal and service. Some preliminary testing has been done in regard to thermal insulating properties and depressurization in FRG but apparently at reduced size and velocity, and substantially more test and development work must be done before a design can be selected.

The FRG design of the hot gas duct has been finalized to the extent of general configuration; that is, the use of concentric hot/cold ducting with cold gas cooling of the structure has been decided on, but final materials selection and details of attachments and expansion devices are not yet firm. One duct design concept, which was for lower temperature use (700°C), was not actively pursued.

The FRG design appears to be a logical extension of previous gas reactor design experience in both FRG and elsewhere. FRG experience at similar temperatures was achieved with the AVR, where the coolant actually attains a temperature of 950°C. In the HTR test reactors (Dragon and Peach Bottom) and in the Windscale AGR, coaxial ducts are used and cooled in counterflow by cold gas. However, the maximum helium temperatures are only 750°C, and extrapolating to the present design concept is difficult. In addition, these ducts have small dimensions compared to the large-scale plants.

In the THTR (300 MWe), hot gas is transported by a hot duct between the reactor core and the heat exchangers, with insulation within the ducts made of metal foils. Since all the primary circuit components are within one large cavity in the prestressed concrete vessel, the problem of insulating the vessel itself against hot gas does not occur.

h. Reactivity Control System

Several reactivity control systems have been proposed and are being developed in FRG by HRB, KFA and GHT.

Table 6-5 summarizes the reactivity control systems as currently designed by HRB, KFA and GHT. The following table illustrates the different means which the three companies have chosen to meet the FRG licensing requirement for two independent shutdown systems. Both HRB and GHT have specified the first shutdown system as the primary facility for fast shutdown, long-term shutdown, and for the control of the power level and distribution during normal reactor operation. An alternate shutdown system consisting of the KLAK absorber spheres is utilized for emergency shutdown. The alternate system is planned to be used only 2 to 4 times over the life of the plant. The HRB and GHT control systems are in contrast to the KFA concept in which two different control drive mechanisms are used to satisfy the requirements for independent systems. At present the HRB control design is the reference system for the HHT plant and the GHT design is the reference system for the PNP plant. The KFA control system currently serves as a backup concept.

Table 6-5
CONTROL CONCEPT SUMMARY

	Total System	**First Shutdown System**	**Second Shutdown System**
HRB (HHT reference)	108 core rods	108 lifting rods (pneumatic drives)	KLAK*
	42 reflector rods	42 reflector rods (gravity/electric drive)	
KFA	168 core rods rods	36 lifting rods (penumatic drives)	132 rotating rods (spindle drives)
GHT (PNP Reference)	156 core rods	156 rotating rods (spindle drives)	KLAK*

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*The control rod system and the KLAK system are capable of independently shutting down the reactor.

Control Rod System - Two generic issues concerning the lifting rods are the forces of insertion into the pebble bed and the quality of position indication. The first issue has been resolved to the satisfaction of the three FRG companies through scaled experiments done by HRB. The experiments were performed in 1:6 scale model with 198 rods and a bed of graphite spheres. The analysis of these tests indicated that the insertion forces were similar to those of THTR and that the forces were within acceptable limits. The issue of rod position indication still remains open, especially for the pneumatic drive. Development work is being done in this area by GHT.

The rotating rod with a spindle drive has been completely designed, and several components of this drive have been tested — in particular, the spindle and spindle nut, the bearings, and the penetration through the PCRV for the drive shaft. One complete rotating control rod has been ordered for a feasiblity test which is scheduled to begin in 1979 at KFA. Coincident with this feasilibity test, a detailed design of a prototype rotating rod will begin. The prototype design is scheduled for testing during 1981. Thus, further design and a good deal of testing are scheduled for this rod design, which would be completed at the earliest by 1982. The rotating rod has a major advantage over the lifting rods because its use results in no net compression forces in the pebble bed, and it does not require the use of ammonia injection. Also the rotating rod has no difficulty in achieving the required accuracy in position indication.

The KLAK system proposed for the PNP second shutdown system is similar to an emergency shutdown system used in the GA HTGR. In the GA system, small, high-absorbing balls held in containers are released to fill channels in the prismatic blocks. If it is available, data from the development of the GA system would be useful in developing the KLAK system. At present, the KLAK system is in an early design stage. The inlet system has been considered; however, a means of modifying the core bottom to facilitate removal of the KLAK has not been considered in detail. The behavior of the KLAK balls in the pebble bed has been studied. Also, the earthquake behavior of KLAK is being studied at the University of Aachen. However, the entire KLAK system will have to be tested once it is completely designed.

The development needs of the KLAK system would be significantly changed if the system were used as an emergency shutdown system with a probability of use of approximately 10^{-6} per reactor year. Using the KLAK in this manner has been proposed for the unified control concept. Due to the low probability of usage, no special modifications of the core bottom for removal of the KLAK would be required. Also, complete periodic testing of the system would not be needed as it would be if the KLAK system were used as a second shutdown system. Elimination of the requirement for complete testing of the system could be an advantage in overall plant availability.

When the control rods are inserted into the core, they will be exposed to high temperatures. It is well known that most alloys lose toughness and ductility after extended exposure at elevated temperatures, particularly between 550-950°C. Strength may be either increased or decreased due to such exposures. Of the proposed top candidate alloys, Incoloy 800 should be affected the least. In this regard, Inconel 625, although considerably stronger than Incoloy 800, suffers considerable loss of toughness and ductility (particularly lower temperature toughness and ductility - below about 450°C after exposure between 650-800°C), and the properties of Inconel 519 would be expected to fall somewhere between those of Incoloy 800 and Inconel 625. These materials appear to carburize extensively, which can significantly reduce toughness and ductility. Neutron exposure, particularly to fast neutrons, also results in loss of toughness and ductility.

Because of the low oxygen potential (or more simply, very low ratio of oxidizing to reducing species) of reactor-purity helium, the types of oxides formed on most alloys at high temperatures in air are not stable. For this reason most contact between alloys will be actual metal-to-metal contact which, under conditions of high temperature, high contact stresses, and long times, can result in self-welding. When relative motion is required under such conditions, severe galling or seizing (and self-welding) is possible.

Nuclear Instrumentation-Design of a nuclear instrumentation system for a large PBR has proven to be problematic, due to the difficulty involved in using in-core detectors. Hence, a design which uses ex-core detectors is being pursued. Presently, the system utilizes detectors between the side reflector and thermal shield to measure the leakage through the side reflector and provide an indication of the axial power distribution. The radial power distribution is monitored by fast flux detectors located in the upper reflector. The fast flux is measured because the thermal flux in the upper reflector is not representative of the power distribution due to the effect of the empty space between the core and the top reflector.

The purpose of nuclear instrumentation is to provide information for the reactor protective system, in addition to the measurements necessary for reactor control during all normal operating or transient conditions. In the case of large pebble bed cores with low power density and an OTTO fuel cycle, new techniques are needed for measuring the flux distributions, because in-core measurements are difficult in a pebble bed, and large cores have a tendency toward xenon oscillations. In the pebble bed core, axial xenon instability is limited by the relatively low core height, and by the highly peaked axial power distribution. However, analysis has shown that azimuthal oscillations can occur because of the very loose coupling of the local power distribution over the large core diameter. Hence, an essential task for the instrumentation is the detection of radial and azimuthal xenon oscillations in the upper half of the core.

6.3.3 Status of Auxiliary Systems

A brief discussion of the status of the components that comprise the auxiliary system is given in the following sections.

a. Fuel Handling

All the components required for the fuel handling system are well known and tested in the FRG from AVR operational experience and from the THTR development program. Also, the fuel handling systems for the large PBR operating on an OTTO fuel cycle is simpler than the systems required for the AVR and THTR, since it does not recirculate fuel elements. Thus, the fuel handling system could be built and is available at this time. Consequently, there should be no need for development of these components other than the development of manufacturing capabilities within the U.S.

b. After-Heat Removal

These systems, as presently envisioned in FRG, for PNP and HTR-K are founded on sound engineering principles. The design used for the HTR-K system has been widely used for other gas-cooled reactors both in the FRG and the U.S. Therefore, it does not seem to offer significant developmental uncertainties. The PNP design has not as yet been qualified against the spectrum of plant accidents possible for the PNP concept. The new dimension of transients, although potentially severe, does not appear to pose unsolvable developmental problems. The FRG reference concepts incorporate one more level of redundancy than required in the U.S. The design and testing of an after-heat removal system may be desirable.

c. Gas Purification

Extensive experience has been developed in the U.S. for purification of helium coolant from the design and operation of the Peach Bottom and Fort St. Vrain HTGRs. However, the limited data on the PNP purification plant has given rise to some concerns. The experience to date in helium purification indicates that the equipment necessary to provide 100% helium volume turnover per hour will be extensive and costly. The need for a low tritium level in the synthetic natural gas, and the desire for low hydrogen content in the primary helium, both impose increased size requirements on the purification plant. It may be that the purification process planned by the Germans is different from previous plants; however, such information has not been provided. Therefore, the development program should evaluate the economic incentives for providing such a large-capacity purification system. These concerns, however, are not felt likely to block development of an acceptable purification concept.

d. Process Control and Instrumentation

The capabilities for design of the process controls and procurement of the appropriate instrumentation should be available in the U.S. No extensive development should be required in this area. Some of the development tests of the components will require the design and operation of similar process control and instrumentation systems. Depending on the purpose of the process heat plant, which will be defined to supply the market needs, the process controls could be somewhat different for each plant site. Capabilities for designing the process controls and specify the required instrumentation will be developed within the U.S. in parallel with the testing of the various components.

APPENDIX A
MULTIPLEX COST ESTIMATES

A.1 INTRODUCTION

As part of this assessment, a more complete review was made of HTR-Multiplex cost estimates. This review resulted in costs for all systems significantly higher than those reported in Section 2 which were based on previous studies. It is believed that these new Multiplex cost estimates are conservative, especially with respect to the nuclear island and TCP pipeline costs.

The market conclusions remain unchanged: the HTR-Multiplex has large cost advantages over available alternatives in the dispersed industrial heat market (one- and two-shift operations plus peaking and mid-range electric power generation.

A.2 COST ESTIMATES

An HTR-Multiplex system based on the thermochemical pipeline (TCP) as one mode for distributing the product can produce electricity, industrial process steam, and lower grade heat for distric theating at various load factors, in urban locations, with minimum polluting emissions and safety limitations. Some electricity and heat can also be made at the HTR plant. To compare the economic merit of such a system with the competing alternatives for each form of energy delivered requires a consistent and fair framework of economic assumptions.

The Electric Power Research Institute (EPRI) has issued a *Technical Assessment Guide*(A-1) that recommends a consistent set of cost assumptions and economic methodology for studies on alternative power generating systems, so that studies by different contractors of different alternatives can be better compared. General Electric has used this methodology in recent studies both for EPRI and for DOE/NASA(A-2). The Gas Cooled Reactor Associates (GCRA) has indicated they use the EPRI data base and methodology.*

Basically, it is a method of life-cycle costing, so that all costs — investment, fuel, operation and maintenance — are considered over the lifetime of the system, for each competing alternative. Costs incurred in different years are placed on a common base by converting to their present worth in the initial year of operation. Capital costs incurred before operation are discounted forward to the base year; all costs incurred after the base year are discounted back to this year. This requires assumptions on the inflation and net escalation scenarios of all cost components over the period from the present to the end of the system lifetime, eg 30 years after the base year.

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*Private communication by Edward Sproat, III of GCRA.

For simplicity, since a wide range of alternative scenarios are credible, the EPRI *Technical Assessment Guide* assumes a continuing basic inflation of 6 percent/annum from now on. This applies to all cost components unless a specific net escalation, as a percentage greater than or less than the basic inflation, is specified. Net escalation scenarios for the various fuels are given in the EPRI Technical Assessment Guide; capital equipment and operation and maintenance (O&M) are assumed to have no net escalation.

The cost of money at different risk levels, common and preferred stocks and bonds, is a function of the current and expected inflation. Consistent with the above scenario, the EPRI Technical Assessment Guide suggests a discount rate of 10 percent.

It is convenient for utilities and other users to consider annual costs, not just a lump sum representing the present worth in the year of initial operation. Although actual costs will escalate over the life of the plant, it has become accepted practice to convert present worth of all the annual costs over the plant life to an equivalent constant, uniform or "levelized" stream of annual costs. To convert capital costs into such a stream, a fixed charge rate (FCR) is derived that considers the typical utility financial practices and some continuing charges that are capital dependent, such as insurance and property taxes. For the above scenario, FCR = 0.18. Since some components of O&M are fixed costs rather than variable costs (related to plant output level), and these average about 2-3 percent of the annual fixed charges, we include them in the FCR, making it 0.186.

Annual fuel costs and annual variable O&M costs are levelized by a factor dependent on the inflation rate, the discount rate, the plant life, and the net escalation rate assumed for the fuel. Even with no net escalation, i.e. 6 percent inflation only, the fuel price in the 30th year will be 5.74 times that in the initial year. The levelized cost of fuel is intermediate between these extremes, specifically 1.89 for 6 percent inflation, 10 percent discount rate, and 30-year life.

It should be emphasized and understood that this methodology makes the cost used for fuel look almost twice as big as that used in studies that use a "current" fuel cost rather than levelized. Consequently, the cost of energy forms such as heat and electricity look higher. However, by using levelized costs not only for the system being studied but also for the alternatives considered, comparability is assured.

This Appendix will emphasize the application of an HTR to drive a thermochemical pipeline to deliver product gas 100 km (or more) away for conversion to heat and electricity. A 3000 MW_t HTR is assumed which produces 1800 MW output as the difference in energy content between the gases entering and leaving the reformer. Some electricity generation is required from steam generator output for the internal needs of the system: circulators, reformer compressors, and miscellaneous parasitic power. This totals 100 MW_e. Some steam is required by the reformer heat exchange processes; it is extracted from the electric power cycle at a high temperature and returned to the power cycle at a lower temperature. The energy from this steam charged against the reformer plant, 318 MW_t, gives it an efficiency of 85 percent. In the German KFA reformer/electricity plant design on which this was based, the total electricity produced is

453 MW_e or the net electric output is 353 MW_e. While it may be possible to redesign the electric cycle to produce no net electric power (100 MW_e gross) and still meet the steam needs of the reformer plant without reducing its efficiency, we assume here that a credit may be taken for the electric output capability so that the TCP product costs can be separately evaluated.

A.2.1 Cost of HTR for TCP Application

There have been a number of studies of high temperature gas-cooled reactors in the past, which arrived at estimated plant costs for process heat and for electric power applications. These, and several recent estimates on PWR plants, will be compared as a starting point for deriving a best estimate of the HTR-Multiplex plant here analyzed. As these were done in different years, all are converted to beginning-of-the-year 1979 dollars (1/79$) using actual inflation indices. As they also have different reactor sizes, all are scaled to 3000 MW_t. The Direct Cost level (equipment, plus on-site labor and materials) is the preferred level for comparison. Several were given in Base Costs, which include indirect costs and some contingency allowance. These were reduced by 35 percent to approximate direct cost. Table A-1 presents the comparison.

The table shows that even when converted to a constant size and to 1/79$ there is a significant range in the estimated cost. For plants that have not yet been built, this is to be expected. The most recent, #4 and #7, are identical in rating and probably represent the same data base. The difference, $533 M versus $566 M, represents the uncertainties in converting from base cost to direct cost. Similarly, the differences between the GCRA and EPRI estimates on a PWR represent uncertainties in converting from base cost to investment cost. Investment or total cost includes interest during construction, owner's costs (such as site acquisition), spare parts, and other miscellaneous items not generally included in base costs.

With the cooperation of United Engineers and Constructors, an estimate of the cost of an HTR-Multiplex was made based on the data used in #7. This is described as #8, with $616 M direct costs.

The method used was to consider the changes required to modify the steam cycle version to a version in which a reformer plant replaced much of the steam generators, to drive a TCP. Table A-2 indicates the major additions in cost (+) and deletions (−) to the various major cost accounts.

First, the net electric output is decreased from 1332 MW_e to 353 × 3360/3000. Allowing for the electricity consumed within the plant, the Turbine Island components are reduced by 62 percent. Part of the structures, related to the turbine building are also reduced, but a reformer building is added, adequately containing the heat exchangers, compressor, and associated piping which are the part of the reformer plant external to the nuclear containment. Part of the steam generators are replaced by reformers, inside the primary containment. As for reformers operate at very high temperature, require duplex tubing for isolation of the helium and product gas streams, and contain catalyst, their cost per MW thermal exceeds the cost of steam generators replaced. The $40 M net in Account 22 represents this difference. Also, the PCRV and secondary containment required to include the reformers within primary

Table A-1
COMPARISON OF HTR AND PWR PLANT COSTS

	Source	$ Used	Input MW$_t$	Cost $M	Normalized Direct, 3000 MW, 1/79$
1.	GE VHTR Process Heat (as adjusted by ORNL) (ORNL, 1975)	7/74	3000	488	668
2.	GA VHTR Process Heat (as adjusted by ORNL) (ORNL, 1975)	7/74	3000	462	633
3.	GE HTR-Chemical Heat Pipe (ESTD, 1976) (and internal GE notes)	1/76	3000	626	773
4.	HTGR-SC (GCRA, 1979)	1/79	3360 (1330 MW$_e$)	780 (Base)	533
5.	PWR (GCRA, 1979)	1/79	3817 (1274 MW$_e$)	740 (Base)	463
6.	PWR (EPRI TAG)	12/77	2960 (1000 MW$_e$)	830 (Investment)	424
7.	EEDB-HTR-SC (UE&C, 1979)	1/78	3360 (1330 MW$_e$)	557	556
8.	#7 Modified to HTR-Multiplex	1/79	(3000 1800 product 353 MW$_e$)		616

containment are estimated to be larger and more complex. Finally, the reformer heat exchangers and compressor are included as a separate account, since the conventional accounts do not assign a number for chemical process adjuncts.

As indicated, the direct costs total to $45 M more than the HTR-SC (#7 in Table A-1). Converted to 3000 MW$_t$ and 1/79$ gives the estimate of $616 M, which will be used herein.

To simplify treatment of separate products, electricity and TCP product, a credit is taken for the 353 MW$_e$ net electric output, as a fraction of the cost of #7 (603 × 353/1330) or $160 M. The remaining $456 M represents a direct cost for the plant producing 1800 MW of product gases (MW$_p$).

As there are uncertainties in the cost estimates, this value will be used as the base case and a high-low range carried through the subsequent analysis. Table A-3 describes the steps in determining the delivered cost of heat, starting with the direct cost of the HTR-Multiplex plant.

Table A-2
REVISION OF EEDB-HTR-SC (3360 MW_t 1330 MW_e TO PBR-TCP (3000 MW_t 1800 MW_p 353 MW_e

Account	EEDB*	Modifications	Net $M	Scale to 3000 MW_t (exponent)	 $M
20 Land	2.0	None	2.0	0.8	1.8
21 Structures		− 8 Turb Bldg + 5 Sec. Contain. +15 Ref. HX Bldg	104.0	0.8	95
22 Nuc Plant Equipment		+35 More Complex PCRV +40 net (Reform-SG)	339.0	0.45	322
23 Turb PE		−78	48	0.8	44
24 El PE		−26	16	0.8	15
25 Misc		None	12	0.8	11
26 Cooling		−12	7	0.8	6
27 Reformer PE	--	+74 Reformer HX & Compressor	74	--	74
1/78$	557				$ 569 M
		+45 Net			
1/79$	603				
Red. to 3000 MW	557	Credit for 353 MW_e $\left(\frac{353}{1330} \times 603\right)$ =			−152
					456

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*Use only total, details not yet released by DOE.

The low and high estimate columns in Table A-3 are roughly 10 percent lower and 20 percent higher, and reflect the range of earlier estimates. Investment costs, adding the indirect costs, contingencies, interest during construction, and other costs of acquisition are on the average 2.12 times the direct cost. The unit capital cost of $546/$kW_p$ compares to the $900/$kW_e$ for a PWR.

Considering both the fixed and variable costs of a system and how they compare to those of alternatives is best done in terms of the unit cost of product in dollars per megawatt hour (same as mills per kWh). The fixed charges are found in line 5. This component is a function of the capacity factor, the fraction of the rated capacity for 8760 hours a year that is actually utilized. For a lower capacity factor the cost must be allocated over fewer MWh_p, hence the fixed charge per unit is higher.

The levelized fuel costs are indicated in line 6 in terms of dollars per million Btu. A net escalation for nuclear fuel costs is postulated for nuclear fuel based in part on a rising cost of U_3O_8. The impact of rising costs of U_3O_8 will be different for different fuel cycles, with and without plutonium recycle, and

Table A-3
DELIVERED COST OF HTR-MULTIPLEX
(Assumes: January 1979 dollars; 3000 MW_t input; 1800 MW to TCP, 353 MW_t net)

		Base	Low	High
1.	Direct Cost of HTR-Multiplex Plant: $M	616	550	710
2.	Credit for 353 MW_e: $M	−160	−150	−150
		456	400	560
3.	Investment Cost: $M	963	848	1187
4.	Unit Capital Cost: $/kW_p$ (product)	536	471	659
5.	Fixed Charges per MWh_p (FCR = 0.186 inc. Fixed O&M; CF = 0.80)	14.30	12.50	17.50
6.	Fuel Costs per MBtu (levelized, 30 years, no recycle, 1995 startup)	(1.60)	(1.40)	(1.75)
7.	Fuel Costs per MWh_p (efficiency 0.85)	6.42	5.62	7.03
8.	Variable O&M: $/MWh_p$	1.44	1.40	1.50
9.	Total Cost at Pipeline Inlet: $/MWH_p$	22.16	19.52	26.03
10.	Added Costs per 100 km Pipeline	3.70	3.70	3.70
11.	Added Costs for Methanators (FCR = 0.186, CF = 0.80)	1.41	1.41	1.41
12.	Delivered Cost of TCP Product as Heat $/MWh_t$	27.27	24.65	31.15
		(8.00 $/MBtu)		

for the LWR and HTR. The EPRI levelizing assumptions $1.60 per MBtu in 1/79$ for 1995 startup, 30-year levelizing, and no recycle. The lower and higher values of a range are assigned to the low and high columns. Converting these to the $/MWh_p$ including the 85 percent thermal efficiency of the reformer cycle gives line 7. The amount of O&M added is at best a first-order estimate, taken as somewhat higher than the LWR. The summation of these cost elements gives the total cost of product in $/MWh_p$ at the pipeline inlet.

A.2.2 Pipeline Costs

The thermochemical pipeline required consists of three parallel pipes. The largest carries the syngas from the reformer, a mixture predominantly of hydrogen and carbon monoxide, but with 11 percent unconverted methane and 9 percent carbon dioxide at the temperatures and pressures considered optimum and achievable in the reformer. The second pipeline returns the methane, with about 13 percent hydrogen remaining, unconverted in the methanator. The third pipeline returns the condensed water generated in the methanation plant. Return transmission of water is optional, an alternative to water consumption at one end and water disposal at the other. It represents only 5 percent of pipeline cost. The first two pipes represent 61.3 and 33.3 percent respectively.

One hundred kilometers of pipeline are assumed. For longer (or shorter) distances the cost component given, \$3.7/MWh_p, varies linearly. In practice the pipeline will be a branching network, with one or two pipelines leaving the HTR plant and branching to increasingly smaller pipe diameters and capacities as individual industrial users or dispersed electric plants are approached. The effect per km of reduced pipeline capacity is to increase the cost in \$/MWh_p.

A rough approximation to the multiplier over the range of 100 to 1800 MW_p transmitted is $3.70 \times (0.89 + 110/P)$ \$/MWh_p, where P is the capacity in MW_p. In other words, the parentheses above equals 1.0 at 1000 MW, decreases only to 0.95 at 1800 MW, but increases to 1.11 at 500 MW and to 1.99 at 100 MW. There is little advantage to pipelines larger than 1000 MW, but for local pipes carrying less than 100 MW the cost per km more than doubles. For any branching network the equivalent number of km of 1000 MW_p pipeline and its cost can be found.

A.2.3 Methanator Costs

Methanators, at the terminations or intermediate points on pipelines, are the means of converting the syngas in the outgoing pipeline into methane, water, and heat. Relativelty small modules, eg 50 MW_t of heat output, can be used without a cost penalty over the \$1.41/MWh_p indicated in line 11. For much smaller sizes the cost will rise significantly so that the cost increment per MWh_p will be roughly twice as great for a 6 MWh_p methanator.

The delivered cost of heat as methanator output is \$27.27/MWh_t or \$8.00/MBtu (line 12) for the base case. For the alternative assumptions in the low and high cases it is 7 percent less and 18 percent more.

The methanators are nominally designed for operation at 40 bars pressure and 427 °C, adequate for supplying saturated steam by heat exchangers at any desired pressure. With multiple stages of methanation, as may be the preferred design, the initial stages can superheat and reheat steam to 538 °C without impairing the final conversion rate to methane. Use may be made of the heat output to generate industrial process steam, to produce steam for dispersed electric generation, or for both of these in a cogeneration, or Combined Heat and Power mode.

A.3 COMPARISON WITH THE COST OF ALTERNATIVES

For the production of heat as process steam, the cost of heat from the methanator terminals of a HTR-Multiplex system will be compared with heat from oil-fired boilers, coal-fired boilers, and electrode boilers powered by electricity from LWRs. The same economic methodology and inflation scenario will be used: the EPRI Technical Assessment Guide (TAG) levelized fuel costs appropriate to each fuel for a 1995 start of operation, and for nominal sizes of 100-300 MW$_t$, which in some cases may be a composite of smaller modules.

For oil-fired boilers the capital cost is low, but the fuel cost is high. In 1/79\$, at the 0.80 capacity factor assumed in the HTR-Multiplex analysis, the fixed charges on an oil-fired boiler are \$1.58/MWh. At a boiler efficiency of 0.85 the levelized cost of low sulfur residual oil is \$31.80/MWh$_t$. These total \$33.38/MWh$_t$. This is about 25 percent more than the base case cost of steam from the methanators of the HTR-Multiplex system and slightly more than the high case.

For coal-fired boilers, a 1977 study(A-3) was used as a basis for small plants (50 KPPH to 400 KPPH; approximately 18 MW$_t$ to 140 MW$_t$). It compared High Sulfur and Low Sulfur coal fuels with conventional and atmospheric fluidized bed (AFB) boilers. A conventional stoker with high sulfur coal and flue gas desulfurization is estimated as \$7.7 M in 1Q '75\$ for 100 KPPH (thousand pounds of steam per hour). This is for steam conditions of 600 psi/750 °F, 200 °F feedwater, and 82 percent boiler efficiency. Including fuel cost, O&M and boiler feedwater equipment (not in above capital cost) the cost of steam is given as \$5.19 KPPH or \$14.74/MWh$_t$.

The capital costs include contingency (20 percent) but not other components of base and investment cost; the fuel costs are not levelized. Converting the capital costs to 1/79\$ and further increasing them by 40 percent for the added equipment and the conversion to investment cost gives \$14 M capital cost, or a fixed charge of \$10.62/MWh$_t$.

The cost of fuel in 1/79\$ is \$1.15/MBtu (Midwest) in 1979 in 1/79\$, but with the EPRI TAG escalation assumptions the levelized cost in 1/79\$ for a 30-year life with 1995 start of operation is \$3.05/MBtu. At 82 percent boiler efficiency the fuel charges are \$12.71/MWh$_t$. Levelized O&M including consummables for the flue gas desulfurization is \$7.85/MWh$_t$. These total \$31.18/MWh$_t$, again somewhat higher than the base case HTR-Multiplex but less than the high case.

As given in the Exxon report(A-3), both low sulfur coal and use of AFB technology reduce this cost. They estimate the following percentage decreases: high sulfur coal with AFB, 7 percent; low sulfur coal with AFB, 15 percent; low sulfur coal with conventional stoker, 19 percent. Thus in low sulfur coal regions much of the above indicated margin disappears.

Electrode boilers, like natural gas or the TCP delivery of energy, is a clean, non-polluting method of generating steam in urban areas where oil and coal may not be feasible. Electrodes in slightly conducting water generate the desired saturated steam at 100 percent boiler efficiency. The capital cost is low and may be considered as \$32/kW$_e$. At CF = 0.80, the fixed charges are \$0.85/MWh$_t$.

However, the "fuel cost" is that of delivered electricity, generation plus transmission. The exemplar electric generating sources, the EPRI TAG PWR in line 6 of Table A-1, and the GCRA HTGR-SC in line 4 give a generated cost of electricity with the assumptions on fuel costs and levelizing used herein of \$46.84/MWh$_e$ and \$38.23/MWh$_e$ respectively. This is for a capacity factor of 0.8 as used for the

\$46.84/MWh$_e$ and \$38.23/MWh$_e$ respectively. This is for a capacity factor of 0.8 as used for the HTR-Multiplex. While considered reasonable for PBR with on-line refueling, this may be 5-10 points high for a PWR or a prismatic HTGR. This would increase the fixed charges by about \$2/MWh$_e$. Electrical transmission (overhead) was estimated in another study(A-4) to be about half the cost of TCP transmission, so an addiitonal \$2/MWH$_e$ should be added. This gives a range for electrode boiler heat costs of \$42 to \$52/MWh$_t$. The HTR-TCP has a clear advantage over this mode of heat generation.

A.3.1 Lower Capacity Factors

The preceding analysis assumed no storage and assumed a high capacity factor, as is attainable in industries operating essentially continuously, 24 hours a day and over 300 days a year. Varying load demands, determined by one- or two-shift operation, by seasonal variations, or by market factors determining percent of capacity used, will increase the fixed charges per MWH$_t$, both for the HTR-Multiplex and for the alternative ways of meeting the demand.

The HTR-multiplex system is capital intensive. Of the \$27.27/MWh$_p$ base case cost, 70 percent is capital cost and only the \$7.86/MWh$_p$ for fuel and O&M is a variable cost. For capacity factors other than 0.80, the delivered cost of heat would be

Cost of Heat in \$/MWh

CF	TCP	Oil	Coal
1.00	23.39	33.06	29.06
0.80	27.27	33.38	31.18
0.60	33.74	33.91	34.72
0.40	46.68	34.97	41.80
0.20	85.50	38.14	63.04

It is obvious that the advantage of the HTR-Multiplex disappears if it is used at a capacity factor less than 0.6. While a mix of low capacity factor loads with non-coincident peak loads could mitigate this problem somewhat, the possibility of storage of the pipeline gas should be explored. Natural gas pipelines provide some flexibility to meet varying loads by pipeline packing, i.e., variation of the pressure in the pipeline, and also store natural gas in underground reservoirs for seasonal variations.

A.4 PIPELINE STORAGE

Pipeline packing for the closed loop TCP system would consist of increasing the storage of high energy gas, the $CO + 3H_2$ syngas, during off-peak hours by putting more gas in at the reformer than is passed through the methanators. The pressure will rise in this pipe. Similarly, since more methane is removed from the second pipe than is received from the methanators, the pressure in this pipe will decrease. During peak load hours, the energy release rate in the methanators will exceed the 1800 MW$_p$ rate of generation at the reformer. The pressure in the syngas pipe drops, and may drop below the nominal pressure of 40 bars (600 psi) while the methane pipe presure exceeds the nominal pressure.

A pressure swing of 20 to 60 bars can be considered. Both pipelines must be redesigned for safe operation at the 60-bar pressure. This increases the pipeline cost per 100 km by 40 percent of \$1.48/MWh$_p$. The diameter and velocity of the resdesigned pipeline indicate that the syngas pipe contains 4.84 hours of product at 60 bars while the methane pipe contains 6.55 hours of product. With a swing from 20 to 60 bars in the syngas pipe, the available surplus or storage is 3.23 hours of product. For this amount of storage, the methane pipe need only swing from 25 to 55 bars.

In order to have both reformer and methanator continue to operate at 40 bars, to avoid control complexities, compressors are neded at each end of each pipeline, with ratings of 20 bars on the syngas pipeline and of 15 bars on the methane pipeline. At the methanator the compressors must handle the peak mass flow; at the reformer they handle the nominal 1800 MW$_t$ (1396 Mg/hr mass flow). If the rated peak flow is 1.5 times the average, it can deliver peak rating for 6.26 hours. If it is 2.0 times the average, the storage would be depleted in 3.23 hours. With a 1.5 times rating, the cost of compressors would increase the pipeline cost by another \$1.15/MWh$_p$.

A further cost is the power required for each of these compressors. Unlike the compressors provided to overcome frictional pressure drops, which operate continuously, the storage compressors each operate over limited portions of the charge/discharge cycle, and vary in power input required as the pipeline pressures change from 20 to 40 bars and from 40 to 60 bars. For a useful cycle in which the pipeline delivery is at the 1.5 times rate (2700 MW$_p$) for 6.3 hours and below average, at 1480 MW$_e$ for the remaining 17.7 hours, the cost of power adds another \$0.65/MWh$_p$.

These total to \$3.28/MWh$_p$ for pipeline-packing storage of 3.26 hours. The amount of storage can be increased by adding to its length or diameter, i.e., for each added 100 km of pipeline or its equivalent as parallel pipelines or a larger diameter add \$5.18/MWh$_p$ of the 60-bar pipeline and gain 3.26 hours of storage. The required compressors would not be increased by this expansion of storage but the power required would increase. The alternative of increasing the swing to a higher pressure would require more compressor cost as well as a more costly pipeline.

The storage described permits serving an overall load demand with capacity factors down to 0.53 at a cost of \$32.45/MWh$_p$, compared to \$37/MWh$_p$ when done by load following at the reformer. This provides a further margin over oil- and coal-fired boilers.

A.5 MULTIPLEX DISPERSED GENERATION ESTIMATES

While the HTR-Multiplex system has an advantage over alternatives for production of heat alone at dispersed sites, the emphasis of the Multiplex concept is on synergistic benefits from producing both heat and power at dispersed sites. First we look at the cost estimates of producing electricity alone at such sites.

A.5.1 Dispersed Electric Generation

The application of the HTR-Multiplex production of electricity near the load centers, in urban and industrial areas has potential advantages over central and dispersed generation with alternative fuel sources.

There are the advantages over other dispersed forms of generation that emissions from coal- and oil-fired boilers are eliminated. There are no stack losses as a source of inefficiency. The capital costs of the Boiler Island (the cost components associated with the fuel storage and use, boiler, piping, stack, and cleanup) are reduced. There are the advantages over central electric generation that the cost of transmission and distribution is reduced (a trade-off with the cost of the TCP) and that the environmental problems of transmission rights of way are reduced (underground TCP versus overhead electric). There are the further advantages when both electricity and heat are cogenerated that the equivalent efficiency of electric generation is increased, the amount of thermal emissions (waste heat) is reduced, and the heat produced is near the market, reducing distribution costs.

The output of the methanator can be designed for production of steam at conditions as high as 16 MPa/538 °C/538 °CRH. Since there are no stack losses, electricity can be generated at an efficiency over 40 percent. The economic optimum for dispersed generation may be at a lower pressure and temperature, such as 12.4 MPa/510/510 °C.

Two cycling coal plants for the above two conditions were compared with each other and with base load high sulfur coal (HSC) plants in a recent study(A-4). The cycling coal plants, designed for mid-range and peaking capacity factors are rated 512 MW_e net output; the base load plants are rated 794 and 1252 MW_e net. On the same economic assumption basis used earlier in this section their capital costs are as given in Table A-4.

The top line gives the plant direct costs in millions of 1974 dollars; converting to total costs and dividing by the net output in kW gives the second line. Note that for the high pressures and 1000 °F, the first three, the unit cost increases and efficiency decreases as size is reduced. The last column, with lower

Table A-4
COMPARISON OF COAL-FIRED GENERATING PLANTS

	1232 MW_e 3500 psi	**794 MW_e 3500 psi**	**512 MW_e 2400 psi**	**512 MW_e 1800 psi**
M$ (direct cost) 1/79$	459	327	247	209
Total Investment Cost $/$\text{kW}_e$	805	889	1042	885
Plant Efficiency	0.374	0.360	0.357	0.331
Turbine Island Only $/$\text{kW}_e$	374	396	456	398
Cycle Efficiency	0.423	0.408	0.404	0.375

pressure and temperature has a significantly lower unit cost at a penalty of 8 percent in heat rate and efficiency. The lower capital cost makes it more attractive than the other cycling coal plant for low capacity factor operation.

The cost of the Turbine Island portions of these plants can be separated from the Boiler Island, or steam-supply-related costs by collecting the cost accounts into these two categories, allocating between them the miscellaneous costs that serve both. The unit costs of the Turbine Island are shown in the fourth line. The efficiency from steam supply to electric output is higher than that in line 3 by the inverse of the 0.883 boiler efficiency assumed (i.e., stack losses).

Dispersed electric generation using methanators as a steam supply can be analyzed by combining the fixed charges on the Turbine Island with the fixed and variable charges of the HTR-Multiplex system. As a specific example we analyze the 512 MW$_e$ Turbine Island in column 4, 12.4 MPa/510/510°C. At 0.186 fixed charge rate and CF = 0.80, the fixed charges add $10.56/MWh to the cost of electricity. Using as the fuel cost the Base Case delivered cost of HTR-Multiplex heat and the efficiency in Table A-4 we have for 0.80 capacity factor a fuel cost of $73/MWh$_e$, or a total of $83/MWh$_e$.

The coal-fired cycling plant of the same rated capacity and steam conditions has a 0.80 CF a fixed charge of $23.49/MWh$_e$. As discussed under coal-fired boilers, the levelized cost of coal is $3.05/MBtu. At 0.331 efficiency, the fuel cost component is $31.45/MWh$_e$, and levelized O&M including consummables is $19/MWh$_e$. These total $73.94/MWh$_e$.

At lower capacity factors for mid-range and peaking, the cycling coal plant electricity generation costs are $81.77/MWh$_e$ at CF = 0.60, $97.43/MWh$_e$ at CF = 0.40, and $144.41/MWh$_e$ at CF = 0.20.

The combination of methanator and Turbine Island capital costs would have much less variation with capacity factor. If the HTR-Multiplex capital cost components continued to be operated at CF = 0.80 the "fuel cost" would remain constant at $70/MWh$_e$ (transferring the methanator cost, which does vary with capacity factor, to the Turbine Island costs). The electricity generation costs would be $89/MWh$_e$ at CF = 0.60, $98/MWh$_e$ at CF = 0.40, and $127/MWh$_e$ at CF = 0.20. At low capacity factors the HTR-Multiplex can beat the cycling coal plant for dispersed generation.

However, the HTR-Multiplex system cannot continue to operate at CF = 0.80 when the loads served are peaking and mid-range, unless:

- There is sufficient load pattern diversity among all TCP product users.
- There are users on an interruptible load basis, or
- There is storage.

There is more likely to be load diversity if there are many small loads than if they are as big as the 12 MW$_e$ plant used as an example. The unit cost of both the Turbine Island and the Boiler Island will in-rease with smaller ratings; the modular methanators remain constant in unit cost until sizes below say 40 MW$_t$ are considered. Cost variation with size may be approximated either as a constant plus linear term or as an exponent representing the slope of the variation of cost with size. Using an exponent of 0.7 for capital cost (−0.3 for unit costs) gives a factor of increase of unit cost ($\$/kW_e$) from 512 MW_e down to say 50 MW_e of 2.01. This will raise the cost of electricity for both the cycling coal and the HTR-Multiplex-Turbine Island plants considered, but will affect the former more. For example, at a capacity factor of 0.4 the cost of electricity is \$116/MWh_e for the HTR-Multiplex and \$145/MWh_e for the cycling coal plant. In this case the HTR-Multiplex has a significant advantage over cycling cost plants even at capacity factors above 0.40, but both compare poorly with larger central plants, base load or cycling coal.

Storage by pipeline packing was examined earlier in this section. It will permit the HTR-Multiplex to operate continuously at full rating in the presence of variable demand (not completely neutralized by diversity). It adds \$3.28/MWh_p, or \$8.75/MWh_e to the cost of heat and electricity respectively. Adding \$8.75 to the \$98.64/MWh_e previously calculated for CF = 0.40 gives \$107.34/MWh_e, or 9 percent less than the \$116/MWh_e above.

A.5.2 Dispersed Heat and Power Generation

The heart of the HTR-Multiplex concept[A-5] is that the thermochemical pipeline can deliver clean energy to urban and industrial areas to flexibly make electricity, steam, and HTW for district heating. Among the non-economic advantages are:

- Underground pipeline transport is more environmentally acceptable than overhead electric transmission and is much cheaper than underground electric transmission for moderate power levels.
- The primary power plant, an HTR, can be located many tens of kilometers from load centers, avoiding the remaining emissions problems with dispersed coal-fired boilers for process steam and electricity generation.
- The cogeneration of electricity and heat provides the thermodynamic/cogeneration advantages of using less fuel than that required to produce separately electricity and heat. The savings of oil and gas now conventionally used for heat production in the densely populated areas by industrial and the residential/commercial sector is particularly attractive.
- The long distance transport by the TCP to load areas means that the distribution of electricity and heat from the dispersed centers need be at most a few miles or km.

Let us look now at the economics of such combined heat and power production.

When thermal energy is extracted from a steam turbine cycle, as steam or HTW, the electric output of the plant is reduced. The amount of electricity reduction per megawatt (thermal) extracted depends on the location(s) of the extraction. That is, there is more loss of electricity if prime steam from the steam supply is extracted, before it has generated any electricity, than if low pressure steam or HTW at a lower temperature and pressure than the steam supply is extracted. This ratio, of electricity lost to thermal energy made available, in MW_e/MW_t, may be called the equivalence ratio. For prime steam it is the p

turbine cycle efficiency, or about 0.4. For HTW for district heating, at temperatures under 100 °C, it may be under 0.10.

In a recent study of thermal energy storage and transport(A-4) a number of cases of thermal extraction of steam and HTW, at different temperatures, were analyzed with the aid of the computer programs of GE's Large Steam Turbine Division, to get realistic estimates of this equivalence ratio for both a coal-fired plant with high pressure, high temperature steam supply, and for a Light Water Reactor, with lower pressure and temperature operation. The more efficient high temperature and pressure cycle of course loses more electricity per megawatt thermal extracted, but is most comparable to the cycle used in the previous pages with the TCP steam supply. From the several cases studied and thermodynamic analysis of the rationale for interpolating/extrapolating, the equivalence factor estimates shown in Figure A-1 are derived. These are for HTW efficiently extracted from the feedwater heating train at appropriate points. Up to the boiler inlet feedwater temperature the available temperature can be essentially any point on the solid part of the curve, with a little trim heating with higher temperature extraction heat if the desired temperature does not match conventional temperature levels between feedwater heaters. The boiler inlet temperature is about 260-285 °F in conventional designs but could be as high as 315 °C with redesign. Alternatively, some prime steam can be diverted to raise the HTW up to 340 °C if this is desired.

Large quantities of steam can be easily extracted only at several points, without major turbine system redesign: the inlet to the HP turbine (prime steam), the inlet to the IP turbine, the inlet to the LP turbine, and at the LP turbine outlet. The latter is only useful if the turbine system is designed to have a high backpressure turbine, i.e. usually above atmospheric pressure, so the steam is at a useful temperature and pressure. The analysis indicated that steam extraction gave almost the same equivalence factors versus temperature as the HTW solid curve temperature range. The dotted line indicates a more uncertain range for steam and very high pressure HTW, connecting to the known equivalence factor for prime steam.

The economic significance of the equivalence factor is great. In principle, the costs of electricity generation (in $\$/MWh_e$) must be recovered from revenues received from electricity customers. The lost revenues from reduced electric output during cogeneration of heat and power must be recovered by revenues received from the larger quantities of thermal energy extracted, so the cost of heat is proportional to both the cost of electricity and the equivalence factor ($\$/MWh_t = F_e \times \$/MWh_e$).

From the preceding discussion of dispersed electric generation, cost of electricity from an HTR-Multiplex-steam turbine system at 512 MW_e and CF = 0.80 is $\$84.56/MWh_e$. If HTW is extracted from the steam turbine at 177 °C and returned to the feedwater heating train at 81 °C, the extracted thermal energy costs only $\$16.32/MWh_t$ to recover the lost revenue. This compares very favorably to the $33 and $\$31/MWh_t$ for oil- and coal-fired boilers.

However, not only must the cost of heat compete with the local cost of heat but the cost of electricity must complete with the delivered cost of base load electricity which is less than the $\$84.56/MWh_e$.