sacrifice on the part of consumers - adjusting thermostats for 80° in the summer and 65° in the winter, observing the 55 mph speed limit; the other is built in - better insulation of homes, cogeneration of electricity and heat. The President's Council on Environmental Quality has found that energy use need increase only 10 to 15% by the year 2000, instead of more than doubling as projected by a number of other estimates, "with a determined national effort to conserve energy(3-4)."

Opportunities for conservation in the electric-utility sector are seldom included, except with regard to higher-efficiency generation of electric power, in estimates of conservation opportunities. However, cogeneration of power and heat by utility plants offers a greater opportunity for energy conservation than does cogeneration by industrial plants, which usually is included. Further, it is only utility plants that can use nuclear fuel for cogeneration to displace oil and gas. The HTR-Multiplex can make major contributions toward implementing conservation through cogeneration by utility plants, as seen from Tables 3-5, 3-6, and 3-7, which are taken from or based on the Department of Energy's Monthly Energy Review.

Table 3-5. Energy consumption by economic sector (quads).

Year	Coal	Natural Gas	Petroleum	Hydro-electric	Coke	Electricity		Total Energy Use
						Consumed	Associated Losses	
Industrial[*]								
1973	4.38	10.40	7.22	0.03	(0.01)	2.34	5.56	29.92
1974	4.05	10.01	6.43	0.03	0.06	2.34	5.67	28.59
1975	3.79	8.53	5.93	0.03	0.01	2.30	5.61	26.21
1976	3.77	8.77	6.52	0.03	0.00	2.53	6.14	27.76
1977	3.61	8.64	6.74	0.04	0.02	2.64	6.43	28.11
1978	3.43	8.29	6.76	0.04	0.13	2.73	6.77	28.15
Residential and Commercial[†]								
1973	0.29	7.63	6.05			3.49	8.30	25.75
1974	0.29	7.52	6.06			3.47	8.42	25.76
1975	0.25	7.58	5.84			3.58	8.73	25.98
1976	0.24	7.87	6.45			3.73	9.06	27.34
1977	0.23	7.46	7.14			3.93	9.59	28.36
1978	0.27	7.68	7.17			4.08	10.10	29.29
Transportation[‡]								
1973	0.00	0.74	18.13			0.01	0.03	18.93
1974	0.00	0.69	17.68			0.02	0.04	18.41
1975	0.00	0.60	17.87			0.02	0.04	18.52
1976	0.00	0.56	18.80			0.02	0.04	19.41
1977	0.00	0.54	19.48			0.01	0.04	20.07
1978	0.00	0.54	20.02			0.02	0.04	20.61

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[*] Construction, manufacturing, agriculture, and mining establishments.

[†] Housing units, non-manufacturing business establishments, health and educational institutions, and government office buildings.

[‡] Private and public passenger and freight transportation; government transportation, including military operations.

Source: U.S. Department of Energy, Energy Information Agency, *Monthly Energy Review*, May 1979.

Table 3-6
ENERGY CONSUMPTION BY ELECTRIC UTILITIES (QUADS)

Year	Coal	Natural Gas	Petroleum	Hydro-electric	Nuclear	Other	Total
1973	8.63	3.75	3.43	2.98	0.91	0.05	10.74
1974	8.54	3.52	3.29	3.28	1.27	0.06	19.94
1975	8.79	3.24	3.09	3.19	1.90	0.07	20.28
1976	9.72	3.15	3.41	3.03	2.11	0.08	21.51
1977	10.26	3.29	3.82	2.48	2.70	0.08	22.64
1978	10.37	3.29	3.91	3.11	2.98	0.07	23.72

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*Includes bituminous coal, lignite, and anthracite coal.
+Includes geothermal power and electricity produced from wood and waste.

SOURCE: U.S. Department of Energy, Energy Information Agency, *Monthly Energy Review*, May 1979.

Table 3-7. National energy consumption (quads) based on the EIA *Monthly Energy Review*, May 1979.

Electric Generation and Distribution and Associated Losses Shown Separately				Electric Generation and Distribution Losses Included in End Use		
	Q		%		Q	%
	(Elec)	(Heat)			(E + H)	
1974				1974		
Res./Comm.	11.9	13.9	19	Res./Comm.	25.8	35
Industrial	8.0	20.6	28	Industrial	28.6	39
Transportation	0.1	18.4	25	Transportation	18.4	25
Elec: Delivered	5.8					
Elec: Losses	14.1					
		19.9	27			
		72.8	100		72.8	100
1975				1975		
Res./Comm.	12.3	13.7	19	Res./Comm.	26.0	37
Industrial	7.9	18.3	26	Industrial	26.2	37
Transportation	0.1	18.5	26	Transportation	18.5	26
Elec: Delivered	5.9					
Elec: Losses	14.4					
		20.3	29			
		70.7	100		70.7	100
1976				1976		
Res./Comm.	12.8	14.6	19	Res./Comm.	27.3	37
Industrial	8.7	19.1	26	Industrial	27.8	37
Transportation	0.1	19.4	26	Transportation	19.4	26
Elec: Delivered	6.3					
Elec: Losses	15.2					
		21.5	29			
		74.5	100		74.5	100
1977				1977		
Res./Comm.	13.5	14.8	19	Res./Comm.	28.4	37
Industrial	9.1	19.0	25	Industrial	28.1	37
Transportation	0.1	20.0	26	Transportation	20.1	26
Elec: Delivered	6.6					
Elec: Losses	16.1					
		22.7	30			
		76.5	100		76.5	100
1978				1978		
Res./Comm.	14.2	15.1	19	Res./Comm.	29.3	38
Industrial	9.5	18.6	24	Industrial	28.1	36
Transportation	0.1	20.6	26	Transportation	20.6	26
Elec: Delivered	6.8					
Elec: Losses	16.9					
		23.7	30			
		78.0	100		78.0	100

Table 3-5 gives energy consumption by economic sector for the years 1973-1978. For each of the three sectors - industrial, residential/commercial, and transportation - the amount of fuel consumed is shown, by type of fuel, together with the amount of electricity consumed and the associated losses (cycle inefficiencies plus transmission and distribution losses). Note that in 1978, industrial sector energy consumption was about 30% gas, 24% petroleum, 12% coal, and 34% electricity (consumed plus associated losses). The losses associated with electricity consumed were equal to the total petroleum consumed. Similarly, in the residential/commercial sector, the losses associated with electricity consumed exceeded consumption of either gas or oil.

Table 3-6 shows the makeup of the energy used for generation of electricity by utilities. Power plants consumed almost three times as much coal as other sectors combined, and less than one-third as much gas and oil as the industrial and residential/commercial sectors.

Table 3-7 summarizes energy use by sectors in two ways: with electricity consumed (delivered) shown separately, in the left-hand column, and with electric generation and distribution included in the end uses, in the right-hand column. Again, considering only 1978, note in the left-hand column that the losses associated with electric generation and distribution, 16.9 quad, are comparable in magnitude to the total use of heat in either the residential/commercial or the industrial sector. (Feedstocks for the industrial sector are included in "heat" consumption.) The loss of 16.9 quad is equivalent to about 8.5 million barrels of oil per day, which by coincidence is roughly the amount of oil now imported. While losses associated with electric generation and distribution obviously cannot ever be reduced to zero, the magnitude of the losses gives some appreciation of the conservation opportunity available.

From Table 3-7 it is seen that 2.49 units of heat were lost for each unit of electrical energy delivered in 1978. If the TCP energy from an HTR-Multiplex is used for on-site cogeneration, this ratio can be reduced substantially. Most of the energy not used for generating electricity will be used to produce useful heat for industry and residential/commercial applications. For illustration, suppose and HTR-Multiplex and pipeline produce and deliver, as TCP energy, 80% of 1000 MWt of nuclear heat input to the Multiplex, and that an on-site turbogenerator can produce electricity at 32% efficiency when supplied with 1000°F steam from a methanator and required to discharge steam at 350°F. (The efficiency would be about 40% if the steam were discharged at normal cold-condensing temperature, around 100°F.) The electric output is 0.32 x 800 MWt = or 256 MWe. The exhaust steam contains 544 MWt of heat that is useful for industrial process heat, hot tap water, space heating, and absorption-cycle air conditioning. To be charged to the electrical output of 256 MWe is 64 MWt of loss in the Multiplex and pipeline. This 64 MWt is to be compared to the 636 MW of loss found by considering the 1978 national ratio of electricity lost to electricity delivered shown in Table 3-7. This is a 10:1 reduction in losses.

Extrapolating the results given above to the year 2010 gives a rough estimate of energy conservation opportunities available through implementing the HTR-Multiplex. Assuming a growth rate of 2.5% per year in electrical demand, 14.3 quads of delivered electricity will be needed in the year 2010. If HTR-Multiplexes generate about 15% or 2.1 quads of the 14.3 quads demand, the associated losses would be about 0.5 quad instead of 5.2 quads. Further, there would be roughly 4 quad of heat produced from nuclear fuel which could be utilized for industrial or residential/commercial applications. Energy conservation would then be savings of about 2.4 million barrels of oil equivalent per day, plus fuel substitution benefits of about 2 million bbl/day (equivalent) satisfied with nuclear fuel instead of oil and gas.

3.5 Financial Risk Considerations

Allowing or forcing the price of oil in the United States to approach the world price is seen as a way to reduce consumption. Deregulation of prices is seen as producing very large windfall profits for oil companies producing domestic crude. The Administration has proposed, and the Congress is discussing, a windfall profits tax, from which an energy security fund would be formed. Subsidies that could total more than $200 billion have been proposed, with a number of more modest suggestions being seriously considered, to furnish federal loans, grants, loan guarantees, and price supports for energy supplies that reduce dependence on imports of oil and gas. The financial risks associated with coal liquefaction plants are similar to those encountered by any new energy resource which must compete with imported oil.

For both the production of synthetic fuels (among which TCP gas can be included for purposes of this discussion) and the implementation of the HTR-Multiplex concept, the investments required and the uncertainties or risks involved are so large that the private sector is very unlikely to finance new ventures at anything approaching the rate required to achieve energy independence.

The Comptroller General's May 1979 Report to the Congress, Questions on The Future of Nuclear Power: Implications and Trade-Offs, examines nuclear and non-nuclear options. Oil and gas for generating electricity are assumed in the GAO report to remain at 1976 levels through 1985, then to decrease one-third every 5 years until phased out in 2000. GAO doubts that substantial petroleum generation of electricity could be sustained past the end of the century, "particularly in the face of demands for petroleum from sectors such as transportation, for which there are almost no fuel alternatives".(3-3)

3.5.1 Synthetic Fuels From Coal

Financial support of coal processing plants by the federal government appears necessary principally because coal-derived fuels are so much more expensive to produce than petroleum. Only if world prices remain at a level well above the cost of domestic production, and far above the cost of OPEC production, will coal-derived fuels be competitive in price. Industry is very unlikely to invest the capital necessary to build coal processing plants when faced with the threat of OPEC's dropping the price of oil enough to make coal-derived fuels non-competitive. This threat could be met by some combination of price regulation, relief from domestic anti-trust provisions, imposition of tariffs to keep domestic prices above world prices (and produce additional capital for alternative-energy infrastructure), and government support of the coal processing plants. This suppport might range from government ownership to price supports to protect private-sector owners against loss. It has been observed that price supports instituted by the Congress can be reduced or withdrawn by the same or a subsequent Congress.

3.5.2 Nuclear Plants

Although the HTR-Multiplex appears to have the potential for profitable operation in competition with domestic coal, its profitability may also be vulnerable to downward price adjustments by OPEC.

Federal funding surely will be required to at least prove out full-scale operation of the HTR.

Experience in commercializing light-water nuclear plants is not to be forgotten. The following description is drawn from the Rand report by Perry, et al, published in 1977.(3-5)

The Atomic Energy Act of 1946 created a framework within which the development of nuclear power in the United States could proceed. By 1953, the equipment manufacturers and elements of the AEC that had been involved in the various tasks of reactor development generally agreed that commercialization required little more than solving definable engineering problems and convincing American utility firms that the nuclear power era had begun. But "costs and returns were uncertain, owing to the immaturity of reactor technology. Although notably nervous about the possibility of nationalized nuclear power, neither utilities nor manufacturers were interested in financing open-ended nuclear projects".

Various kinds of government support were proposed under the AEC's Power Reactor Demonstration Program announced in January 1955. Under a cost and risk sharing approach, a number of small plants were built. However, "attempts to insure the development of a healthy, diversified nuclear industry, to induce large and small producers and utilities of all sizes to participate, failed. The failure occurred not merely because nuclear reactors were expensive to develop, or plants were costly, but because few developers or users were either accustomed to or financially able to assume the considerable risks of developing, building, and operating nuclear reactors of uncertain profitability. In no instance, before 1963, was any power reactor built without at least some direct or indirect federal subsidy".

3.6 REFERENCES

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3-5 Perry, Robert, et al, *Development and Commercialization of the Light Water Reactor, 1946-1976*, R-2180-NS, NTIS PB 825 919; Prepared for the National Science Foundation under Contract OEP-7521596; The Rand Corporation, Santa Monica, California, June 1977.

SECTION 4

THE THERMOCHEMICAL PIPELINE

4.1 INTRODUCTION

4.1.1 The Thermochemical Pipeline Concept[+]

The closed-loop chemical systems described in this section involve transporting and/or storing thermal energy by the use of reversible chemical reactions. The name given to this concept is the Thermochemical Pipeline (TCP)[*] and it is illustrated in Figure 4-1. In the TCP, the primary thermal energy is converted to a chemical form by a catalyzed endothermic chemical reaction. The sensible heat required to heat the reactants from the ambient temperature to the reaction temperature is provided by a countercurrent heat exchange with the products leaving the reactor. Thus, the transport/storage of fluids takes place at ambient temperatures rather than elevated temperatures, thereby eliminating excessive thermal loss. The invested energy is recovered by reversing these steps by a countercurrent heat exchange between reactor inlet and outlet streams and a catalyzed exothermic chemical conversion to reproduce the original chemicals.

The overall process is a closed cycle, the only exchanges with the environment being thermal and mechanical energy. By removing the chemicals from the catalyst bed, it is possible to prevent the reversal of the conversion in the

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[+] This overview of the Thermochemical Pipeline is a combination of new material and material condensed from Reference 4-1.

[*] This concept has also been referred to in the past as the Chemical Heat Pipe (CHP).

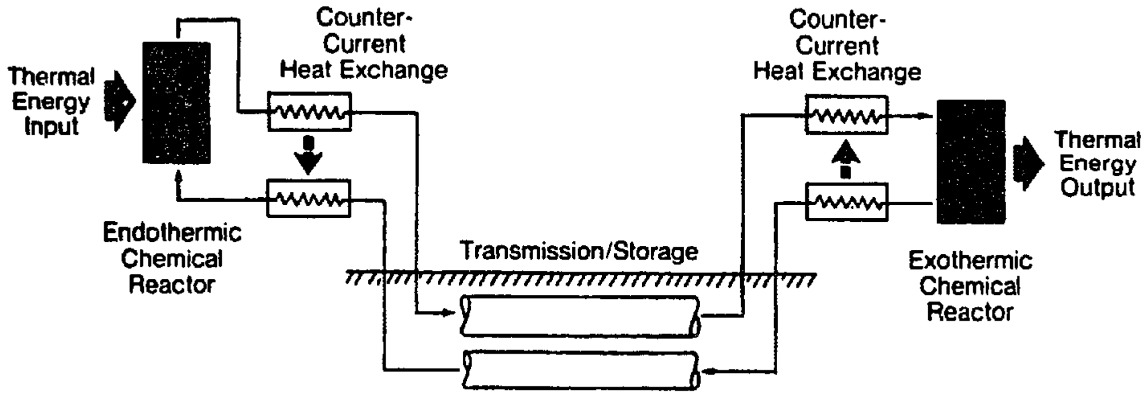

Figure 4-1. THERMOCHEMICAL PIPELINE CONCEPT

endothermic reactor during the subsequent cooling. Thus, the high energy products do not spontaneously release the stored energy at ambient temperature, in spite of the fact that this change is highly favored thermodynamically. This advantage of being able to "freeze in" the stored energy in a metastable state makes chemical conversions different from simple thermal absorption processes (phase change, heat of solution, sensible heat). For example, a mixture of carbon monoxide and hydrogen at ambient temperature does not spontaneously undergo a reaction to produce CH_4 and H_2O with a large release of thermal energy, in spite of the fact that this change is highly favorable thermodynamically. The same is not true for phase change processes; for example, steam cannot be kept in a highly subcooled (metastable) state without a spontaneous condensation and the release of stored energy.

This major difference can be readily expressed quantitatively by the large activation energy requirements for the chemical reaction, and allows for a convenient way to "turn off" the chemical change once the chemicals are out of contact with the reaction catalysts. The same is true for the exothermic change; the reaction is held off until the chemicals are heated to desired temperatures, then proceeds as they are introduced into the catalyst bed. Were it not for this convenient control, it would be impossible to contemplate a TCP.

4.1.2 TCP Concept In Relation To Energy Conservation and Transport

The concept of thermal energy transmission and storage through thermochemical pipelines has major implications for many important energy issues. Among these are prime fossil fuel substitution, energy conservation by cogeneration, storage, and reduction of oil imports.

In recent years, clean fossil fuels have become increasingly limited in availability. There have been numerous studies to predict resources of oil and natural gas. While there is considerable uncertainty as to the range of validity of different methods for estimating undiscovered natural resources(4-2), there is a general agreement among all such studies that the projected overall energy demand exceeds the anticipated fossil energy supplies, and even the most optimistic projections concede that clean fossil fuels will provide an ever decreasing fraction of the nation's needs(4-3).

A review of past energy consumption patterns(4-4) shows that approximately 75% of the total energy needs in the U.S. are met by the use of petroleum products and natural gas. The only possible way in which these needs can continue to be met by prime fossil fuels is through massive imports. Either for balance of trade or security reasons, this "heavy foreign imports" scenario is generally judged neither desirable nor viable in the long run. The only other alternative for the near future, apart from a drastic reduction in energy consumption, appears to be a switch to alternative sources of energy consisting predominantly of coal and nuclear power. Solar, wind, and geothermal sources will play a larger role when they become technically available and economically attractive. Since there are considerable differences in the ease of substitution among the various uses of prime fuels, an examination of the nature of problems associated with this shift in energy sources from oil and gas to coal and nuclear power is in order.

The largest use of oil and gas (approximately 33% of total) is in the transportation sector, almost exclusively in the form of petroleum products. While there is a growing effort on storage batteries for electric vehicles, it is highly unlikely that in the near future petroleum products can be successfully replaced in this end use. Therefore, the main efforts to reduce consumption in this sector have been restricted to conservation measures (lower speed limits, car pools) and higher efficiencies (smaller, high-mpg cars).

The next largest use of oil and gas (approximately 30%) is in the industrial sector. While 4.4% of this 30% is in the form of chemical feedstocks, and very difficult to replace, the rest is consumed as fuel to provide either steam or process heat. In this category, the usage for steam is much easier to substitute than that for process heat, because the latter is often closely coupled to (and often in direct contact with) the actual processing. A first estimate of the needs and consumption of energy in the industrial sector has resulted in the realization that not only is there lack of an accurate, detailed breakdown of consumption or requirements by size, temperature levels, geographic location, and fuel type, but also there are discrepancies and inconsistencies among all major studies that have attempted even a gross characterization of industrial energy consumption. These difficulties notwithstanding, industrial process steam users may be categorized into four broad groups:

- Users that require process steam in quantities large enough to accommodate on-site coal handling and cleanup equipment to produce process steam at competitive costs. The problem of prime fuel shortage is less severe for these users, for they can readily switch to a less expensive and more abundant source (e.g., coal) if they have not already done so. Some of these users are even large enough to justify a captive nuclear plant (LWR) to supply steam. Examples are attempts by Dow Midland, U.S.A.(4-5) and BASF, West Germany(4-6) to license an LWR to provide process steam. Though it is not possible to state the size beyond which on-site generation of steam from coal combustion is the preferred, economic route, a reasonable guess would place the critical size in the vicinity of 200 MWt. Currently, this would account for approximately 20% of the fuel consumed for process steam.

- Users that are not large enough to afford large, capital-intensive coal-handling facilities but still require sufficient quantities of steam on a continuous basis to be unable to pay an excessive price for it. If they are denied the fuel currently used, they must either rely on imported oil or switch to more expensive alternatives such as electrode- or SNG-fired boilers. Currently, these users account for approximately 55% of the total fuel consumption for process steam.

- Users that need process steam on a part-time basis (either one-or two-shift operation every day). These users will be unable economically to use coal for their needs and will require some source they can utilize when needed (imported oil or SNG) or must transfer the storage need inherent in their consumption pattern to others (e.g., to utilities by the use of electrode boilers).

- Seasonal users of process steam. In the absence of a seasonal energy storage device, it appears inevitable that these users will be dependent on prime fuels at any cost, whether imported oil, coal-derived clean fuels, or electricity.

The use of oil and gas in the residential and commerical sectors (approximately 28% of total oil and gas consumption) is predominantly for providing space heat and hot water. These two uses alone add up to 25% of the total of 28% for these two sectors(4-4). In principle, any source of low-grade heat can be substituted for this usage of prime fuel. The potential benefits of using powerplant reject heat for this purpose are increasingly recognized in Europe. Whether or not district heating concepts are equally viable in this country is a matter of debate. At present, these consumers are considering primarily oil (domestic or imported), coal-derived clean fuels, heat pumps, or low-temperature solar heat for their future needs.

The last category of oil and gas users (approximately 10% of the total) comprises the electric utilities(4-4). The rising fuel costs have led them to an increasingly intensive search for alternatives. The two main approaches to obviate the need for oil and gas are load management by regulation or pricing policies and the use of energy storage to deliver peak electricity from base load generation. While considerable work needs to be done in these areas, the problem of reducing peak electrical consumption or accommodating it with sufficient storage does not appear insurmountable in comparison with the problems described earlier.

To summarize this discussion of substitutions for prime fossil fuel, the major problem stems from the fact that, in the past, prime fossil fuels have provided a unique source of energy that could be stored inexpensively and used when and where needed, in quantities small or large, for a broad range of end uses. Their potential substitutes (coal or nuclear energy) simply cannot match these attributes. As a result, there will be a significant mismatch of characteristics of future energy sources and demands. The mismatch is most evident in three areas:

- Sources are large in capacity (with typical sizes approximately 3000 MWt). The demands are in smaller quantities and distributed. This requires a method for distributing the energy among various users that share the same source.
- Sources are capital-intensive and uneconomical unless operated on a continuous, round-the-clock basis. The demand is often time-dependent or intermittent. This requires storage devices to buffer the primary source against demand fluctuations.

- Siting requirements dictate that the sources be located at some distance (typically approximately 160 km) away from population centers, where the demand is. This necessitates a transmission of energy from the source to demand centers.

In light of these mismatches, one can readily appreciate why considerable emphasis has been placed on such programs as coal conversion to produce synthetic fuels that match the versatility of oil and gas, energy storage (electrical and thermal) to modify load characteristics to suit baseload sources, and conservation, especially in those end uses where prime fuel use is seemingly inevitable (e.g., residential space heating, transportation).

It would be extremely useful if the primary energy from nuclear or coal plants could be transmitted and distributed to these users. The central thrust of the thermochemical pipeline concept is to develop this ability to transport useful thermal energy. Moreover, it will not only enable a significant substitution for oil and gas by alternative sources but also provide an opportunity to achieve a significant energy conservation by combined energy delivery.

4.1.3 Importance of Thermal Transport

At present, two other secondary energy carriers are electricity (derived from nuclear or fossil steam) or SNG or liquid fuel (derived from coal). The thermochemical pipeline concept provides a third option through conversion to chemical reaction energy.

It is important to compare these possible alternatives on the basis of overall energy efficiency achievable in each case. A proper methodology for such a comparison is based on thermodynamic principles governing energy conversion and transfer in combination with practical realities of the relevant technologies. This comparison has been carried out in detail(4-1). The important conclusions from that analysis can be summarized as follows:

The TCP is a unique combination of a chemical process technology and an

- The first step in the conversion of primary heat with a high exergy ratio* is invariably the degradation of this heat to a lower exergy ratio due to the materials limitations (e.g., corrosion) of the equipment used.
- The subsequent conversion of this heat with an intermediate exergy ratio into a secondary energy carrier with a high exergy ratio (electricity, SNG, hydrogen) results in a large portion of the heat being rejected at the conversion site (i.e., a low first-law efficiency).
- The use of this secondary energy carrier with high exergy ratio to produce process steam with an intermediate or low exergy ratio, using current thermodynamically inefficient methods, invariably results in a loss in the potential work available from the secondary energy (a low second-law efficiency).
- The combination of these two steps invariably results in both heat rejection at the conversion site and loss of available work (exergy) at the user site (low first- and second-law efficiencies).

- The use of a secondary energy carrier with an intermediate exergy ratio (lower than the degraded primary energy but higher than the final process steam exergy ratio), such as the TCP, will generally result in a higher first-law efficiency and a higher-second law efficiency.
- Using these arguments as a basis, it becomes clear that if primary thermal energy needs to be converted, transported, and reconverted to deliver lower grade heat, it is inefficient to generate an intermediate energy carrier with a very high exergy ratio.

These conclusions are graphically illustrated by Figure 4-2. If, in the production of process heat, a comparison is made between the transport of thermal energy from a High Temperature Gas-Cooled Reactor (HTR) through a TCP and by the generation of electricity, an overall energy efficiency of approximately 85% is obtained in the first case versus 36% in the second.

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*The exergy ratio for heat equals the maximum work available from the heat divided by the quantity of heat (W/Q). See Appendix 2 of Ref. 4-1 for a detailed explanation of this concept.

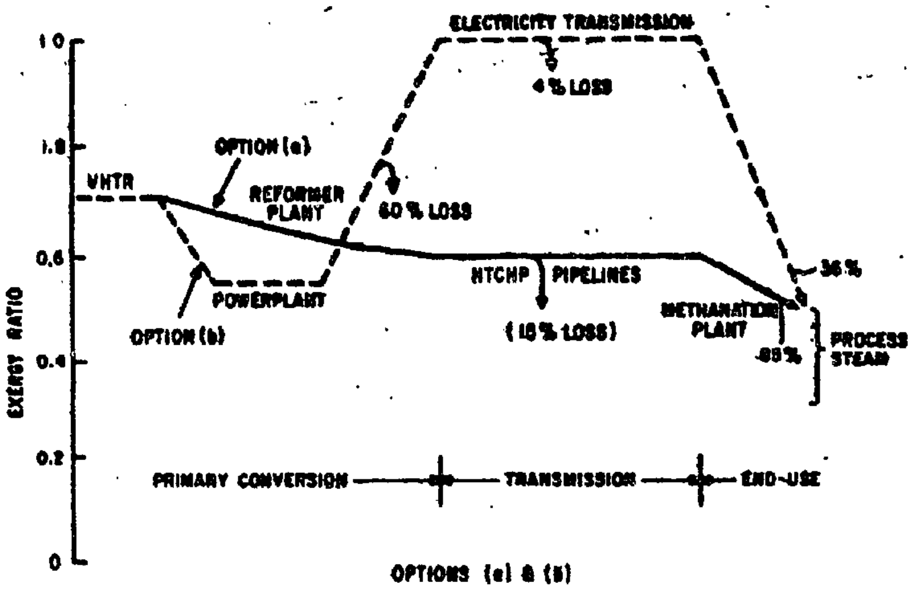

Figure 4-2. EFFICIENCY IN ENERGY CONVERSION

These thermodynamic arguments show why serious consideration must be given to the TCP concept of heat transport. However, thermodynamics itself is not sufficient to evaluate the real economic viability of such a scheme. The major consideration in the viability of the TCP concept is the price of the delivered energy and its relationship to alternative energy sources. Preliminary investigations(4-1) have shown that the TCP system can deliver heat economically. However detailed attention must be paid to the efficient design of the chemical and power plants comprising the system.

In addition to the advantage of thermal transmission from a conversion thermodynamics viewpoint, other advantages to TCP transmission of energy should be pointed out. Since the transmitted energy has an exergy ratio less than that of the primary source energy, the undesirable thermal rejection at the primary site is eliminated. As long as a thermodynamic "uphill" climb is avoided there is no need to reject any low-grade energy at the source. This advantage can have important consequences in siting of the primary source and its effect on the environment. Similar advantages accrue at the delivery site where there are no unwanted products in addition to work and heat.

An enormous conservation potential is offered by thermal transmission combined with total delivery systems. Since the primary energy is at a temperature much higher than that of the demand, the option exists with thermal transmission of expanding high-pressure steam in a back-pressure turbine before delivering the process steam to the users (e.g., cogeneration). In thermodynamic terms this is equivalent to splitting the original thermal energy into two parts -- one at a much lower exergy ratio (exhaust, low-pressure steam), and the other at a much higher exergy ratio (electricity or shaft work). This method of generating electricity has been recognized generally as the most efficient since it constitutes complete utilization of energy and exergy. In view of the mismatch between future supply and demand of energy discussed earlier, it is necessary that thermal energy be transported and distributed to make possible a large-scale combined delivery system.

To illustrate the enormous conservation potential of this combination, consider the following example. If low-temperature industrial process heat ($9x10^9$ GJ/yr), residential/commercial space heat ($13.6x10^9$ GJ/yr), and electrical needs were to be supplied by the use of a combination of remotely located central sources, thermal transport, and cogeneration, aproximately $11x10^9$ GJ/yr would be saved (mostly in the form of oil and gas). The reduction in total energy consumption would be equivalent to approximately $5x10^6$ barrels of crude per day, a figure close to the current level of total foreign oil imports. It should be emphasized, however, that while the advantages of combined generation of low-pressure steam and electricity are not new to industries, they can only be available in the future (from coal and nuclear sources) in the majority of cases if thermal transmission is a reality.

In the preceding discussions, thermal transmission has been treated generically, regardless of how it may be achieved in reality, since the advantages mentioned do not depend specifically on these details. For example, if it were practical simply to transport steam as such in pipelines over a distance of approximately 160 km, all of the advantages mentioned above could be realized immediately. In reality, however, because of excessive thermal losses it is not practical to pipe steam over distances exceeding a few kilometers*.

An interesting point is, in cogeneration systems without the TCP, coal and nuclear energy cannot be used directly to supply the on-site heat source (except in large baseload situations). Total energy systems, then, would use more prime fuel at the industrial site, with large savings in fuel consumption in the utility sector. Unfortunately, since industrial steam is generated mainly from oil and gas, this would lead to an increased consumption of prime fuels with large reductions in utility fuel use (mainly coal and nuclear, in the near future). Thus, the net effect of a total energy system may well be an overall reduction in energy consumption but with an actual increase in prime fuel use and savings in coal and nuclear use. This is the exact opposite of the required fuel substitution described earlier. Therefore, a major advantage of thermal transmission is that it enables the use of more abundant future resources (coal and uranium) for supplying industrial heat needs currently served by oil and gas, and much more importantly, offers a possibility for enormous savings through total energy systems.

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*Recent studies on thermal transmission as steam or high-pressure hot water in insulated ducts (Ref. 4-7,4-8) indicate that there is a tradeoff between capital cost (insulation) and operating costs (thermal losses). This may lead to a reassessment of the distance beyond which pure thermal transport would be uneconomical; the current estimate for this distance is 50 km.

4.1.4 Historic Background and Current Status Of TCP Work

West German Program: While there are occasional references in the patent literature to the use of endothermic decomposition reactions for storing and transporting heat(4-9), the earliest proposals to use reversible chemical reactions for thermal energy transport as envisioned in the TCP concept originated in West Germany in the past decade(4-10,4-11). The proposed concepts were aimed at transporting thermal energy from a high-temperature, gas-cooled pebble bed reactor (PBR) under development at Kernforschungsanlage (KFA); the application was specifically aimed at providing low-grade thermal energy (as hot water) to a district heating network.

In the initial phase, chemical reactions involving dehydrogenation of ethane and propane were suggested; however these were found generally unsuitable for cyclic operation. The subsequent work with steam reforming of methane has been much more promising and as a result almost all of the West German TCP work is aimed at using this reaction coupled to a PBR. The end use of the methanation thermal output was originally space heat. Until recently, the German work has been directed primarily toward lower temperature designs for the methanator.

The West German program, generally referred to as Nukleare Fern Energie (NFE), is the largest TCP research effort in the world*. To include the acronym of their first experimental test facility for helium-heated reformers (einzelrohrversuchsanlage or "EVA") their proposed scheme has become known as the EVA-ADAM concept. The single tube refromer experiments on EVA have demonstrated the technical feasibility of the convectively heated reformer concept. Plans call for the continuation of the experimental phase in the form of a 30-tube reformer bundle linked to a matching methanator to form the first "test loop". In the meantime, numerous design analyses have been performed and potential problems in the primary interface between the reactor helium and the reformer(4-12) have been investigated.

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*The NFE program is a part of the general Nuclear Process Heat Program (PNP), which also includes nuclear coal gasification projects. The dominant project in NFE is the closed-loop demonstration with a 30 tube reformer (SUPER-EVA).

To summarize, the main thrust of the West German program is centered around the PBR and its use for process heat. The helium-heated reformer is considered by them to be a key component for a variety of process heat applications, including NFE, and a large fraction of their efforts is devoted toward establishing the technical feasibility of couplinga methane reformer to the PBR. Their primary interest in end-use application has been district heating; more recently, they have also been considering cogeneration and process steam applications as potential end uses for methanation heat.

General Electric Program: General Electric became interested in the KFA EVA-ADAM concept early in 1974. Preliminary analyses showed that the concept was inherently much more versatile than contemplated earlier. Several new concepts were developed incorporating different reactions, combined storage/-transmission applications, distributed electrical generation, on-site storage for peak electricity, application to process steam delivery, and LWR topping. The generalized concept was named Chemical Heat Pipe and later changed to the Thermochemical Pipeline at DOE's request.

DOE (then ERDA) became interested in the TCP concept in 1975 and funded General Electric's Corporate Research and Development (CR&D) to study the overall concept, chemistry that could potentially be used in TCP applications, possible heat sources, the potential market, and the position of the TCP concept in the overall U.S. energy management program. The conclusions from this contract are documented in the final contract report(4-1) from which the majority of the material presented here is extracted.

General Electric Corporate Research and Development has identified the cyclohexane dehydrogenation/benzene hydrogenation set of reactions as the best chemistry for matching the temperature levels currently available (LWR, solar, coal) but was unable to obtain funding to undertake an experimental evaluation of high temperature, high selectivity catalysts to carry out the desired reactions. Currently work at CR&D is being carried out under this DOE contract aimed at identifying more details of the methane reforming reaction and, in addition, the application of the HTR to other synfuels processes.

Other TCP-Related Projects in the U.S.: Over the last several years, there have been several groups in the United States pursuing a variety of TCP-related proposals. It would be fair to generalize that virtually all of them have in mind the use of reversible reactions for on-site collection and storage of thermal energy from a focused solar collector.

1. The SOLCHEM process proposed at the Naval Research Laboratories(4-13) aims to use the SO_2/SO_3 reaction system for harvesting and storing thermal energy from collectors for delivery to a steam generator. The main role of the TCP in this application is collection of thermal energy from several focused collectors and transportation over a short distance to a central site where the large-scale storage function is provided by a molten salt storage device. Recent work has concentrated in the use of the methane reforming reaction using CO_2 instead of water as the CO reactant.
2. Stevens Institute of Technology(4-14) has investigated the use of hydrogenation/dehydrogenation reactions, once again to store solar energy from focused collectors. The use of cylindrical, focused collectors is contemplated, with chemical reactions taking place in the solar tube heated at the focal line.
3. The work at the University at California at Berkeley has been aimed at the use of SO_2/SO_3 reaction for on-site storage of thermal energy from a focused solar tower source(4-15).
4. The efforts at the University of Houston have been directed at the solar "power tower" concept. The chemical reactions proposed consist of salt decompositions and methanol synthesis(4-13).
5. The work reported by Rocket Research Corporation(4-16) involves a survey of potential chemical reactions and their thermochemical and thermodynamic properties. The work has been aimed at identifying promising candidates, but not in connection with any specific end-use.
6. Open-loop applications of methane-based TCP have been investigated at the Institute of Gas Technology under DOE funding(4-17). This project is aimed at near-term (open-TCP) and mid-term TCP application. This concept will be discussed in more detail later in this section of the report.

7. Colorado State University is currently investigating the use of the high pressure ammonia dissociation/synthesis reactions for collection of solar energy in distributed collectors(4-18).

4.2 TECHNICAL ANALYSIS OF THE TCP

4.2.1 Introduction

In this section of the report an overview of the technical design work carried out previously(4-1) will be given as concerned with possible high temperature gas cooled reactor applications. First, possible chemistries and their interaction with the type and quality of thermal sources available will be briefly discussed. Then, the importance of the efficiency design of TCP will be demonstrated by the examples of the interaction of the TCP system with an associated power plant and the use of a two phase heat exchanger (referred to as the mixed feed evaporator, or MFE) in the reformer plant for the methane based TCP. Next, a summary of system efficiencies and the impact of changes in the various design parameters will be given. Finally, the summary will contain the key technical issues identified in the previous technical work.

4.2.2 Chemical Reactions and Thermal Sources

General Chemical Requirements for TCP Applications: Any chemical reaction that is to be considered for large-scale use in the storage or transport of energy through the TCP must meet many requirements. Among them are:

- The main energy-carrying reaction should exhibit fairly complete reversibility with few side reactions, since reactants and products must be recycled many times without undue losses.
- It should be possible to control reaction kinetics easily and sharply by the use of catalysts. Homogeneous reaction rates (without catalysts) should be very low, to inhibit chemical changes during heating and cooling.
- Chemicals and materials of construction should be inexpensive and abundant.
- Enthalpy change associated with the principal reaction should be large enough to achieve high energy densities and reasonable flowrates.

- Thermodynamic equilibria and overall reaction rates should be suitable in the temperature range of the input thermal source and the output end use.
- Working materials should present no safety hazard.

None of the many proposed reactions can satisfy all of these requirements; only a few satisfy most of them. In previous studies(4-1) a high priority has been placed on the technical feasibility of cyclic operation, adequate experience in large-scale operation, efficiency of operation, and reasonable economics. Variations in energy densities have a smaller influence on overall economics than, say, a necessity to use expensive alloys to overcome corrosion.

Thermodynamic Considerations: While it is true that candidate reactions must satisfy thermodynamic requirements, this is not a sufficient condition. The thermodynamics of a particular reaction are a good initial criterion for evaluating the potential of a particular reaction for TCP application, but other criteria must also be used to screen those reactions which are thermodynamically favorable. Remarks in this section deal only with thermodynamic issues.

In preliminary screening of reaction candidates, a simplified thermodynamic test is often used to estimate the temperature range for useful operation and the energy density. For example, if at ambient temperature T_0, the standard enthalpy and entropy of the forward, endothermic reaction step are $\Delta H^o(T_0)$ and $\Delta S^o(T_0)$ respectively, then the ratio $T^* = \Delta H^o/\Delta S^o$ is the "seesaw", or change-over temperature, roughly indicating the temperature levels of operation, since the Gibbs energy change (ΔG^o) equals approximately zero at T^*. The ability to change ΔG^o from some negative value (endothermic reaction favored) to positive value (exothermic reaction favored) by a given temperature change requires a large entropy change (ΔS^o) during the reaction.

The requirement for high ΔS^o has some important implications. One may deduce immediately that the class of reactions that lead to an increase in moles in the gas phase during the endothermic reaction should meet this requirement easily. By contrast, it would be surprising if reactions where both reactants and products are liquids or solids should prove to be good candidates. This general conclusion is consistent with reported lists of candidate reactions(4-16,4-19). One may conclude, therefore, that a good chemical reaction should have a large ΔH^o for high energy density, a large ΔS^o for easy reversal of the reaction without a large temperature drop, with the restriction of an appropriate value of T* midway between the thermal source temperature and the end-use temperature.

In many instances some of the reactants or products are either liquids or solids at ambient temperature and system pressure, but the reaction takes place entirely in the gas phase. As a result, the overall change is a combination of a chemical reaction and phase changes. In these cases, the requirements for ΔG^o apply to the gas phase reaction but the transported energy (ΔH^o) is that corresponding to the reaction written with appropriate condensed species. The exergy[*] associated with the transported energy is also that corresponding to the condensed species. For example, in the methane reforming reaction the effective exergy-to-energy ratio is lower for pipeline conditions (liquid H_2O) than for the gas phase reaction; this is also reflected by a change in T* by approximately 200°K between the reactions occuring with either liquid water or steam.

Such considerations play an important role in internal heat exchange and overall system efficiency. In addition, they may lead to erroneous conclusions about the adequacy of thermal source temperatures. In the case of methane reforming, T* of approximately 750°K for pipeline conditions with water should not be interpreted as indication that sources above this temperature will be adequate for the reaction. The relevant reaction T* (with steam) is approximately 960°K and indicates that a source temperature in excess of 1000°K is needed.

[*] Exergy is defined as the maximum work available from the chemicals; see Reference 4-1, Appendix 2 for a more detailed discussion of exergy.

Any TCP application based on a chemical reaction that is not currently carried out on a large scale will need more elaborate laboratory tests, catalyst development, pilot plant operation, design scaling, and finally industrial demonstration. Even for those candidate reactions that are currently practiced industrially on a large scale, there will be a development time because of major differences in application between TCP and chemical manufacture. It seems certain that any TCP system implemented will be the one that is least esoteric and closest to current industrial know-how.

TCP/HTR Candidate Reactions: The four reactions that appear to be of most interest for possible utilization for energy storage and transport are methane reforming, sulfur trioxide decompsition, phosgene decomposition, and ammonia decomposition.

Methane Reforming: The two reactions that have been proposed are:

$CH_4 + H_20 \rightleftarrows CO + 3H_2$ $\Delta H^o_{298} = 206.2$ kJ/mole
$T^* = 1059K$

$CH_4 + CO_2 \rightleftarrows 2CO + 2H_2$ $\Delta H^o_{298} = 247.3$ kJ/mole
$T^* = 962K$

The former was proposed in West Germany(4-10), and the latter was suggested as a possible alternative during earlier studies at General Electric Corporate Research and Development Center(4-19,4-20). In actual practice the water-gas shift reaction

$CO + H_2O\,(g) \rightleftarrows CO_2 + H_2$ $\Delta H^o_{298} = 41.2$ kJ/mole
$T^* = 979K$

occurs simultaneously.

Sulfur Trioxide Decomposition:

$2SO_3(g) \rightleftarrows 2SO_2(g) + O_2$ $\Delta H^o_{298} = 98.3$ kJ/mole
$T^* = 1036K$

This reaction has been proposed in recent studies(4-13,4-15) for energy storage. Its major advantage is a higher theoretical storage density resulting from the ability to store SO_2 and SO_3 as liquids. Storage densities achieved in actual designs depend on the overall conversion as well as energy efficiencies of various conversion steps. This system however suffers from corrosive chemicals and lack of industrial experience in the high temperature endothermic SO_3 splitting reaction and currently is not as attractive as the well known methane reforming system.

Phosgene Decomposition(4-21):

$$COCl_2(g) \rightleftarrows CO + Cl_2 \qquad \Delta H^o_{298} = 109.6 \text{ kJ/mole}, \quad T^* = 801.1K$$

The source temperature requirements are lower for this reaction than for the others in this class. In view of the toxicity and corrosion potential, this system may not be suitable for transport, and therefore is probably not a promising TCP candidate. The interesting aspect of this system is the possibility of a hybrid electrochemical/thermal system where the stored energy can be recovered electrochemically at a low temperature as partly heat and partly electricity; or conversely, the input step can be carried out by the use of off-peak electricity.

Ammonia Dissociation:(4-18)

$$2NH_3 \rightleftarrows N_2 + 3H_2 \qquad \Delta H^o_{298} = 46.2 \text{ kJ/mole}, \quad T^* = 466K$$

This reaction has a significantly lower T* than others mentioned here. However, this reaction temperature parameter can be increased by increasing the reaction pressure making the reaction amenable to HTR applications. However, the high pressures required (300 bar) would make interfacing the chemical system directly to the primary coolant in the HTR virtually impossible. In addition, the traditional ammonia synthesis reaction requires high temperatures for reasonable reaction rates and the conversion becomes thermodynamically limited to 15 to 20% per pass. The unreacted nitrogen and hydrogen must be separated from the ammonia product and recycled to the synthesis reactor. The inherent irreverisiblities of this separation would undoubtedly decrease the system efficiency; however, a more detailed design would have to be undertaken to determine realistic efficiencies for this chemistry.

Thermal Sources: This study is confined to HTR applications; therefore, other chemistries better suited for lower temperature sources (coal, solar, and LWR's) will not be discussed and neither will the characteristics of those sources. The major concern in evaluating the potential application of the TCP concept to HTR driven systems is peak reaction temperature achievable in the chemical reactor. This temperature is determined by two factors: 1) the peak He coolant temperature at the outlet of the reactor core, and 2) the temperature drop required to transfer the heat between the primary coolant and the chemical process stream. The two core designs under development, the prismatic core (HTGR) design by General Atomic, and the Pebble Bed (PBR) core design by West Germany may differ in practically achievable primary coolant outlet temperatures because of the fuel and core design. The HTGR outlet temperature has been demonstrated at approximately 750°C (1025°K), while the PBR outlet temperature has been demonstrated at 950°C (1225°K). As will be shown in more detail in the following section, the advantage of higher outlet helium temperature is increased conversion in the reformer. This increased conversion results in both increased efficiency and more attractive economics. However, materials for reformer construction have not yet been identified which can withstand the reducing environment of the helium coolant at the 950°C outlet temperature for 30 years as required for licensing. Therefore, the tradeoff associated with He outlet temperature is one of lower efficiency but known materials of construction for the lower temperature versus higher efficiency but unknown materials of construction for the higher temperature.

4.2.3 Importance of Design Efficiency in TCP Applications

Traditionally, chemical processes have been designed to maximize profits. The design criteria usually include maximum conversion to desired product, minimum capital investment, and minimum operating cost. Many times, these design criteria are directly opposed to each other, and the dominating factor is determined as the criteria which has most influence on the profitability of the chemical process.

The TCP is a unique combination of a chemical process technology and an energy delivery system. The TCP involves large chemical plants but has no expensive chemical product to sell. Instead, its only product is energy; energy that must be transported at a minimum cost and supplied to the user on a reliable basis. Costs in the TCP system are determined by the energy efficiency and the capital costs of the system. In general, increased capital investment can lead to increased efficiency. However, there is some optimum design where an increase in capital does not lead to an economical reduction in the irreversibilities of the system and, hence, in increased efficiency. This optimum design will change as the cost of energy (relative to capital) changes. Thus, the design that may be attractive today will become unattractive in the future when the cost of energy increases and it becomes an economic necessity to increase the efficiency of the system.

In the past, the analysis of the TCP(4-1) has concentrated on maximizing both first and second law efficiency, e.g., delivering as many units of energy as possible and maximizing the quality of the energy delivered. This has resulted (as shown in Figure 4-3) in a design that involves an intricate system of heat exchange in the reformer plant designed not only to recovery as much "waste heat" as possible from the system but also to recovery as much available work as possible from the waste heat. The system shown in Figure 4-3 serves a dual purpose, firstly, to transfer heat available from the product stream to the reactant stream to provide preheat and, secondly, to provide a means of interchange of heat with a companion power plant to match not only the heat requirements of the process but also to match the temperature levels of the heat transferred.

This is shown diagramatically in Figure 4-4 which shows the heat liberated from the product stream as it is cooled vs. temperature along with the heat required by the reactant stream vs. temperature as it is heated to the inlet temperature of the nuclear reformer. Three separate areas are shown, (1) the blank area where the two duties overlap indicating a match in both duty and temperature level, (2) the cross hatched area where there is an excess of heat available (at that temperature) from the product stream, and (3) the dotted area where there is insufficient heat available from the product stream to provide the heating requirements of the reactant stream. Traditionally, the high temperature heat surplus would be used to supply a portion of the thermal deficit of the reactant stream and the remainder of the deficit would be made up from the high grade (temperature) heat available from the reactor source. In addition, the low grade thermal surplus would be rejected to cooling water. This solution would close the energy balance on the plant, but would result in an inefficient system.

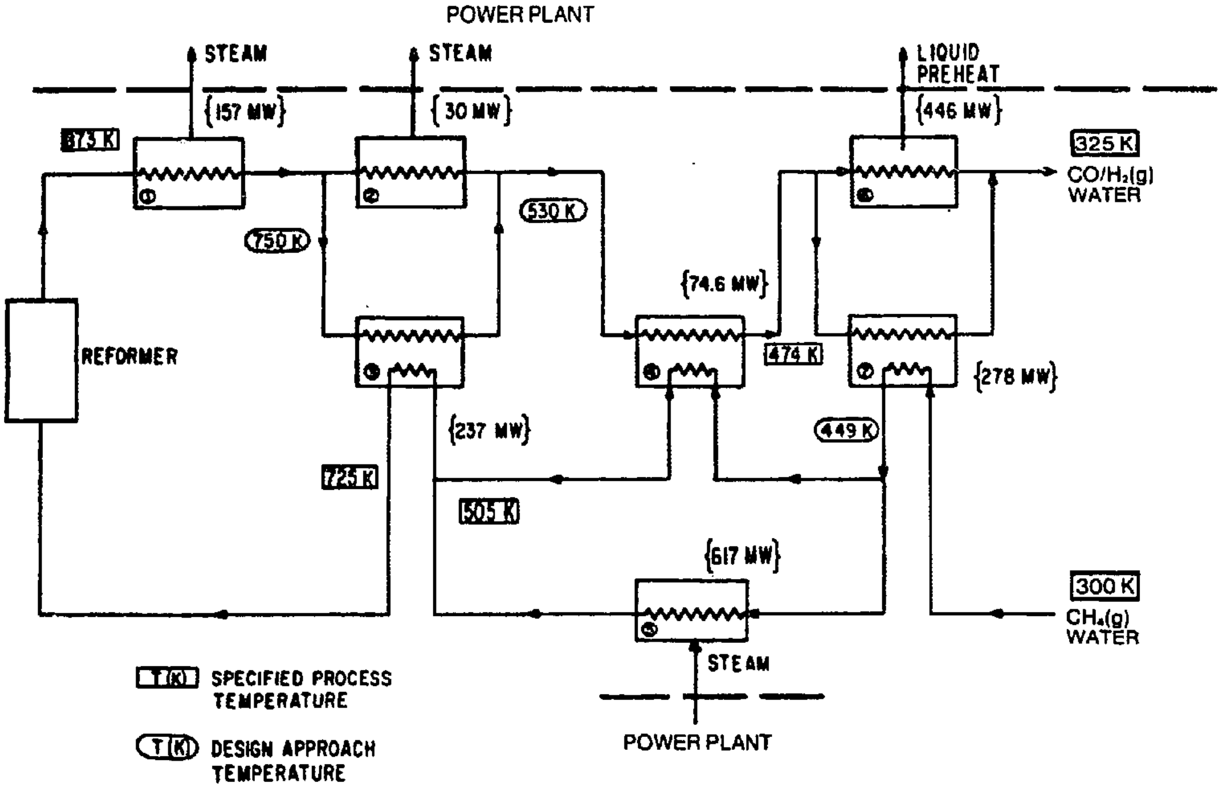

Figure 4-3. HEAT EXCHANGER TRAIN (1000 MW REFORMED)

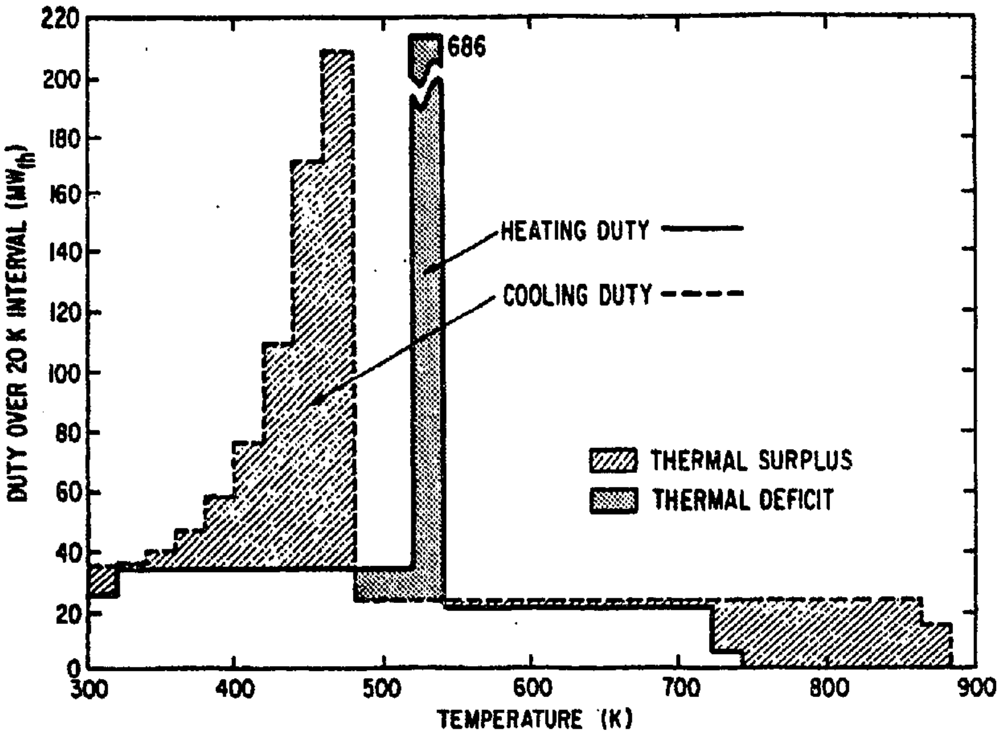

Figure 4-4. THERMAL DUTY FOR PROCESS GAS HEATING AND COOLING IN TCP REFORMER PLANT

In an efficient design, the excess high grade thermal energy from the chemical process would be used to superheat the steam in the power cycle this this accompanying the TCP endothermic chemical plant. The excess low grade thermal energy would be used as boiler feed water preheat in the power cycle, and the intermediate grade thermal deficit in the chemical system would be made up by extraction steam from the turbine in the power plant. Approximately 85% of the available work from the thermal excess can be utilized in the power cycle and taken as a credit. Conversely, about 85% of the available work in the extraction steam would be lost from the power cycle and would be taken as a debit.

Another portion of an efficient design is the elimination of irreversibilities that detract from the efficiency of the system. An excellent example of this approach is the use of the Mixed Feed Evaporator, or MFE, in the reformer plant. The large spike representing a thermal deficit in Figure 4-4 results from the evaporation of the feed water at the 40 atmosphere operating system as shown in Figure 4-5. The following discussion demonstrates the importance and reasoning behind an alternate approach of mixing the water and gas streams before this evaporation takes place.

When the reaction between CH_4 and H_2O proceeds to form CO and H_2 two things happen: H_2O is consumed, and the mole fraction of H_2O also decreases. The consequence of the former is that condensation heat of the effluent is less than the corresponding evaporation heat for the feed; the consequence of the latter is that the temperature level of the condensation heat is lower than that for the evaporation because of the effect of H_2O partial pressure. However, this effect does not imply an irreversibility; exactly the converse takes place upon methanation, where both moles and mole fraction of H_2O increase, and so do the condensation heat and its temperature level.

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*Exergy again refers to the available work; see Reference 4-1, Appendix 2 for a detailed explanation of this concept.

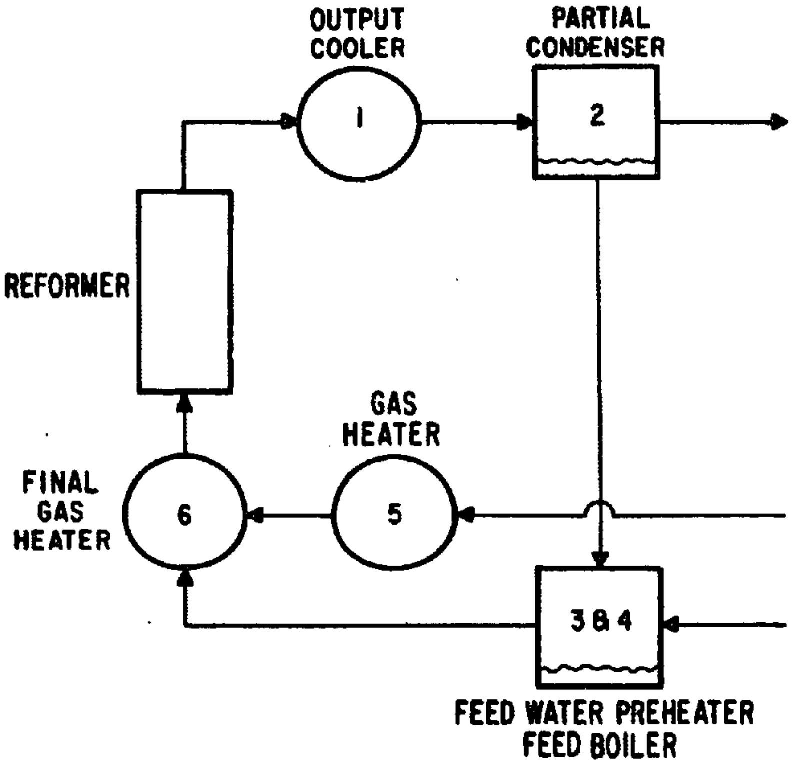

Figure 4-5. SCHEMATIC FLOW DIAGRAM OF REFORMER PLANT IN TCP

The manner in which reactor feed streams are prepared, on the other hand, can lead to significant irreverisibilities. When two gaseous streams from the boiler and the gas preheater are mixed they undergo irreversible mixing. The thermodynamic price for this is not immediately evident until the process is "undone" by separating out the water molecules in the partial condenser. For the example case reformer plant, this irreversible entropy of mixing and the associated exergy* loss have been calculated(4-1) and show an exergy loss of 39.5 MW due to the feed steam/gas mixing. There is a great incentive to eliminate this loss and thereby improve the overall process efficiency. A modification in the process of preparing the feed stream has been developed to eliminate this irreversible step.

In the reformer plant, the requisite amounts of H_2O are introduced into the gas stream by first evaporating (boiling) it separately and then mixing the two gas streams. As a result, the heat of evaporation would be required at the boiling point corresponding to the total system pressure of 40 bars. If, instead, the gas and liquid streams are introduced together in a heat exchanger and allowed to heat up with a continued evaporation process (much like humidification) progressive amounts of H_2O can be introduced into the gas stream at much lower temperatures. The net effect is to approach conditions that are exactly the reverse of partial condensations; at each point in the heat exchanger, the amount of water vapor in the gas stream corresponds roughly to the equilibrium vapor pressure at the local temperature.

The reformer plant process flowsheet with this modification, as shown in Figure 4-6, is only slightly different from that in Figure 4-5. The two units -- feed boiler and gas heater -- are now combined into one MFE. As mentioned earlier, this does not alter the total heating duty; however, the temperature levels at which heat is needed is changed. This shows up most clearly in the histogram shown in Figure 4-7.

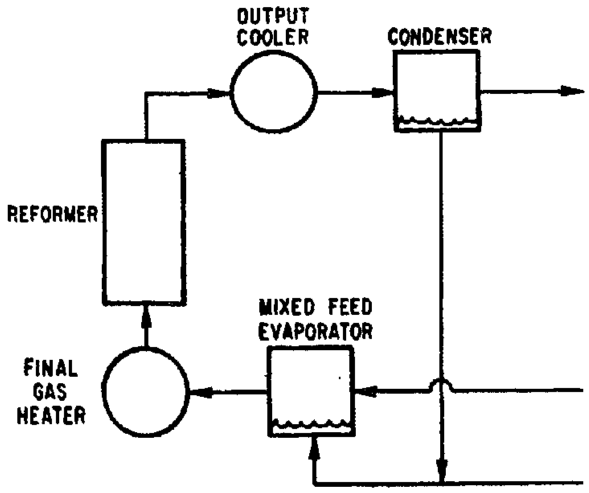

Figure 4-6. SCHEMATIC FLOW DIAGRAM FOR REFORMER PLANT IN TCP WITH MIXED FEED EVAPORATOR

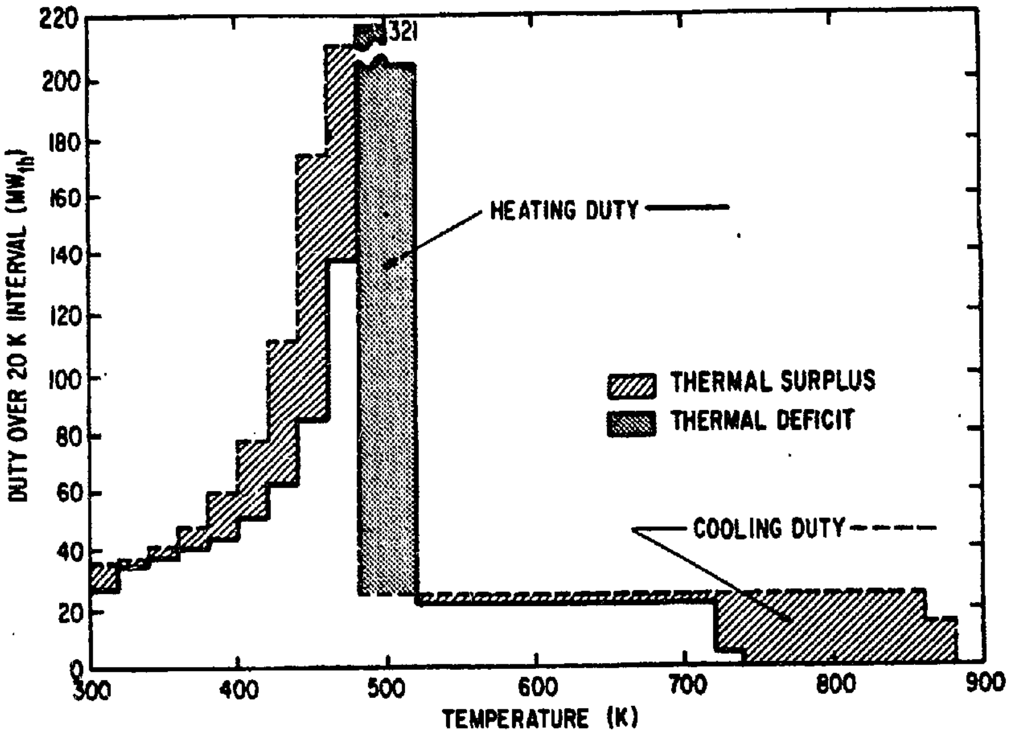

Figure 4-7. THERMAL DUTY FOR PROCESS GAS HEATING AND COOLING IN TCP REFORMER PLANT WITH MIXED FEED EVAPORATOR

A comparison with Figure 4-4 shows that the large spike at 524°K is no longer present; the same duty, instead, is spread over a range of temperatures from 505°K to 300°K resembling qualitatively the partial condenser. The overall affect on the heat exchanger is a reduction in the amount and the temperature level of the thermal deficit that must be made up by extraction of steam from the powerplant. The magnitude of the feedwater preheat is also reduced. This reduction in heat exchanges with the powerplant occurs because some of the condensation heat can now be used to evaporate water in a partial condenser/MFE combination. The impact of these modified exchanges can be evaluated by computing the exergy associated with thermal exchanges. The exergy trades show a surplus of 24.3 MW. Thus, the difference in the net exergy flow with and without MFE is 40 MW (for 1000 MW_t transported), which is comparable to the irreversible mixing loss.

These examples of powerplant interchange and mixed feed evaporator are presented to emphasize the importance of efficient design in the TCP system. Again, it must be emphasized that this system is a synthesis of chemical plants and a thermal conversion plant and that these types of concepts must be adopted to insure that the system operates at its potential efficiency.

4.2.4 Variation of System Characteristics with Changing Design Parameters

In the discussion of thermal sources it was mentioned that two different types of HTR's are in the development phase and that there are advantages and disadvantages to the current outlet temperature achievable by each. One potential problem mentioned previously is the practical limit in helium outlet temperature, as determined by reformer construction materials and reactor type, available for supplying heat to the process gas stream. In addition to this variation in the source temperature, changes in the process stoichiometry and pressure can be implemented in order to increase the methane conversion and, hence, the amount of gas that must be transported in the pipeline decreases. In each case where these modifications are made, not only is the conversion affected but process equipment sizes are also affected. For instance, an increase in steam to methane ratio increases the conversion of methane but also increases the heat duty in the heat exchange for heating and cooling the process gas streams. This increase in duty requires additional heat exchange area and results in increased capital investment for the larger equipment. Thus, changes in design parameters will result in variation of economic and efficiency measures. It is not obvious "a priori" whether any particular change in the process design will positively or negatively affect these measures of the quality of the process.

In this section, a brief discussion will be given to show the effect of some design variations on the process economics and efficiency. No detailed optimization for the TCP system has been undertaken as yet; however, the analysis presented here gives a first order estimation of the positive and negative effects of changes in the design parameters on the system economics and efficiency.

Reactor Outlet Temperature: The effect of the process gas peak reaction temperature on important process parameters is shown in Table 4-1. The process gas peak reaction temperature is related to the reactor outlet temperature by the temperature drop through the heat exchanger from the helium coolant to the process gas stream. The positive effect of increased reactor outlet temperature is obvious from the data presented in Table 4-1. Both the efficiency and economic measures for the TCP system are more favorable as the reactor outlet temperature is increased. There is obviously incentive to increase the peak process temperature; however, not shown in this table is the effect of the difficulty of obtaining materials of construction for use at higher temperatures.

H_2O Recycle Rate: As shown in Table 4-2, increasing the H_2O recycle enhances the conversion because of a higher H_2O/CH_4 ratio, thereby increasing the energy density and reducing the transport flow rates. The disadvantage of a high H_2O recycle is the resultant increase in heat exchange load and associated exergy losses. The indications are that the disad vantages outweigh the advantages. As a result, it is generally advantageous to use lower H_2O/CH_4 ratios. It is difficult to specify the optimum H_2O recycle; however, it appears that for the basic design it would be advantageous to use a lower H_2O recycle. The lower H_2O/CH_4 ratio might well be set by the fact that if there is not sufficient steam preent, the methane will decompose and deposite carbon on the catalyst making it ineffective. In cases where a high conversion is obtained, a reduction in H_2O recycle is an important way of trading back some of the conversion for a reduction in equipment duties and irreversibilities.

Choice of Operating Pressure: The selection of reformer pressure may be dictated by interface problems and materials capability at elevated temperatures.

Table 4-1

Process Characteristics and Economics of Thermochemical Pipeline as a Function of Peak Reformer Process Temperature

Basis: $H_2O/CH_4 = 3{:}1$; constant methanation conditions; mixed feed evaporator; 40 atm pressure; 1000 MW_{th} transported at pipeline conditions
160 km transmission distance

Reformer Temperature K (C)	CH_4 Conversion %	Reformer Heat Duty MW_{th}	Reformer Heat Exchange Duty MW_{th}	Efficiency* Index	Incremental Cost Above HTR Heat Cost \$/GJ	Total Delivered Heat Cost \$/GJ (HTR Heat @ \$2.50/GJ)
1200 (925)	81.9	981	904	88.1	1.26	3.76
1100 (825)	64.8	1003	1239	85.7	1.47	3.97
975 (700)	33.3	1106	2360	80.2	2.10	4.60

*Efficiency index defined as $\frac{\text{Heat Delivered at Methanator}}{\text{Heat Consumed at Reformer + Equivalent Thermal Energy for all work requirements}}$

Table 4-2

Process Characteristics and Economics of Thermochemical Pipeline as a Function of Water Recycle Rate at the Reformer

Basis: 1100°K peak process temperature; constant methanation conditions; mixed feed evaporator;
40 atm pressure; 1000 MW_{th} transported at pipeline conditions
160 km transmission distance

H_2O/CH_4 Ratios	CH_4 Conversion %	Reformer Heat Duty MW_{th}	Reformer Heat Exchange Duty MW_{th}	Efficiency* Index	Incremental Cost Above HTR Heat Cost $/GJ	Total Delivered Heat Cost $/GJ (HTR Heat @ $2.50/GJ)
2:1	49.4	1010	1075	87.5	1.45	3.95
3:1	64.8	1003	1239	85.7	1.47	3.91
4:1	69.8	1005	1413	84.2	1.52	4.02

*Efficiency index defined as $\frac{\text{Heat Delivered at Methanator}}{\text{Heat Consumed at Reformer + Equivalent Thermal Energy for all work requirements}}$

This limitation aside, Table 4-3 shows that a lowering of pressure, especially in cases with low conversions, is generally beneficial in reducing equipment duties and flow rates. The analysis used as a basis for the data presented here tends to underestimate the negative effect of lower pressure operation -- a result deriving from assumptions about work needed to overcome equipment pressure drops and about pipeline cost dependence on the pressure. The major effect of low-pressure operation of the reformer is in reducing correspondingly the transmission and methanation pressures. This is so because it is preferable to operate the whole system at lower pressures than to take a large penalty in compression energy that cannot be recovered. The disadvantages of a low-pressure operation of reformer and methanation plants is that reactor volumes become large, heat transfer coefficients may suffer, and the pressure drops in the reactors heat exchangers are more energy consumptive. At low pressures, the transportation involves large pipelines and increased friction. Unfortunately, the assumptions of the transportation analyses used here do not penalize low-pressure operation sufficiently; the material cost, which is dominant for large pipes, remains approximately the same; the increase in diameter is compensated by a reduction in wall thickness. This will not be true in the limit, as some criterion other than hoop stress will determine wall thickness. Thus the pipeline compressor work is underestimated and, in view of the assumptions, so is the work to overcome equipment pressure drops. A full accounting of low-pressure designs would involve extensive equipment design and optimization which has not been undertaken.

4.2.5 Summary

In this section the possible chemistries for the HTR driven TCP have been reviewed and the methane reforming system has been studied in detail. The importance of efficient design has been emphasized and the effect of variation in design parameters has been briefly reviewed. Low reaction pressures and low H_2O/CH_4 appear to be favored at lower reaction temperatures where conversions are low. In general, more detailed calculations are required before the optimum design can be identified.

Table 4-3

Process Characteristics and Economics of Thermochemical Pipeline as a Function of Reformer Reactor Pressure

Basis: $H_2O/CH_4 = 3:1$; constant methanation conditions; mixed feed evaporator;
1000 MW_{th} transported at pipeline conditions
160 km transmission distance

Reactor Pressure (bars)/ Temperature (K)	CH_4 Conversion %	Reformer Heat Duty MW_{th}	Reformer Heat Exchange Duty MW_{th}	Efficiency* Index	Incremental Cost Above HTR Heat Cost $/GJ	Total Delivered Heat Cost $/GJ (HTR Heat @ $2.50/GJ)
40/1100	64.8	1003	1239	85.7	1.47	3.97
25/1100	70.3	988	1058	87.5	1.36	3.86
40/975	33.3	1106	2360	80.2	2.10	4.60
10/975	57.6	994	1300	86.0	1.54	4.04

*Efficiency index defined as $\frac{\text{Heat Delivered at Methanator}}{\text{Heat Consumed at Reformer + Equivalent Thermal Energy for all work requirements}}$

4.3 THE OPEN THERMOCHEMICAL PIPELINE

4.3.1 Introduction

When the TCP concept is examined in detail, it is apparent that the closed loop could be opened, as shown in Figure 4-8 . That is, heat could be as chemical energy via an endothermic reaction, the products of that reaction transported to a user site, an exothermic reaction carried out to release the stored energy, and then the products of that reaction could be sold at the user site. In this section, a brief overview of this concept will be given concentrating on possible reaction systems and discussion the advantages and disadvantages of the idea.

4.3.2 Possible Reaction Systems

There are two key constraints on choosing reaction systems for the open TCP. First, the chemical raw materials for the endothermic reaction must be readily available at the centrally located thermal energy source. Since they will not be returned to the central site as in the TCP, the source of these raw materials will be consumed. Secondly, the product of the exothermic reaction must be one that can be readily sold for profit or disposed of economically. The two most likely hydrocarbon materials that could be used at the central site are natural gas and coal. The other candidate material would be water.

If natural gas were employed, it could be reformed to synthesis gas, or "syngas", transported and then an exothermic reaction carried out at the user site. Exothermic reactions that could be coupled to a syn gas pipeline will be discussed in the next section. For natural gas reforming, nuclear heat could be used to supply the energy requirements of the reaction and thus about one fifth of the total heat content of the delivered heating value of the syn gas would be nuclear derived.

The other source of carbon that appears attractive is coal. In Section 5 of this report, it is shown that nuclear energy can be used to gasify the coal. With a syn gas product, coal consumption can be reduced by approximately 40% by the use of the HTR. The syn gas product from the gasifier could then be transported and reacted to yield process heat, cogenerated eelctricity, peak electricity and a salable chemical product.

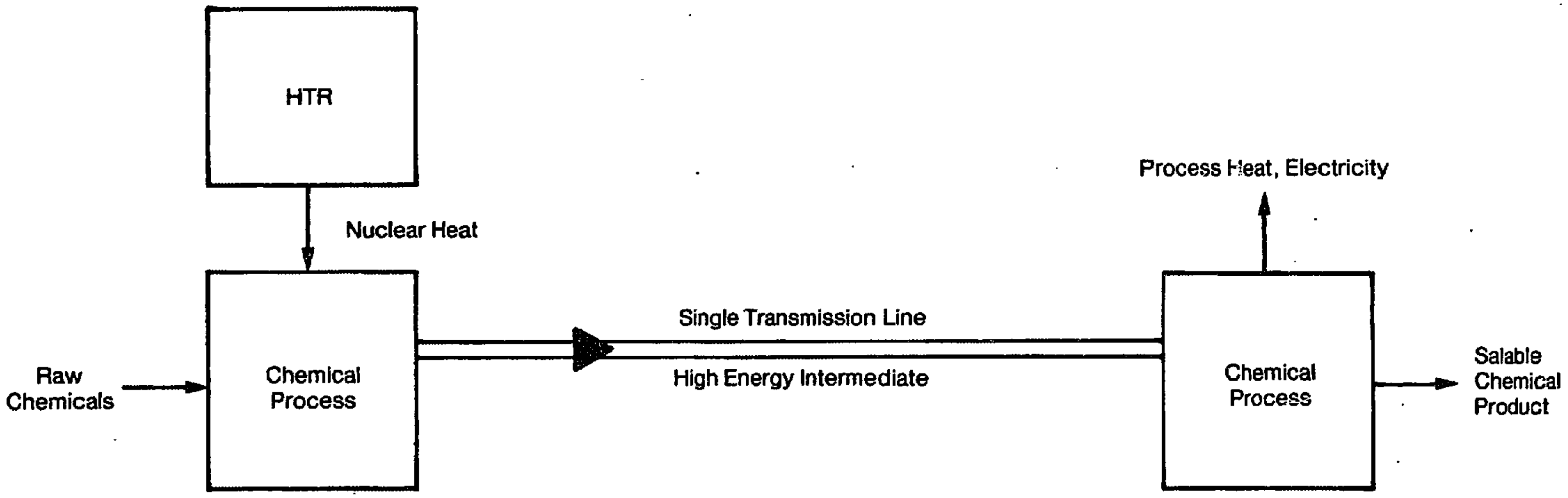

Figure 4-8. THE OPEN THERMOCHEMICAL PIPELINE CONCEPT

The other raw material for use in the open TCP concept that has received much attention is water. In this case, the water would be split into hydrogen and oxygen with the hydrogen shipped to the user site and the oxygen vented or sold as a byproduct. The nuclear energy required to decompose the water could be supplied as electricity or thermally to drive what is commonly referred to as a Thermochemical Hydrogen Process. Previous studies have shown(4-1) that the hydrogen option would operate a significantly lower efficiency than the carbon based options. However, if the hydrogen could be used in a thermodynamically efficient manner at the user site, the efficiency of the process could be improved. To date, little attention has been paid to the hydrogen utilization at the user site; therefore, current efficiencies are significantly below those theoretically obtainable.

4.3.3 Possible User Site Reactions and Chemical Products

Most discussions to date of the open TCP(4-17) concept have centered on the use of the synthesis gas intermediate as a reactant in a methanation process to form Substitute Natural Gas (SNG) and product heat. The SNG would then be sold into the residential pipeline grid and the product heat used to produce process steam, cogenerated electricity, or peaking electricity.

As shown in Figure 5-12, the synthesis gas intermediate could also be used in a variety of other reactions and end uses other than the production of SNG. These reactions are all exothermic and liberate heat when carried out. In particular, methanol synthesis, Fischer-Tropsch gasoline synthesis, and ammonia synthesis would all be net energy producers if efficient processes could be designed to utilize the available heat of reaction. It must be realized that the temperature history in the methanol synthesis and Fisher-Tropsch reactors might well be highly important to the product distribution. A scheme to extract high quality thermal energy from these reactions might alter the product distribution such that undesirable byproducts would be formed yielding an uneconomical process. The idea of using chemical reactions other than methanation has received little attention to date and requires further investigation to prove its merits.

The other option in the open TCP concept is the water splitting, hydrogen intermediate process. Most discussions of this concept center on combustion of the hydrogen to yield heat and form a water product which would be discarded. The advantage of this reaction system would be the production of the benign water as a product. The combustion of hydrogen is thermodynamically inefficient since materials are not known which can withstand the high theoretically obtainable temperatures and, at the same time, extract the potentially available work from the combustion gases. An alternative would be to convert the hydrogen in a fuel cell to yield electricity and heat from the inefficiencies. This concept needs to be studied in much greater detail if it is to prove efficient and economical.

4.3.5 Advantages and Disadvantages of the Open TCP Concept

The obvious advantage of the open TCP system is that it offers a means of transporting the non-fossil nuclear energy supply by the HTR at a remote site to a series of users located in one or more sites. It also offers a unique means of combining coal gasification, an HTR, process heat delivery, and chemicals production into an efficient, integrated system. The centrally located endothermic reactor could be large and base loaded (operating continually) to take advantage of the capital investment in the equipment and the economy of scale. The pressure in the pipeline could be varied to provide storage of the intermediate energy carrier and, thus, supply energy at the user site on a one or two shift basis at a very economical price(4-1).

The disadvantages of this scheme are concentrated in the piping of the synthesis gas (or hydrogen) intermediate. The increase in the number of moles (from 1 to 4) of gas during methane reforming would require additional pipeline volume and compression power as compared to natural gas transmission. This cost is approximately two to three times as much per unit distance when compared to natural gas transmission. This means that it would probably be uneconomical to ship the synthesis gas over a long distance (1000 miles) as is done with natural gas. However, for ranges up to 200 miles, the affect of the transportation cost on the overall economics would not be overwhelming.

The transmission of hydrogen and carbon monoxide presents some obvious hazards. Government regulations and extensive safety studies are not yet available and will be required before the system can be implemented.

4.4 SUMMARY

In this section of the report, an overview of the closed-cycle TCP and open-cycle TCP concepts have been given. The importance of the TCP concepts discussed here is the ability to store and transport non-prime fossil energy to industrial users to produce process steam, peak electricity, or cogenerated steam and electricity on a periodic basis. These concepts offer an efficient means of economically delivering energy derived from a large remotely located site to industrial users located up to 200 miles from the source.

These concepts offer advantages of reduced fossil fuel consumption through substitution by nuclear energy and by conservation by high efficiency. Both concepts are in embryonic stages of development but it is clear that operating parameters such as peak process temperature, reaction pressure, and reaction stoichiometry will have a major impact on both process efficiency and process economics. More detailed design and economic investigations must be carried out to identify the optimum process configuration.

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SECTION 5
NUCLEAR HEAT UTILIZATION IN COAL PROCESSING AND CHEMICALS PRODUCTION

5.1 INTRODUCTION

In this section of the report, the application of nuclear heat from a High Temperature Gas Cooled Reactor (HTR) to coal treatment to produce liquid and gaseous fuels and chemicals will be investigated. The incentive behind this investigation is the fact that coal appears to be one of the sources of carbon* that will be readily available over an extended (100-300 years) period in the future and that there is a possibility to substitute, with more economical nuclear energy, a portion of the coal energy consumed in the conversion processes to liquids or gases. This substitution would result in a less expensive product, reduced coal consumption (with resulting extension of reserves), and reduced emissions of effluents and waste products such as CO_2, SO_2, NO_x, particulates, trace toxic elements, and ash.

The objectives of this study are (1) to briefly summarize the types of gasification processes available for consideration, (2) to select possible gasification processes in which nuclear energy could be utilized, (3) to compare nuclear and conventional heat supplies for these processes, and (4) to analyze the application of the optimum nuclear process to the production of various chemicals and fuels from coal.

Comparison of different technologies that are not yet commercial is difficult at best. In this analysis, certain common bases are used to facilitate comparisons. Specifically, the assumed feed to all processes, Illinois No. 6 coal, was chosen because of its "workhorse" characteristics in the coal economy. It is a bituminous coal which is not as easily gasified as lignite; however, it does have a higher carbon content per pound which requires a lower feed rate than lignite. The analysis of the Illinois No. 6 coal used here is given in Table 5-1$^{(5-1)}$. Most conclusions in this analysis drawn from a comparison of treatment processes utilizing Illinois No. 6 are also valid for processes with other types of coal feeds**. A total coal feed rate of 12,000 tons per day was chosen because it is

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*The other sources being heavy oil and oil shale.
**Illinois No. 6 is equivalent to Kentucky #11 and Indiana VI B

Table 5-1

Assumed Composition of Illinois #6 Coal

Proximate:	Moisture*	4.2 wt%
	Ash	9.6
	Fixed Carbon	52.0
	Volatile Matter	34.2
Ultimate (DAF)	C	77.26 wt%
	H	5.92
	S	4.29
	O	11.14
	N	1.39

High Heating Value (HHV) 12235 Btu/lb

Low Heating Value (LHV) 11709 Btu/lb

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*Moisture may be as high as 11% in some instances.

approximately the size of one large U.S. coal mine. Each individual gasifier is sized at a coal rate of 2000 short tons/day (ST/D). Thus, each of the plants would require six gasifiers. In the past, higher rates have been used to take advantage of larger nuclear reactors. Preliminary analyses have shown that, in almost all applications other than electricity production, heat usage rates are significantly below the "standard" 3000 MW level usually chosen for nuclear reactors. In light of this and other factors such as siting, transportation, and technology nucleation, the 12,000 tons per day is used as the standard feed rate.

Following this introduction, Section 5.2 describes three basic types of conventional coal gasification processes, the application of nuclear heat to two of these processes, and then compares the conventional and nuclear processes. Section 5.3 describes the application of a catalyzed coal gasification system and a nuclear heat source to supply hydrogen and thermal energy to processes for the production of hydrogen, coal liquids, ammonia, steel, and, briefly, other applications. Finally, Section 5.4 summarizes the major ideas developed here and offers conclusions as to the technical, economic, and environmental viability of using nuclear heat in the processing of coal.

5.2 COAL GASIFICATION PROCESSES

5.2.1 Introduction and Process Selection Criteria

A wide variety of coal gasification processes are currently under development in both the public and private sectors. For this analysis, three general types of processes were chosen based on the type of gasifier used. The three chosen are entrained bed, fixed bed, and fluidized bed gasifiers. The process selection based on the type of gasifier was made in order to identify the best type of gasifier for nuclear heat applications and to understand the process changes that would have to be undertaken to utilize thermal energy from a nuclear reactor in coal gasification. Several different companies are working on each type of gasifier; however for this study the Texaco entrained bed, the Lurgi fixed bed, and the Exxon fluidized bed processes were chosen as typical examples of the three types of gasifiers. This choice was made based upon the availability of literature data and does not imply that other processes could not be used.

Different gasification concepts are compared in a uniform processing plant design as shown in Figure 5-1. After gasification, the raw gas is cooled and the acid gases (H_2S and CO_2) are removed. The hydrogen to carbon monoxide ratio is adjusted to 3.1:1 by the water gas shift reaction and the CO_2 formed in this conversion is removed. Methane is then formed by a methanation reaction and the product substitute natural gas (SNG) is dried and compressed to pipeline pressure. The operating conditions for the gasifiers and the raw gas treatment were taken from the literature. In the case of the Exxon fluidized bed reactor design, the product treatment is modified to account for the greater amount of methane formed in the reactor. Since the references used describe diverse applications of gasification systems (fuel gas production, different coal feeds), not all process operations are optimally designed for this study. This approach was taken, however, in order to keep the comparison between the various gasifiers using both coal and nuclear heat sources on a consistent basis. The assumptions used in designing the gasification process are summarized in Table 5-2

In the first portion of this section, the characteristics of the conventional Texaco entrained bed gasifier are described. These characteristics include a process description, a simplified flow sheet, and the material balance. Next, the process design characteristics are given for a Lurgi fixed bed reactor system using

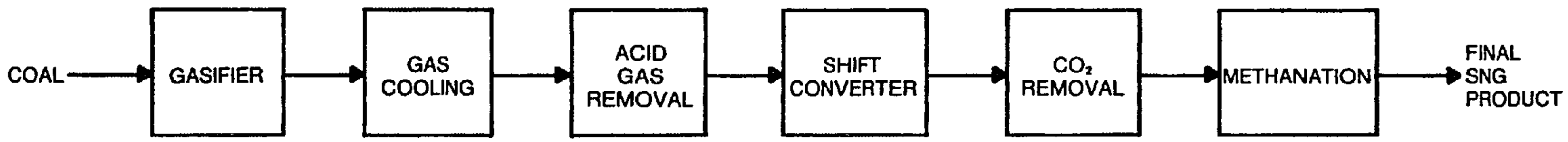

Figure 5-1. SCHEMATIC FLOW DIAGRAM FOR GASIFICATION PROCESS

Table 5-2

Assumptions Used in Gasification Process Design

o Oxidant Composition	O_2 98 mole % N_2 1.5 mole % Ar 0.5 mole %
o Quench	- All tar and oil condensed - Gas cooled to 100°F - Pressure drop 13 psi - Gas saturated with water - All ammonia removed
o Acid Gas Removal	- 98% of CO_2 removed - 100% of H_2S removed - Outlet at 100°F saturated - 8 psi pressure drop - All COS hydrolyzed
o Water Gas Shift Converter	- Shift reaction in equilibrium at 900°F - H_2/CO ratio 3.1:1 - 23 psi pressure drop - Output cooled to 100°F
o CO_2 Absorber	- 98% CO_2 removed - Outlet gases at 100°F saturated
o Methanation Reactor	- Water gas shift and methanation reactions in equilibrium at 791°F - Outlet gases cooled to 100°F - 20 psi pressure drop
o Final Product	- Dried to trace of water - Compressed to 1015 psia

both coal and nuclear heat sources. Then a comparison of the conventional and nuclear heated Exxon catalytic gasifier systems is made. Finally a summary of the concepts developed in this section is given.

5.2.2 Texaco Entrained Bed Reactor(5-1,5-2)

5.2.2.1 Process Description

The key element of this process (Figure 5-2) is the oxygen blown entrained bed gasifier in which coal is fed as a water slurry (66.5 wt % coal) from the top cocurrently with the oxygen. The gasifier is a refractory lined carbon steel shell

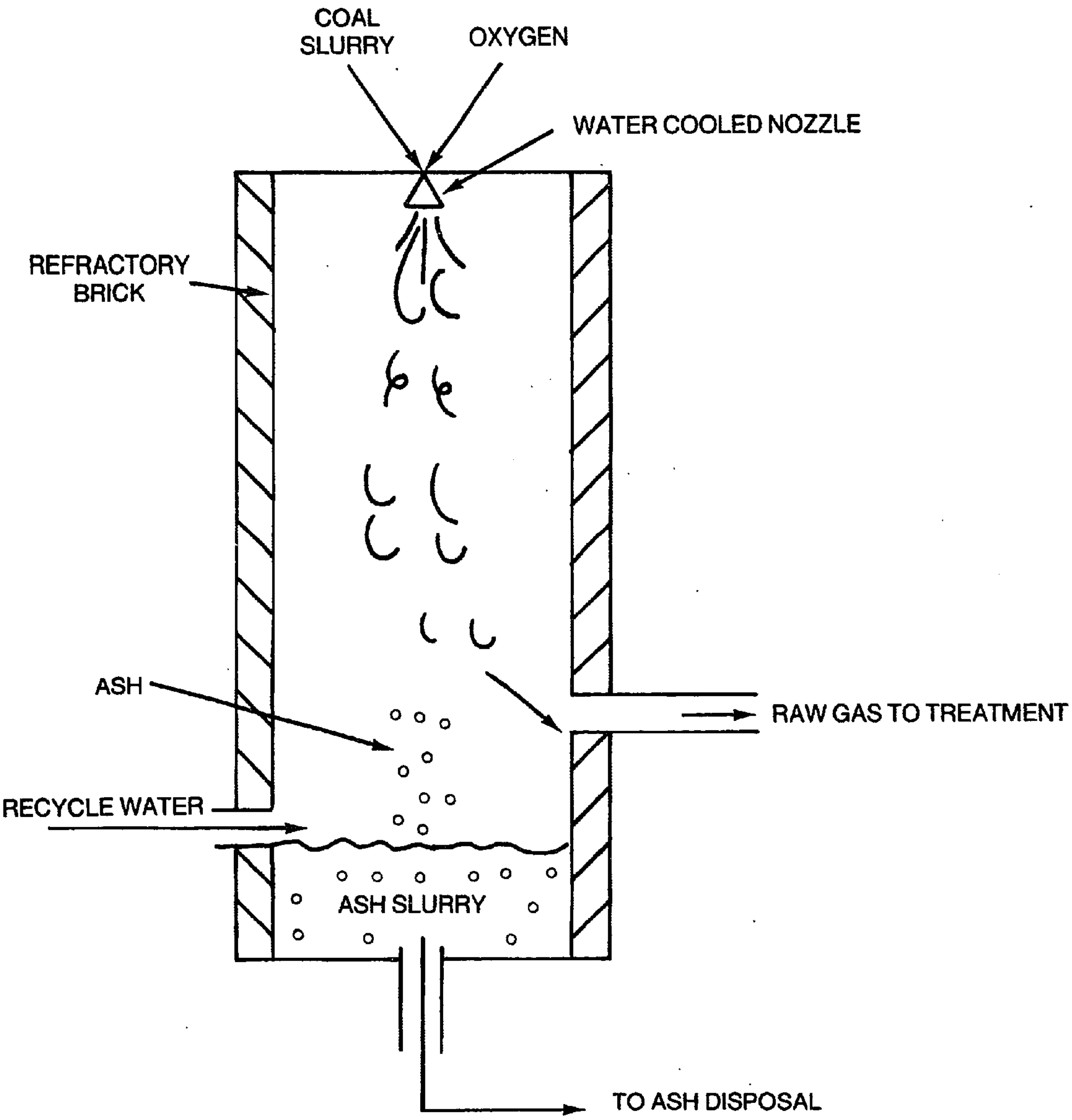

Figure 5-2. TEXACO ENTRAINED BED GASIFIER

which operates at 600 psig with a process gas outlet temperature of from 2300°F to 2600°F. The slag formed from the ash drops into quench water at the bottom of the reaction vessel and is removed as a slurry to dewatering and disposal. The product gas contains mostly steam, hydrogen, carbon monoxide, and carbon dioxide as shown in Table 5-3. Very little methane, tar, pitch, or heavy oils are present due to the high reaction temperature.

Table 5-3
PRODUCT COMPOSITION FROM TEXACO ENTRAINED BED GASIFIER

Component	Mole Fraction
H_2O	17.88
CO_2	8.71
H_2	28.84
CO	42.45
CH_4	0.08
N_2	0.66
Ar	0.12
H_2S	1.01
COS	0.06
NH_3	0.19

The coal feed is ground, weighed, and slurried with water prior to injection into the burner at the top of the gasifier as shown in Figure 5-3. The oxygen fed with the coal in the water cooled burner is obtained from a standard air separation plant. Air is compressed, cooled, and expanded to provide the liquid feed to an oxygen/nitrogen distillation column. The liquid products are heat exchanged with the incoming air and then are vented (N_2) or fed to the process (O_2). The oxygen feed is 98% O_2, 1.5% N_2, and 0.5% Ar.

The effluent gas is cooled in a waste heat recovery unit which generates steam at 1520, 440, and 50 psig. Additional cooling is accomplished by feed water and fuel gas preheat. The gas is then washed with water to remove the ammonia which is recycled to the reactor where it is eventually decomposed to N_2 and H_2.

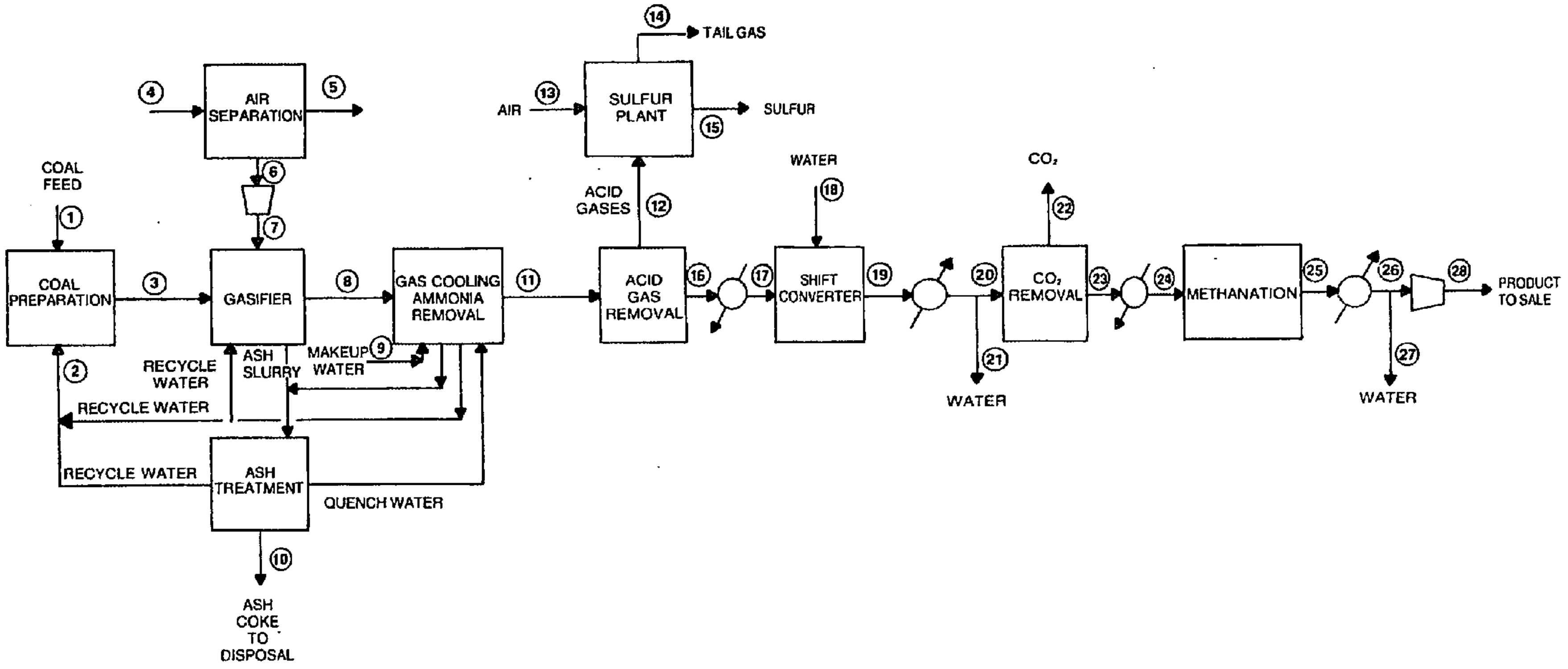

Figure 5-3. TEXACO OXYGEN BLOWN ENTRAINED BED GASIFIER

The process flow diagram varies downstream of the raw gas cleanup depending on the final use of the product gas and the operating pressure of the reactor. For use as an industrial fuel, the only cleanup operation is the removal of sulfur (as H_2S) from the gas stream. The fuel gas is compressed prior to combustion if it is to be used in combined cycle production of electricity. For other applications, both H_2S and CO_2 are removed from the raw gas and it is then fed to a shift converter (Fig. 5-3). In this converter, excess water is added to the gas and it reacts catalytically with the carbon monoxide present to produce hydrogen and carbon dioxide. The ratio of H_2 to CO in the final product is adjusted by the amount of steam added and the reaction temperature. The stoichiometry for substitute natural gas (SNG) production requires a 3:1 H_2 to CO ratio. In the process studied here, the H_2/CO ratio is assumed to be 3.1:1 in order to shift the equilibrium of the methanation reaction towards a low residual carbon oxide concentration after the methanation reaction that follows. The CO_2 generated in the shift reaction is removed by a hot carbonate absorption system. The H_2/CO gas mixture is then catalytically reacted in a methanation unit to yield methane and water. The water is condensed out and the high Btu Substitute Natural Gas (SNG) is compressed to pipeline pressures and transmitted to the user.

5.2.2.2 Material Balance

The material balance for the Texaco Entrained Bed Reactor System is shown in Table 5-4 with the stream numbers matching those given in Figure 5-3. This material balance is based on the reactor effluent composition from the literature(5-1) and the design assumptions given in Table 5-2. It should be emphasized that the process design developed here is preliminary in order to quickly estimate the process characteristics. More complete designs (using different design constraints) are available in the literature. A summary of the important process characteristics is given in Table 5-5.

5.2.2.3 Discussion

The design characteristics of the entrained bed gasifier present several problems when attempting to couple it to a nuclear thermal source. The high temperature at the outlet of the entrained bed gasifier is impossible to achieve in a system in which the thermal energy is supplied by a HTR. In addition, a nuclear heated entrained bed gasifier would require that heat be transferred across a heat transfer surface. At the inlet of the gasifier, the coal/water slurry would not

Table 5-4

Material Balance for Texaco Entrained Bed Gasifier Process

Flow Rates - Short Ton/Day

Stream No.	Temp. °F	Pressure psia	MAF Coal	H_2O	CO_2	H_2	CO	CH_4	O_2	N_2	Ar	Ash	H_2S	COS	NH_3	S
1	70	14.7	1724.0	84.0								192.0				
2	70	14.7		879.7											7.3	
3	140	625	1724.0	963.7								192.0				
4	70	14.7		120.6					1776.5	5773.9	106.1					
5	70	14.7		120.6					133.3	5751.7	95.7					
6	70	14.7							1643.2	22.2	10.4					
7	300	720							1643.2	22.2	10.4					
8	2450	614.7		696.3	828.6	125.7	2568.0	2.8		40.1	10.4		74.2	7.8	7.3	
9	70	14.7		317.0												
10	70	14.7		128.0								192.0				
11	100	547		5.6	828.6	125.7	2568.0	2.8		40.1	10.4		74.2	7.8		
12	120	22		0.7	812.0								74.2	7.8		
13	70	14.7							37.0	121.7						
14	200	14.7		40.0	817.7					121.7						
15	250	14.7														73.9
16	100	539		4.9	16.6	125.7	2568.0	2.8		40.1	10.4					

Table 5-4 (continued)

Stream No.	Temp. °F	Pressure psia	MAF Coal	H_2O	CO_2	H_2	CO	CH_4	O_2	N_2	Ar	Ash	H_2S	COS	NH_3	S
17	825	539			16.6	125.7	2568.0	2.8		40.1	10.4					
18	477	550		1493.2												
19	900	527		522.0	2401.2	234.9	1052.9	2.8		40.1	10.4					
20	100	527		6.8	2401.2	234.9	1052.9	2.8		40.1	10.4					
21	100	527		515.2												
22	120	22		1.6	2353.2											
23	100	519		5.2	48.0	234.9	1052.9	2.8		40.1	10.4					
24	752	519		5.2	48.0	234.9	1052.9	2.8		40.1	10.4					
25	791	499		674.5	57.2	9.4	0.5	602.0		40.1	10.4					
26	100	499		trace	57.2	9.4	0.5	602.0		40.1	10.4					
27	100	499		674.5												
28	70	1015		trace	57.2	9.4	0.5	602		40.1	10.4					

Table 5-5

Process Characteristics for the Texaco Entrained Bed Gasification System

Coal Consumption		2000 ST/D*
Total Btu in Coal Feed (HHV)		4.894×10^{10} Btu
Product Gas Composition (Vol. %)		
CH_4	83.07	
CO_2	2.88	
CO	0.04	
H_2	10.27	
N_2	3.27	
Ar	0.58	
Product Heating Value		872 Btu/SCF
Total Gas Flowrate		34.3×10^6 SCF/D
Total Btu in Product Gas		2.992×10^{10} Btu/D
CO_2 Emissions		3170.9 ST/D
SO_2 Emissions (Based on 1 lb $SO_2/10^6$ Btu (HHV) of coal feed)		24.47 ST/D
Water Consumption		317.0 ST/D
Energy Efficiency**		0.611

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*Short Tons/Day

**Btu in product gas/Btu in coal feed

efficiently contact the heat transfer surface and as a result the heat transfer surface would have to be extremely large and, hence, very expensive. To make matters worse, the hottest portion of the exchanger surface, near the gasifier outlet, would be in contact with an environment containing a significant amount of corrosive ash material. A combination of erosion and corrosion under these circumstances would make fabrication of the heat exchanger apparatus designed for a 30 year life required for nuclear applications virtually impossible.

Considering these factors, the application of a nuclear heat source for the supply of the thermal energy requirements for the entrained bed gasification system appears virtually impossible. Hence, the design investigation for the coupling of these processes will not be undertaken in this study.

5.2.3 Lurgi Moving Bed Gasifier(5-1,-3,-4)

5.2.3.1 Process Description for Conventional Oxygen Blown Lurgi Gasifier

The Lurgi gasifier is a high pressure (615 psig) moving bed reactor (also referred to as a fixed bed) shown schematically in Figure 5-4. The solid phase (coal) moves slowly downward by gravity flow, countercurrent to the upward flow of gas. As the coal descends, it passes through zones of increasing temperature in the bed. First, it is preheated, dried and devolatilized in a rapid process by the upflowing hot gases.

The devolatilization reactions are

$$\text{Coal} \xrightarrow{\text{heat}} \text{Coal Char} + CH_4 + H_2O + H_2 + CO + CO_2 + \text{tars} + \text{oils}$$

The amount of devolatilization is a function of the temperature, pressure, composition of the carrier gas, and the heating rate. A slight additional amount of methane may also be produced due to the high pressure methanation of CO. The rate of this reaction is extremely low and the reaction is only very slightly catalyzed by the impurities present in the coal, hence, the product gas is not in equilibrium with respect to the methanation reaction. The devolatilization gas products make up less than 10% of the total gas produced. Devolatilization, which occurs at temperatures ranging from 1150°F to 1600°F, is followed by gasification of the resulting char. Carbon monoxide and hydrogen are formed by carbon/steam reduction in this oxygen-free zone.

The reactions are:

1) $C + H_2O = CO + H_2$	endothermic
2) $C + CO_2 = 2CO$	endothermic
3) $C + 2H_2 = CH_4$	exothermic

The heat of reaction of the endothermic reactions is supplied by combustion in the bottom zone of the bed. Carbon reacts with oxygen producing CO/CO_2 at a

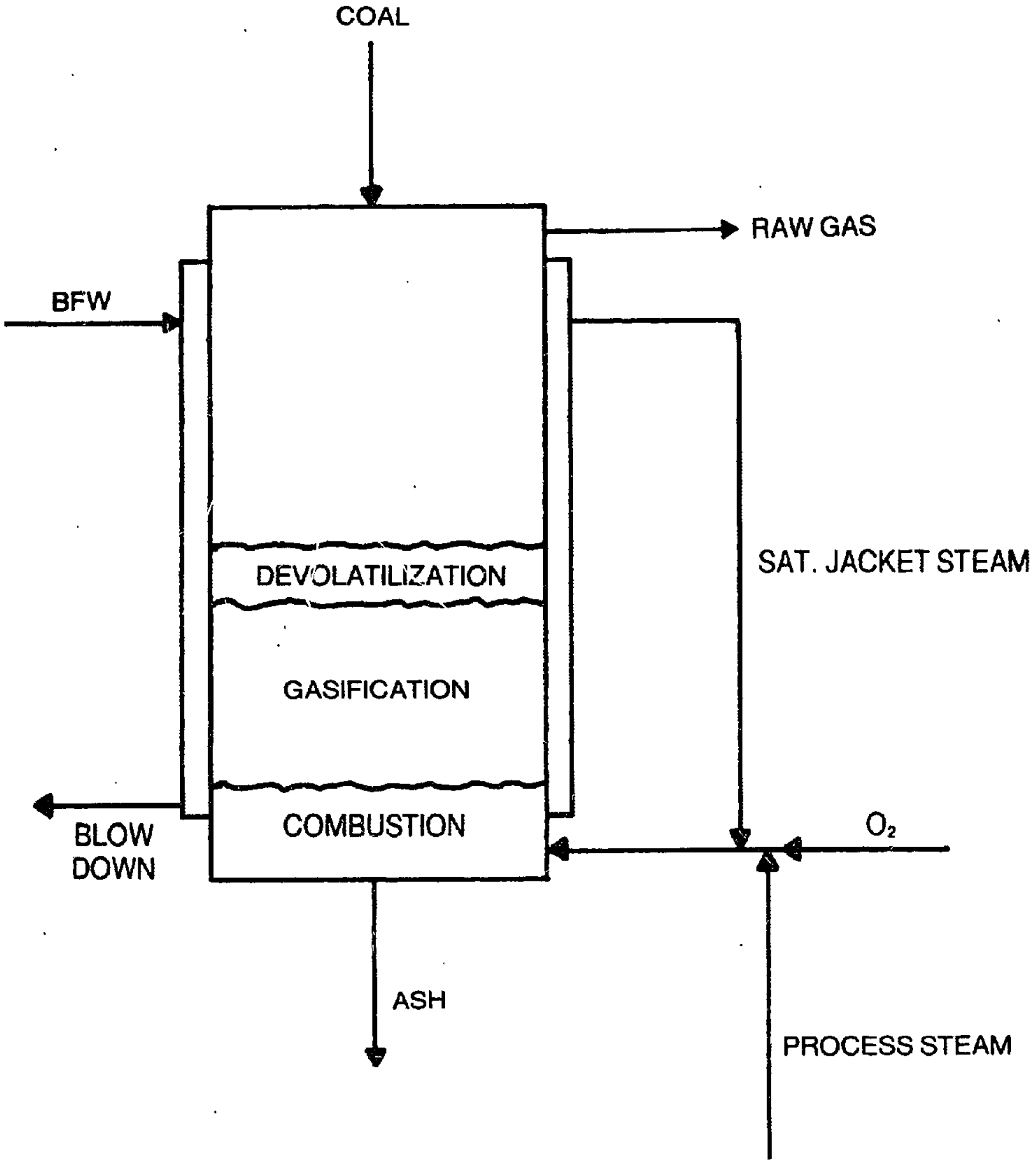

Figure 5-4. LURGI OXYGEN-BLOWN MOVING BED GASIFIER

maximum temperature (2200°F) in the combustion zone. The maximum temperature must be great enough to ensure complete gasification. The location of the maximum temperature is also important. A temperature profile across the reactor bed is shown schematically in Figure 5-5. The peak temperatures indicates the location of the combustion zone. The resultant ash from the combustion zone falls through the revolving grate at the bottom of the reactor, through an accumulator, and is sluiced to a disposal basin. The rate of ash disposal is adjusted to the gasifier load. The raw gas leaves the gasifier at approximately 1000°F, 615 psia. The raw gas then flows to the spray cooler. Its composition is:

	mole %
H_2O	55.5
H_2	18.0
CO	7.8
CO_2	12.4
CH_4	3.4
O_2	0
N_2	0.1
H_2S	0.3
NH_3	0.1
Ar	0.0
tar and oil	2.3

Prior to being fed to the gasifier, the coal is screened, crushed, and stored in piles or silos. The oxygen fed to the bottom of the gasifier is obtained from a standard air separation plant. The nitrogen byproduct from the air separation is vented.

The raw gas treatment portion of the process consists of tar and oil separation, acid gas removal, shift conversion followed by further CO_2 removal and finally methanation as shown in Figure 5-6. From the gasifier, the raw gas flows to the spray cooler where heavy tars are condensed out of the gas stream at 390°F. The gas is cooled to 100°F in a waste heat recovery system where remaining tars and oils are condensed. The tars and oils in the bottoms from the spray cooler and

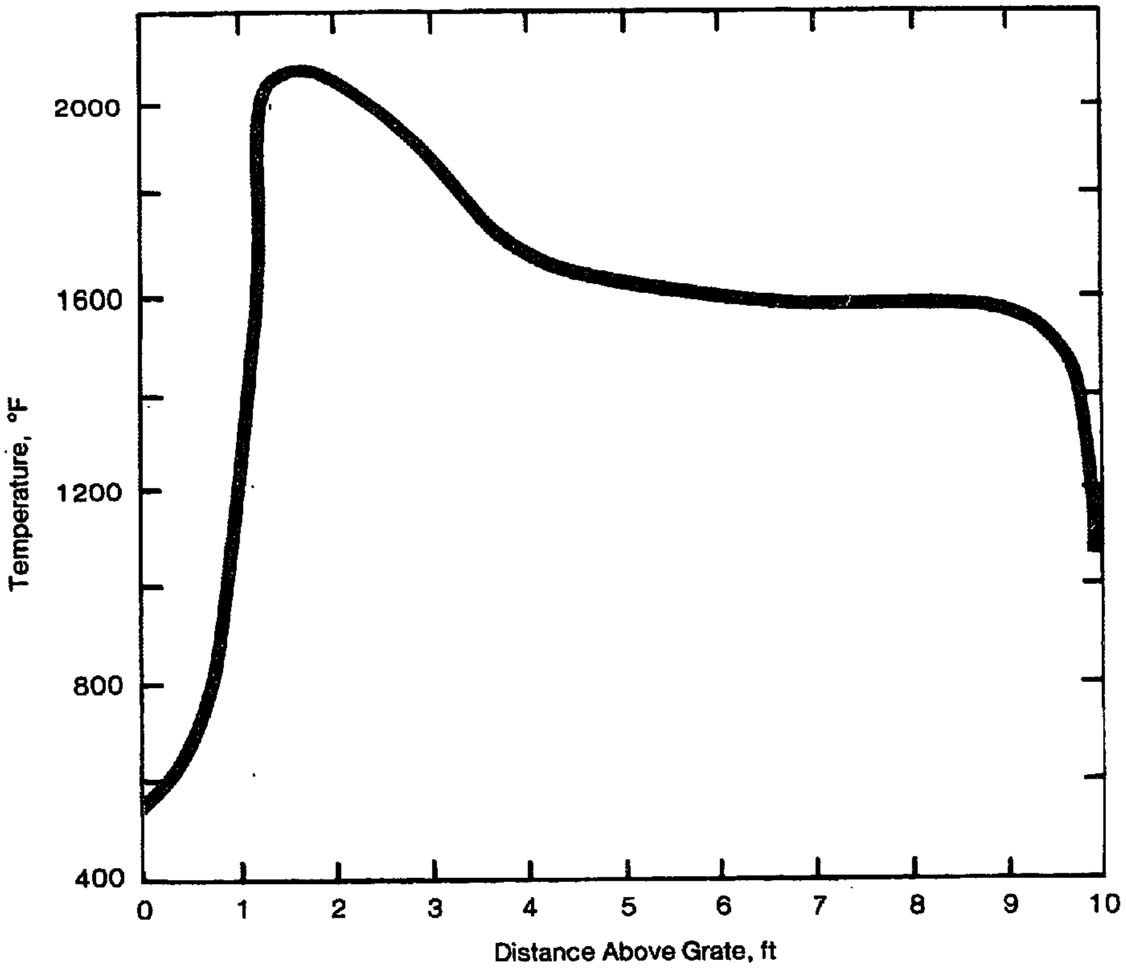

Figure 5-5. TEMPERATURE PROFILE IN LURGI OXYGEN BLOWN GASIFIER

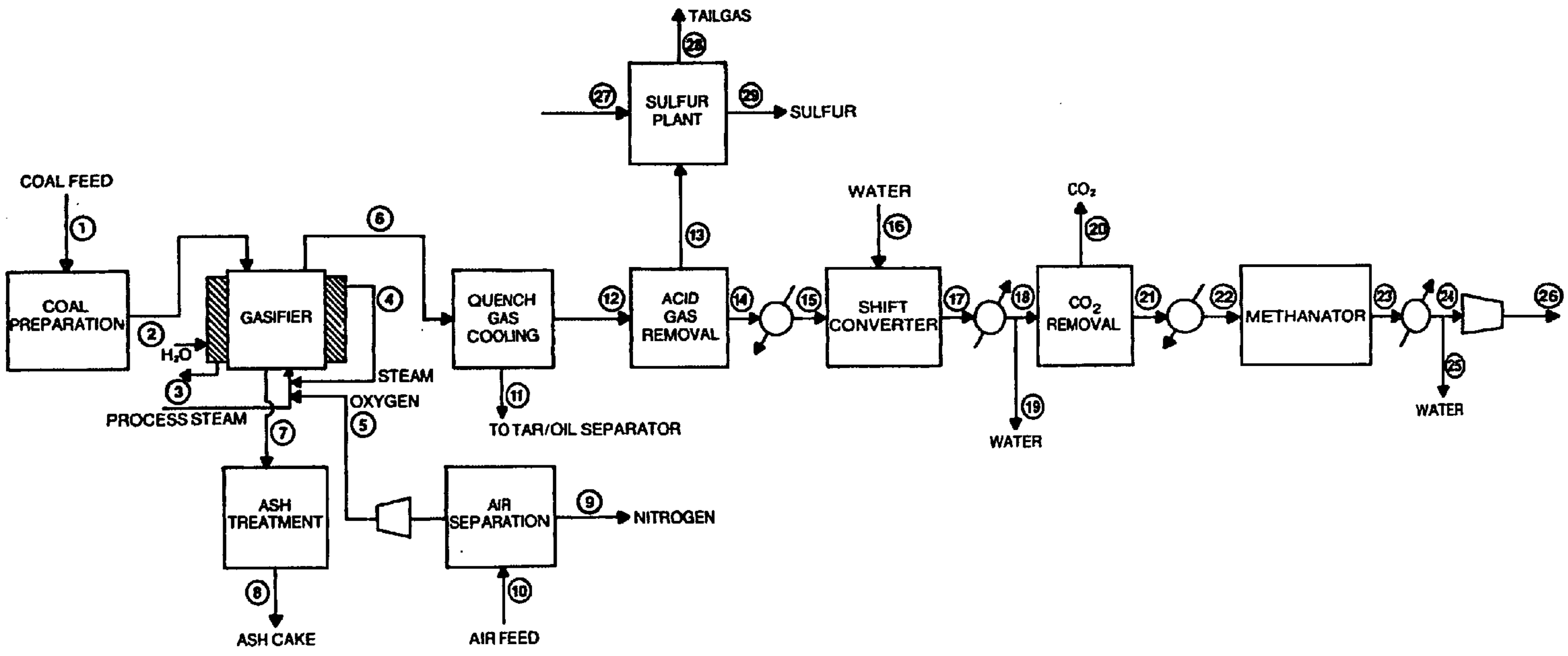

Figure 5-6. LURGI OXYGEN-BLOWN MOVING BED GASIFIER

heat recovery systems are separated from the water phase in a tar-oil separation unit. The tar-oil liquor is recycled to the spray cooler as the coolant.

The gas stream from the quench system is sent to the potassium carbonate absorption-regeneration unit. The hydrogen sulfide and 98% of the carbon dioxide are absorbed via the following reactions:

$$CO_2 + H_2O + K_2CO_3 \rightleftharpoons 2KHCO_3$$

$$H_2S + K_2CO_3 \rightleftharpoons KHS + KHCO_3$$

The removal of sulfur compounds from the gas prevents poisoning of the nickel catalyst in the shift conversion and methanation units that follow. The acid gases are recovered by flashing the rich solution from the bottom of the absorption unit to low pressure and then heating it to drive off the acid gas remaining in the solution. The lean solution from the bottom of the stripping tower is cooled and pumped to the pressure of the adsorption tower. The acid gas stream is sent to a Sulfur Recovery Unit (Claus system). All the H_2S is assumed to be converted to sulfur for the material balances presented here.

After acid gas removal, the H_2/CO ratio is adjusted to 3.1:1 by the water gas shift conversion reaction:

$$CO + H_2O \xrightarrow{\text{Ni catalyst}} CO_2 + H_2$$

The water gas shift reactor feed is preheated to 825°F at a system pressure of 594 psia. The outlet composition of the gas is determined by the equilibrium composition of the reaction at the outlet temperature, 900°F. The temperature increases because the reaction is slightly exothermic (-9.8 Kcal). The exit gas is cooled to 100°F and processed through another potassium carbonate unit to remove 98 wt % of the remaining CO_2.

Finally the gas is preheated to 751°F and methanated. The exit temperature is 791°F. The product gas is cooled, dried and pressurized to pipeline pressure (1015 psia). The resulting composition is:

mole%	
H_2O	trace
CO_2	1.2
CO	0.02
H_2	9.5
CH_4	88.0
inerts	1.29

5.2.3.2 Material Balance

The material balance for the oxygen-blown Lurgi Gasifier is presented in Table 5-6. The stream numbers correspond to the stream numbers in Figure 5-6. The raw gas composition, temperature, heating value and flow rate from the gasifier were calculated theoretically. The gasification calculations are based on equilibrium shift reaction, and the devolatilization calculations are based on distillation data found in the literature(5-5). The output parameters are a function of the steam and oxygen feed rates, the fixed carbon to oxygen ratio, the blast temperature, and the gasifier heat loss. The values of the feed rates, blast temperature and gasifier heat loss were proportionately scaled up to 2000 ST/D coal feed from Westfield Illinois #6 coal gasifications experiment(5-4). The rest of the material balances are consistent with the assumptions stated in Table 5-2.

Important process characteristics for the Lurgi gasifier are shown in Table 5-7.

5.2.3.3 Process Description for Lurgi Nuclear Moving Bed Gasifier

The heat required for coal gasification in the Lurgi Nuclear Moving Bed Gasifier is provided by nuclear heat instead of by coal combustion and, thus, combustion coal is made available for gasification. The oxygen supported combustion in a conventional Lurgi gasifier is replaced by a hypothetical nuclear heat exchanger. The structural challenges of such a heat exchanger are not addressed in this report; however, many of the problems associated with the tranfer of high temperature heat across an exchanger surface, as mentioned in the section on the entrained bed gasifier, would be present in a Lurgi type design. In order to evaluate the incentives for developing the materials that would allow heat

Table 5-6

Lurgi Oxygen Blown Moving Bed Gasifier: Material Balance

Flow Rates - Short Ton/Day

Stream No.	Temp. °F	Pressure psia	MAF Coal	H_2O	CO_2	H_2	CO	CH_4	O_2	N_2	Ar	Ash	H_2S	NH_3	T&O	S
1	70	14.7	1724	84								192				
2	70	14.7	1724	84								192				
3	660	376.7		4480.7												
4	490	376.7		1203.3												
5		615							1027	13.7	6.5					
6	998	615		4527	2473	163	991	248		12.35	6.5		45	7.5	136.8	
7												192				
8	70	14.7										192				
9	70	14.7		75.26					83.3	0.05	0.68					
10	70	14.7		75.26					1110.4	3554.6	58.88					
11				4521.6										7.5	136.8	
12	100	602		5.4	2473	163	991	248		12.32	6.5		45			
13	120	22		1.54	2423.5								45			
14	100	594		3.86	49.5	163	991	248		12.32	6.5					
15	825	594		3.86	49.5	163	991	248		12.35	6.5					
16				196.7												
17	900	582		130.6	352.4	177	798	248		12.35	6.5					

Table 5-6 (continued)

18	100	582		4.15	352.4	177	798	248		12.35	6.5				
19	100	582		126.4											
20	120	574		0.17	345.3										
21	100	574		3.97	7.10	177	798	248		12.35	6.5				
22	752	574		3.97	7.10	177	798	248		12.32	6.5				
23	792	554		502	25	9	0.3	698		12.35	6.5				
24	100	554		1.53	25	9	0.3	698		12.35	6.5				
25	100	554		500.5											
26	100	1015		trace	25	9	0.3	698		12.35	6.5				
27	70	14.7							21.4	74.5					
28	200	14.7		25.4	2424										
29	250	14.7													42.6

Table 5-7

PROCESS CHARACTERISTICS FOR THE LURGI OXYGEN BLOWN GASIFICATION SYSTEM

Coal Consumption		2000 ST/D
Total Btu in Coal Feed (HHV)		4.894×10^{10} Btu/day
Product Gas Composition (Vol %)		
CH_4	88.08	
CO_2	1.2	
CO	0.02	
H_2	9.5	
N_2 Ar	1.29	
Product Heating Value		919.6 Btu/SCF
Total Gas Flow Rate		3.756×10^{7} SCF/day
Total Btu in Product Gas		3.454×10^{10} Btu/day
CO_2 Emissions		2769.3 ST/D
SO_2 Emissions (Based on 1 lb $SO_2/10^6$ Btu (HHV) of coal feed)		24.47 ST/D
Water Consumption		791 ST/D
*Energy Efficiency		0.71

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*Btu in product gas/Btu in coal feed

to be transferred directly into a steam gasifier, a detailed analysis is made here to evaluate the concept. It is assumed for the sake of comparison, that a black box heat exchanger provides the required amount of heat.

Nuclear heat eliminates the need for an air separation plant and produces less CO_2 and H_2S. In this section, a hypothetical Lurgi Nuclear Moving Bed Gasifier system is described in order to provide a basis to determine the coal savings, efficiency improvements, and general process changes when compared to the conventional Lurgi gasifier. Process changes were made where required as shown in Figure 5-7. The comparison is based on a by-unit material balance of each system. The material balance closes to within 5-10%. The discrepancy is due to the estimation of devolatilization products in the gasifier. The overall process steam and power usage (Table 5.9) was estimated by direct scale down of a similar plant design of a conventional process by the U.S. Department of Interior, Bureau of Mines$^{(5-3)}$.

The pressure assumed for the gasifier is 615 psia. This pressure is approximately equal to the helium coolant from the nuclear heat source (40 atm) to prevent creep failure in the heat exchanger interface. The steam feed temperature is set at 1520°F (the approximate temperature of the gasifier-side of the nuclear heat exchanger). The steam rate is adjusted to produce a H_2/CO ratio of approximately 3.1 in the raw gas and is 95% of the steam rate in the oxygen blown gasifier. This eliminates the need for a shift converter and the second CO_2 absorber downstream of the gasifier and allows the methanation unit to run at about twenty pounds pressure higher than in the conventional design. If an equivalent amount of steam were used in the nuclear gasifier, the H_2/CO ratio would be 3.3. As expected, the CO_2 concentration level decreases when combustion heat is replaced by nuclear process heat. Because the shift reaction is assumed in equilibrium, a decrease in the CO_2 level causes a simultaneous H_2 concentration increase. The raw gas is quenched and cooled as in the O_2 case. It is assumed that all of the tar, oil and NH_3 is removed. Then the product stream passes through the acid gas removal unit where 100% H_2S and 98% CO_2 removal occurs. A Claus Unit converts the H_2S to elemental sulfur.

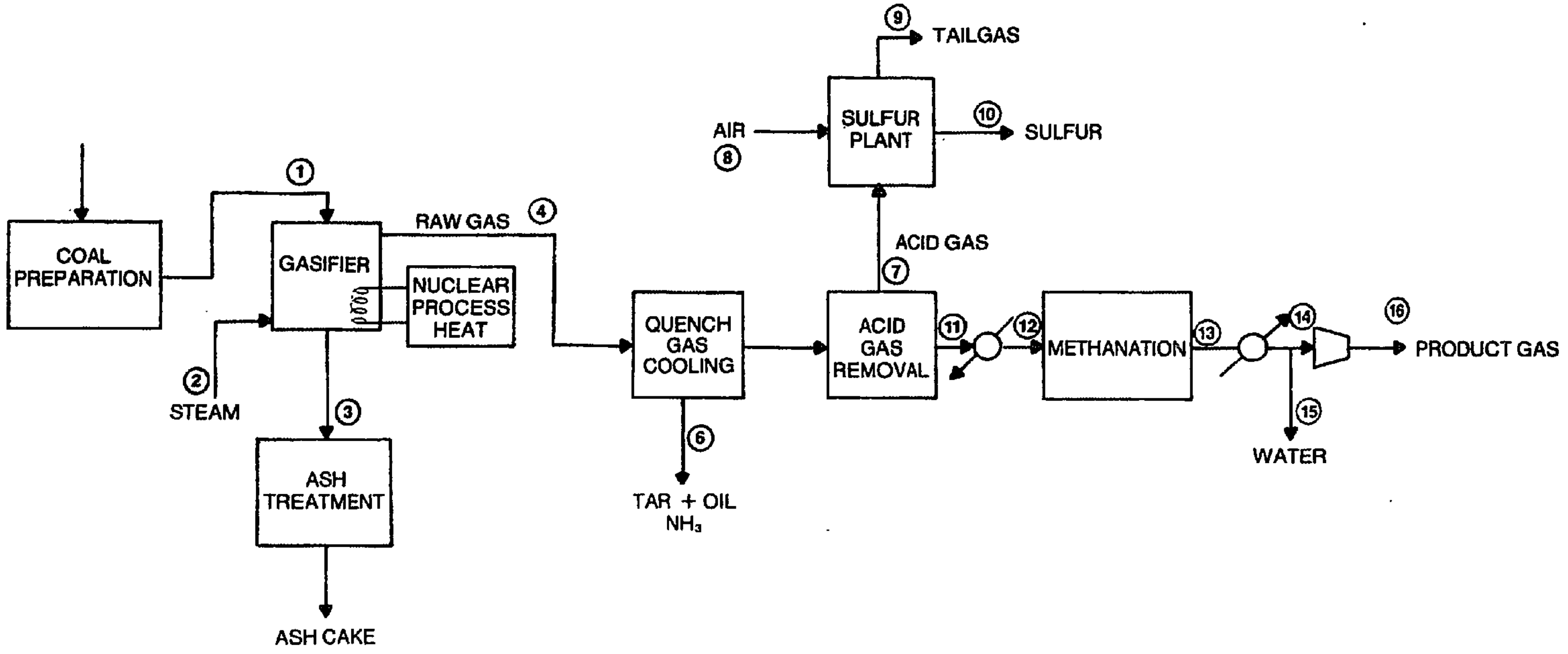

Figure 5-7. LURGI NUCLEAR MOVING BED GASIFIER

5.2.3.4 Material Balance

The material balance for the Lurgi nuclear gasification system is given in Table 5-8. It is assumed for this analysis that all of the H_2S is converted to sulfur. In reality, up to one pound SO_2 per million BTU coal feed may be emitted to the atmosphere as in the conventional design.

5.2.3.5 Discussion

Based on equal coal feed rates, the nuclear gasifier requires 5% less steam to produce a raw gas H_2/CO ratio of 3.16 as compared to the conventional Lurgi design. The raw gas would be rich in H_2 and CO and lean in O_2, N_2, Ar, H_2O and CO_2 as compared to the conventional process. The combustion heats of the nuclear and oxygen based raw gas are 6.17×10^{10} Btu/day and 4.51×10^{10} Btu/day respectively. The nuclear Lurgi gasifier process characteristics are presented in Table 5-9.

The coal feed to one 2000 ST/D oxygen-blown gasifier would have to be increased by 738 ST/D (37%) to produce a heat content equivalent to that of a nuclear gasifier.

On an equivalent Btu/hr basis, it is estimated that the nuclear gasification system would require 20% less total process steam than the conventional gasifier. This includes steam for all process units, the oxygen plant (in the O_2 case) and the power plant. The total coal usage is 44% lower for the nuclear case, including both the coal needed to produce steam and used as feed to the gasifier. Approximately 24% of the difference is a result of nuclear power generated steam and 20% is a result of nuclear supplied heat in the gasifier. Overall approximations for electric power, cooling water and raw water usage for the two processes are compared in Table 5-10. In all three categories, the nuclear process has a lower usage level due to the elimination of the oxygen plant, shift converter and CO_2 absorber. Because the oxygen blown case has a 44% higher coal usage rate, the SO_2 and CO_2 emissions are also 44% higher. A comparison of these emission rates is also presented in Table 5-10 for the two processes.

It must be recognized that the analysis presented here assumes that the nuclear heat can be transferred directly into the gasifier vessel through a heat exchanger surface of some undetermined material. It cannot be emphasized too

Table 5-8

Lurgi Nuclear Moving Bed Gasifier

Flow Rates - Short Ton/Day

Stream No.	Temp. °F	Pressure psia	MAF Coal	H_2O	CO_2	H_2	CO	CH_4	O_2	N_2	Ar	Ash	H_2S	NH_3	T&O	S
1	70	14.7	1724	84								192				
2	1520	615		4263												
3												192				
4	1113	615		1795	2120	275	1218	246					44	7	137	
5	100	602		3	2120	275	1218	246					44			
6	100	602		1792										7	137	
7	120	22			2070								44			
8	70	14.7							21	73						
9	200	14.7		23	2077											
10	250	14.7														13
11	100	594		5.7	44	275	1218	246								
12	752	594		5.7	44	275	1218	246								
13	792	574		790	41	13	0.4	942								
14	100	574		2	41	13	0.4	942								
15	100	574		788												
16	100	1015		trace	41	13	0.4	942								

Table 5-9
PROCESS CHARACTERISTICS FOR THE LURGI NUCLEAR GASIFICATION SYSTEM

Coal Consumption		2000 ST/D
Total Btu in Coal Feed (HHV)		4.89×10^{10} Btu/day
Product Gas Composition (vol %)		
CH_4	88.64	
CO_2	1.4	
CO	0.02	
H_2	9.94	
N_2 Ar	0	
Product Heating Value		926.1 Btu/SCF
Total Gas Flow Rate		5.04×10^{7} Btu/day
Total Btu in Product Gas		4.67×10^{10} Btu/day
CO_2 Emissions		2077 ST/D
SO_2 Emissions (Based on 11B $SO_2/10^6$ Btu (HHV) Coal Feed)		24.47 ST/D
Water Consumption		1475.8 ST/D
Energy Efficiency		0.954

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*Btu in product gas/Btu in coal feed.

strongly that the corrosion problems associated with the contact of the alkali metal compounds in the coal ash with this heat exchange surface at the elevated temperatures required for gasification are currently infeasible. No metallic material known today can withstand this type of environment for the 30-year lifetime required for licensing. In addition to the chemical environment, the moving bed type gasifier has rotating internals that expedite the flow of coal down through the bed and the flow of ash out the bottom of the gasifier. The mechanical clearances required by such devices would make the design of the heat exchangers even more difficult. Assuming that these problems could be overcome, the exchanger would have to be designed such that there was intimate contact between the surface and all portions of the bed. A design configuration to achieve this intimate contact that would not impede the flow of coal down through the bed is difficult to imagine.

Table 5-10
COMPARISON OF OXYGEN BLOWN AND NUCLEAR HEATED LURGI GASIFICATION SYSTEMS

Basis: Equivalent Product Heating Value

4.7×10^{10} Btu/day

	Oxygen Blown	Nuclear Heated
Coal Consumption in Gasifier, ST/D	2738	2000
Electric Power*, MW	2.0	1.5
Cooling Water, GPM	48,600	28,600
Raw Water, GPM	3,300	2,400
SO_2 Emissions, ST/D	46	25
CO_2 Emissions, ST/D	3,900	2,200

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*Assumes steam drive in oxygen plant.

In general then, it appears as if the moving bed gasifier presents insurmountable problems for coupling to an external heat source such as an HTR.

5.2.4 The Exxon Fluidized Bed Catalytic Gasification Process

5.2.4.1 Process Description for Conventional Process

The Exxon Catalytic Gasification process is a developmental process which differs from competing coal gasification processes in that it uses a K_2CO_3 catalyst to promote low temperature gasification (down to 700°C (1300°F)). As shown in Figure 5-8, it employs a fluidized bed gasification system that operates in a well mixed mode approaching isothermality, the fluidizing gas being steam and recycle hydrogen and carbon monoxide. The effect of the K_2CO_3 is to catalyze the methanation reaction such that at the reactor outlet the methanation reaction and the shift reaction are in thermodynamic equilibrium given optimum operating conditions(5-6). These reactions are:

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

$$CO + H_2O \quad CO_2 + H_2$$

The overall chemistry of the reaction is approximately(5-7)

$$\text{Coal} + H_2O = CH_4 + CO_2$$

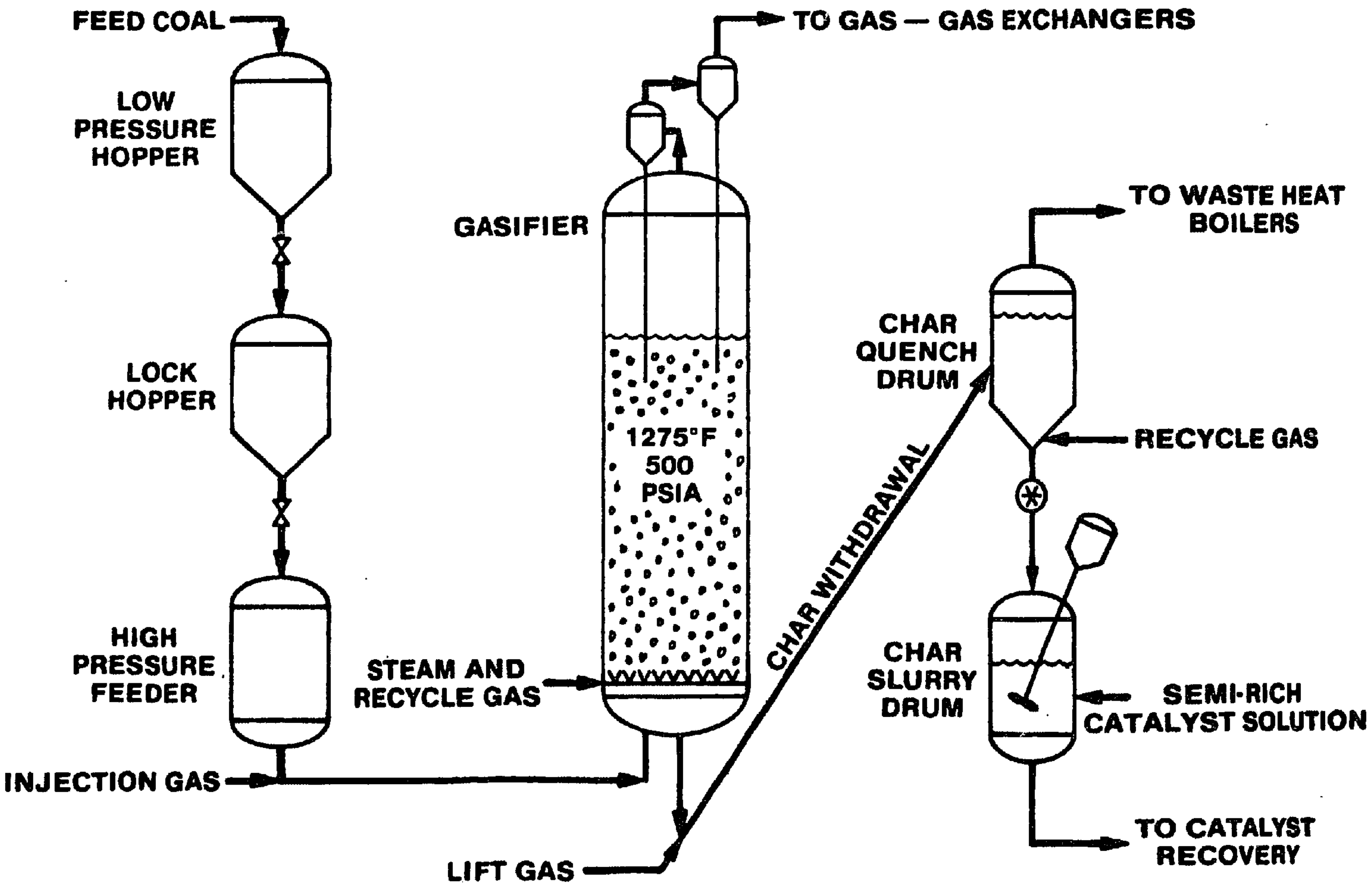

Figure 5-8. EXXON CATALYSED FLUID BED GASIFIER (5-7)

The equilibrium constants for the shift and methanation reactions and the material balance around the gasifier can be used to estimate the product distribution from the gasifier.

Because of the catalytic action of the K_2CO_3 and the lower gasification temperature, a much higher percentage of methane is formed in the gasifier (approximately 15% CH_4 by volume) than in steam gasification processes. In effect, the methanation reaction that is carried out downstream of the gasifier in steam gasification processes is carried out in the catalytic fluidized bed gasifier along with the steam gasification reaction. The heat released from the methanation reaction is used to supply the heat required for the endothermic steam gasification reaction. As shown schematically in Figure 5-9, the effluent from the gasifier is first cooled and the excess steam is condensed. Then the CO_2 is removed from the raw gas by a conventional acid gas removal process. A cryogenic distillation is carried out to separate the product methane (SNG) from the hydrogen and carbon monoxide which are recycled to the gasifier.

Thus, the process:

(1) Produces methane directly as its product

(2) Requires no reaction heat be supplied directly to the gasification vessel.

(3) Requires preheat, catalyst recycle drying, recycle H.P., and separation work.

Experimental work has been limited to a 25 lb/hr pilot plant operating at between 100 and 500 psi(5-8,-9). Exxon is currently funded through FY 1983 for this pilot plant work and estimates that they could convert the current "Synthane" plant to a 93 ton/day catalytic system for $150M(5-9). The consequences of this early stage of development are twofold:

(1) No realistic flowsheet for the process now exists giving power heat and material flows.

(2) A vigorous incentive could include the aims of the current HTR project into the development program.

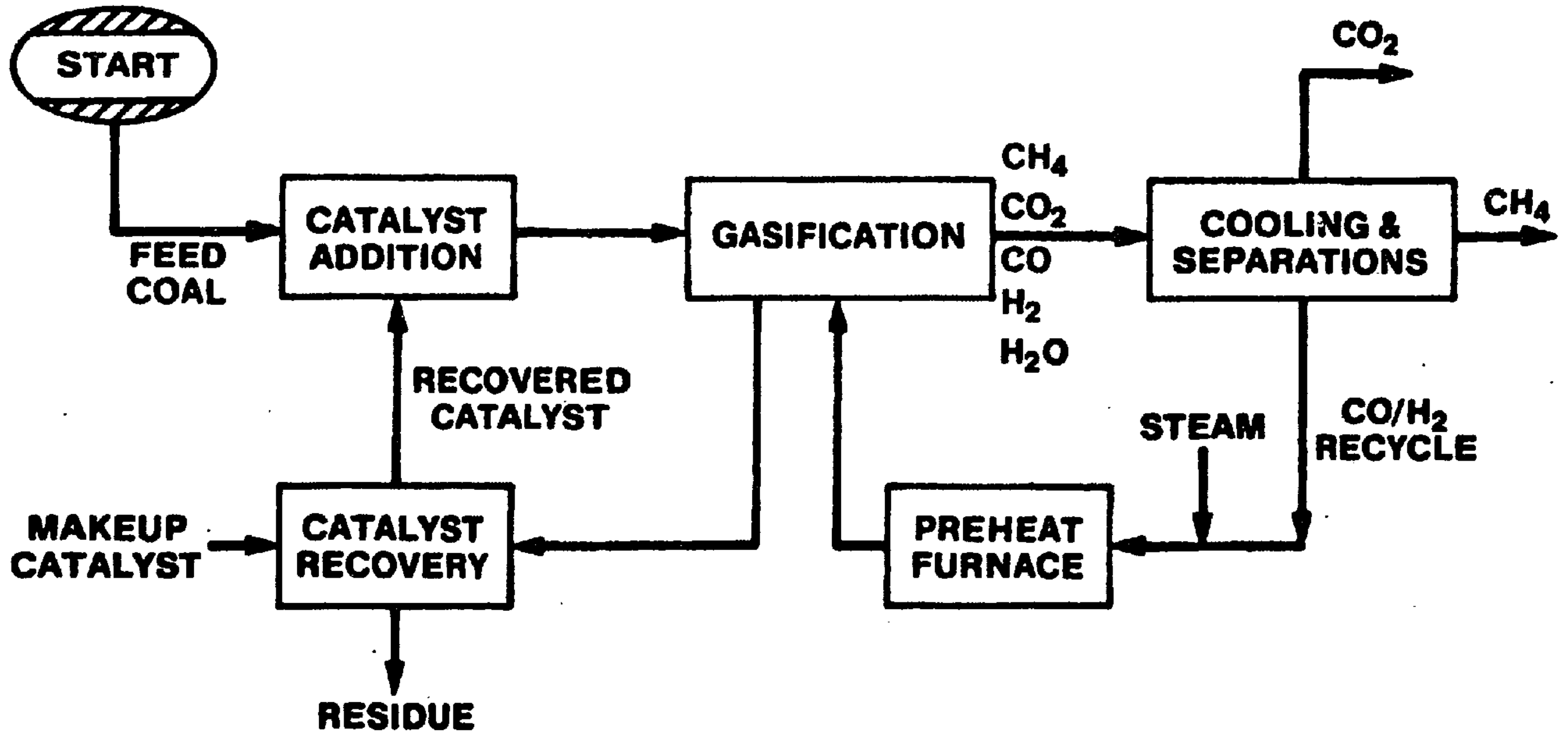

Figure 5-9..EXXON CATALYTIC GASIFICATION PROCESS (5-7)

5.2.4.2 Conventional Catalyzed Fluid Bed Gasifier Material Balance

Exxon(5-7) quotes actual data for a fluidized catalytic bed coal gasifier operating at 1330°F and 100 psig. An approximate mole balance for 2000 ST/D feed is given in Table 5-11 below with the stream numbers agreeing with those shown in Figure 5-10. SNG production is approximately 3.5 ton moles/hr (6.5×10^{10} Btu/day HHV) for the consumption of 2000 ST/D of Illinois #6 coal. Two cautions are necessary: 1) it is emphasized that the SNG production is approximate for this case and should not be directly compared to the other carefully computerized cases and 2) at the conditions chosen (i.e., near experimental test conditions) the preheat requirement for this case is 218 MW compared with 48 MW in a comparable nuclear case. Some fraction of the 218 MW could be recirculated to the process; just how much is to be determined by thorough process design. The CO_2 effluent is approximately 1.5 ton moles/hr (66 tons/hr) and the S release is 8.6 tons/day into Claus plant (elemental S removal process).

Table 5-11
APPROXIMATE PRODUCT STREAMS IN ST MOLES/HR
FOR 2000 STD ILLINOIS #6 COAL FEED TO CONVENTIONAL
CATALYZED FLUID BED COAL GASIFIER

Stream	1	2	3	4	5	6
CH_4	-	-	-	3.5	-	3.5
H_2O	9	-	9	7	7	-
CO	-	2.5	2.5	2.5	-	-
CO_2	-	-	-	1.5	1.5	-
H_2	-	15	15	15	-	-

Gasifier @ 720°C and 100 psig (9)

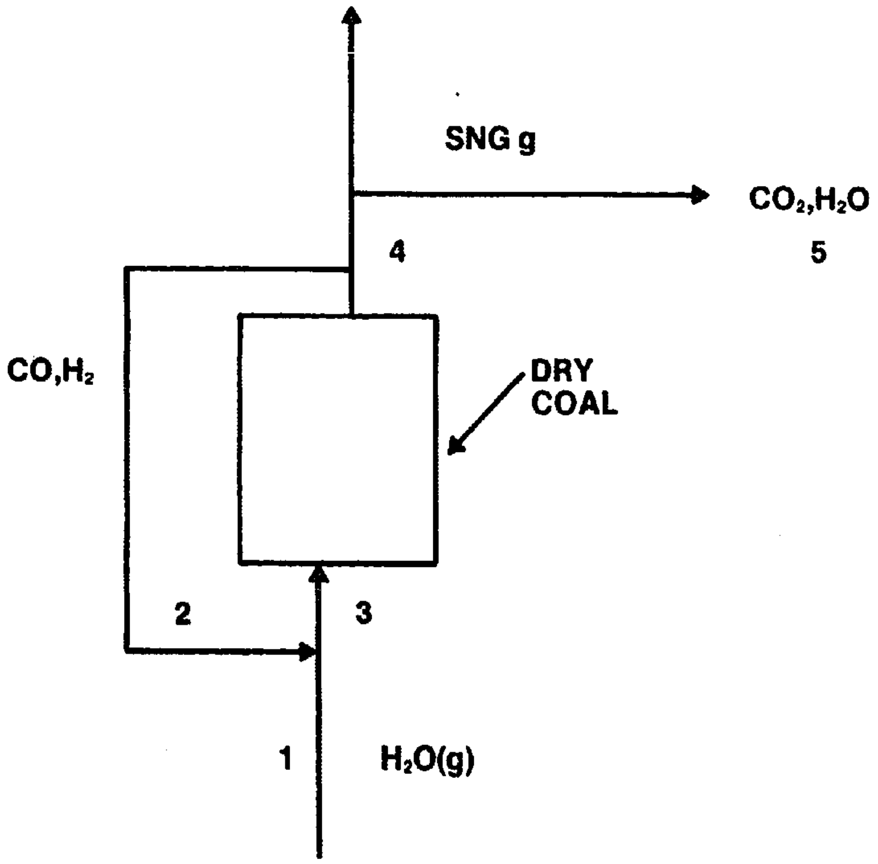

Figure 5-10. CONVENTIONAL CATALYTIC FLUIDIZED BED COAL GASIFIER

The conventional plant would require virtually no energy input to the gasifier but would require energy for coal drying (4 wt.% moisture is assumed here), catalyst recovery from solution, recirculation HP, and work of separation of approximately 30 MW_e for the methane cryogenic recovery.

5.2.4.3 HTR Coupled Catalyzed Fluid Bed Coal Gasifier

The Exxon process operates within the temperature range of a Pebble Bed Reactor (PBR) or a Prismatic HTR. The disadvantages of the existing process are twofold with respect to nuclear process heat (NPH) applications because of the increased reaction rate of the methane formation reaction due to the presence of the catalyst.

(1) Only preheat is needed for the reaction (at approximately 100°C-150°C above reaction temperature in the Exxon design).

(2) The product is SNG and not syn gas (a mixture of hydrogen and carbon monoxide).

A modified Exxon process would use the HTR to reform the methane to syn gas and thus absorb additional NPH at temperatures to 950°C (825°C in the reformer) yielding a syn gas product instead of SNG. The syn gas product would contain approximately 18% more energy per unit of carbon consumed than a SNG product. By using the nuclear heated reformer, this additional energy could be derived from nuclear source rather than from additional coal consumption. The process is shown in Figure 5-11 with the various recycles needed.

Methane obtained from the gasifier and from incomplete conversion in the reformer and water are pre-mixed, evaporated and preheated prior to being fed to the nuclear heated reformer. After the catalytic reaction in the reformer, the exiting gases containing CH_4, CO, CO_2, H_2 and $H_2O(g)$ are cooled, the excess steam is condensed and the CO_2 removed by absorption. The reforming process requires a large excess of steam. Therefore a large quantity of heat is available from the cooler condensers. Typically for each MW transferred into the reformer, another MW is used in the preheat, evaporation, cooling and condensation steps. Clearly, these steps need to be properly matched for both thermal open (MW) duty and temperature levels. The rich liquor from the absorption step for the CO_2 must be regenerated which requires substantial low temperature (less than 250°C) steam.

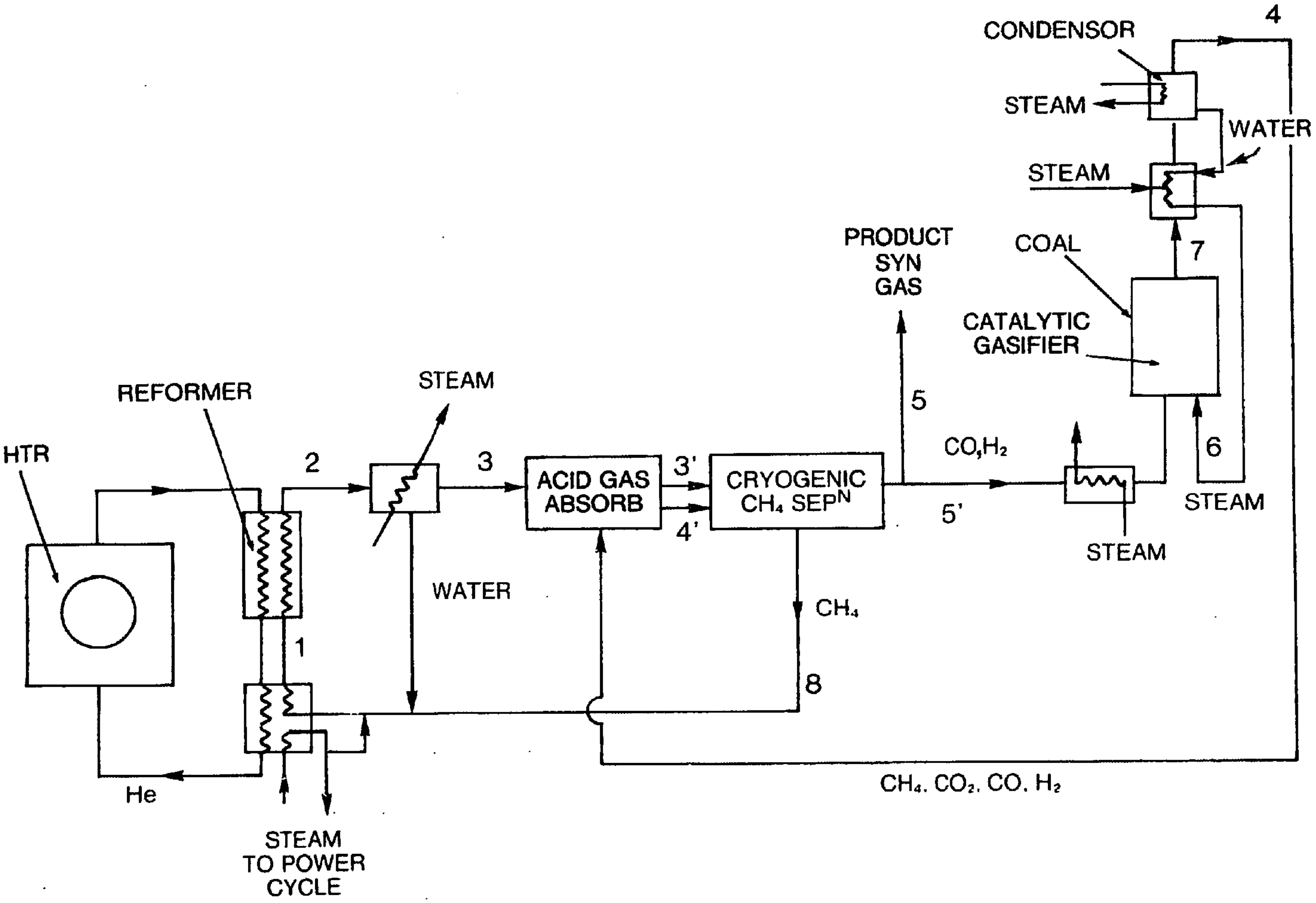

Figure 5-11. HTR/CATALYTIC GASIFIER SYMBOSIS FOR SYNTHESIS GAS PRODUCTION

After the CO_2 is removed the gases are chilled to -150°C to remove the methane. This step is also energy consumptive. A 20% separation efficiency is typical for this type of separation. The effluent gas of CO and H_2 is mixed with the water feed and then preheated to reaction temperature (approximately 700°C or greater) and passed into the gasifier. The steam flow must be carefully monitored to assure that direct hydrogenation and steam gasification occur to the extent to make the overall gasifier heat balance athermal, or slightly exothermic (This avoids adding heat to the coal/K_2CO_3 catalyst reaction at $T = 700^{\circ}C$, a step resulting in severe corrosion problems(5-10). The products from the gasifier contain CH_4, CO, CO_2, H_2 and $H_2O(g)$.

The stoichiometry of the modified (nuclear) catalyzed fluid bed gasifier process analyzed here differs from that used in the Exxon experiments in two ways: 1) the process explored here is carefully designed to be athermal (the ratio of H_2O to H_2 in the feed is not necessarily typical of those experiments) and 2) the computer model has been forced to ensure that the gasifier consumes all of the coal it possibly can and produces the highest possible CH_4 fraction. This is mathematically assured by continuing the calculation until approximately 1% of the solid carbon from the coal is lost in the ash. In this situation, the utilization of feed steam is essentially 100% so that the effluent gasifier gases are low in steam (and in hydrogen via the assumed equilibria). A highly desirable product gas consisting of approximately 65% methane and approximately 35% carbon monoxide would be formed under these conditions. Some equivocation is necessary since kinetic arguments, not addressed here, would indicate a very slow reaction corresponding to very low concentrations of the reacting species. In other words, at the stoichiometry chosen here (i.e., constrained by athermality and maximum carbon consumption), the reactor may not achieve the assumed exit equilibria or alternatively a very large gasifier would need to be built to achieve equilibrium. These considerations will be deferred and equilibrium effluent conditions will be assumed.

The effluent gas is cooled and its condensate is fed with additional steam to provide steam for the gasifier. The steam and condenser requirements should be heat duty and temperature matched. CH_4, CO_2, CO and H_2 are fed to the CO_2 absorber. It would also be necessary to remove the coal sulfur as H_2S and COS at this time. For each 1% by wt. of S in each 1000 ST/D of coal consumed the removal of H_2S amounts to 0.44 ST moles/hr resulting in 10 ST/D of elemental S at the final Claus plant. The scrubbed effluent from the acid gas removal process is then fed to the cryogenic separation unit. The methane produced is fed to the reformer and the hydrogen and carbon monoxide are recycled to the gasifier. Not shown is the catalyst recycle which consumes heat to evaporate water from the extract/solution for the K_2CO_3 recovery. The energy analysis has not yet traced the various matches of steam and condensate through the loops.

5.2.4.4 Material Balance For Nuclear Catalytic Gasification System:

The mass balances are given in Table 5-12 below, and, for example, the calculation does show that 2000 ST/D of coal will require 215.6 MW_{th} of nuclear heat at 950°C (825°C in the process reformer) and will produce 6.254×10^{10} Btus/day (HHV) of combustible energy. The separation work for the cryogenic methane separator is about 32 MW_e. In this case, the gasifier has been assumed to operate at 700°C, the specified minimum in the quoted publications(5-6,-7). To determine if there is potential advantage of raising this temperature (at the cost of HTR heat) a further set of data were run resulting in Table 5-13. Here the processing of 2000 ST/D of coal results in the consumption of 220.7 MW of HTR heat, produces 6.266×10^{10} Btu/day but requires an additional 4 MW_e of separator work (36 MW_e total). Finally in Table 5-14, the case of a lower temperature HTR is explored using 700°C as the peak reformer temperature and 700°C in the gasifier. For 2000 ST/D coal consumption, the nuclear heat conversions of 181 MW lead to 40 MW_e for separator work and a product of 6.260×10^{10} Btu in combustion heat. In addition, 40 MW_t is required to preheat the gasifier feed materials to 700°C.

The reader is cautioned from concluding from these data which of these processes is "best". For example case III requires an additional 105 MW transferred from the steam bottoming cycle over Case I. In general the reformer operates better at temperatures in excess of 800°C and the gasifier at 700°C (or lower if kinetically possible).

5.2.4.5 Discussion

The catalytic fluidized bed gasification system offers several unique advantages for application of nuclear supplied thermal energy to coal processing. To

Table 5-12: Product Streams in ST moles/hr*
for 2000 ST/D Illinois #6 Coal Feed
to Exxon Gasifier Operating @ 700°C

STREAM	1	2	3	3'	4	4'	5	5'	6	7	8
CH_4	5.472	1.936	1.926	1.926	3.546	3.546	-	-	-	3.546	5.472
H_2O	16.415	11.239	-	-	-	-	-	-	0.564	trace	-
CO	-	1.910	1.910	1.910	1.784	1.784	2.971	0.722	-	1.784	-
CO_2	-	1.631	1.631	-	trace	-	-	-	-	trace	-
H_2	-	12.267	12.267	12.267	0.004	0.004	7.654	4.618	-	0.004	-

Nuclear Heat @ 950°C	215.6 MW (Reformer Heat @ 825°C)
Btu/day (HHV) in syn gas	6.254×10^{10}
Work in Cryogenic Separation	32 MW_e

*Short ton moles/hr.

Table 5-13: Product Streams in ST moles/hr
for 2000 ST/D Illinois #6 Coal Feed
To Exxon Gasifier Operating at 825°C

Stream	1	2	3	3'	4	4'	5	5'	6	7	8
CH_4	5.604	1.972	1.972	1.972	3.631	3.631	-	-	-	3.631	5.604
H_2O	16.811	11.510	-	-	-	-	-	-	0.480	trace	
CO	-	1.956	1.956	1.956	1.754	1.754	2.942	0.768	-	1.754	-
CO_2	-	1.670	1.670	-	0.001	-	-	-	-	0.001	-
H_2	-	12.563	12.563	12.563	0.068	0.068	7.703	4.928	-	0.068	-

Nuclear Heat @ 950°C (Reformer Heat @ 825°C)	220.7 MW
Btu/Day (HHV) in Syn Gas	6.266×10^{10}
Work in cryogenic separation	36 MW_e

Table 5-14: Product Streams in ST moles/hr for 2000 ST/D Illinois #6 Coal Feed to Exxon Gasifier @ 700°C

Stream	1	2	3	3'	4	4'	5	5'	6	7	8
CH_4	7.009	3.462	3.462	3.462	3.546	3.546	-	-	-	3.546	7.006
H_2O	35.044	28.785	-	-	-	-	-	-	0.563	trace	-
CO	-	0.827	0.827	0.827	1.353	1.353	1.893	0.287	-	1.353	-
CO_2	-	2.712	2.712	-	0.001	-	-	-	-	0.001	-
H_2	-	13.345	13.345	13.345	0.013	0.013	8.732	4.626	-	0.013	-

Nuclear Heat @ 825°C (Reformer Heat @ 700°C)	181 MW
Btu/day (HHV) in Syn Gas	6.260×10^{10}
Work in cryogenic separation	40 MW_e

produce methane, nuclear energy could be used to supply preheat for the reaction gases, energy for recovery of the catalyst, and energy for the cryogenic separation of the product methane. When used in conjunction with a steam reformer, the catalytic gasification system could also use nuclear energy to supply a portion of the heat content of the product syn gas. The most important characteristic of this system is the fact that no heat need be transferred directly into the gasifier vessel; thus, the problems of designing a heat exchanger to operate in a corrosive environment are completely eliminated.

This system is under development and not all of the process characteristics are available in the open literature. However, estimates were made which indicate that a combination of a nuclear heat supply, a steam reformer, and a catalytic gasification system producing a syn gas product would require approximately 40% less coal per unit energy output than a conventional catalytic gasifier producing a SNG product. Thus, both emissions and coal consumption would be reduced by approximately 40%.

5.2.5 Summary of Applications of Nuclear Heat to Coal Gasification Processes

In this section, the application of nuclear heat to coal gasification has been investigated. None of the three types of gasification vessels lends itself directly to the application of nuclear heat to coal gasification because of the severe problems associated with transferring the heat across a surface exposed to an extremely corrosive environment. However, a nuclear heat source might be employed to supply energy to peripheral processes such as the steam supply, the oxygen plant or the raw gas treatment processes in all of these existing gasification systems. Most of these applications utilize lower temperature heat which could possibly be supplied by a Light Water Reactor. By using a High Temperature Reactor in the catalytic gasifier reformer system, additional nuclear heat could be incorporated into the product heating value and, perhaps, the other steam requirements could be supplied more efficiently. When both the addition of nuclear energy to the product heating value and the potential replacement in peripheral processing are considered, approximately 40% of the coal required in the gasification process could be substituted by nuclear heat. Thus, coal consumption and emissions of CO_2, SO_2, and other undesirable effluents could be reduced by approximately 40%.

Three major portions of the nuclear heated catalytic gasification system are yet to be fully developed. The gasifier itself has only been run at a bench scale level. Experimental results show that the catalyst appears effective. However, a broad program is needed to prove the effects of scale-up and to determine if there are more efficient and economic catalysts which might be employed. The reformer presents a major materials problem in that materials of construction that will survive the reducing environment for the 30 year period required for nuclear licensing have not yet been identified. Work is currently underway in both the U.S. and the FRG aimed at identifying candidate materials for this application. Experimental work in the FRG on the Pebble Bed Reactor and development work at General Atomic Company on the prismatic core design have laid the basis for application of the HTR to nuclear process heat applications. If a major program in Nuclear Process Heat is undertaken in the U.S., each of these three areas must be included in the investigation.

5.3 APPLICATIONS OF COAL/NUCLEAR DERIVED HYDROGEN TO CHEMICALS PRODUCTION

5.3.1 Introduction

The gasification of coal by the catalytic gasification system under consideration in the previous section could be an integral part of a variety of schemes for the production of synthetic fuels, hydrogen and chemicals and for the delivery of process heat and electricity to the industrial sector and SNG to the residential sector via a Thermochemical Pipeline (TCP). These applications are exemplified in Figure 5-12. It is easily seen that syn gas (CO/H_2 mixtures) is a flexible resource. It may be incorporated in an open TCP, i.e., distributed as syn gas, remotely methanated for cogeneration, process steam or heat and for SNG production (This concept is described in more detail in Section 6 of this report). The syn gas product may be shipped to a Fischer-Tropsch plant for conversion to methanol or synthetic fuels (e.g., gasoline) or may be used in direct reduction of iron ore to sponge iron. By operating shift and methanation reactors, hydrogen gas of known

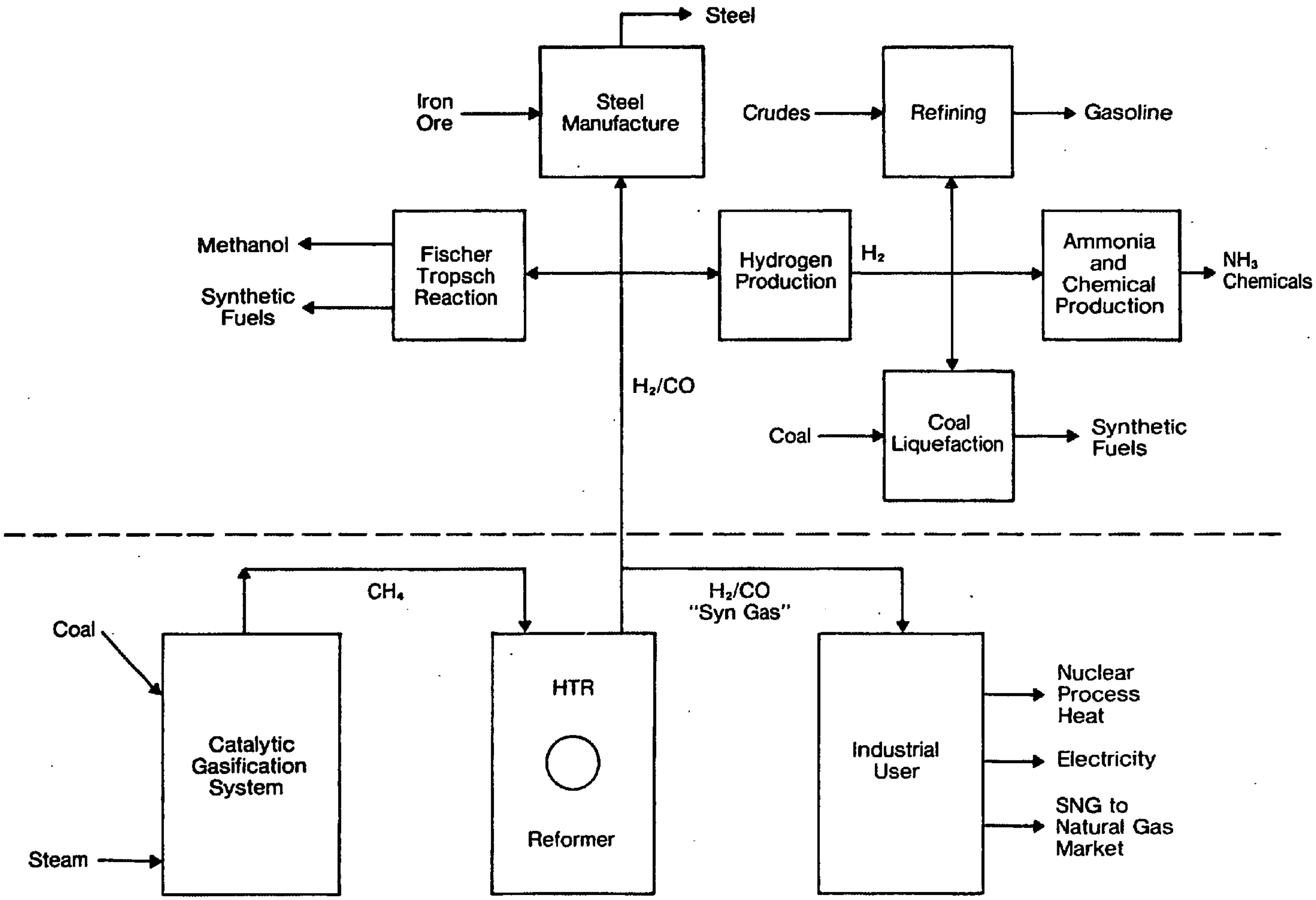

Figure 5-12. ENERGY, FUELS AND CHEMICAL COMPLEX BASED ON HIGH TEMPERATURE HEAT SOURCE FOR COAL GASIFICATION

purity could be a product which might be used in a variety of high H_2 demand processes, e.g., ammonia and fertilizer, coal liquefaction and refining of crudes (from petroleum*, kerogen** or coal***) to gasoline or other light products.

The design of processes for these applications differ markedly. For example, the purity of the hydrogen demanded is critical in determining steam This requirements. In addition, these processes all have large energy demands for process steam and electricity in excess of the direct NPH indicated in Figure 5-12. These power demands can also be met by an HTR. It would be imperative to maximize the energy economy of an HTR/chemical complex as previously discussed by supplying the process steam and electricity as a "bottoming cycle" of the HTR. All of the above chemical processes will produce "waste heat" because of mismatches between demand and cooling streams. The effective utilization of these waste heat streams would be as boiler preheat in the "bottoming cycle" of the HTR. Failure to incorporate these features would lead to poor overall efficiencies and large cooling tower construction whose purpose would be to dissipate thermal energy which would be otherwise useable.

It has not been possible to consider all of the processes shown in Figure 5-12 in detail for lack of time. However, a discussion of coal liquefaction, ammonia manufacture, and iron ore reduction using hydrogen from the catalytic gasifier system will be given in some detail. Other possible applications for HTR heat utilization will be discussed very briefly.

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*Petroleum refiners generally use about 10% of their feedstocks for processing. Light oil products are approximately $CH_{2.2}$ in overall stoichiometry.

**Kerogen is the organic phase of shale. It contains less H_2 that petroleum, approximately $CH_{1.5}$ overall.

***Coal has very limited hydrogen. It requires large quantities of H_2 to upgrade. Coal is approximately $CH_{0.8}$.

5.3.2 Coal Liquefaction Using Nuclear Heat

5.3.2.1 Introduction

Hydrogen gas must be added to raw coal to produce liquid products. Therefore, it is possible to consider the use of an HTR, a coal gasifier and a liquefaction plant to produce liquid fuels as shown schematically in Figure 5-13.

It is necessary to ask a vital question: Is the savings in combusted coal sufficient to justify the HTR concept over traditional coal gasification for H_2 production? To explore this question we will consider three prototypical processes for liquefying coal. These processes which are all technological leaders and in a high state of development, are:

1. Solvent Refined Coal (SRC)
2. Hydrogenated Coal (H-Coal)
3. Exxon Donor Solvent (EDS)

Processes (1) and (2) are described in detail in Reference 5-11 and (3) appears in References 5-12 and 5-13. These processes have been chosen respectively to illustrate increasing hydrogen consumption. Those processes which require more hydrogen will require more HTR heat. However, the quality of the product is directly proportional to the hydrogen consumption. Thus, the SRC process favors manufacture of boiler fuel and the EDS process favors gasoline and other light petroleum products. As indicated in Figure 5-13, the liquefaction process has two steps, hydrogen manufacture and coal liquefaction. Before investigating these liquefaction processes in detail, the production of a "liquefaction grade" hydrogen product from the catalytic gasifier/nuclear reformer system will be briefly discussed.

5.3.2.2 Production of "Coal Liquefaction Grade" Hydrogen

For certain applications pure hydrogen is not needed; a key process for which this is true is the liquefaction of coal. The literature cited in the previous section indicates that a gas stream containing approximately 5% CO/95% H_2 is sufficient for liquefaction . The flowsheet for a process to produce this quality of hydrogen is shown in Figure 5-14. In Figure 5-14, the process is initiated with the product gas from the nuclear reformer, a mixture of CO, CO_2, H_2, H_2O and CH_4.

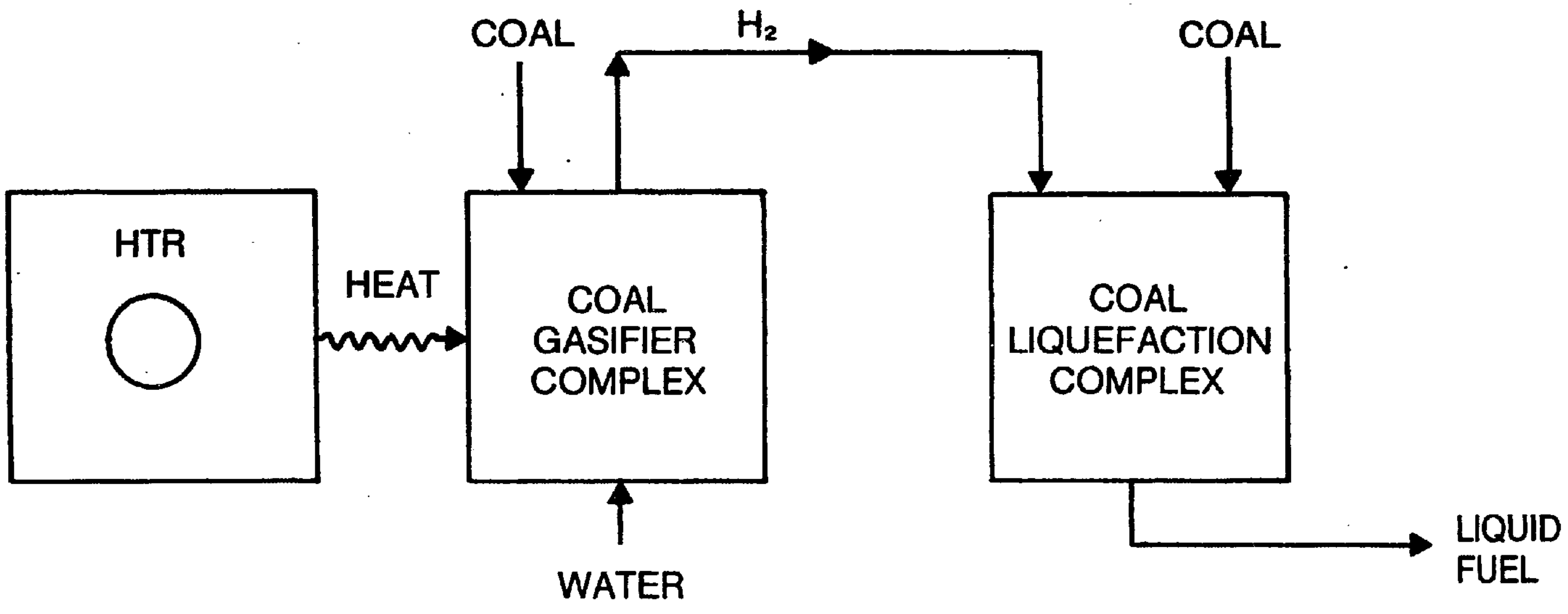

Figure 5-13. HTR/COAL LIQUEFACTION COUPLING