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An Evaluation of Nonelectrical Energy Transmission Options

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AN EVALUATION OF NONELECTRICAL ENERGY TRANSMISSION OPTIONS

T. K. Stovall J. G. Delene
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ABSTRACT

The possibility of substituting nonelectrical energy forms (generated at central nuclear plants) for prime fuels (gas and oil) is evaluated in this study. Steam and combustible gases are considered appropriate energy forms for delivery to industrial users. The energy transport mechanisms evaluated include molten salt, hot oil, open- and closed-loop chemical heat pipes, and hydrogen. Nonnuclear options including medium-Btu gas and hydrogen (both from coal) provide a comparative base for nuclear alternatives.

All economic evaluations are presented in constant 1980 dollars for plant start-ups in the years 1990 and 2000. The delivered-energy cost is provided for each option as a function of delivery distance for a common set of economic ground rules. Environmental, institutional, and other intangible factors are discussed for each energy transport option and for very high temperature reactors.

1. EXECUTIVE SUMMARY

1.1 Introduction

1.1.1 Purpose and scope

The industrial sector of the United States consumed 9.1 EJ (8.6×10^{15} Btu) of natural gas and 8.0 EJ (7.6×10^{15} Btu) of petroleum during 1979.[1] This study examines the possibility of substituting nonelectrical energy forms, generated at central nuclear plants, for these prime fuels. By considering energy transport distances up to 320 km (200 miles) the study goes beyond localized distribution systems that serve a limited number of customers within a short distance ($\leqslant$2 km). This extended range is expected to increase the load on the central energy supply system and to provide economies of scale to those smaller industrial users located along serviceable routes.

Prior assessments of this subject have included General Electric Chemical Heat Pipe (CHP) Studies,[2] Oak Ridge National Laboratory/General Atomic Company (ORNL/GA) Molten Salt Studies,[3] Brookhaven National Laboratory Hydrogen Studies,[4] and an ORNL High-Temperature Energy Transport Systems Study.[5] The purpose of this evaluation is to use the results of these prior studies, incorporate the most recent technological and cost information, and place the various energy alternatives on an equal basis for comparison.

This comparison was then used to understand the relative benefits of the energy transport alternatives powered by either high-temperature gas-cooled reactors (HTGRs) or light-water reactors (LWRs). The results are also used to indicate possible directions for future research and development.

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1.1.2 Market

The industrial market review of petroleum and natural gas uses includes process heat, steam, feedstock, and mechanical drive applications. Process heat and steam uses are specified based on required temperatures and pressures.

Not many industrial plants were found to be big enough to directly support large-scale energy sources such as nuclear reactors. However, geographical areas are identified that contain large numbers of industrial users. These areas will be particularly amenable to a central energy supply and distribution system.

This study does not project changes in industrial energy use caused by conservation efforts, nor does it quantify any future market for the potential synthetic fuel industry. These changes, along with the trend toward industrial cogeneration, could have a substantial impact on the industrial energy temperature profile. However, these changes are unlikely to diminish the market for the types of energy systems considered in this study.

1.1.3 Alternatives considered

The energy transport mechanisms considered in this report include molten salt, hot oil, hydrogen, and a CHP. Nonnuclear alternatives including coal-derived hydrogen and medium-Btu gas (MBG) are considered for comparative purposes. Steam is not considered because of its limited transport range. Coal-fired steam generation on site is a viable industrial option but is not considered here because of necessary site-specific modifications and environmental concerns.

Distribution distances as great as 320 km (200 miles) are examined to allow placement of the energy supply system remote from industrial population centers. These longer distances can also be appropriate for widely scattered groups of industrial users. The systems are scaled for delivery of 800 to 1200 MW(t) at the user's site.

The energy transport alternatives are evaluated based on their delivery of either steam or combustible gases. The costs therefore include heat exchangers to transform sensible heat or chemical energy to steam at the plant boundary. The costs of interior plant modifications are not included.

All economic analyses are based on the energy delivered at the plant boundary, even for gases that could be used as feedstock. No premium is given for the delivery of high-temperature steam vs low-temperature steam. The calculation of such a premium would involve an analysis of available alternative fuels for each industry's high-temperature energy needs. Such analysis was beyond the scope of this study. Given this limitation, these cost treatments were chosen to provide a consistent basis for comparison between the various energy delivery alternatives.

1.2 Results

1.2.1 Description of systems

Molten salt and hot oil systems. The molten salt and hot oil systems are each composed of two heat exchangers and a closed-loop pipeline. The fluids pick up heat from intermediate heat exchangers (IHXs) located at a nuclear plant and are then transported through insulated pipelines to heat exchangers located at the user's plant boundary. Sensible heat is here removed from the fluids to provide steam to the user. The relatively cooler fluids are then pumped back through insulated return pipelines to the nuclear plant.

Energy delivery temperatures are determined by the fluid and pipeline material properties and range from about 200 to 500° C.

Hydrogen. Four methods of hydrogen production are examined. Three of these methods use electricity generated at a nuclear plant. These electrolysis methods include the traditional potassium hydroxide (KOH) electrolyte, the solid polymer electrolyte (SPE) under development at General Electric, and SPE with storage using off-peak power rates. The production of hydrogen from coal is also analyzed.

The hydrogen distribution system consists of traditional pipelines similar to those commonly used by the natural gas industry.

CHP. The CHP system is based on methane-steam chemical reactions. Methane and steam are mixed at high temperatures in a reformer. The resulting endothermic reaction absorbs heat from a nuclear reactor through an IHX. Heat is recovered from the product gases, and they are transported at ambient temperature to the user's plant site. These gases are there exposed to a catalyst in a methanator and the reverse reaction, an exothermic reaction, takes place. The released heat is then transferred to water to produce steam for the user's plant.

The gases (again at ambient temperature) are then piped back to the reformer plant (closed cycle) or sold to the user as a fuel gas (open cycle).

MBG. Medium-Btu gas is a synthetic gas with a heating value between 9.3 and 13 MJ/scm (250 and 350 Btu/scf). Available commercial processes for the production of MBG from coal include fixed-bed gasifiers, fluidized-bed gasifiers, and entrained-flow gasifiers.

Gas delivery systems are based on traditional natural gas pipeline systems. This study examines three transmission pressures: 0.9, 1.8, and 3.6 MPa (130, 260, and 520 psia).

1.2.2 Assessment of applicability

The applicability of any energy transport system to any given user group will depend on several parameters. The major parameters to consider include temperature, distance, load curve (as a function of time), feedstock market, and alternative energy costs.

Molten salt can provide high-temperature energy for near and intermediate distances (up to ~120 km). Hot oil does not provide temperatures as high as molten salt but is economical for greater distances (up to ~200 km).

Medium-Btu gas or hydrogen could be chosen when there is a combined need for fuel and feedstock. The gaseous fuels are also good choices for loads that vary with time. Because the pipeline itself provides a moderate amount of storage (through gas compression), the supply system need not be exactly synchronized with the load. Medium-Btu gas and hydrogen are especially suitable for long-distance transport, because their costs are not strongly affected by distance. A summary of possible end uses is shown in Fig. 1.1.

1.2.3 Institutional and technical concerns

It would be to industry's advantage to continue to treat its energy needs as current costs and to avoid the capital charges related to the installation of new energy generation facilities. On the other hand, setting up a utility-like energy distributor will create other capital requirements problems. However, the capital can probably be made available if the market exists on a favorable economic basis.[6]

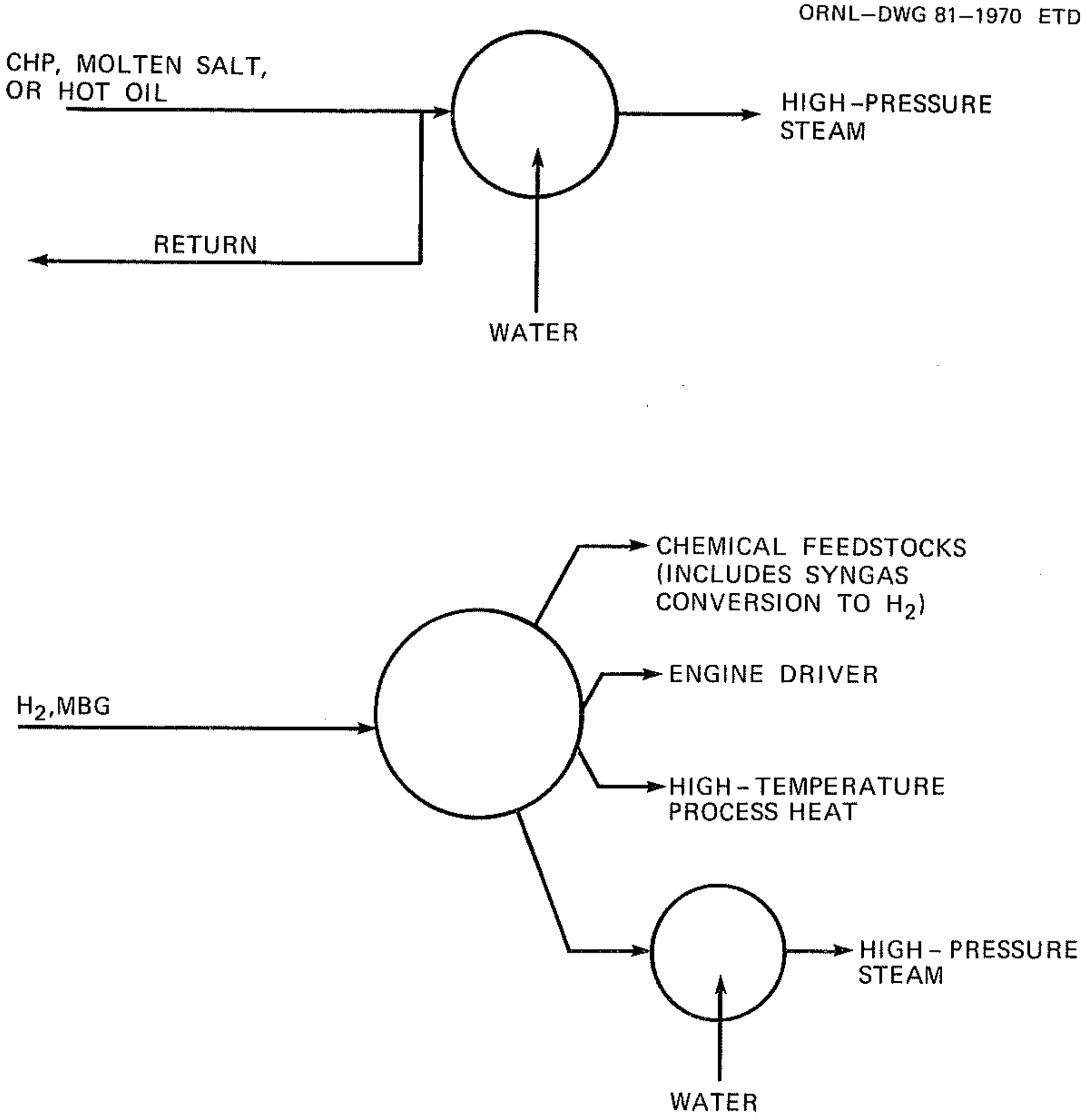

Fig. 1.1. Energy end uses.

Acceptance of long-distance transport systems by industry will depend on cost and reliability. It may be necessary to provide extensive reliability studies, pilot operation, and back-up systems to gain the support of industrial users. Even with such assurances, demonstration of decided cost advantages or lack of alternative supplies will be necessary.

Hydrogen. The major hazard associated with pipeline transportation of hydrogen gas is leakage followed by combustion. Because of its small molecular size, hydrogen is more likely to leak than any other combustible gas.

A hydrogen pipeline network is in use in the Rhine-Ruhr area in Germany. Their operating experience has been quite good over a period of 30 years, with some fires but no explosions or hydrogen embrittlement problems.

CHP. The CHP requires a very high temperature gas-cooled reactor (VHTR) to supply the necessary high temperatures (>800° C). A number of technical and institutional barriers must be overcome before commercial deployment of the VHTR.

The coupling of a VHTR and a CHP introduces many new concepts to the already complicated task of licensing nuclear power plants in the United States. Acceptance must be obtained for the following regulatory and institutional issues:

1. CO-H_2/CH_4 proximity to a nuclear power plant (potential effects of explosion),
2. trace radioactive contamination (tritium) of the process gases,
3. CO-H_2 transmission pipelines, and
4. coupling with industrial processes.

Other institutional issues include (1) deployment of a relatively new reactor technology, which involves an enormous investment by both suppliers and utilities; (2) national energy policy hostile to new nuclear energy development over the past decade; (3) the regulatory environment pertaining to nuclear energy, which has become increasingly onerous during the past decade; (4) the flexibility that allows public utilities to assume major developmental risks, which has been diminished by politically based public utility commissions; and (5) the fact that the licensing basis for VHTRs is yet to be developed in the context of current requirements.

Development and characterization of metallic and ceramic materials for long-term operation in the VHTR environment is required for applicable code qualification and design verification.

In current reformers, cycling the reformer tubes over an extended temperature range in a short time interval is catastrophic because of extreme metal expansion and contraction stresses placed on the tubes and the structural degradation of catalyst pellets under such cycling. This limitation must be considered when coupling a reactor to a reformer.

Pipeline leaks of CO/H_2 and CH_4 gases can pose dangers similar to those described for a hydrogen pipeline system. CH_4 pipelines are routinely accepted and have been used for ~40 years.

MBG. The main environmental concerns regarding coal gasification include particulates, ash, nitrogen and sulfur oxides, and organic condensates. When used with available control technology, however, coal gasification is one of the cleanest ways of using coal today. Medium-Btu gas pipelines pose leakage hazards similar to those discussed for hydrogen pipelines.

Molten salt and hot oil. The operation of molten salt and hot oil pipelines gives rise to several areas of concern. Pipeline rupture and a fluid spill would result in property and environmental damage and create a potential threat to persons in the area. However, if a buried pipeline is considered, the potential danger is diminished except at pump stations. Spilled salt tends to freeze and its spread would be limited. Hot oil spreads further and is more difficult to clean up. The hot oil is also likely to burn.

Barriers to the commercial deployment of HTGRs are very similar to those faced by VHTRs, discussed previously. The molten salt and hot oil systems require an HTGR energy source to provide the necessary high temperatures.

1.2.4 Economic analysis

Capital investments. The capital investments for each system, shown in Table 1.1, range from \$200 to \$750/kW(t). The hydrogen systems are found at the very top and at the very bottom with an SPE system requiring the least capital investment and a KOH system the most. The hot oil, molten salt,

Table 1.1. Capital investments required per unit of energy delivery capacity (1980 dollars)[a]

System	Unit investment [$/kW(t)]
Molten salt/ss	430
Molten salt/cs	350
Hot oil	300
CHP, closed loop	700
CHP, open loop	650
SPE hydrogen	200
SPE hydrogen with storage	550
KOH hydrogen	750
Hydrogen from coal	690
MBG (at 0.9 MPa)	430

[a] Based on 1000-MW(t) system, transport distance of 80 km (50 miles), 2000 start-up.

and MBG systems rank in the bottom half, while the CHP and hydrogen (except SPE) systems are clearly more expensive.

Fuel costs. Prediction of future costs for energy resources is difficult because of the current state of uncertainty concerning fossil fuels. In this study, all fuel costs are levelized for plant start-up dates of 1990 and 2000, assuming utility financing conditions. The resulting energy costs are shown in Tables 1.2 and 1.3.

Table 1.2. Electric generation costs[a]

[mills/kWh(e)]

	System		
	PWR[b]	HTGR[b]	Coal[c]
1990 start-up			
Capital	13.3	14.1	11.0
O&M	2.6	2.4	3.5
Fuel	8.4	7.7	22.6
Total	24.3	24.2	37.1
2000 start-up			
Capital	14.8	15.6	12.2
O&M	2.6	2.4	3.5
Fuel	9.4	8.5	27.5
Total	26.8	26.5	43.2

[a] Based on utility economic assumptions. Levelized to remain constant in dollars of 1980 purchasing power and to increase at the general rate of inflation in nominal (dollars in any given year) terms.

[b] Based on a 1200-MW(e) plant.

[c] Based on an 800-MW(e) plant.

Table 1.3. Nuclear process heat costs[a]
($/GJ)[b]

	Reactor system		
	PWR[c]	HTGR[d]	VHTR[e]
	1990 start-up		
Capital[f]	2.18	2.62	2.87
O&M[g]	0.40	0.39	0.38
Fuel[h]	0.85	0.95	1.05
Total	3.43	3.96	4.30
	2000 start-up		
Capital	2.39	2.86	3.13
O&M	0.40	0.38	0.39
Fuel	0.96	1.06	1.16
Total	3.75	4.30	4.68

[a] Based on utility economic assumptions, 1000 MW(t). Levelized to remain in dollars of constant 1980 purchasing power and to increase at the general rate of inflation in nominal (dollars in any given year) terms.

[b] Includes IHX, but excludes electric generation.

[c] Process steam temperature 340° C (650° F).

[d] Process steam temperature 540° C (1000° F).

[e] Process steam temperature 760–870° C (1400–1600° F).

[f] Scale factor to other sizes is −0.52:

$$\text{Cost} = [\text{cost at } 1000\ \text{MW(t)}] \times [\text{size}/1000\ \text{MW(t)}]^{-0.52}$$

[g] Scale factor to other sizes = −0.7.

[h] Scale factor to other sizes = −0.1.

System Comparisons. Figure 1.2 shows the delivered-energy costs for each energy transport system over distances up to 320 km (200 miles) for a 2000 start-up. These costs include capital charges, operation and maintenance (O&M), and fuel costs. The economic assumptions are shown in Table 1.4.

The results given in Fig. 1.2 show that molten salt and hot oil systems offer energy at the lowest cost for distances up to ~120 km. These costs range from ~$5 to $9/GJ. Hot oil costs ~$8 to $10/GJ between 120 and 200 km. Beyond ~200 km, the hydrogen and MBG are the lowest cost alternatives at ~$10 to $13/GJ. The traditional KOH electrolytic hydrogen and CHP systems cost ~$15 to $23/GJ.

Although there is room for further optimization of each approach, the results presented here are believed to provide a good basis for comparison.

1.3 Conclusions and Recommendations

1.3.1 Molten salt and hot oil

Molten salt and hot oil systems are shown to be competitive for short and intermediate distances (<120 km). The high temperatures available with these sensible heat carriers are appropriate for a large portion of the industrial energy market. The chemical and petroleum industries are currently

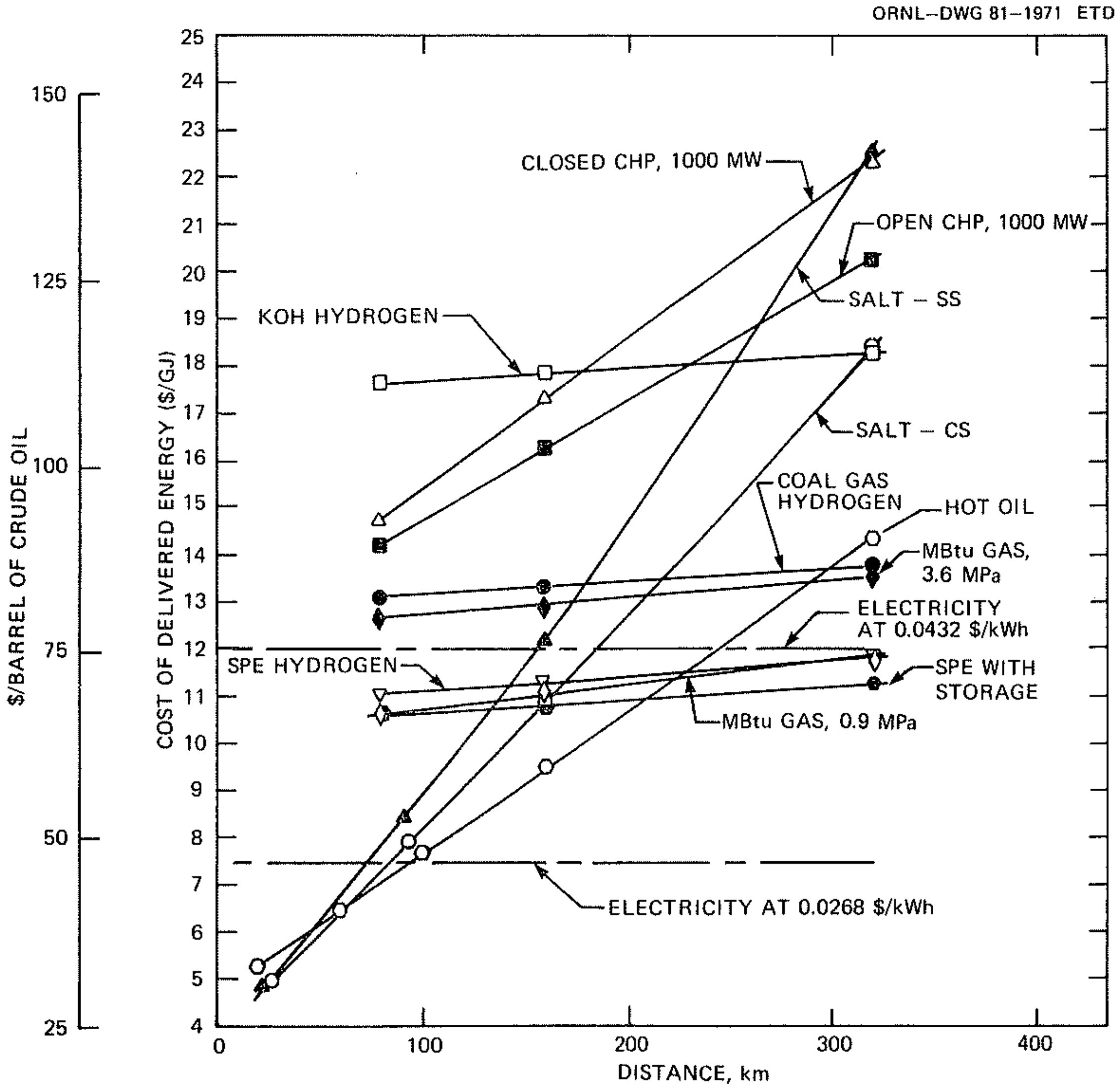

Fig. 1.2. Thermal energy transport options, 1200 MW, 2000 start-up, 1980 dollars.

Table 1.4. Economic factors

General inflation rate, %/year	6.0
Interest rate during construction, %	9.5
Federal income tax rate, %	46.0
State income tax rate, %	3.0
Property tax rate,[a] %	2.5
Insurance rate,[a] %	0.5
Interest rate on bonded debt, %	9.0
Return on equity for utility, %	13.0
Debt/equity capitalization ratio for utility	53/47
Average escalation rate for capital construction, %	7.0
Fixed charge rates for utility, %	17

[a] Based on initial capitalized investment.

consuming large amounts of oil and natural gas to raise steam of varying pressures, including some high-pressure steam that would require the temperatures available with a molten salt system. The industries are mainly located in the central southwest United States. Other high-temperature energy requirements are found in the metals and stone/clay/glass industries.

These concepts merit further consideration. If the barriers to HTGR commercialization are lessened or removed, site-specific applications should be considered.

1.3.2 Hydrogen

The SPE and coal-based hydrogen systems appear to be competitive with coal-derived MBG although coal-based hydrogen is more costly than coal-derived MBG. The hydrogen can serve as both fuel and feedstock to the petroleum and chemical industries. Both of these industries meet ~70% of their steam requirements with natural gas and oil. Hydrogen can be a direct substitute for these prime fuels.

Since SPE systems do not rely on any particular type of power plant, the only impediment to their application is the developmental nature of the SPE technology itself. Once major pilot plants have been constructed to test and prove the concept, distribution systems should be evaluated for specific sites.

1.3.3 MBG

The MBG market is very similar to that discussed for hydrogen in Sect. 1.3.2. It also has feedstock qualities as well as fuel value.

Gasifier technology is fairly well established. This study assumes that some system improvements will be made within the next few years. These systems will require testing, and operational experience is needed for large-scale installations. Medium-Btu gas certainly presents an attractive method of substituting coal for current consumption of oil and natural gas.

1.3.4 CHP

The CHP was the most expensive (based on the cost numbers used here) energy transport alternative considered, which is not surprising, because it delivers only ~1.7 MJ for every standard cubic meter of gas transported. This output can be compared to ~35.5 MJ/scm for natural gas and ~10 MJ/scm for MBG. Unless some major technical breakthrough is discovered, there is no reason for further consideration of the CHP concept.

1.3.5 Summary

The industrial market review included in this report shows that there are several geographical areas where groups of industrial energy users are clustered together. While this study is focused on the industrial market, the concept can also be applied to commercial or residential district heating systems. The selection of a distribution method will be based upon the distribution distance and load size involved. The user's equipment conversion costs will also differ among the various concepts.

There does appear to be merit to the general concept of nonelectrical energy distribution based on the results of this study. The results given in this study indicate that centralized hydrogen distribution systems supplied by nuclear power can be competitive with synthetic gas (MBG) systems for long-distance energy transmission and that molten salt and hot oil can be considered for intermediate distances.

The next logical step in the study process is the detailed examination of candidate distribution systems (molten salt, hot oil, MBG, and hydrogen) laid out to serve existing or potential user groups. This study would extend the previous studies[7] of isolated groupings of users. It would also be desirable to explore the possibilities of using cogenerated energy to serve commercial or residential district heating systems.

Although this study is directed at nonelectrical energy forms, the results show that the direct use of electricity may be a good choice. The concept of electrification deserves serious consideration and further study.

The CHP systems are not economically competitive in this study. Any future work in this area should be directed at finding new technological methods that would improve the economics.

This study does not propose to select a preferred reactor type, and both LWRs and HTGRs are considered. Significant markets appear to be present for either type.

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2. MARKET ASSESSMENT OF INDUSTRIAL ENERGY CONSUMPTION

2.1 INTRODUCTION

To provide a basis for the assessment of the potential for thermal energy transport from a centrally sited nuclear energy facility, a survey was performed to determine the amount and types of energy that are consumed in the various U.S. industrial sectors. Section 2 summarizes this survey.

For the past ten years or more a census has been taken, at two-year intervals, of industrial energy consumption in the United States. These data are compiled and published approximately two years after each census by the Department of Commerce's Bureau of the Census.

The major data base for industrial energy use is that gathered in 1974 (Ref. 1). Currently, the Energy Information Administration is compiling an industrial energy data base, and the Oak Ridge Associated Universities' Institute for Energy Analysis has prepared a similar compilation (based on 1976 data), *Industrial Energy Use Data Book,* which became available in 1980. Pertinent data reproduced from galleys of this publication are included in Appendix A.

Although much is presently being done and has been done in the last five years to conserve energy in all industrial sectors and especially in the more energy-intensive ones, an updated data base is unfortunately not available at this time. In many instances, industry management, particularly in the chemical sector, has chosen to spin off marginal profit lines and use the resulting fuel and feedstocks to manufacture products which offer a higher return on investment or an improved product mix. Some of this information will be forthcoming from the current census data in about two years.

This assessment is presented in three major sections. Section 2.2 summarizes the industrial energy requirements in the United States with a breakdown of process energy, steam and electrical requirements, and the fuels that are being used to provide them. The chemical, petroleum, and pulp and paper industries are closely examined. Section 2.3 presents information on the current production and end uses of hydrogen and projections of possible future markets for hydrogen in the evolving synthetic fuels industries. Section 2.4 identifies possible future markets for MBG as a substitute fuel and feedstock.

2.2 Industrial Energy Requirements

The estimated total amount of process heat used in 1974 was 17.5×10^{15} kJ (16.6×10^{15} Btu).[2] About 59% of this heat was consumed in 78 Standard Industrial Classification (SIC) groups.* Five major industries—paper (SIC 26), chemicals (SIC 28), petroleum (SIC 29), rubber (SIC 30), and primary metals (SIC 33)—utilized 75% of the industrial steam generated.[3]

Industrial process heat requirements in the United States represent ~25% of the total demand for combustible fuels (excluding feedstocks). Approximately one-half of this industrial fuel is required for steam production. The balance is consumed in open-fired kilns, driers, and furnaces (17%); process heaters (15%); gas turbines and internal combustion engines (2%); and miscellaneous other applications. Generally, these latter applications require higher operating temperatures than do steam applications. Reference 4 contains a summary of 25 two-digit SIC groups which cover the most energy-intensive industries (Tables 2.1 through 2.3 and Fig. 2.1). Figure 2.2 is an energy flowchart for a combination of the most energy-intensive industries (SIC 20 through 39).

Table 2.4 shows that industrial high-temperature heat use is centered principally in four major sectors—chemical; petroleum; primary metals; and stone, clay, and glass industries. In the chemical

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*A breakdown of these 78 SIC groups is listed in Appendix A.

Table 2.1. Summary of process heat data base by industry (1974)

SIC	Industry	Process heat included within data base [PJ (10^{12} Btu)]
10 14	Mining	136.22 (129.14)
20	Food and kindred products	336.41 (318.93)
21	Tobacco products	1.43 (1.36)
22	Textile mills	122.78 (116.40)
23	Apparel	
24	Lumber and wood products	181.21 (171.80)
25	Furniture	12.45 (11.8)
26	Paper and allied products	1,152.47 (1,092.60)
27	Printing and publishing	
28	Chemicals	563.44 (534.17)
29	Petroleum products	2,781.16 (2,636.67)
30	Rubber	10.23 (9.70)
31	Leather	2.66 (2.52)
32	Stone, clay, and glass	1,045.24 (990.94)
33	Primary metals	3,979.15 (3,772.42)
34	Fabricated metal products	0.03 (0.03)
35	Machinery	
36	Electrical equipment	1.66 (1.57)
37	Transportation	24.57 (23.29)
38	Instruments	
39	Miscellaneous	
Total		10,351.11 (9,813.34)

Table 2.2. Industrial energy use profile

[PJ · year (1974)]

SIC	Industry (products)	Purchased energy			High-temperature furnace thermal process[a]	Boiler steam/hot water[b]
		Fossil	Electric	Total		
20	Food	739	121	860	333	440
21	Tobacco	14	3	17	4	10
22	Textile mill	222	90	312	50	182
23	Textile apparel	19	21	40	6	15
24	Lumber and wood	146[c]	40	186	82	32
25	Furniture and fixtures	10	14	24		10
26	Paper and pulp	1,193	127	1,320	562	623
27	Publishing	56	32	88	40	20
28	Chemicals	2,281	359	2,640	327	692
29	Petroleum	3,337[d]	85	3,422	1,885	1,278
30	Rubber	153	55	208	79	74
31	Leather	21	5	26	10	8
32	Stone, clay, and glass	1,162	438	1,600	824	391
33	Primary metals	2,028	438	2,466	1,642	491
34	Fabricated metals	195	72	267	209	16
35	Nonelectrical machinery	222	80	302	148	90
36	Electrical equipment	169	85	254	121	53
37	Transportation equipment	269	99	368	243	37
38	Instruments	48	13	61	17	32
39	Miscellaneous	48	14	62	31	17
Total		12,332	2,191	14,553	6,613	4,511

[a] Energy use in production of high-temperature direct heat for furnace processes.

[b] Energy use in production of steam and hot water.

[c] Includes three energy units of wood waste products.

[d] Includes 1700 energy units of process residuals.

Table 2.3. Industrial process energy requirements[a]

[PJ/year (1974)]

SIC	Industry products	High-temperature direct heat furnace process	Hot water/ steam process[b]	Mechanical drives[c]	Other[d]
20	Food	185	300	116	20
21	Tobacco	2	5	3	0.5
22	Textile mills	31	126	61	18
23	Textile apparel	4	12	13	5
24	Lumber and wood	47	16	49	1.6
25	Furniture and fixtures		7	12	1
26	Paper and pulp	340	301	211	16
27	Publishing	21	10	29	3
28	Chemicals	195	359	68	370
29	Petroleum	1130	877	191	6
30	Rubber	44	31	58	5
31	Leather	5	6	4	1
32	Stone, clay and glass	507	88	426	13
33	Primary metals	972	184	148	248
34	Fabricated metals	130	13	40	11
35	Nonelectrical machinery	85	42	45	24
36	Electrical equipment	69	30	71	10
37	Transportation equipment	137	28	76	11
38	Instruments	10	23	5	4
39	Miscellaneous	17	12	10	3
Total		3931	2470	1636	747.1

[a] Does not include stack losses or engine losses associated with furnace, boiler, or engine inefficiencies.

[b] Includes space heating requirements.

[c] Includes air conditioning and refrigeration.

[d] For example, lights, sales, gasoline, or electrolysis.

Table 2.4. Industrial high-temperature heat usage

[PJ/year (1977)]

Industry	Temperature ranges		Plant size distribution	
	<870° C (<1600° F)	>870° C (>1600° F)	Largest	Typical
Chemical	290	310	40	2–6
Petroleum	1700		80	10–50
Primary metals		3000	120	5–50
Stone, clay, glass		110	1	0.1–0.3

and petroleum industries, higher temperatures are employed primarily in hydrocracking and reforming operations, that is, ethylene manufacture, styrene manufacture, catalytically reformed gasoline, and viscosity-breaking of pitch.

One example of high-temperature heat use in the primary metals industry is in the production of metallic iron from iron ore, which is carried out in a blast furnace at 1480°C (2700°F), although substitution of direct reduction using hydrogen at 980°C (1800°F) is sometimes technologically

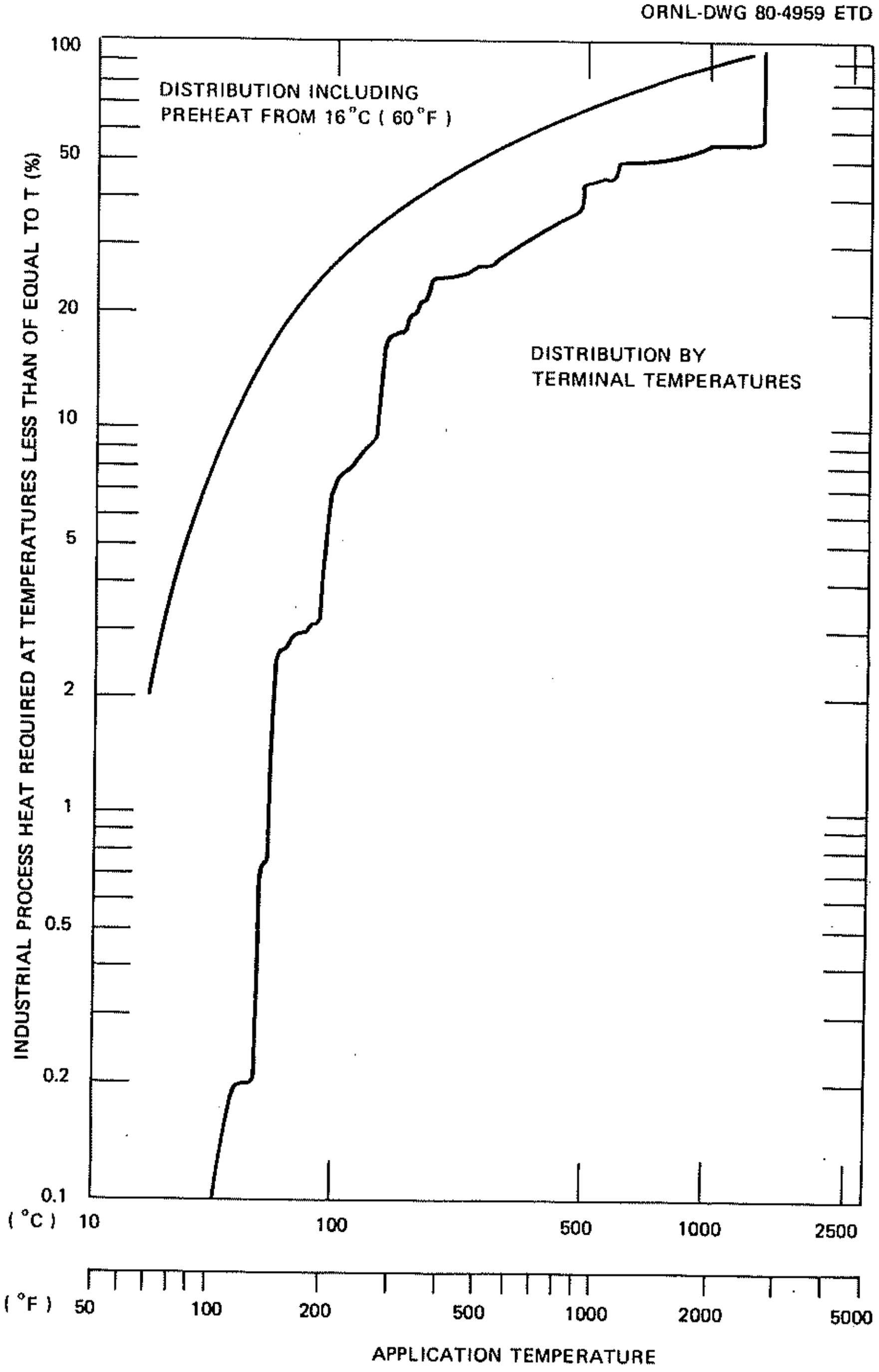

Fig. 2.1. Cumulative distribution of process heat requirements.

feasible. Other high-temperature applications are in glass melting and firing, 1480 to 1590°C (2700 to 2900°F); brick and clay industries, 1370 to 1820°C (2500 to 3300°F); and hydraulic cement calcining, 1260 to 1480°C (2300 to 2700°F). In the latter application, however, the industry is increasingly converting to coal as a substitute for natural gas and oil.

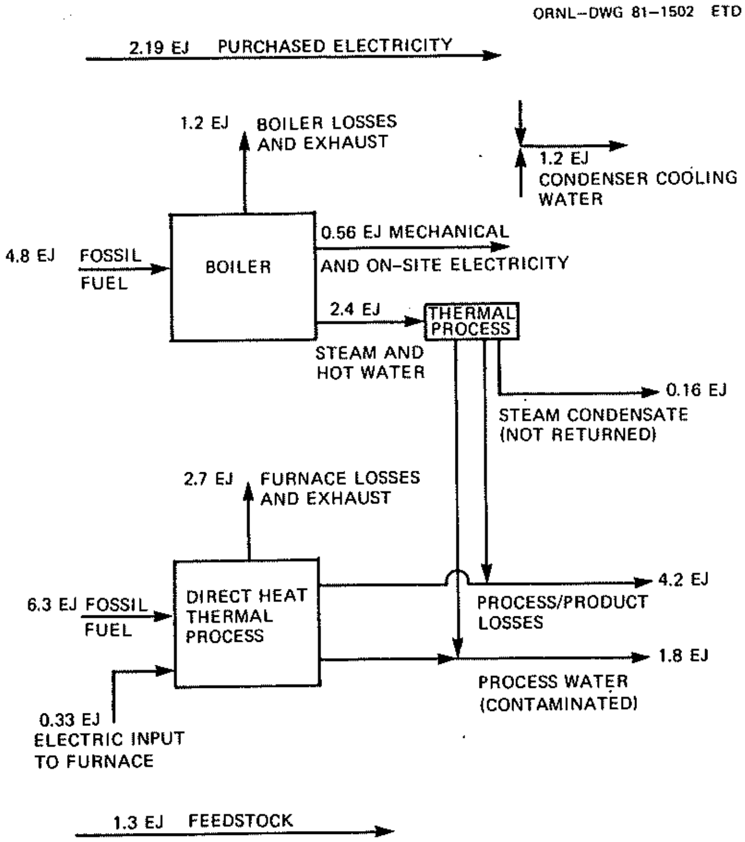

Fig. 2.2. Industrial energy flow, 1974 (includes SIC 20–39). Fossil fuel going into the boiler includes 1700 units of process residuals and 3 units of wood waste. The amount of steam and hot water coming from the boiler includes turbine steam (for thermal process) which was initially used to produce mechanical or electrical energy.

2.2.1 Chemical industry

In the chemical industry, ten major four-digit SIC classifications, shown in Table 2.5, account for 90% of the total energy use in the industry. Within these ten groups are about 1900 plants, over 1300 with 20 or more employees. These 1300 plants account for about 95% of the energy for the ten major classifications. Total fuel consumption in the chemical industry is shown in Table 2.6 (Ref. 5). The fuels

Table 2.5. Chemical product groups

SIC	Chemical product group
2812	Alkalies and chlorine
2816	Inorganic pigments
2819	Industrial inorganic chemicals
2821	Plastic materials and resins
2822	Synthetic rubbers
2823	Cellulosic man-made fibers
2824	Organic man-made fibers
2865	Cyclic intermediates and crudes
2869	Industrial organic chemicals
2871	Fertilizers

Table 2.6. Energy purchased by the chemical industry

[PJ/year (1974)]

Region	States	Number of plants	Purchased fuels					Purchased electric power	Total purchased energy
			Coal	Oil	Natural gas	Other fuel steam	Total fuel		
New England	ME,NH,VT,MA,RI,CT	70		21	11		32	16	48
Middle Atlantic	NY,NJ,PA	280	53	95	42	21	211	95	306
East North Central	IL,IN,MI,OH,WI	230	116	21	106	11	253	158	411
West North Central	LA,KS,MN,MO,NB,ND,SD	60	11		53		63	32	95
South Atlantic	DE,FL,GA,MD,NC,SC, VA,WV	230	58	63	106	11	338	127	465
East South Central	AL,KY,MS,TN	120	85		127		212	285	497
West South Central	AR,LA,OK,TX	200			898	11	909	190	1099
Mountain	AZ,CO,ID,MT,NE,NM, UT,WY	30			53		53	32	85
Pacific	CA,OR,WA	110			85		85	48	133
Total		1330	423	200	1481	54	2156	983	3139

used have significant geographical dependence as a result of the differences in availability and price in each region. Nationally, natural gas provides 47%, coal 13%, oil 6%, and purchased electricity* 31% of the total fuels used in the chemical industry.

Within the U.S. chemical industry, 51% of the total fuel input is used to produce steam. Table 2.7 shows the distribution of fuels for this use. Significant changes in the fuel use pattern undoubtedly will occur because of decreasing availability of natural gas. Many industrial users are switching to oil; and as fuel priorities are established, coal should become more attractive as an industrial boiler fuel.

Table 2.8 shows the quality distribution of steam produced in the chemical industry. Note that the bulk of the steam is produced at much higher pressures (and therefore higher temperatures) than

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*Based on 10,560 kJ of fuel to generate 1 kWh of electricity (10,011 Btu/kWh).

Table 2.7. Fuels utilized for steam generation by the chemical industry

[PJ/year (1974)]

Region	Coal	Oil	Natural gas	Other	Total
New England		21			21
Middle Atlantic	53	53	32	11	149
East North Central	116	11	42	11	180
West North Central	11		32		43
South Atlantic	158	42	53	11	264
East South Central	84		85		169
West South Central			643	11	654
Mountain			42		42
Pacific			63		63
Total	422	127	992	44	1585

Table 2.8. Steam generation in the chemical industry

[PJ/year (1974)]

Region	Steam pressure (kPa)					
	<1030	1030–1710	1710–3080	3080–5825	>5825	Total
Direct firing boilers						
New England		5	5	5	5	20
Middle Atlantic	5	42	21	42	16	126
East North Central	5	48	21	53	21	148
West North Central		11	5	11	5	32
South Atlantic	11	67	40	74	32	224
East South Central	5	48	21	53	21	148
West South Central	26	174	85	190	74	549
Mountain		11	5	11	5	32
Pacific		16	11	16	11	54
Total	52	422	214	455	190	1333
Heat recovery boilers						
Total		21	74	95	21	211

subsequently is required for use as process heat. Although 90% of the steam is produced at pressures above 1380 kPa (200 psi) and 204°C (400°F), more than one-half of the steam is used at temperatures below 149°C (300°F). Table 2.9 presents a regional tabulation of steam utilized in different pressure ranges. Apparently, according to the data shown in Table 2.10, the average is more than two boilers per plant, and larger plants have three or more boilers.

Table 2.9. Process steam utilized in the chemical industry

[PJ/year (1974)]

Region	Steam pressure (kPa)				
	<685	685–1370	1370–3425	>3425	Total
New England	7	13	3	1	24
Middle Atlantic	44	77	22	4	147
East North Central	52	90	25	4	171
West North Central	11	19	5	1	36
South Atlantic	77	135	38	6	256
East South Central	52	90	25	4	171
West South Central	191	334	96	16	637
Mountain	12	19	5	1	37
Pacific	18	32	10	2	62
Total	464	809	229	39	1541

The nonsteam fuel uses in the chemical industry are primarily cracking furnaces for ethylene and other petrochemical manufacture, gas turbines and engines for shaft power, Dowtherm* heaters, and direct-fired kilns or reactors. These applications generally are those that require higher temperatures than are feasible from steam heat (Tables 2.11 through 2.13). Table 2.13 shows the use of miscellaneous fuels and primarily represents process residuals or those fuels that are by-products of the main chemical process. In the chemical industry, these fuel by-products are largely produced in cracking operations, and their use therefore is in these same processes.

The third significant segment of chemical industry energy use is in the production of mechanical work (i.e., shaft power) and in various electrical applications. Although this category includes the use of purchased electricity, which is not included elsewhere, it also includes self-generated electricity and direct steam drive, both representing "downstream" use of the steam generation previously discussed. These uses are shown in Table 2.14.

Under the heading "In-plant generation" in Table 2.14 are two categories, nonsteam drive and steam drive. These categories represent, respectively, (1) self-generated electricity consumption and (2) steam used for applications other than electricity generation. A summary of energy consumption in the chemical industry is given in Table 2.15.

2.2.2 Petroleum refining

Reference 5 presents a summary of energy consumption in U.S. refineries. Energy use per barrel of input in refineries varies widely between plants, covering a range of 0.05 to 0.17 $\mathrm{kJ/m^3}$ (0.3 to 1.0 million Btu/bbl). This variability is not necessarily related to differences in efficiency or conservation practices.

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*Trademark of Dow Chemical Company

Table 2.10. Distribution of plant steam loads in the chemical industry (1974)

Region	Number of plants	Steam generated	Average plant steam load		Steam generated (PJ/year) as a function of plant size [10^3 kg steam/h (10^3lb/h)]							
			PJ/year	10^3 kg/h (10^3 lb/h)	4–6 (<10)	4–11 (10–24)	11–22 (25–49)	23–46 (50–100)	46–114 (100–250)	114–228 (250–500)	228–456 (500–1000)	>456 (>1000)
New England	70	24	0.35	14.5 (32)	6	5	5	4	2	1		
Middle Atlantic	280	148	0.53	22 (48)	8	13	11	11	14	21	21	50
East North Central	230	171	0.74	31 (68)	7	13	11	11	16	19	21	74
West North Central	60	36	0.60	25 (55)	2	3	2	3	4	4	5	12
South Atlantic	230	256	1.12	45.5 (100)	4	9	9	8	24	21	26	153
East South Central	120	171	1.42	59 (130)	2	5	6	5	16	13	21	102
West South Central	200	636	3.17	132 (290)			3	5	24	21	96	485
Mountain	30	37	1.27	50 (110)	1	1	2	2	2	2	4	22
Pacific	110	61	0.53	22.7 (50)	4	5	5	5	6	7	7	20
Total	1330	1540	1.16	47.7 (105)	36	55	55	55	110	110	203	918
Number of plants					620	265	133	97	99	27	32	57
Number of boilers[a]					[b]	330	650	700	800	150		

[a] Estimated from American Boiler Manufacturer Association data for the period 1965 to 1975 (watertube boilers only).

[b] In the smaller ranges, firetube boilers are heavily used and distribution data are not available.

Table 2.11. Natural gas utilized by the chemical industry for nonsteam process heating
[PJ/year (1974)]

Region	Process temperature [°C (°F)]				Total
	40–315 (100–600)	315–540 (600–1000)	540–815 (1000–1500)	>815 (>1500)	
New England					
Middle Atlantic	5		5	5	15
East North Central	11	5	21	5	42
West North Central			5	11	16
South Atlantic	16		11	11	38
East South Central	16		11	11	38
West South Central	26	16	100	58	200
Mountain				11	11
Pacific	5		11	5	21
Total	79	21	164	117	381

Table 2.12. Oil utilized by the chemical industry for nonsteam process heating
[PJ/year (1974)]

Region	Process temperature [°C (°F)]				Total
	40–315 (100–600)	315–540 (600–1000)	540–815 (1000–1500)	>815 (>1500)	
New England					
Middle Atlantic	5		11	11	27
East North Central	5		5		10
West North Central					
South Atlantic	11			11	22
East South Central					
West South Central					
Mountain					
Pacific					
Total	21		16	22	59

Rather, high energy consumption is associated with high processing intensity (such as lubricants, petrochemicals, high gasoline yield fractions requiring high energy consumption) and with crude oil quality [crude with lower American Petroleum Institute (API) gravity requiring more energy to achieve given gasoline and distillate fuel yields than do higher gravity crudes]. The simplest plants consist of a crude oil distillation unit plus auxiliaries such as the boiler, storage, and shipping facilities. The most complex plants include these facilities and also cracking, coking, alkylation, reforming, treating, naphthas and solvents manufacture, aromatics recovery, and lubricants manufacture. Each of these processes uses equipment and energy to increase the average value of the refinery product mix. Apparently, because of the wide range of per barrel fuel consumption in refineries, the relative efficiency of a given plant can be established only by detailed study of its operations.

Table 2.13. Miscellaneous fuels[a] utilized by the chemical industry for nonsteam process heating

[PJ/year (1974)]

Region	Process temperature [°C (°F)] 40–315 (100–600)	315–540 (600–1000)	540–815 (1000–1500)	>815 (>1500)	Total
New England					
Middle Atlantic			5		5
East North Central			5		5
West North Central			11		11
South Atlantic					
East South Central			11		11
West South Central			153		153
Mountain					
Pacific			5		5
Total			190		190

[a] Includes process residuals.

Table 2.14. Electricity and shaft power consumption in the chemical industry

[GWh/year (1974)]

Region	Purchased electricity	In-plant generation: Nonsteam drive	In-plant generation: Steam drive	Total
New England	1,100		100	100
Middle Atlantic	8,800	700	1,100	1,800
East North Central	15,000	700	2,400	3,100
West North Central	2,200	900	300	1,200
South Atlantic	12,900	1,600	3,400	5,000
East South Central	25,600	900	3,100	4,000
West South Central	18,300	9,100	4,000	13,100
Mountain	3,100		100	100
Pacific	4,500	1,100		1,100
Total	91,500	15,000	14,400	29,500

Table 2.16 presents a summary of energy consumption by the largest U.S. refineries. Natural gas is particularly important to southern region refineries, where it represents 51% of the total process energy. On a national basis, natural gas represents 34.3% of total energy consumption.

Nearly 28.7% of the total energy within the industry is consumed in process steam. Approximately 23% of this process steam is produced in direct-fired steam generators, and the remaining 77% is produced in heat recovery boilers. A breakdown of process steam generation from these two sources as a function of pressure is shown in Table 2.17. The fuels utilized for direct-fired boilers are listed in Table 2.18. The distribution of boiler unit capacity vs percentage of total steam generation is shown in Table 2.19; popular unit sizes range from 20,000 to 160,000 kg/h (50,000 to 350,000 lb/h).

Table 2.15. Energy consumption by the chemical industry

(1974)

Type of energy	Consumption [EJ (10^{15} Btu)]	Subtotal
Steam, kPa (psi)		
⩽685 (⩽100)	0.464 (0.439)	
⩽685–1370 (⩽100–200)	0.808 (0.765)	
⩽1370–3425 (⩽200–500)	0.230 (0.218)	
>3425 (>500)	0.040 (0.038)	
⩾5825 (>850)	0.042 (0.040)	
Total	1.584 (1.500)	1.584 (1.500)
Nonsteam, °C (°F)		
40–315 (100–600)	0.100 (0.095)	
315–540 (600–1000)	0.021 (0.020)	
540–815 (1000–1500)	0.370 (0.350)	
>815 (>1500)	0.137 (0.130)	
Total	0.628 (0.595)	2.212 (2.095)
Electricity		
Purchased	0.982 (0.930)	
In-plant generation	0.306 (0.290)	
Total	1.288 (1.220)	3.500 (3.315)

The main waste-heat recovery systems or sources follow.

- *Carbon monoxide boilers.* When coke is burned from catalyst in the catalytic cracking operation, the catalyst attains a temperature of 590 to 650° C (1100 to 1200° F). The hot gaseous product contains 8 to 10% CO, and the gas is burned along with about 10% auxiliary fuel in a waste-heat boiler.
- *Fluid coking plants.* The coking is accomplished by burning some of the coke (in a fluidized condition to heat the coke particles to a high temperature). When oil is sprayed on the hot particles, coking of the oil occurs and additional particles of coke are produced. The hot gases from the coke burner are sent through a waste-heat boiler or are used to heat boiler feedwater.
- *Sulfur recovery plants.* These plants involve a reaction that generates much heat, enough for about a ratio of 3:1 pounds of steam generation to sulfur recovered.
- *Hydrogen manufacture.* This process involves the water-gas reaction at a temperature of about 760°C (1400°F). Heat is recovered from the flue gases used in heating the water (steam) and hydrocarbons to 760°C.
- *Exhaust from gas turbines.* These turbines are used at a very high temperature to generate electricity or shaft power. The hot exhaust is sent, along with regular fuel, to a waste-heat boiler or to the tubestill furnaces used in crude distillation or vacuum plants.
- *Side-circulating reflux.* In order to maintain the heat balance of a fractionating tower, reflux heat must be removed. In very high temperature fractionating towers (e.g., cracking, coking, or vacuum), hot liquid [204 to 316° C (400 to 600° F)] is removed from the side of the tower, cooled, and returned to the tower. In most large operations, the heat is recovered in waste-heat boilers.
- *High-temperature products.* Even after heat exchange with incoming feed, these products at 200 to 370°C (400 to 700° F) are used to a limited extent to generate steam or to heat feedwater.

Table 2.16. Energy consumption of refineries consuming >100 PJ/year[a]

[PJ/year (1974)]

Region	Purchased fuels: Coal and crude	Distributed oils	Residual oils	Natural gas	Liquid petroleum gas	Steam	Total	Refinery gases	Coke (catalyst)[b]	Total fuels	Purchased electricity
New England											
Middle Atlantic		2.5	90	29	3.0	6.3	138	91	36	395.8	20
East North Central		3.2	89	35	6.0		139	169	80	521.2	37
West North Central		0.4	12	28	0.6		42	38	12	133.0	8
South Atlantic		1.3	2	0.6	0.1	7.5	12	20	12	55.5	4
East South Central		0.5	8	28	1.1	2.0	41	41	11	132.6	6
West South Central		21.0	11	748	12.0	16	808	439	144	2199.0	83
Mountain											
Pacific		5.7	22	99	11.2	13	156	162	49	517.9	51
Total		34.6	234	967.6	34.0	44.8	1336	960	344	3955	209

[a] Running at 81.7% capacity (94% is a reasonable maximum).

[b] During 1974 only 1.8 PJ/year of salable coke was burned.

Table 2.17. Steam generation in the petroleum industry

[PJ/year (1974)]

Region	Direct firing [kPa (psi)]				Heat recovery boilers [kPa (psi)]					Total
	<1370 (<200)	1370–3425 (200–500)	3425–6850 (500–1000)	>6850 (>1000)	<685 (<100)	685–1370 (100–200)	1370–3425 (200–500)	3425–6850 (500–1000)	>6850 (>1000)	
Middle Atlantic		10.2	37.6	2.5		3.4	8.6	5.0		67.3
East North Central		16.3	57.3	8.6		8.6	14.5	5.8		111.1
West North Central		5.0	14.9			2.9	2.9	1.4		27.1
South Atlantic							0.3	15.3		15.6
East South Central			12.7			1.3	1.4	2.5		17.9
West South Central		51.2	192.6	12.8		20.5	33.4	12.9		322.4
Pacific		14.7	55.0	3.7		7.0	11.6	4.6		96.6

Table 2.18. Fuels utilized for steam generation[a] by the petroleum industry

[PJ/year (1974)]

Region	Coal	Oil	Natural gas and liquid petroleum gas	Refinery gas (residual)	Coke (residual)[b]	Total
Middle Atlantic		27.7	9.0	25.7	Small	62.4
East North Central		31.9	14.0	55.5	Small	101.4
West North Central		3.9	9.0	11.6	Small	24.5
South Atlantic						
East South Central		1.8	5.8	8.0	Small	15.6
West South Central		8.4	195.4	113.0	Small	316.8
Pacific		8.3	34.0	48.4	Small	91.1

[a] Fuels assumed to be in proportion to total fuels of each variety consumed, based on 81% gross efficiency.

[b] Possibly 10,182 metric tons salable coke burned by entire industry (0.36 PJ/year) in 1974 in all districts. The large catalytic coke consumption was considered to be used for nonsteam heating (and CO-rich gases generating steam in waste-heat boilers).

Table 2.19. Process and other steam utilized by the petroleum industry (including purchased steam)[a]

[PJ/year (1974)]

Region	Steam [kPa (psi)]					
	Process[b]			Other[c]		
	<685 (<100)	685–1370 (100–200)	1370–3425 (200–500)	<685 (<100)	685–1370 (100–200)	1370–3475 (200–500)
Middle Atlantic	22.0	7.2	26.1	10.5	6.7	3.8
East North Central	34.0	11.1	40.0	15.3	10.6	5.5
West North Central	8.2	2.7	9.7	3.9	2.4	1.4
South Atlantic	2.4	0.7		3.0	1.2	0.4
East South Central	5.8	1.9	6.8	2.7	1.7	1.0
West South Central	105.0	33.8	122.5	49.6	31.7	18.0
Pacific	33.0	10.8	39.0	15.5	9.9	5.6

[a] Pressures vary according to need. Actually some of the steam will be furnished from the next higher pressure range, but little steam will be supplied above 3425 kPa (500 psi).

[b] Includes partial pressure reduction, steam stripping, jet ejectors, and process heating or reboiling.

[c] Needs other than for processing such as transfer pumping, pumping semisolids, snuffing or flare, soot blowers, tanks, line tracing, building and shop heating, or fuel atomization. Some of this steam is being used a third time.

The fuels used to supply nonsteam heating requirements are shown in Table 2.20. Electricity and shaft power consumption are shown in Table 2.21. A summary of energy consumption in petroleum refineries is given in Table 2.22.

Table 2.20. Fuels utilized for nonsteam heating[a]

[PJ/year (1974)]

Region	Process temperature[b] (°C)											
	Oils			Natural gas and liquid petroleum gas				Refinery gas (residual)				Coke (residual)[c]
	40–150	150–315	315–540	40–150	150–315	315–540	540–815	40–150	150–315	315–540	540–815	540–815
Middle Atlantic	8.5	7.7	61.0	2.7	2.4	18.9	1.3	7.9	7.3	53.3	3.6	29.0
East North Central	8.2	7.5	59.0	4.0	3.3	24.2	1.7	14.3	13.1	96.4	6.5	58.2
West North Central	1.2	1.0	7.7	2.4	2.2	16.4	1.2	3.3	3.0	21.5	1.5	9.5
South Atlantic	0.3	0.3	2.5	0.1	0.1	1.0		2.1	1.9	13.9	1.0	11.3
East South Central	0.8	0.7	6.3	2.7	2.5	18.9	1.3	3.9	3.6	26.1	1.8	9.4
West South Central	2.8	2.6	20.7	67.5	61.4	454.0	30.7	39.1	35.7	262.6	17.8	115.8
Pacific	2.4	2.2	17.4	9.8	9.0	66.3	4.4	13.9	12.7	93.8	6.3	38.5
Subtotal	24.9	21.2	174.6	89.4	80.9	599.6	40.5	84.6	77.1	567.6	38.6	271.8
Total	2071.2											

[a] Fuels are divided for each region in proportion to the total amounts consumed in that region.

[b] Amount of fuel utilized at each process temperature allocated in proportion to the entire U.S. industrial utilization.

[c] Burned from catalyst in cracking plants.

Table 2.21. Electricity and shaft power consumption in the petroleum industry

[GWh/year (1974)]

Region	Total power consumption	Purchased electricity[a]	In-plant generation (nonsteam)			In-plant generation (steam)[c]	
			Gas turbines		Others (shaft)	Electricity[b]	Shaft
			Electricity[b]	Shaft			
Middle Atlantic	2,650	1,853	12	63	38	152	532
East North Central	4,323	3,450	11	60	36	266	500
West North Central	881	729		9		64	79
South Atlantic	46	431					46
East South Central	813	605	2	6		45	155
West South Central	10,730	7,800	65	288	137	585	1,915
Pacific	5,125	4,765	7	12	10	229	102
Totals	24,568	19,633	97	438	221	1,341	3,329

[a] Data from Bureau of Mines.

[b] Generation of electricity. Data from Department of Commerce, *Oil and Gas Journal* and *AACE Bulletin.*

[c] Substantially no "condensing" operation (estimated 5%); in the "noncondensing" operations, most of the steam is extracted at 0.69 1.38 MPa (100–200 psig) (estimated 55%).

2.2.3 Pulp and paper industry

The pulp and paper industry generally requires two process steam pressures. In an integrated mill, the pulping operation requires temperatures of 180 to 200° C (350 to 390° F). These temperatures establish the need for a high-pressure saturated steam line operating at pressures of 1.03 to 1.2 MPa (150 to 175 psig). Drying temperatures in the paper machine, however, are lower; they are typically about 120 to 150° C (250 to 300° F). This requires a second process steam line operating at pressures of 0.35 to 0.55 MPa (50 to 80 psig). Overall steam usage at the lower-pressure level is typically two to three times greater than that at the high-pressure level.

Process steam use in the industry was determined from the energy consumption data base shown in Table 2.23. The listed steam energies include contributions from both purchased and self-generated fuels. The data base fuel energies were reduced by a British thermal unit equivalent of the listed in-plant electricity and also by the natural gas and residual oil used for direct-fire applications. A boiler efficiency of 85% was applied to obtain the steam contribution from purchased fuels. Boiler efficiencies of 70% were used for bark and 60% for spent liquor to determine the self-generated contributions.

While the process steam is used at relatively low pressures, much of the steam is produced at higher pressures for combined-cycle mechanical or electrical power generation. The distribution of steam-raising capacity as a function of boiler pressure is shown in Table 2.24. These data were established by a survey of the boiler information contained in Lockwood's[6] and Post's[7] directories.

The pulp and paper industry's consumption of electrical power per ton of product has increased over the years, and the use of higher boiler pressures (greater in-plant generation) has followed this trend. Information from the American Boiler Manufacturer's Association is presented in Table 2.25, showing new boiler installations in the pulp and paper industry during the period 1965 through 1975. The largest capacity addition occurred in the 5.8 to 8.6 MPa(g) (850 to 1250 psig) range, and significant capacity was added at pressures greater than 8.6 MPa(g) (1250 psig). Comparison of the new

Table 2.22. Energy consumption by petroleum refineries (1974)

Type of fuel	Consumption [PJ (10^{12} Btu)]	Cumulative Subtotal
Steam, kPa (psi)		
Direct firing		
<1370 (<200)		
1370–3425 (200–500)	97.4 (92.2)	
3425–6850 (500–1000)	370.1 (350.5)	
>6850 (>1000)	26.5 (25.1)	
Total	494.0 (467.8)	
		494.0 (467.8)
Heat recovery boilers		
<685 (<100)		
685–1370	43.5 (41.2)	
1370–3425	72.6 (68.7)	
3425–6850	47.5 (45.0)	
>6850		
Total	163.6 (154.9)	
		657.6 (622.7)
Process and other (includes purchased)		
<685	308.9 (292.5)	
685–1370	131.2 (124.2)	
1370–3425	282.3 (267.3)	
Total	722.4 (684.0)	
		1379.9 (1306.7)
Electrical generation, kWh/year		
Nonsteam (756 × 10^6)[a,b]	8.0 (7.9)	
Steam (4,670 × 10^6)[a]	49.3 (46.7)	
Purchased (19,633 × 10^6)	259.4 (245.7)	
Total (24,568 × 10^6)	316.7 (299.94)	
	Total	1696.6 (1606.6)

[a] In-plant.
[b] Such as gas turbines.

installation capacity with the present total capacity indicates that about one-half of the present capacity was installed during the past ten years.

If the British thermal unit equivalent of the listed in-plant electricity is retained in the data calculations, resultant steam energies are representative of the as-generated boiler conditions (at boiler pressure rather than at process-use pressure). These steam generation energies are shown in Table 2.26, categorized by major fuel used for steam raising.

Amounts of fuel energies consumed for steam generation are summarized in Table 2.27. These figures are the data base values reduced only by the fuels used for direct-fire applications. (The British thermal unit equivalent of in-plant electricity is retained, and boiler efficiencies are not applied.) Relative use of the different purchased and self-generated fuels varies from region to region, depending on fuel availability and costs and the dominant type of papermaking process. On a national basis, the fuel value of the self-generated fuels represents ~50% of the total energy used for steam raising.

Table 2.23. Process steam use in the pulp and paper industry

(1975)

Region	Range of total energy consumed (PJ/year)	Category total (PJ/year)	Regional percent	United States percent	Mill average	
					Energy (PJ/year)	Steam (10^3 kg/h)
New England	<1	12.6	16.0	1.1	0.29	15.9
	1–10	32.7	40.0	2.8	2.4	136.4
	>10	35.9	44.0	3.1	6.6	372.7
Total		81.3	100.0	7.0	9.29	525.0
Middle Atlantic	<1	13.7	22.0	1.2	0.27	15.5
	1–10	48.6	78.0	4.2	2.0	113.6
	>10					
Total		62.3	100.0	5.4	2.27	129.1
East North Central	<1	16.9	13.0	1.5	0.32	17.7
	1–10	98.2	74.0	8.5	1.8	100
	>10	16.9	13.0	1.5	6.3	354.5
Total		137.3	100.0	11.5	8.42	472.2
West North Central	<1	1.2	6.0	0.1	0.43	24
	1–10	18.0	94.0	1.6	2.2	122.7
	>10					
Total		19.0	100.0	1.7	2.63	146.7
South Atlantic	<1	2.3	0.7	0.2	0.19	10.5
	1–10	60.2	17.3	5.2	3.2	177.3
	>10	285.1	82.0	25.0	9.7	545.5
Total		348.5	100.0	30.4	13.09	733.3
East South Central	<1	3.2	2.0	0.3	0.30	16.4
	1–10	73.9	46.0	6.4	3.5	195.5
	>10	84.5	52.0	7.3	9.0	500
Total		158.4	100.0	14.0	12.8	711.9
West South Central	<1	1.3	0.7	0.1	0.48	26.4
	1–10	43.3	25.3	3.8	3.0	163.6
	>10	126.7	74.0	11.0	7.9	441
Total		169.0	100.0	14.9	11.38	631
West	<1	5.7	3.0	0.5	0.25	14.1
	1–10	105.6	59.0	9.1	2.9	159
	>10	67.6	38.0	5.9	8.5	472.7
Total		179.5	100.0	15.5	11.65	645.8
United States	<1	57.0	5.0	5.0	2.53	140.5
	1–10	485.8	42.0	42.0	21.0	1168.1
	>10	512.5	53.0	53.0	48.0	2686.4
Total		1055.3	100.0	100.0	71.53	3995.0

Table 2.24. **Total boiler capacity in the pulp and paper industry (for 1975)**

(10^9 kg/year)

Region	Pressure [kPa (psig)]								Total
	<1030 (<150)	1030–1710 (150–250)	1710–2400 (250–350)	2400–3080 (350–450)	3080–3150 (450–600)	3150–5820 (600–850)	5820–8560 (850–1250)	>8560 (>1250)	
New England	8.2	6.8	0.3	2.1	4	19	11	13	64.4
Middle Atlantic	10.9	14	0.7	14	1.3	24	11		75.9
East North Central	18.2	16	8.6	11	15	41	4.1	8.6	122.5
West North Central	0.9	4.1				11		7.3	23.3
South Atlantic	7.3	1.9	2.7	1.7	18	73	72	4.2	180.8
East South Central	1.5	11		4.3	2	28	43		89.8
West South Central	3.9	4	4.1		8.2	25	40	13	98.2
West	5	27	12	5.9	11	50	5.5		116.4
United States	55.9	84.8	28.4	39	59.5	271	186.6	46.1	771.3

Table 2.25. **New boiler installations in the pulp and paper industry**

(1965–1975)

Operating pressure [kPa (psig)]	Number of boilers	Capacity (10^9 kg/year)
<1030 (<150)	43	6.8
1030–1710 (150–250)	173	40
1710–2400 (250–350)	60	18
2400–3080 (350–450)	51	31
3080–4110 (450–600)	33	26
4110–5820 (600–850)	117	96
5820–8560 (850–1250)	90	123
>8560 (>1250)	26	46
Total	593	387

Table 2.26. Steam generation in 1975 by the pulp and paper industry

(PJ/year)

Region	Range of total energy consumed	Major fuel: Coal	Residual oil	Distributed oil	Natural gas	Bark and hogged wood	Spent liquor	All fuels
New England	<1	0.23	13.0	0.07		0.004		13.3
	1–10		31.0	0.26		1.8	3.5	36.6
	>10		28.0	0.06		2.4	13.0	44.0
Total		0.23	72.0	0.39		4.2	16.0	93.9
Middle Atlantic	<1	2.3	11.0	0.003	0.31		0.13	14.0
	1–10	16.0	26.0	1.6	1.4	1.4	7.7	53.0
	>10							
Total		18.0	37.0	1.6	1.7	1.4	7.8	67.0
East North Central	<1	6.3	2.6	1.3	6.5	0.03		17.0
	1–10	52.0	9.7	0.97	35.0	2.3	9.6	110.0
	>10	8.0		1.1	6.5	0.52	5.3	21.0
Total		67.0	12.0	3.4	48.0	2.9	15.0	148.0
West North Central	<1		0.24		0.96			1.2
	1–10	2.9	2.7	0.05	10.0	1.5	3.2	20.0
	>10							
Total		2.9	3.0	0.05	11.0	1.5	3.2	21.0
South Atlantic	<1	0.72	1.3	0.03	0.28			2.4
	1–10	12.0	21.0	0.72	7.0	5.6	20.0	66.0
	>10	39.0	110.0	1.6	21.0	32.0	130.0	330.0
Total		52.0	130.0	2.3	29.0	37.0	150.0	398.0
East South Central	<1		2.2	0.21	0.69			3.0
	1–10	6.4	13.0	1.1	21.0	7.1	33.0	81.0
	>10	11.0	16.0	0.51	16.0	9.5	42.0	96.0
Total		18.0	32.0	1.9	37.0	17.0	75.0	180.0
West South Central	<1				0.61			1.2
	1–10		4.9	1.1	15.0	3.5	19.0	43.0
	>10		26.0	0.46	47.0	15.0	52.0	140.0
Total			31.0	1.6	62.0	19.0	71.0	184.0
West	<1		0.43	0.22	3.1	0.67		5.5
	1–10		21.0	1.6	26.0	12.0	48.0	110.0
	>10		11.0		18.0	5.5	34.0	66.0
Total			32.0	1.8	47.0	18.0	82.0	182.0
United States	<1	9.5	30.0	1.9	12.0	0.70	0.13	47.0
	1–10	90.0	130.0	7.5	120.0	35.0	140.0	520.0
	>10	58.0	190.0	3.7	110.0	65.0	270.0	700.0
Total		158.0	350.0	13.0	242.0	101.0	410.0	1267.0

Table 2.27. Energy consumed in 1975 for steam generation by the pulp and paper industry

(PJ/year)

Region	Range of total energy consumed	Major fuel						All fuels
		Coal	Residual oil	Distributed oil	Natural gas	Bark and hogged wood	Spent liquor	
New England	<1	0.27	15.0	0.08				16.0
	1–10		36.0	0.30		2.5	5.8	45.0
	>10		33.0	0.07		3.4	21.0	58.0
Total		0.27	85.0	0.46		5.9	27.0	120.0
Middle Atlantic	<1	2.7	13.0		0.36		0.21	16.0
	1–10	18.0	30.0	1.9	1.6	2.0	13.0	67.0
	>10							
Total		21.0	43.0	1.9	2.0	2.0	13.0	83.0
East North Central	<1	7.4	3.1	1.6	7.7	0.4		20.0
	1–10	62.0	11.0	1.1	42.0	3.0	16.0	140.0
	>10	9.4		1.3	7.6	0.72	8.9	28.0
Total		78.0	14.0	4.0	57.0	3.8	25.0	180.0
West North Central	<1		0.28		1.1			1.4
	1–10	3.4	3.2	0.05	12.0	2.1	5.3	26.0
	>10							
Total		3.4	3.5	0.05	13.0	2.1	5.3	27.0
South Atlantic	<1	0.84	1.6	0.04	0.32			2.8
	1–10	14.0	24.0	0.85	8.2	7.6	34.0	89.0
	>10	46.0	130.0	1.9	25.0	42.0	200.0	460.0
Total		61.0	160.0	2.7	34.0	50.0	250.0	550.0
East South Central	<1		2.5	0.24	0.81			3.5
	1–10	7.5	15.0	1.3	24.0	10.0	55.0	110.0
	>10	14.0	19.0	0.60	19.0	13.0	70.0	140.0
Total		21.0	37.0	2.2	44.0	23.0	120.0	250.0
West South Central	<1							
	1–10		5.8	1.3	18.0	4.6	31.0	61.0
	>10		31.0	0.54	55.0	21.0	86.0	200.0
Total			37.0	1.8	73.0	25.0	120.0	260.0
West	<1		0.59	0.25	3.7	0.94		6.6
	1–10		25.0	1.9	31.0	17.0	80.0	160.0
	>10		12.0		21.0	7.9	57.0	95.0
Total			38.0	2.2	56.0	26.0	140.0	260.0
USA	<1	11.0	36.0	2.2	15.0	0.98	0.21	68.0
	1–10	110.0	150.0	8.8	140.0	49.0	240.0	690.0
	>10	69.0	220.0	4.4	130.0	88.0	460.0	970.0
Total		190.0	410.0	15.0	280.0	140.0	700.0	1700.0

Natural gas is used in the industry for boiler fuel, direct-fire hot air dryers, and space heating. The relative use of natural gas was established by an API survey in mid-1975 covering gas use during the period November 1974 through March 1975. Results of the survey pertaining to gas consumption are shown in Table 2.28. While these results are based on wintertime use only, the data are believed to be the best available and representative of the present average consumption patterns for natural gas.

Direct-fire use of residual oil occurs primarily in lime kilns as part of the recovery process for pulping chemicals. Energy consumption for this use was estimated to be directly proportional to the spent-liquor self-generated energy listed in the data base. Lime kiln operation typically requires about 2 GJ (2×10^6 Btu) of fuel energy per ton of paper product. Because the corresponding spent-liquor energy generation is about 20 J/kg (20×10^6 Btu/ton), the overall direct-fire residual oil consumption was estimated at 10% of the spent-liquor value. Results are shown in Table 2.29.

Extensive low-temperature air ventilation is used throughout the mills to control humidity, to cool operating equipment, and to maintain satisfactory working conditions. Some paper drying is accomplished using hot air at temperatures of about 320° C (600° F). Exit air temperatures can be as high as 230° C (450° F); but economizers are usually employed for air preheating, and final exhaust temperatures are less than 150° C (300° F).

Table 2.28. Natural gas use in the pulp and paper industry from November 1974 to March 1975

Region	Natural gas use (%)		
	Boiler fuel	Direct fire	Space heating
New England		99	1
Middle Atlantic	31	68	1
East North Central	86	13	1
West North Central	67	33	Negligible
South Atlantic	76	24	Negligible
East South Central	75	21	4
West South Central	78	21	1
West	75	19	6
United States	76	22	2

Table 2.29. Direct-fire residual oil use in the pulp and paper industry

(1975)

Region	Residual oil use (PJ/year)
New England	2.9
Middle Atlantic	1.4
East North Central	2.6
West North Central	0.6
South Atlantic	26
East South Central	13
West South Central	13
West	15
United States	75.0

Most paper drying is performed on steam-heated cylinders. The water evaporated from the paper (typically 0.45 kg of water vapor diluted by 4.55 kg of air) discharges from the dryer section at temperatures of 95 to 150° C (200 to 300° F). This waste heat, however, is partially recovered through the use of air economizers and water heaters, and the final exhaust temperature is significantly reduced. Generally, most waste-heat sources in the mill are relatively low temperature and are best utilized by heat recovery through air or water heat exchangers.

Energy conversion factors used in this section (Sect. 2.2.3) are shown in Table 2.30.

Table 2.30. Energy conversion factors suggested by the American Paper Institute

Type of fuel	Conversion factor [GJ (10^6 Btu)]
Purchased fuels	
Coal, 1 short ton	27.7 (26.2)
Residual oil (#5,6), 1 bbl	6.6 (6.3)
Distributed oil (#2), 1 bbl	6.3 (6.0)
Natural gas, 10^3 ft^3	1.09 (1.03)
Other	
Steam (no condensate return), 10^3 lb	1.4 (1.3)
Steam (condensate return), 10^3 lb	1.2 (1.1)
Liquid propane gas, 1 gal	0.097 (0.092)
Self-generated fuels	
Bark (50% moisture), 1 short ton	9.5 (9.0)
Hogged wood (50% moisture), 1 short ton	8.1 (7.7)
Spent liquor (oven dry basis), 1 short ton (solids)	14.2 (13.4)[a]
Electricity, 10^3 kWh	
Purchased and self-generated Hydroelectric power	3.6 (3.4)

[a] 6700 Btu/lb.

2.3 Hydrogen

Today the largest industrial application of hydrogen is as a chemical feedstock for production of several important chemicals.[8] As shown in Table 2.31, ~95% of all hydrogen consumed for industrial purposes is consumed as chemical feedstock. The bulk of chemical hydrogen consumption is in one of three main industries: (1) ammonia manufacture, (2) petroleum refining and petrochemicals manufacture, and (3) methanol production. Minor quantities of hydrogen are also consumed in metals treating, edible oil production, and other applications.

Technically, there are three general grades of hydrogen: (1) pure hydrogen (99.8%), (2) hydrogen that is 99.8% to ultrapure, and (3) hydrogen with less than 10 ppm impurities (ultrapure).

Appendix B contains descriptions of the methods of hydrogen production in common use today.

Table 2.31. Recent U.S. hydrogen consumption

(10^9 scm)

End use	1968	1971	1972	1973
Ammonia	24.5	28.3	30.7	31.3
Refinery operations	22.2	37.8	38.8	40.6
Methanol	4.11	5.41	6.4	8.58
Hydrogenation of oils	0.31	0.34	0.40	0.42
Miscellaneous[a]	7.25	4.93	5.27	5.69
Total	58.3	76.8	81.6	86.7

[a] Electronics and electrical, space, other industrial, and other chemicals; for example, hydrogen peroxide, alcohols.

Source: H. Kelley and E. A. Tasimann, *Hydrogen Tomorrow...Demands and Technology Requirements,* report of the Hydrogen Energy Systems Technology (HEST) Study for NASA, Jet Propulsion Laboratory Report JPL 5040-1, December 1975.

2.3.1 Economic aspects

The manufacture of hydrogen is largely an integral part of two other chemical manufacturing processes, the production of ammonia and methanol. In these operations, hydrogen is manufactured and used as a mixture with other gases, nitrogen in the case of ammonia synthesis and carbon monoxide in the case of methanol synthesis. Nevertheless, the manufacturing processes are essentially identical, except that the gas mixtures must be purified to produce pure (95 to 100%) hydrogen.

The steam reforming process is by far the most important hydrogen manufacturing process; next in importance is the hydrocarbon partial oxidation process. Natural gas, liquefied petroleum gases, and propane may all be used as feedstocks, according to availability and cost. An important advantage of the partial oxidation process lies in its versatility with respect to feedstocks, particularly its adaptability to fuel oil.

Of the minor hydrogen manufacturing processes, ammonia dissociation (cracking) is most widely used. Several companies furnish push-button plants for producing the nitrogen-hydrogen mixture that results; the nitrogen content is not detrimental in such applications as annealing of steel, powder metallurgy, and atomic-hydrogen welding.

Large quantities of hydrogen are produced as a by-product of catalytic reforming of petroleum stocks. Some of this hydrogen is used in the refinery for hydrotreating, some for hydrocracking processes, and some for ammonia manufacture. Considerable amounts are flared or burned for fuel.

Other uses include hydrogenation of edible oils to produce shortenings and margarines, many of which are low in cholesterol; hydrogen is also used in large electric generators to reduce windage losses and to transport heat. Liquefied hydrogen is used as a rocket fuel; it has also become the commercial source of the ultrapure hydrogen required in some segments of the electronics industry.

2.3.2 Production

Because hydrogen is a by-product of so many chemical processes, statistics on its production and use are difficult to interpret; hydrogen production is sometimes reported, sometimes not, depending on whether it is (1) used as produced or isolated for sale or (2) wasted or burned as fuel.

The reporting basis for hydrogen production has changed several times. Beginning in 1960, the reported production excluded hydrogen used on-site for manufacture of ammonia and of methanol, that produced by ammonia dissociation, and that vented or used as fuel. Since 1970, reports also exclude hydrogen produced by petroleum refineries for their own use.

2.3.3 Price

The price of hydrogen has been fairly constant until the recent escalation in natural gas costs. Hydrogen is usually sold on a contract basis, and prices can vary as much as 1000% depending on volume, location, contract length, competitive situation, and feedstock availability. Cylinder price is as much as ten times bulk price.

Reference 9 contains a comparison of three studies made to estimate the cost of transporting hydrogen and natural gas via pipeline. Generally, where pumping costs are equal, hydrogen transmission costs about 1.5 times as much as natural gas transmission (per unit distance and energy).

2.3.4 Major uses of hydrogen as a feedstock

As a feedstock, hydrogen is used in three major sectors: (1) production of ammonia, (2) production of methanol, and (3) petrochemical production and petroleum refining. These processes are discussed briefly in the following sections.

Ammonia production. As shown in Table 2.31, ammonia production represents the largest hydrogen consumption in the industrial sector today (excluding the petroleum industry, which produces hydrogen to meet its requirements). In this sector, hydrogen is used primarily for the production of fertilizer-based products. It is also used as a feedstock for some products of the fiber industry and other chemical processing industries.

Ammonia is produced by the Haber-Bosch process developed in 1913. In this process, a mixture of nitrogen and hydrogen in a 1:3 ratio is passed over an iron-oxide promoted catalyst to produce ammonia by the reaction $3H_2 + N_2 \rightarrow 2NH_3$ (85% yield). The reaction is exothermic, and typical reaction conditions range from 640 to 700 K (700 to 800° F) and from 21 to 35 MPa (3000 to 5000 psig).

Figure 2.3 is a simplified schematic of the ammonia synthesis process. In this process, the synthesis gas (H_2–N_2 mixture) is mixed with a recycle stream and subsequently passes to the ammonia converter. Reactor effluent is then cooled, and ammonia product is recovered. The balance of the reactor effluent is recycled to mix with the fresh feed, although a portion of the effluent is purged from the synthesis loop to remove any inerts (e.g., methane) that may have been introduced in the synthesis gas.

The consumption of ammonia can be broken down as follows:

	(%)
Fertilizer and feeds	76
Fiber and plastics intermediates	9
Explosives	4
Other	11
	100

The largest percentage is tied directly to agriculture, where in one form or another (liquid ammonia, "nitrogen" solutions, ammonium nitrate, or urea) it is used as a fertilizer. Accordingly, the ammonia market shows many of the ups and downs that plague agriculture in general. Excess ammonia manufacturing capacity has been at times a normal condition; yet, in the peak demand season in the spring, storage tanks become empty and localized shortages occur.

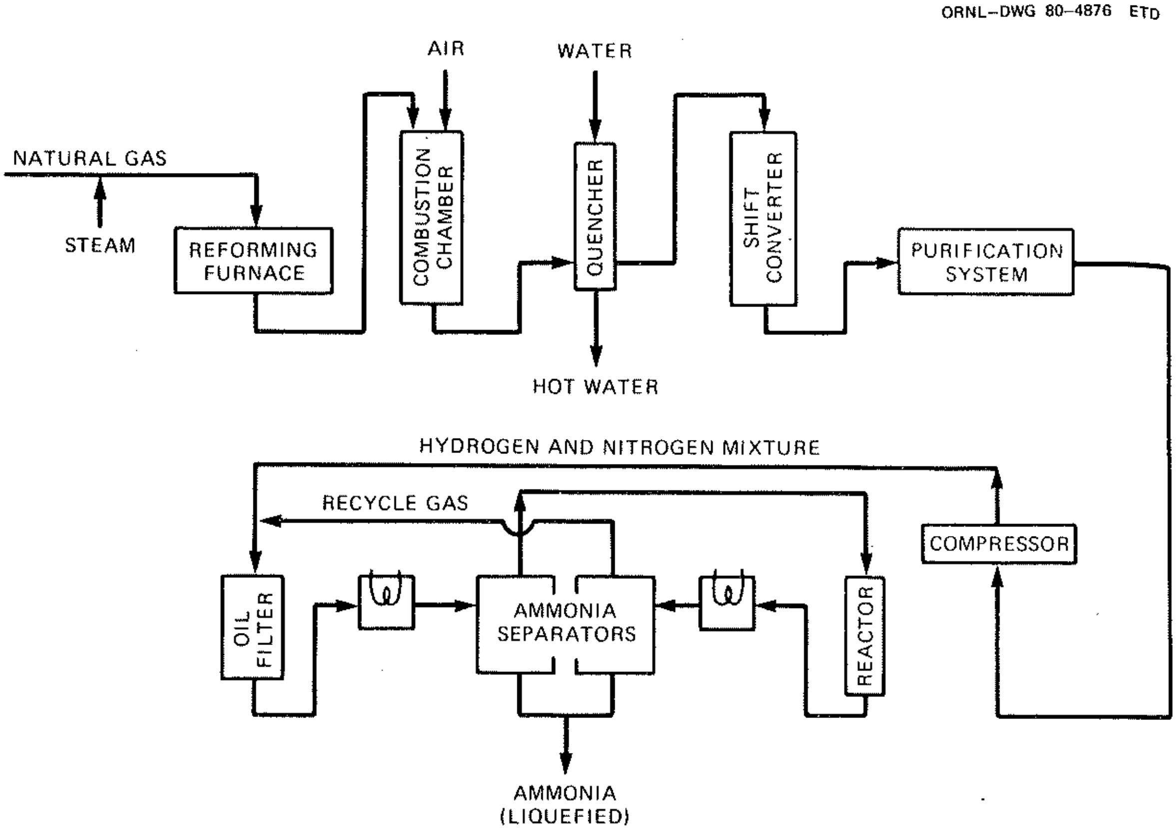

Fig. 2.3. Ammonia synthesis process.

Production and price variation data for ammonia are shown in Figs. 2.4 and 2.5, respectively. Although overshadowed by agricultural consumption, ammonia use for chemical manufacture is also extensive. The largest industrial use is in fibers, plastics, and coating resins; these uses should expand. Ammonia is also the raw material for nitric acid production. Anhydrous ammonia is also an excellent commercial refrigerant. Its high latent heat of vaporization, chemical stability, low corrosivity, and low friction losses account for its attraction in this application.

Methanol production. Methanol is synthesized by the reaction of hydrogen (H_2) and carbon monoxide (CO) under high pressures. This reaction is:

$$CO + 2H_2 \rightarrow CH_3OH \; .$$

Regardless of the source of the CO–H_2 mixture, the ratio is adjusted so that approximately the theoretical ratio is obtained (two volumes of H_2 to one volume of CO). The mixed gases are compressed in multistage compressors to pressures of 20 to 35 MPa (3000 to 5000 psi) and heated in heat exchangers by the reaction gases. The heated gases pass through a copper-lined steel converter containing a mixed catalyst of the oxides of zinc, chromium, manganese, or aluminum; for example, zinc oxide with 10% chromium oxide. The temperature of the reaction is maintained at ~300°C by proper heat removal from the exothermic reaction (24,620 cal/g-mole of methanol). The converter must be heated to initiate the reaction; but, once started, it is self-supporting. The temperature is kept constant by proper space velocity and heat interchange.

ORNL–DWG 80–4872 ETD

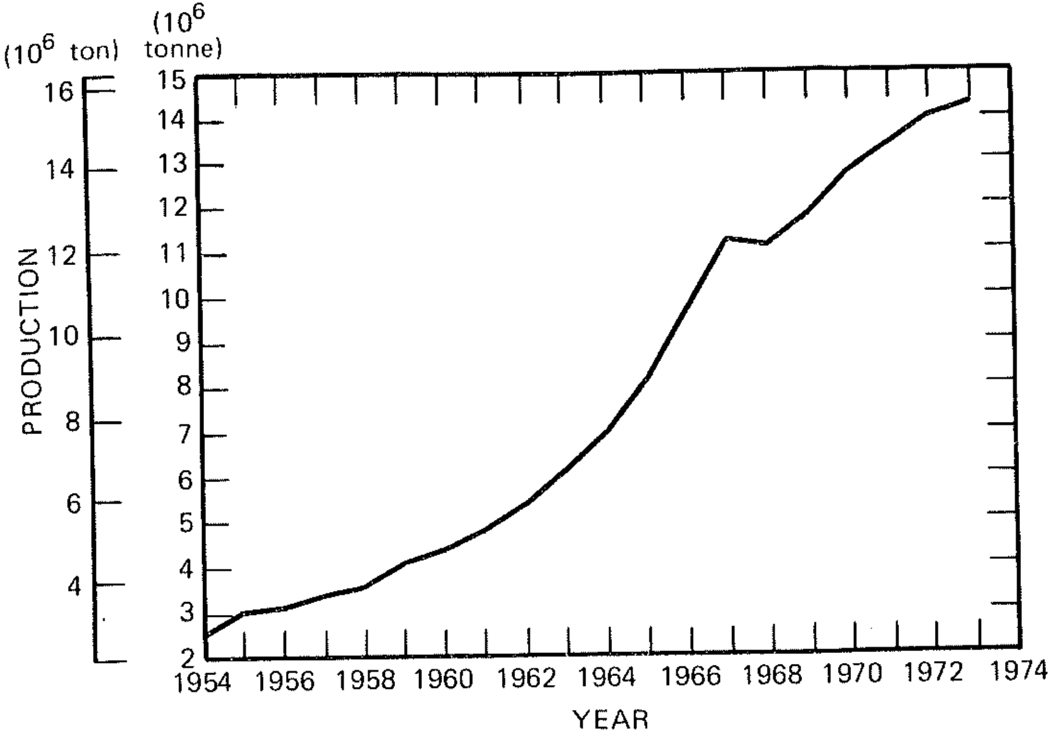

Fig. 2.4. Production—ammonia.

ORNL–DWG 80–4873 ETD

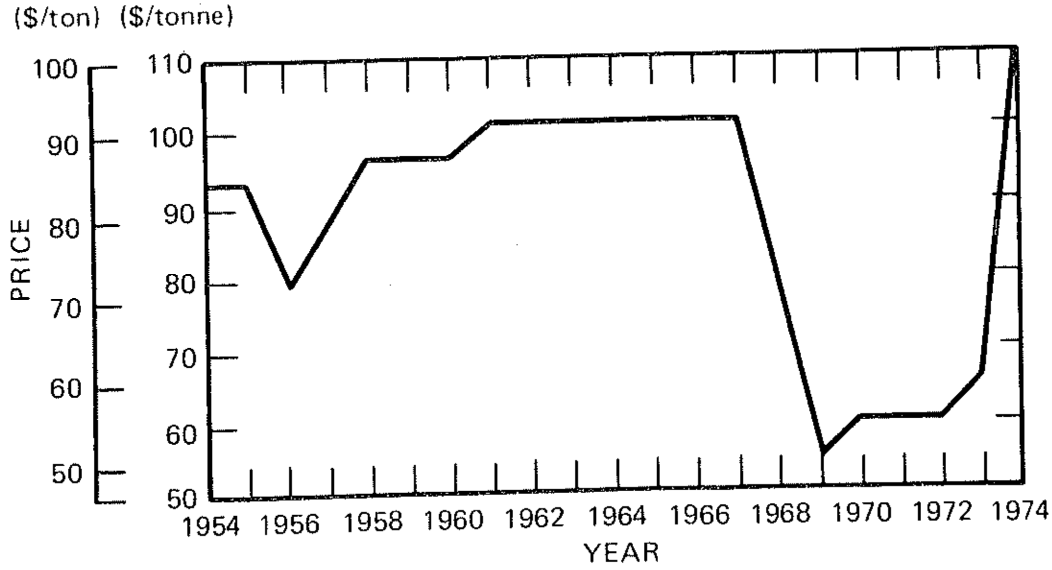

Fig. 2.5. Price—ammonia.

The methanol-containing gases leaving the reactor are cooled by the reactants in heat exchangers and condensed under full operating pressures. The pressure is released, and the cool liquid methanol (0 to 20° C) is run off; it may be further purified by distillation. Residual gases are returned to the system for reprocessing. Possible accumulation of inert gases is eliminated by the purging of part of the recycle gases. Using this process, methanol of 99% purity is obtained, and the equilibrium yield is better than 60%. By-products are 1 to 2% dimethyl ether and 0.3 to 0.5% higher alcohols. Useful catalyst life is 9 to 12 months.

A few plants use CO_2 instead of CO. The methanol reaction in this case is

$$CO_2 + 3H_2 \rightarrow CH_3OH + H_2O\,.$$

In the CO_2–H_2 process, the crude methanol product containing about 25% water is sent to a continuous refining column, where an overhead cut, consisting of ethers and some methanol, is removed. The bottoms are sent to a second column, where methanol is removed overhead and water is removed as bottoms. A side cut, which is removed from near the bottom of the column, consists of 30% methanol, 50% higher alcohols including ethyl alcohol, and 20% water.

Most of the synthesis gas plants in operation until a few years ago were based on high-pressure synthesis using electrically powered reciprocating compressors. The most recent advance in methanol production processes is toward high-volume plants utilizing large steam-driven compressors operating at lower pressures. In this low-pressure process, a copper-based catalyst is used with pressures of 5 to 10 MPa (50 to 100 atm). The high activity of the catalyst allows the reaction to take place at temperatures between 250 and 270° C. This new process is more efficient because of lower by-product formation, lower energy costs, and lower investment and maintenance requirements. End use products of methanol are shown in Table 2.32; production and price variations are shown in Figs. 2.6 and 2.7, respectively.

Table 2.32. Methanol use pattern

End use	Percent
Formaldehyde	45
Dimethyl terephthalate	10
Methyl methacrylate	8
Methylamines	4
Methyl halides	4
Acetic acid	4
Solvents	10
Miscellaneous and exports	15
Total	100

If the methanol-based fuel cell were to become an economic reality, a multiple increase in production would be required. Replacement of 2.5% of existing diesel engines burning distillate with methanol-based fuel cells probably would double U.S. consumption of methanol.

Synthetic methanol plants vary in size from 75×10^6 to 950×10^6 L annual capacity. More than 85% of recent world expansion in methanol production is based on Imperial Chemical Industry's low-pressure technology. Capital investment costs for new plants using this process range from \$14 million (plant capacity 360 metric tons/d) to \$23 million (1170 metric tons/d).

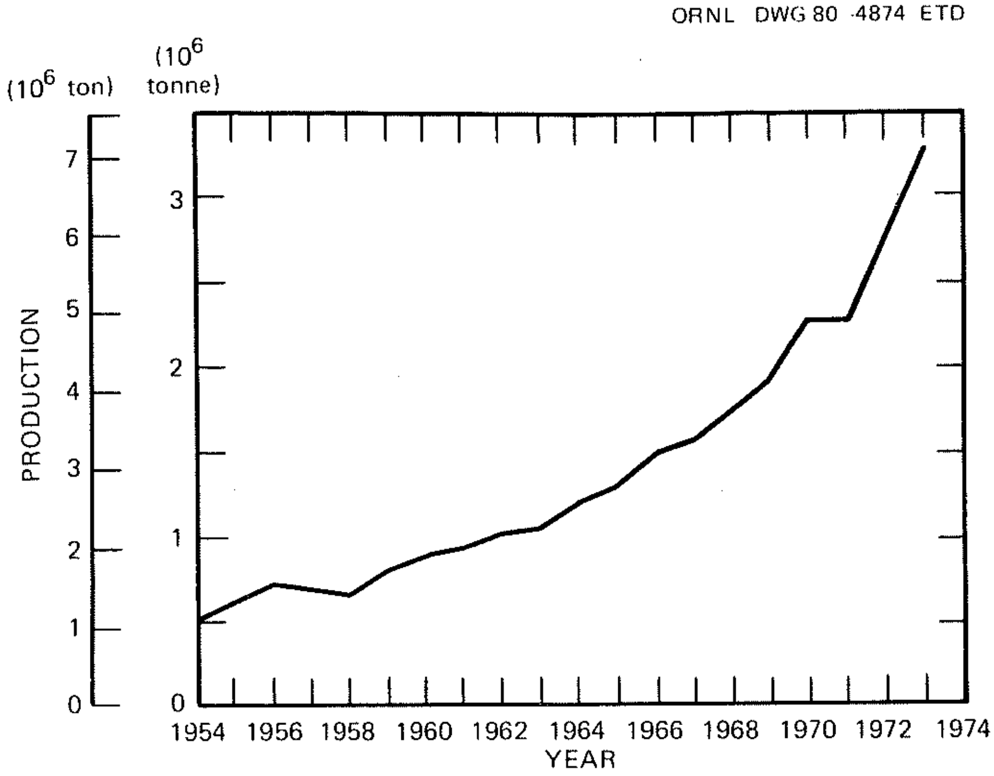

Fig. 2.6. Production—methanol.

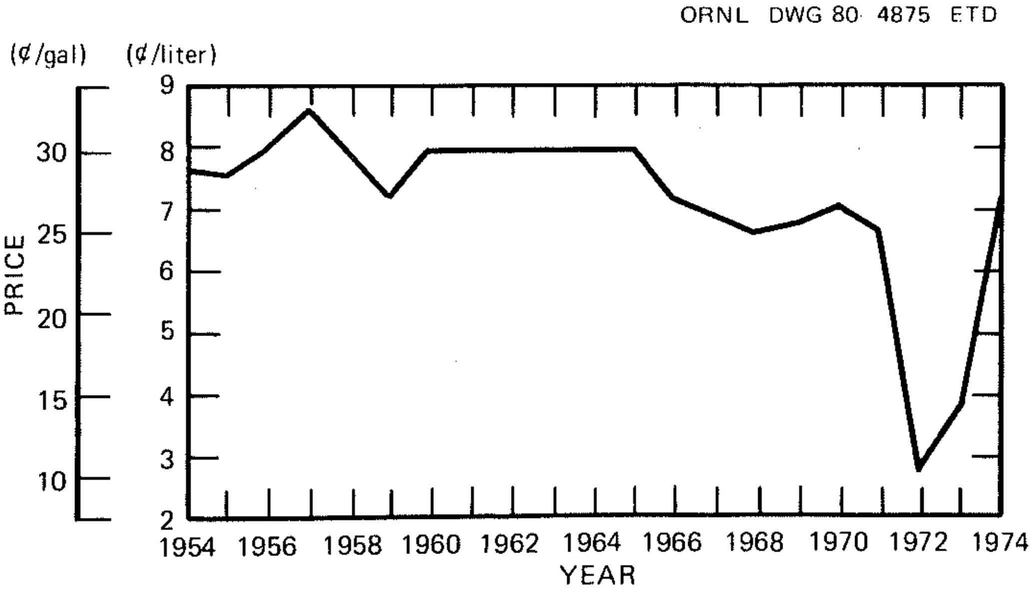

Fig. 2.7. Price—methanol.

Petroleum refining. Hydrogen is consumed in the petroleum refining industry in several processing options. The total hydrogen consumed by these operations makes the petroleum refining industry the largest consumer of hydrogen today.

In the petroleum refining industry, hydrogen is used primarily to achieve one of three processing goals:

1. removal of sulfur from refinery products;
2. conversion of heavy oil fractions to lighter, more valuable oil fractions; and
3. chemical conversion.

The refining industry uses hydrogen in the (1) hydrotreatment process, (2) hydrocracking process, (3) hydroalkylation process, and (4) other miscellaneous applications. A description of each follows.

Hydrotreating process. Hydrotreating processes exist for all petroleum fractions within a refinery. In all cases, a petroleum feedstock is combined with hydrogen and passed over a catalyst at elevated temperature and pressure. The extent of hydrotreating depends on feedstock type, reaction severity, and catalyst.

The primary goal of hydrotreating is usually one of hydrodesulfurization. Hydrogen reacts with the petroleum fraction and with organic (R, R', R" in the following equation) sulfur compounds to produce hydrogen sulfide:

$$H_2 + HS(R) + R'SR''_2 \rightarrow 2H_2S + RH + R'H + R''H \, .$$

The gaseous H_2S is removed from the light hydrocarbon gases present during hydrodesulfurization by a suitable solvent solution (e.g., monoethanolamine). Similarly, hydrogen can react with nitrogen-containing compounds to form ammonia and with O_2 to form H_2, which are subsequently removed. For sufficiently severe hydrotreating conditions, hydrogenation of unsaturated hydrocarbons may also occur.

Low-severity operations are usually associated with hydrotreating of naphtha fractions. In such operations, hydrogen consumption may range from 1.4 to 5.7 scm (50 to 200 scf) H_2 per barrel of feed. Besides deslfurization, such treatment provides improvement in color and stability. High-severity operations are associated with hydrodesulfurization of heavy petroleum fractions high in sulfur content (e.g., hydrodesulfurization of vacuum gas oil and atmospheric residuals) and may result in a hydrogen consumption in the range of 17 to 28 scm (600 to 1000 scf) per barrel of feed.

Hydrotreating in refinery operations can be used to improve petroleum fraction qualities prior to further processing or to improve the qualities of a finished product. Processes for which hydrotreating is used to improve feedstock quality include catalytic reforming and catalytic cracking. Hydrotreating of catalytic reformer feedstock removes sulfur, nitrogen, and metal compounds that could subsequently poison a reforming catalyst. Similar benefits occur in treating catalytic cracking feedstocks. Hydrotreating of finished products is used to sweeten these products (i.e., remove sulfur compounds), improve stability (by hydrogenating certain unsaturates), and improve color properties of the products. Hydrotreating is also growing in use as more and more olefins plants are built. Olefins plants, which produce mainly ethylene and propylene, will also produce other liquid products that may contain acetylene or diolefins material. During storage, these compounds tend to polymerize and form sludge. Hydrotreating is used to hydrogenate such compounds, converting them to more stable paraffinics.

Hydrocracking process. Hydrocracking operations within a refinery are used to convert heavier, less profitable liquid fractions into lighter, more profitable fractions. An example of this process is the conversion of vacuum gas oils and vacuum residuum into lighter distillate fractions. Hydrocracking is basically a higher-severity version of hydrotreating. In such reactions, pressures may range from 10 to 20 MPa (1500 to 3000 psig), as compared with the lower pressures of hydrotreating.

In hydrocracking, hydrogen is used to saturate the cracked products produced under the severe operating conditions of the hydrocracking operations. In addition to cracking and saturation of hydrocarbons, the hydrogen simultaneously desulfurizes the hydrocarbon feedstock. Hydrocracking is thereby able to produce a low-sulfur, lighter product from heavy hydrocarbon feeds.

Hydrodealkylation process. Hydrodealkylation is the selective hydrogenation of aromatic compounds that leaves the benzyl or other aromatic ring structure unaltered. High-purity benzene, toluene, and naphthalene are produced in this way. For the production of benzene, a feed (toluene, xylenes, C_9, and aromatics) together with a hydrogen-containing stream are heated to the reaction temperature and passed over a dealkylation catalyst. Benzene and heavier aromatics are condensed, and lighter vapors may be purified and/or recycled. Benzene is purified in a benzene fractionator to produce the desired specification benzene. Other hydrogen-consuming chemical conversions are the saturation of aromatics to produce high-quality jet fuels or to produce high-purity hexane from benzene.

Miscellaneous applications. Application of hydrogen is found in many other industries, from food processing to electronics to glass production. Consumption in these industries, however, is low in comparison with the previously discussed industries.

Unsaturated organic oils of soybeans, fish, cotton seeds, corn, peanuts, and coconuts are hydrogenated, stabilizing these oils and improving odor, color, and consistency. Inedible oils and greases are hydrogenated for the production of soap and animal feed. Lubricants for use in food-processing equipment are produced by hydrogenation and purification of various greases to rigid specifications.

Much chemical synthesis not directly related to petrochemicals requires hydrogen. Hydrogen peroxide, hydrogen chloride gas, uranium hydride, and many organic chemicals all use hydrogen. Additionally, the pharmaceutical industry uses hydrogen extensively to manufacture certain drugs.

Hydrogen is consumed in oxyhydrogen glass cutting and metals welding. It is also used in working leaded glass, wherein the hydrogen content must be carefully controlled to prevent burning of the lead out of the glass.

2.3.5 Usage potential of hydrogen[8,9]

Present hydrogen usage is still insignificant when compared with its usage potential. Some of the more important processes that may be large consumers of hydrogen in the future are described in the following text. A few of these have already been developed on a commercial scale.

Direct reduction of metals. A currently small but potentially large market for future hydrogen usage is the direct reduction of metal ores to the base metal. In such processes, hydrogen or a mixture of hydrogen and carbon monoxide gas (called reducing gas) reacts with the metallic ore. Usually, the reducing gas flows upward through a falling or fluidized bed of metallic ore.

The iron industry has built several processing plants in which iron ore is reduced by H_2 or a mixture of H_2 and CO. Reaction temperatures range from 550 to 780° C (1000 to 1400° F), and pressures range from 0.2 to 1.5 MPa (30 to 220 psi). In these processes, contact between finely divided iron ore particles and the gas causes the following reactions:

$$3H_2 + Fe_2O_3 \rightarrow 2Fe + 3H_2O \,,$$
$$3CO + Fe_2O_3 \rightarrow 2Fe + 3CO_2 \,.$$

Approximately 85 to 90% of the ore can be metallized using these processes. The "sponge iron" product is subsequently briquetted for shipment and storage or used directly in steel making.

A strong incentive for direct reduction of metals has been the development of iron ore reserves in those parts of the world where coal is not readily accessible or where natural gas is relatively cheap. With cheap natural gas, a reducing gas mixture rich in hydrogen can easily be generated by the steam-methane reforming reaction.

Nuclear power for generation of electricity to be used in producing hydrogen by electrolysis may also be considered where iron ore is plentiful but carbon resources are not. Wide acceptance of this technology depends on the economics of hydrogen and on the solution to technical difficulties (agglomeration and product pyrophoricity). Hydrogen reduction for nonferrous metallurgy (copper, zinc, uranium, and lead) is equally feasible but would be less hydrogen consuming. Direct reduction processes offer several advantages such as flexibility of plant location, reduced use of fossil fuels, and reduced pollution problems.

Coal conversion. In the future, if conversion of coal to synthetic natural gas (SNG) or to synthetic petroleum liquids proves feasible, massive quantities of hydrogen will be required. An examination of the chemical nature of coal shows why. Coal has a hydrogen-to-carbon ratio of 25:30. However, SNG has a hydrogen-to-carbon ratio of 3:1, and synthetic petroleum liquids have a hydrogen-to-carbon ratio of 8:10. To achieve these ratios, hydrogen obviously must be added to the coal or carbon must be removed. Most technologies being piloted are based upon hydrogen addition.

Gasification for SNG production. Today many gasification concepts are being evaluated in pilot plants, demonstration units, and full-scale production plants. In all processes, the coal is gasified to produce a maximum amount of methane with the gasification reactor. To produce methane, reactor pressures are kept as high as possible, though they are usually limited by the ability to feed coal to and/or discharge ash from the reactor. The commercially available Lurgi process operates at ~3 MPa (30 atm), while the new processes under development operate at pressures up to 10 MPa (1500 psi). Unfortunately, no one-step gasifier exists that will convert all of the coal directly to SNG (i.e., primarily methane). Subsequent reactions are required, such as the water-gas shift reaction, CO_2 removal, and methanation. This last reaction converts CO to methane by reaction with H_2:

$$CO + 2H_2 \rightarrow CH_4 .$$

Operating conditions for this exothermic reaction are 260 to 427°C (500 to 800°F) and elevated pressures. The reaction is catalyzed by a nickel-based catalyst.

Liquefaction of coal. Developmental processes for the conversion of coal to petroleum liquids involve similar processing steps. Coal is usually mixed in an oil-based slurry, pumped to high pressures [e.g., 10 to 20 MPa (100 to 200 atm)], and heated to 400 to 480°C (750 to 900°F). At these conditions, hydrogen is passed through the coal slurry mixture to hydrogenate the coal slurry and convert coal to liquid. The reaction may or may not be catalyzed.

After reaction, the reactant mixture is separated into a gaseous fuel product, a synthetic crude oil, and the residue ash contained in the incoming coal. Separation of the ash from the petroleum product has proved difficult in earlier tests of coal liquefaction processes because separation conditions make such conventional devices as filters difficult to operate effectively. The most recent method for ash separation being investigated by DOE is the deashing technique developed by the CE Lummus Company. In this method, an antisolvent is used to effect the separation of hydrocarbon product from a stream rich in ash content. The major advantage of this process is that it requires no mechanical parts to effect separation.

Another method of coal liquefaction involves the gasification of coal first. The synthesis gas (mixture CO + H_2) can then be reacted to produce a synthetic gasoline. This is currently practiced by the SASOL Corporation of South Africa. An alternative to this method is a conversion of the synthesis gas to methanol as previously described. The methanol may then be used directly as a replacement for gasoline in cars of the future, or it may be converted to gasoline via a catalytic process being developed by the Mobil Corporation.

Oil shale liquefaction. As in conversion of coal to petroleum liquids, hydrogen can be used to upgrade the liquefied product by suitably adjusting the carbon-to-hydrogen weight ratio. Reaction conditions would be similar to those of coal liquefaction.

Tar sands upgrading. In certain parts of the world (e.g., Canada, Venezuela), large reserves of tar sands exist. These reserves are various mixtures of sand, clay, minerals, water, and crude bitumen, a substance similar to asphalt. To obtain synthetic oil from tar sands by modern techniques, the sands must first be mined. The bitumen is then separated from the sand before it undergoes a series of processing steps that convert it to a synthetic oil. A key processing step will be one similar to hydrocracking, during which the bitumen is subjected to high pressures and temperatures and reacted with hydrogen. Under these conditions the bitumen is cracked to yield a lighter product that can be sold as synthetic oil. If the mineral content of the tar sands has sufficient activity to catalyze the reaction, another catalyst may not be required.

Ethylene glycol production. Researchers, in an attempt to extend the usage of coal resources, are currently examining the potential for coal-derived chemicals. One petrochemical product, ethylene glycol, is currently derived from petroleum resources that may some day be based on hydrogen. Ethylene glycol is used in the production of man-made polyester fibers; it is also used almost exclusively in the production of automotive antifreeze products. In current technology, ethylene glycol is produced by the reaction of oxygen with ethylene to produce ethylene oxide. The oxide is hydrolyzed to form the glycol ($C_2H_6O_2$):

$$C_2H_4 + 1/2O_2 \rightarrow C_2H_4O \, ,$$
$$C_2H_4O + H_2O \rightarrow C_2H_6O_2 \, .$$

In the future, coal may be gasified to form a synthesis gas rich in CO. After undergoing a shift reaction to alter the H_2-to-CO ratio, the synthesis gas would be subjected to specific reaction conditions [e.g., pressures in excess of 50 MPa (7400 psi)] so that the following reaction occurs:

$$3H_2 + 2CO \rightarrow C_2H_6O_2 \, .$$

At this time, the reaction is still in the laboratory testing stage.

Cryogenic applications.[10] Liquid hydrogen has the second lowest boiling point of any liquid. Until recently no materials had been found that were superconducting above 20.4 K, so using liquid hydrogen to cool superconducting equipment was not possible. Recently, however, the discovery was made that niobium-germanium alloy (Nb_3Ge) superconducts at temperatures up to about 23 K. Thus, widespread hope exists that other materials will be found with even higher superconducting transition temperatures, thereby opening the way to use liquid hydrogen instead of the usual liquid helium as a coolant.

This recent discovery is very important for several reasons. First, more energy is required to liquefy helium than hydrogen. (The theoretical minimum for helium is about twice that of hydrogen.) Second, helium is a rare element mainly found in a dilute (2%) association with natural gas from a few fields. If additional and useful new materials are found that will superconduct in liquid hydrogen, important new uses for the hydrogen will be established. Important synergisms with other aspects of the hydrogen economy would be inevitable.

2.3.6 Industrial hydrogen requirements

Typical hydrogen requirements for current and potential processes are shown in Table 2.33. Table 2.34 presents an estimate of U.S. consumption in conjunction with the above requirements.

Table 2.33. Typical industrial hydrogen requirements

Use	Amount (m^3)	Product (unit)
Ammonia synthesis	2.18 2.50	NH_3 (1 kg)
Methanol synthesis	1.62 2.34	CH_3 (1 kg)
Ethylene glycol[a]	2.6	Ethylene glycol (1 kg)
Petroleum refining (average)	109	1 m^3 of refined crude oil
Hydrotreating		
Naphtha	9	Naptha (1 m^3)
Coking distillate	134	Distillate (1 m^3)
Hydrocracking	267 623	Feedstock (1 m^3)
Coal conversion[a] to		
Liquid fuel	784 1514	Synthetic oil (1 m^3)
Gaseous fuel	1.57	Synthetic gas (1 scm)
Oil shale conversion[a] to		
Liquid fuel	232	Synthetic oil (1 m^3)
Gaseous fuel	1.21	Synthetic gas (1 scm)
Liquid fuel from tar sands	303	Synthetic oil (1 m^3)
Iron ore reduction[a]	0.624	Iron (1 kg)
Hydrogenation of fats and oils	0.109	Product (1 kg)

[a] Future markets, direct-reduced iron is produced now on a subcommercial scale.

Source: J. H. Kelley and E. A. Tasimann, *Hydrogen Tomorrow... Demands and Technology Requirements*, report of the Hydrogen Energy Systems Technology (HEST) Study for NASA, Jet Propulsion Laboratory Report JPL 5040-1, December 1975.

Table 2.34. Estimates of U.S. hydrogen requirements

(EJ)

	1972 1973		2000					
	JPL.[a]	GA[b]	JPL.[a]	JPL[a] maximum	GA[b]	Westinghouse		
						Low	Base	High
Ammonia, methanol, and miscellaneous chemicals	0.49	0.63	2.50	9.77	2.97	0.80	1.55	2.26
Petroleum refining	0.50	0.51	0.82	0.81	0.63	0.61	2.54	2.80
Synthetic fuels			1.81	8.0	7.38	6.15	10.90	14.08
Steelmaking			0.23	0.69	2.23	0.10	0.19	0.38
Transportation, utility, other	0.07	0.60	0.42	4.45	0.39		0.41	1.21
Total	1.06	1.74	5.78	23.72	13.60	7.66	15.59	20.73

[a] Data from California Institute of Technology Jet Propulsion Laboratory, *Hydrogen Tomorrow: Demands and Technology Requirements* (Review Draft), Report 5040-1, Oct. 30, 1975.

[b] Data from General Atomic Company, *Studies of the Use of High-Temperature Nuclear Heat from an HTGR for Hydrogen Production*, Sept. 30, 1975, prepared under contract for NASA.

[c] Data from Westinghouse Astronuclear Laboratory, *Studies of the Use of Heat from High Temperature Production Process* (Draft Final Tasks I and II Report), July 1975.

2.4. MBG

With a heating value between 930 and 1300 kJ/scm (250 and 350 Btu/scf), the synthetic gas MBG has a chemical composition of 35 to 50% CO, 30 to 40% H_2, and 10 to 20% CO_2.

Because of its chemical composition, MBG is suitable for both fuel and feedstock applications in all chemical syntheses fed by natural gas. Using MBG, flame temperatures in the range of 1930 to 2040° C (3500 to 3700° F) can be obtained, equivalent to or slightly higher than those for natural gas. Slight differences in combustion and flue volumes would require minimal, if any, piping and burner adjustments. Because of these favorable characteristics, MBG has a very large potential market if it is priced competitively with alternate feedstock and fuels.

As a feedstock, MBG has few limitations. Its key characteristics are high H_2 and CO content (with varying ratios), a low inert (nitrogen) component, and a low methane concentration (except in the Lurgi process), as shown in Table 2.35.

These characteristics lead directly to several conclusions about desirable applications. The potential of MBG as a chemical feedstock is excellent, particularly for: (1) methanol, which requires an

Table 2.35. Uses of MBG

Characteristics
High H_2 and CO concentration, about 75 ± 13%
Varying H_2–CO ratios
Low inert material concentration
Low methane concentration, except in Lurgi gas
Higher heating values around 1042 kJ/scm (280 Btu/scf)
Advantages
Being transportable in pipelines; multiplant supply possible
Flame temperatures marginally lower or higher than natural gas (±5%) because of high adiabatic temperatures of H_2 and CO
Good value as a chemical feedstock
Applications
All industrial fuel applications
Chemical industry as feedstock
Gas industries for SNG production
All industries for direct firing of finished products, automatic heating systems, firing of flow through radiant burners, indirectly fired furnaces
Iron and steel industries for direct reduction of iron ore, integrated steel mill fuel
Paper and pulp industries for paper and print drying
Glass industry for glass tanks, cutting, annealing and scrapping
Primary metals industry for finishing annealing operations, flat flame burners, partial premix burners

attractive 2:1 ratio of H_2 to CO; (2) ammonia, which requires pure H_2 under pressure;* and (3) urea and related products. Although volume requirements are greater than for natural gas, 37,000 kJ/scm (1000 Btu/scf) pipeline transmission up to 200 miles is believed by a number of suppliers to be economically feasible.

A recent study, made by Booz, Allen and Hamilton, Inc.[11] for DOE, analyzed the potential industrial markets for low-Btu gas (LBG) and MBG. The study included in-depth interviews with potential industrial users, equipment suppliers, and architect/engineering firms and focused on the near-term (1985) potential market. Although their focus was a near-term one, many of Booz, Allen and Hamilton's conclusions apply equally well to this study.

1. MBG can be used widely as both a fuel and a feedstock in the chemical industry. It can readily replace natural gas fuel, according to industrial discussions, in both boilers and furnaces, which account for about 40% of the energy usage in the industry.

2. In feedstock applications, which consume about 55% of the oil and gas used in the petrochemical industry, MBG is a feasible, and sometimes more efficient, substitute for natural gas because of its high H_2 and CO concentration (about 65 to 90%), low methane content, and low diluent content.

3. Likely feedstock applications will, as indicated in industry interviews, include hydrogen, methanol, ammonia, acetic acid, and other commodity chemicals. Chemical plant energy consumption typically ranges from several hundred million to more than 63.6×10^9 kJ/d (60 billion Btu/d). At least 75 major chemical plants consume in excess of 15.9×10^9 kJ/d (15 billion Btu/d), as shown in Table 2.36. As a continuous process industry, load factors in the chemical industry are high, frequently in excess of 80 to 90%.

Table 2.36. Industrial plant consumption of natural gas for fuel and feedstock uses (1974)

Industry	Number of plants	Number of plants with consumption in each range [billion kJ/d (billion Btu/d)]				
		3-17 (3-15)	17-32 (15-30)	35-63 (33-60)	68-95 (64-90)	>95 (>90)
Chemicals	230	154	43	23	8	2
Iron and steel	187	141	20	22	2	2
Paper	727.	718	9			
Petroleum refining	258	201	33	16	0	8
Cement	172	169	3	0	0	0
Aluminum	31	20	7	1	1	2
Total	1605	1403	115	62	11	14

Easily adaptable to integrated steel mill processes, MBG can be applied in most of the steel industry's major energy-consuming processes, as indicated in industrial interviews, because of the similarity of its combustion characteristics with those of natural gas. Medium-Btu gas can be used

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*The Lurgi process produces about 9% methane and needs reforming. Conversely, Koppers Totzek output contains little methane and requires no secondary reforming but needs compression.

directly or mixed with available coke oven gas. Feasible applications include: sintering, coke ovens, steel furnaces, soaking pits, reheating, heat treating and annealing, finishing, and boiler fuel.

Integrated steel mills are generally large and complex enough to accommodate and utilize large quantities of MBG. Plant-purchased fuel consumption usually ranges from 10 billion to 60 billion kJ/d (10 billion to more than 60 billion Btu/d). Nearly 50 steel plants in operation consume more than 15 billion kJ/d (15 billion Btu/d) of natural gas alone (Table 2.36). Energy demands in steel mills are cyclical but nearly continuous. Existing plant gas distribution systems now pipe lower-energy by-product gases, including coke oven gas [about 19 to 21 MJ/scm (500 to 550 Btu/scf)] and blast furnace gas [about 3 MJ/scm (90 Btu/scf)].

Medium-Btu gas can, on technical grounds, substitute for most of the fuel used in the paper industry. It can be used to fire boilers, which account for more than 70% of the energy needed in the industry. Major processes include digesting, evaporating, and pulp drying. This synthetic gas can also be used in several smaller direct heat applications, such as in lime kilns, coating, and specialized dryers.

However, MBG gas use is unlikely to be feasible except in selected large paper or pulp mills because of three factors.

1. Typical plant fossil fuel consumption exceeds 10 billion kJ/d (10 billion Btu/d) only in the larger integrated mills. Much of their requirement is met by burning mill wastes.
2. Only 9 of more than 700 paper industry facilities investigated consumed more than 15 billion kJ/d.
3. Many mills are remote from major industrial areas, reducing the likelihood that MBG would be piped to such locations.

Because refineries are designed for varying fuel quality, MBG can feasibly be used in petroleum refining. It can also replace natural gas in its significant applications in refining. As pointed out in earlier discussions, the combustion similarities in MBG and natural gas indicate that MBG can be readily utilized in the two largest energy-consuming processes, crude distillation and hydrogen production by reforming. It can also replace natural gas in other major processes, for example, delayed coking, straight distillation, and hydrotreating.

Because combustors are generally designed to use fuel inputs of varying qualities and because the gas can easily be mixed with refinery off-gas, both boilers and process furnaces can utilize lower-energy gas. The heat content of refinery gas generally ranges between 34,000 and 52,000 kJ/scm (900 and 1400 Btu/scf), but it can range from as low as 11,000 to as high as 75,000 kJ/scm (300 to as high as 2000 Btu/scf).

More than 50 of about 250 refining facilities examined in one study consumed more than 15 billion kJ/d of natural gas (Table 2.36). Refinery operations are generally continuous on both a daily and seasonal basis, although the mix of fuels used may vary by season.

For the cement industry, the use of MBG is feasible, according to industry interviews. It can be used in the industry's rotary kilns, which account for ~80% of the industry's energy use, to fire the raw materials to produce clinker. Typical plant fuel consumption averages from 5 billion to 12 billion kJ/d. As shown previously in Table 2.36, only 3 of 172 cement plants surveyed indicated natural gas usage in excess of 15 billion kJ/d.

Cement industry applications of MBG will be limited by the industry's growing use of direct coal combustion. Direct coal firing can be quite attractive because of favorable coal prices and the fact that combustion products—including ash and sulfur—are absorbed in the product. The industry is now reliant on coal for more than 70% of its energy requirements, and the current trend is toward greater coal dependence.

Fabricated metal products and motor vehicle industries can use MBG. Subject to several limitations, MBG can replace current natural gas applications in most direct heat and boiler processes in these industries. Although MBG's similarity to natural gas in flame characteristics indicates that its potential uses are wide, retrofitting of combustion equipment may be necessary in tempering and other processes, according to industry interviews. Average plant fuel consumption is relatively small—generally from a few billion kilojoules per day to less than several hundred million kilojoules per day—and exceeds 10 billion kJ/d only in larger automobile manufacturing plants and foundries. According to industry interviews, load factors of 40 to 50% are common in most plants. Major processes where MBG could be feasible, however, include melting, metal recovery, forging, heat treating, coating, painting, and decorating.

Medium-Btu gasification can technically supplant most natural gas applications in the glass industry because MBG combustion characteristics are similar to those of natural gas. About 70% of the industry's fuel consumption is used for glass melting, which is reported to require minimal retrofitting to utilize MBG. Other thermal processes, including forming and annealing, may require more significant burner adjustments.

The brick and structural clay tile industry can also feasibly apply MBG. It is compatible with energy-consuming processes in this industry. The predominant use of energy in the brick industry is for firing structural clay products in kilns. Although some flame control requirements must be met to ensure proper brick coloring, industry interviews indicate that MBG would fulfill these specifications. However, the industry's investigations of, and initial movement toward, direct coal combustion indicate that the feasible market for gasification may be diminishing.

The primary aluminum and nonferrous metals industry can realistically consider MBG in most energy-consuming applications. This gas can replace natural gas in fuel-fired processes such as melting and holding, heat treating and annealing, and fabrication.

Although the industry's largest energy-consuming process—alumina reduction—relies mostly on electrolytic smelting, MBG could be used to displace fuels now used to self-generate electricity. Electricity accounts for about 68% of the aluminum industry's energy use. About 8 to 10% of this electricity is self-generated. Of 31 aluminum plants surveyed (shown previously in Table 2.36), 11 used more than 15 billion kJ/d of natural gas alone. Industry energy demands are cyclical but nearly continuous.

The food industry can use MBG widely to replace natural gas. Among major energy-consuming processes in the industry, MBG can be used in dehydrators, dryers, singeing, roasting ovens, and boilers. Most food processing facilities consume less than 1 to 2 billion kJ/d.

In the textile industry, MBG can feasibly replace natural gas and other fuels. Major technically feasible applications could include fueling dryers and boilers. Only the very largest textile mills consume as much as 8 billion kJ/d. Mill operations are not generally of a continuous process nature through the year.

The potential market for MBG as a chemical feedstock is large. Major markets would be in the production of hydrogen, ammonia, methanol, and acetic acid. The production of hydrogen, ammonia, and methanol are discussed in Sect. 2.3. Acetic acid can be produced by the carbonylation of methanol, $CH_3OH + CO \rightarrow CH_3COOH$; however, the preferred reaction involves the oxidation of acetaldehyde. In any event, the production of acetic acid using methanol is accounted for in figures for the production of methanol.

The feasible market for MBG gasification consists of thousands of industrial plants nationwide. The DOE has identified 3500 major fuel-burning installations nationally (Table 2.37), including over 6000 individual units with fuel-burning capacities exceeding 2.4 billion kJ/d.

Medium-Btu gas can be used in more than 550 plants nationwide. As shown in Table 2.38, 558 industrial establishments consume more than 7.4 billion kJ/d of oil and gas.

The potential displacement of fossil fuel with MBG in 550 plants could total more than 1.6 EJ/year (Table 2.37). Within the 20 largest Standard Metropolitan Statistical Areas (SMSAs), an average of >470 billion kJ/d in fossil fuels was consumed for industrial use (Table 2.39). Assuming use of MBG in 300 multiplant systems nationally, >3 EJ/year of fossil fuel could be displaced by MBG.

Table 2.37. Major fuel-burning installations

Federal region	Number of installations
I (CT, MA, ME, NH, RI, VT)	142
II (NJ, NY, Puerto Rico, Virgin Islands)	324
III (DE, DC, MD, PA, VA, WV)	395
IV (AL, FL, GA, KY, MS, NC, SC, TN)	522
V (IL, IN, MI, MN, OH, WI)	883
VI (AR, LA, OK, NM, TX)	482
VII (IA, KS, MO, NE)	180
VIII (CO, MT, ND, SO, UT, NY)	147
IX (AZ, CA, HI, NV)	284
X (AK, ID, OR, WA)	141
Total	3500

Source: *M.F.B.I.: Complete Printout of FEA C-602-5-0*, Federal Energy Administration, Office of Data Service, Feb. 14, 1977.

Table 2.38. Number of major energy-consuming plants[a]

Industry	Number of establishments[a]
Chemicals	120
Paper	87
Petroleum	83
Primary iron and steel	61
Cement	43
Primary aluminum	8
Others	156
Total	558

[a] Number of plants with oil and gas consumption greater than 7.4 billion kJ/d.

Source: *Low-Btu Gas Industrial Application Analysis*, Systems Consultants, Inc., 1978 (based on 1975 survey).

Table 2.39. Industrial consumption of purchased fuels for 20 largest metropolitan areas[a]

(1976)

Standard metropolitan statistical area	Purchased fuel consumption (billion kJ/d)	Major energy-consuming industries
Houston, Tex.	1817	Chemicals; petroleum
Chicago, Ill.	769	Primary metals, food
Beaumont-Port Arthur-Orange, Tex.	851	Petroleum, chemicals
Pittsburgh, Pa.	768	Primary metals
Gary-Hammond-East Chicago, Ill.	690	Primary metals
Detroit, Mich.	594	Primary metals
Baton Rouge, La.	607	Chemicals
Philadelphia, Penn.-N.J.	521	Primary metals, petroleum chemicals
Los Angeles-Long Beach, Calif.	389	Petroleum; stone, clay, glass, primary metals
Cleveland, Ohio	379	Primary metals
Lake Charles, La.	400	Chemicals, petroleum
Galveston-Texas City, Tex.	396	Chemicals, petroleum
San Francisco-Oakland, Calif.	357	Petroleum, chemicals
St. Louis, Mo.-Ill.	310	Primary metals; stone, clay, glass, chemicals
New York, N.Y.-N.J.	196	Food, chemicals, paper
New Orleans, La.	277	Chemicals; petroleum; stone, clay, glass
Newark, N.J.	225	Chemicals
Buffalo, N.J.	208	Primary metals, chemicals
Steubenville-Weirton, Ohio-W. Va.	238	Primary metals
Corpus Christi, Tex.	244	Chemicals, petroleum

[a] SMSAs ranked by largest consumption of purchased fuels and electric energy.

Source: *Annual Survey of Manufacturers*, U.S. Department of Commerce, Bureau of the Census, 1976.

One penalty when substituting MBG for natural gas is the added cost of compressing and pipelining a greater volume of gas to deliver the same number of British thermal units. The typical cost range for delivering natural gas by pipeline is 1.3 to 2.0¢/GJ per 100 km (2 to 3¢/10^6 Btu per 100 miles) as compared with 4.4 to 6.6¢/GJ per 100 km (7 to 10¢/10^6 Btu per 100 miles) for MBG.[12]

In summary, if MBG were substituted for methane in the various sectors discussed above, the market potential could be as great as 4.8 EJ/year (4.5 quads/year).

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3. TECHNICAL ASSUMPTIONS AND DESCRIPTIONS

3.1 Introduction

Section 3 contains descriptions of each energy transport system considered in this study. All technical assumptions that led to the cost elements reported in Sect. 4 are also discussed in this section.

Piping and pumping calculations common to the various systems are given in Appendix D. A 70% annual system capacity factor was assumed in all cost calculations. A temperature sensitivity study for hot oil and molten salt used a 90% capacity factor.

3.2 CHP

3.2.1 System description

A CHP consists of three basic components: an endothermic chemical reaction (with heat supplied by a VHTR), a transport system, and an exothermic chemical reaction (at the end use). For this study, these components, which are shown in Fig. 3.1, are a steam reformer, a pipeline, and a methanator, respectively. An open-cycle CHP is like the closed-cycle CHP except the methane is not returned to the reformer plant. Instead, methane is purchased and used at the reformer plant and sold after the methanation process.

The endothermic reactions that take place in the reformer in the presence of a catalyst are:

$$CH_4 + H_2O \rightarrow CO + 3H_2\,, \qquad \Delta H = +205.2\ \text{kJ/mol}\,,$$
$$CH_4 + 2H_20 \rightarrow CO_2 + 4H_2\,, \qquad \Delta H = +163.3\ \text{kJ/mol}\,.$$

The reverse reactions, releasing heat for possible process heat applications, take place in the methanator. Water is removed from the gas streams to reduce pumping costs before the gases are transmitted. This approach requires a supply of makeup water at the reformer plant.

3.2.2 Reformer[1]

A typical design for a helium-heated reformer tube is shown in Fig. 3.2 (Ref. 2). Preheated steam and methane (in ratios of 2:1 to 3:1) enter the top of the reformer tube at a temperature of ~450° C and enter the pigtail at the bottom of the tube at a temperature of ~830° C (~1530° F), thereby transferring

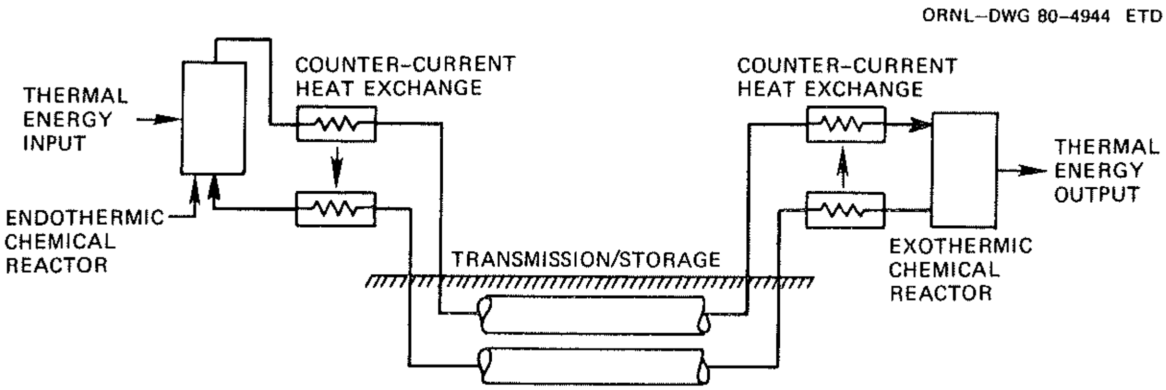

Fig. 3.1. Chemical heat pipe.

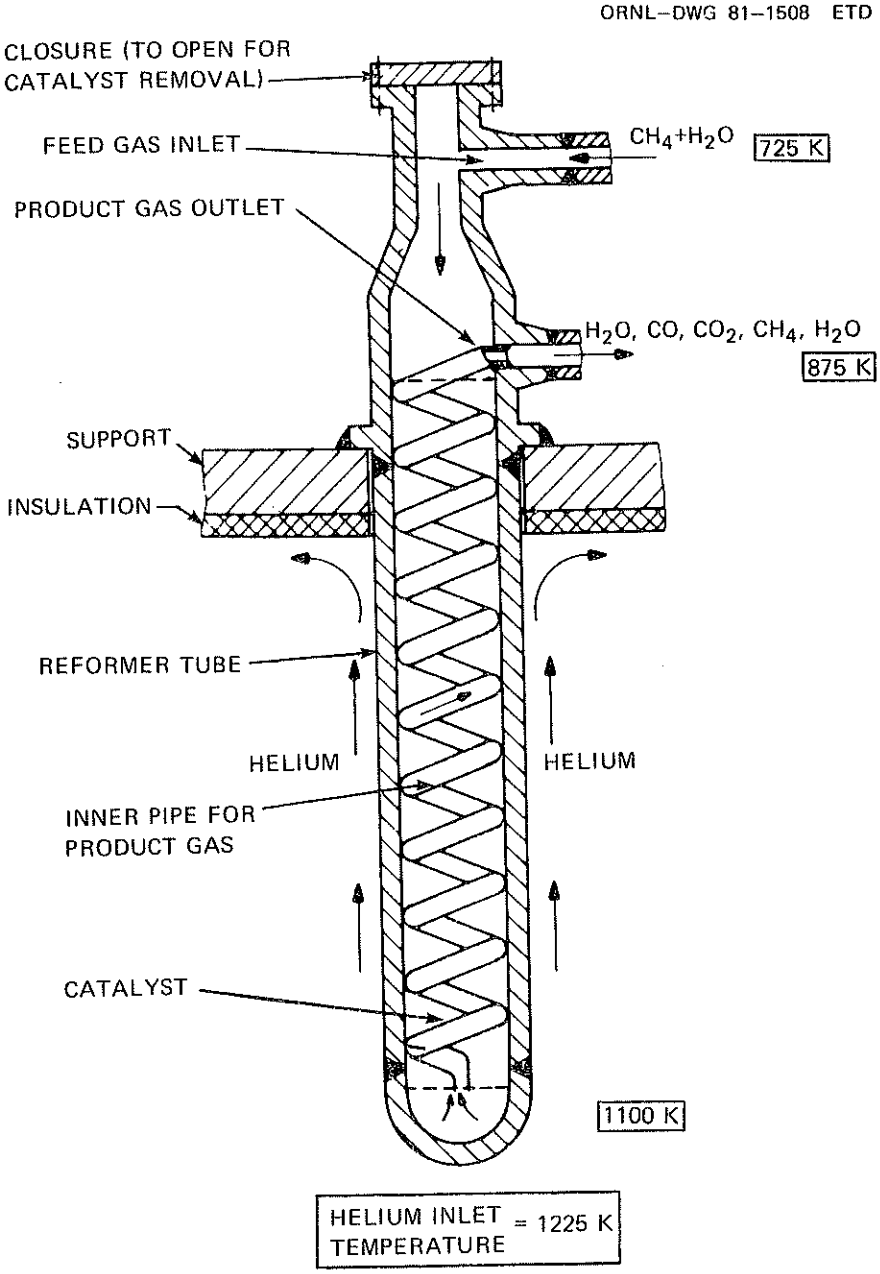

Fig. 3.2. KFA design of helium-heated reformer tube.

part of the sensible heat back to the catalyst bed across the pigtail wall. Hot helium from the nuclear core enters at a temperature of ~950° C (~1740° F) at the bottom of the reformer tubes, flows upward around the tubes, and exits from the reformer at a temperature of 600 to 700° C (1110 to 1290° F).

On the basis of the experiments conducted at *Institut für Reaktorentwicklung, Kernforschungsanlage (KFA), Jülich, Federal Republic of Germany (FRG),*[1] the expected performance of a helium-heated reformer can be summarized as follows.

1. The chemical conversion is limited by heat transfer and not by kinetics. For all practical purposes, thermodynamic equilibrium is reached for both reactions at all points in the catalyst bed.
2. Overall heat transfer coefficients are ~350 $W/m^2 \cdot {}^\circ C$ (~60 $Btu/h \cdot ft^2 \cdot {}^\circ F$), and the average heat flux over the tube is 64 kW/m^2, which is nearly identical with that for conventional reformers (~20,000 $Btu/h \cdot ft^2$).
3. Exit gas composition corresponds to thermodynamic equilibrium at 830°C (1530°F) and 40 bars. No appreciable changes in composition occur in the pigtails as the exit gas stream is cooled to 600°C (1110°F).

These performance characteristics led to a set of design parameters for a nuclear reformer:

- tube dimensions—100-mm ID, 15-mm wall thickness, 10-m length;
- design heat flux—63.1 kW/m^2.

Three half-sized reformers were assumed for reliability.

3.2.3 Methanator

Cold reformer gases are transported to the methanation plant where they are catalytically converted to methane and steam according to the following reactions:

$$CO + 3H_2 \rightarrow CH_4 + H_2O \, , \qquad \Delta H = -205.2 \text{ kJ/mol} \, ,$$
$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O \, , \qquad \Delta H = -163.3 \text{ kJ/mol} \, .$$

The size of the methanator is based on a space velocity of 1500 $scm^3/h \cdot m^3$ (Ref. 1). A feed gas flow of ~1750 scm^3/h produces 1 MW of product heat.[3] These factors size the methanator at ~1.17 m^3 (of catalyst) per MW. Three half-sized methanators were used for reliability.

The methanation plant consists of a methanator train with gas coolers between methanators. These gas coolers are heat exchangers that produce process steam or hot water as they control the temperature of the methanation gases.

3.2.4 Heat exchangers[1,3]

The reformer plant heat exchangers were assumed to have an average heat flux of 32 kW/m^2 (10,000 $Btu/h \cdot ft^2$) based on a temperature difference of 20°C (36°F). These heat exchangers had a duty approximately equal to the delivered heat.

The methanator plant heat exchangers were divided into two types, gas preheat and steam generation. The gas-preheat exchangers had a heat flux of 5.7 kW/m^2 (1800 $Btu/h \cdot ft^2$), and the steam generators had a heat flux of 27 kW/m^2 (8000 $Btu/h \cdot ft^2$). These heat fluxes are based on mean temperature differences of 20°C (36°F) for gas preheat and 60°C (108°F) for steam generation. These temperature differences are appropriate when the product steam has a maximum temperature of 350°C (660°F). The gas-preheat duty was ~60 kW/MW of delivered heat, and the steam-generator duty was equal to the delivered heat.

3.2.5 Auxiliary power: production and consumption

A steam-generator and a turbine-generator set are assumed to be located at the reformer plant. This combination increases overall efficiency by using the heat remaining in the helium (after the reformer) to produce electricity (Fig. 3.3).

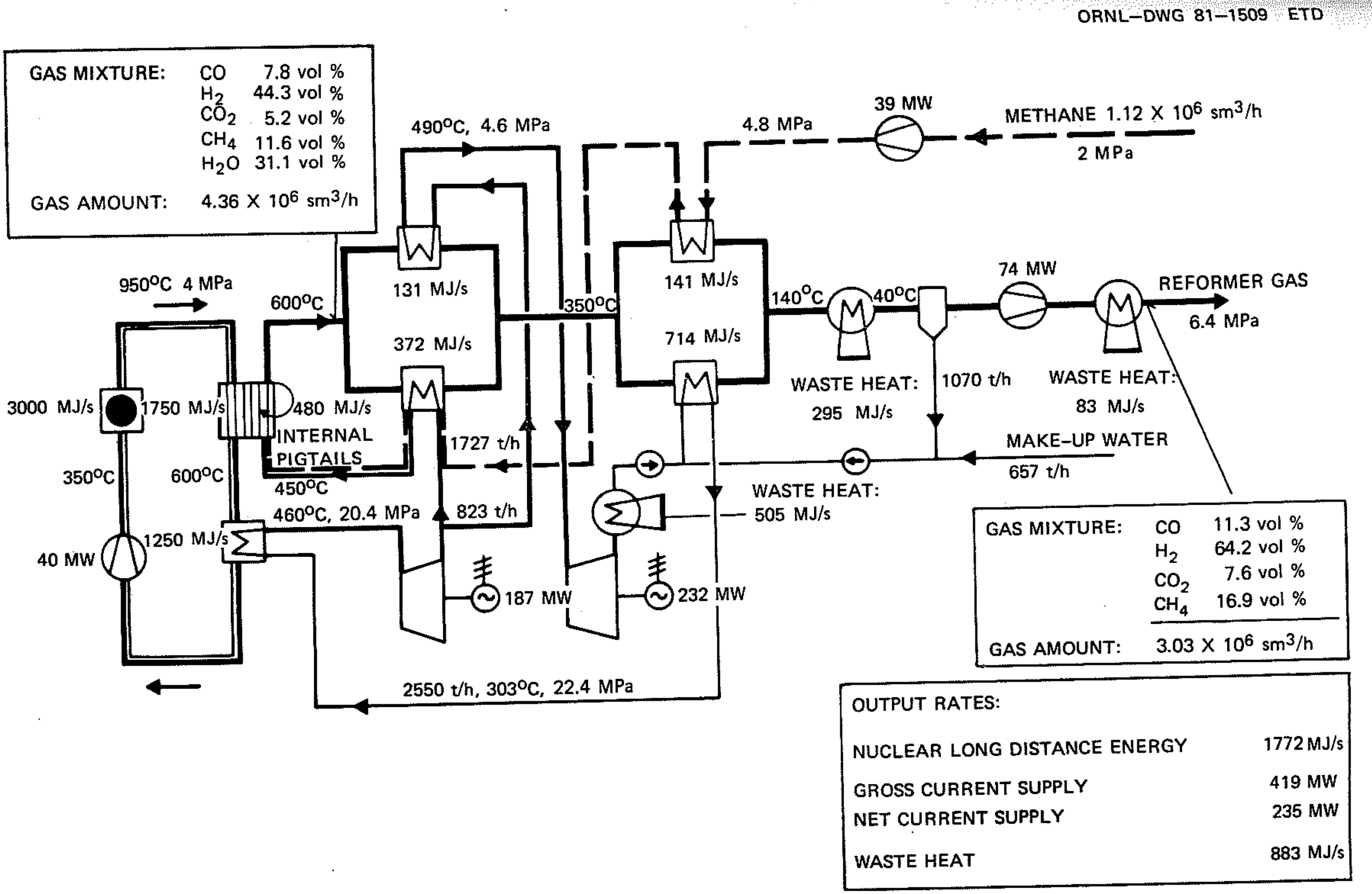

Fig. 3.3. KFA reformer plant details.

The size of the turbine-generator output was approximately one-fourth of the delivered heat. Plant auxiliary power consumption was estimated to be ~16% of this power generation.

Pressure losses for the reforming and methanation processes were assumed to equal those losses calculated for a 160-km (100-mile) closed-cycle pipeline.[1] Appendix D presents a general discussion of pressure loss calculations. The process and pipeline pumps were electrically driven.

3.2.6 Pipe sizing

A rough estimation of piping and pumping costs was done to determine appropriate pipeline diameters. The maximum diameter considered was 1.2 m (48 in.) because of commercial restrictions. A double pipeline was used for the CO–H_2 transmission, with each of the pipes 1.2 m (48 in.) in diameter. Table 3.1 lists the diameters used for the CH_4 return line in the closed-cycle CHP.

The pipeline flows were determined by scaling the KFA pipeline flows[3] by the ratios of the heat outputs considered in this study to the KFA heat output. Table 3.2 gives the pipeline gas flows.

Table 3.1. CH_4 pipeline diameters

Cycle power (MW)	Distance [km (miles)]	Diameter [m (in.)]
800	80 (50)	0.86 (34)
	160 (100)	0.86 (34)
	320 (200)	0.91 (36)
1000	80 (50)	0.91 (36)
	160 (100)	0.91 (36)
	320 (100)	1.0 (36)
1200	80 (50)	1.0 (40)
	160 (100)	1.0 (40)
	320 (200)	1.1 (42)

Table 3.2. Pipeline gas flows

Output (MW)	CO–H_2 [scm/s (lb/s)]	CH_4[a] [scm/s(lb/s)]
800	397 (1054)	146 (219)
1000	493 (1317)	183 (274)
1200	596 (1581)	220 (329)

[a] Closed cycle only.

3.2.7 Process heat

In the reforming process, VHTRs were used to provide heat input and to raise steam. The reactor capacity was assumed to be 10 to 15% above the required heat input. This led to the use of a 1600-MW(t) reactor for the delivery of 800 MW(t) at the methanator, a 2000-MW(t) reactor for the delivery of 1000 MW(t), and a 2400-MW(t) reactor for the delivery of 1200 MW(t).

3.2.8 Methane inventory

The methane inventory for the closed-cycle CHP was calculated by multiplying the contents of the CH_4 pipeline by two to account for the $CO–H_2$ pipeline inventory.

The open-cycle CHP treats methane as a pass-through item. In this study, the assumption was made that the methane would be sold at the methanator for the same price paid at the reformer. This treatment places the burden of the pipeline costs on the thermal energy product and transports the methane "free of charge." This open-cycle treatment has one other drawback. As shown in Sect. 4.1, methane costs are expected to equal oil costs after the year 1990. This means that toward the end of the plant's lifetime (i.e., after 2020) methane may be very costly.

3.2.9 Cycle efficiency

The efficiencies of the CHP cycle were calculated as the net energy output divided by the energy input. Energy input is equal to the output of the VHTR. Net energy output was equal to

$$\text{Net energy out} = Et + Ee - Ep - Ea ,$$

where

Et = thermal energy delivered,
Ee = electrical energy produced,
Ep = pumping energy consumed,
Ea = plant auxiliary energy consumed.

3.3 MBG from Coal

3.3.1 Process description[4]

Available commercial processes for the production of MBG are based on three technologies: (1) fixed-bed gasifiers, (2) fluidized-bed gasifiers, and (3) entrained-flow gasifiers.

In the first process, sized coal is fed into the top of a fixed-bed gasifier and moves slowly downward in a bed maintained at a constant depth of about 1 m on a grate. Steam and air or oxygen pass upward. The residual material is discharged at the bottom of the bed as dry ash or slag, depending on the mode of operation. In the two-stage configuration, part of the gas, containing no tars or oils, is withdrawn hot from the middle of the gasifier. The remaining gas, containing tars and oil, leaves the gasifier from the top. Low-temperature fixed-bed gasifiers are made by Lurgi and others.

Fluidized-bed gasifiers, such as those marketed by Winkler, are similar to fixed-bed gasifiers except the velocity of the air- or oxygen-steam mixture is high enough and the coal pieces are small enough for the coal to be suspended in the upward flowing gas stream. The gas velocity is low enough for the coal to stay in the vessel.

Entrained-flow gasifiers are similar to fluidized-bed gasifiers except that pulverized coal is injected into the high-velocity concurrently flowing stream of air or oxygen and steam. The gas velocity is so high that the coal is carried along with the gas stream. A separator removes the residual particles from the product gas. Since the stream is concurrent, reaction rates decrease with displacement through the gasifier; therefore, high temperatures are required to keep reactor size reasonable. Koppers-Totzek (K-T) entrained-flow gasifiers are made by Krupp-Koppers of West Germany.

3.3.2 Transmission pressures and burner retrofit

Three transmission pressures were considered in this evaluation: 0.9, 1.8, and 3.6 MPa. These pressures were chosen to show the sensitivity of cost to pressure and do not pertain to any particular gasification method or end-use combustion equipment.

Retrofit requirements of various combustion systems were not analyzed. These retrofitting requirements vary for different technologies and processes and for different plant configurations within any technology.

Diluents, such as nitrogen, carbon dioxide, and water vapor, present in manufactured gases result in lower heating values and lower flame temperature than natural gas. In the case of MBG, however, high concentrations of CO and H_2, with adiabatic flame temperatures higher than methane, result in a flame temperature close to or higher than that of natural gas, depending on the specific MBG composition.[4]

Although there are large differences in heating values between the manufactured gases and natural gas, the differences in heating values among the air-fuel mixtures at the burner are much smaller. These close values are achieved, however, by substantially changing the volumes of the MBG-air mix relative to methane-air. Higher flame temperature and lower flue gas volume in the MBG-air case may actually lead to an increase in the rating of a boiler.[4]

Another probable consequence of different air-to-fuel ratios is the required introduction of mechanical changes in the burner system to achieve flame stability from changed gaseous flow rates.[4]

3.3.3 Plant sizes

The size of the MBG gasification plants evaluated in this study provided an equal energy output for comparison with other energy transport options. While this choice does allow easy comparison, it may cause the MBG to appear more costly than it should. An optimum-sized MBG plant will likely be much larger and more economical than the MBG plants considered in this study.

3.4 Hydrogen

Four hydrogen production processes were considered: coal gasification, potassium hydroxide electrolysis, SPE, and SPE with storage. Each plant was scaled for outputs of 800, 1000, and 1200 MW. For hydrogen gas, these energy outputs are equivalent to 5.74, 7.18, and 8.62×10^6 scm/day. Other system specifications are given in Table 3.3.

Pipeline pressure for the hydrogen systems was assumed to be 3 MPa (450 psia) to correspond with the process pressure of the coal gasification system. Although this may not be an optimal transport pressure, it will not have a major impact on the final cost because the transport costs represent only ~4% of the total delivered energy costs.

3.4.1 Coal gasification[5]

Hydrogen is produced from coal by reforming, oxidation, and shift reactions similar to the production of MBG gas:

$$CH_{0.8} + H_2O \rightarrow CO + 1.4H_2 \,, \qquad (1)$$

$$CH_{0.8} + 1.2O_2 \rightarrow CO_2 + 0.4H_2O \,, \qquad (2)$$

$$CO + H_2O \rightarrow CO_2 + H_2 \,. \qquad (3)$$

Table 3.3. Hydrogen system specifications

System	Delivery capacity [MW(t) (10^6 scm/d)]	Pipeline diameter [m (in.)]	Energy output [10^6 GJ/year (10^6 scm/year)]	Electrical generation plant size [MW(e)]	System efficiency[a] (%)
Coal gasification	800 (5.74)	0.66 (26)	17.66 (1470)		46.3
	1000 (7.18)	0.71 (28)	22.075 (1840)		46.3
	1200 (8.62)	0.76 (30)	26.40 (2200)		46.3
KOH electrolysis	800 (5.74)	0.66 (26)	17.66 (1470)	1050	75.7
	1000 (7.18)	0.71 (28)	22.075 (1840)	1320	75.7
	1200 (8.62)	0.76 (30)	26.40 (2200)	1590	75.7
SPE electrolysis	800 (5.74)	0.66 (26)	17.66 (1470)	1030	77.6
	1000 (7.18)	0.71 (28)	22.075 (1840)	1290	77.6
	1200 (8.62)	0.76 (30)	26.40 (2200)	1550	77.6
SPE electrolysis with storage	800 (5.74)	0.66 (26)	17.66 (1470)	2470	77.6
	1000 (7.18)	0.71 (28)	22.075 (1840)	3100	77.6
	1200 (8.62)	0.76 (30)	26.40 (2200)	3720	77.6

[a] Hydrogen system only; it does not include electrical generation efficiency.

The overall reaction for producing hydrogen from coal is therefore

$$CH_{0.8} + 0.70O_2 + 0.60H_2O \rightarrow CO_2 + H_2 . \qquad (4)$$

For coal feed, the gasification reactor is generally operated at 1480° C (2700° F). The heat required for the gasification reaction [Eq. (1)] is supplied by burning part of the coal with oxygen [Eq. (2)]. When SNG is the desired product, coal is usually gasified at a lower temperature than that required for hydrogen production. A higher overall thermal efficiency occurs in the SNG operation.

The K-T process is an entrained-flow gasifier such as described in Sect. 3.3.1. Finely ground coal is gasified with oxygen and steam at ~1480° C (2700° F) and atmospheric pressure to produce synthesis gas, a mixture of H_2, CO, CO_2, H_2O, H_2S, COS, and SO_2. These gases are compressed to ~2 MPa (300 psig) and then processed for CO shift conversion to H_2 and CO_2 and for H_2S removal. Other high-temperature gasifiers are also available.

The investment and operating costs for the K-T gasification process could be reduced by operating the gasifier at 3 MPa (450 psig). This would reduce the physical size of the gasifier vessels and the costs of gas compression. Pressurized operation of the gasifier would, however, require facilities for forcing the pulverized coal feed into the 3-MPa (450-psig) gasifier vessel and removing the ash from this vessel. In this study, the investment and operating costs for the pressurized gasification method have been estimated. To provide the economic advantages over the K-T process assumed in this study, the new coal process should have the following characteristics:

1. Gasification would be conducted at 2.8 to 3.4 MPa (400 to 500 psig).
2. A single gasifier would convert 2500 tons/d of coal to high yields of synthesis gas with essentially zero methane content.
3. The gasifier would operate above 1090° C (2000° F) to minimize methane production.
4. The gasifier would provide over 95% conversion of coal to synthesis gas.

Koppers and Shell are jointly developing a gasifier that will achieve these objectives. Texaco is also working on a pressurized gasifier.

The coal gasification plants used for these economic studies are essentially self-contained facilities that also include oxygen plants.

3.4.2 KOH electrolysis

The electrolysis of water to produce hydrogen is a mature technology and has been a common industrial practice since 1940. In this process, direct current is passed between iron or nickel-plated iron electrodes in a solution consisting of 10 to 25% caustic soda or potash. Only distilled water is added to the electrolyte because the solute is not consumed. Operating voltage is ~2 V; current density is about 650 A/m^2. Electricity required is 4950 kWh or 18 GJ/100 m^3 of hydrogen (140 or more kWh/1000 ft^3) plus one-half that volume of oxygen (if the oxygen is recovered).

The hydrogen is then compressed to 3 MPa (450 psig) for comparison with the other hydrogen production processes.

3.4.3 SPE[5,6]

A new electrolysis process is being developed by General Electric. This process employs a solid polymer electrolyte as a replacement for the traditional aqueous KOH solution. The solid polymer electrolyte has ion exchange properties, and, with the addition of water, hydrogen ions flow through the electrolyte and are converted at the cathode to hydrogen gas.

The SPE technology has demonstrated outstanding performance and operating characteristics in systems developed for relatively small aerospace and military applications. These characteristics include the following:

1. significantly higher cell efficiency than conventional electrolyzers, resulting in lower power consumption per unit of gas generated;
2. higher current density capability, resulting in lower capital cost, size, and weight for the electrolysis modules;
3. the electrolyte chemically bound in the polymer chain, resulting in a system with no free corrosive liquids to be concerned with during design, assembly, operation, or maintenance of the system; and
4. a solid electrolyte making possible greater simplicity in the system design as well as improved reliability and safety.

3.4.4 SPE with iron-titanium hydride storage[7]

In this process, electric energy is assumed to be supplied during off-peak hours and used to produce hydrogen that is stored as a hydride of an iron-titanium alloy. During peak-load periods, this hydrogen is released by heating the hydride. The following description, from "Hydrogen for Energy Storage: A Progress Report of Technical Developments and Possible Applications,"[7] is for a 50-MW system.

During the hydrogen charging phase, high-voltage three-phase ac power from the utility power network is transformed and rectified by solid state rectifiers to dc power. This dc power is supplied to water electrolyzers capable of producing 0.4 kg/s of H_2 (3,250 lbm/h of H_2) and 3.3 kg/s of O_2 (25,800 lbm/h of O_2), maximum. During normal operation, 0.35 kg/s of H_2 (2,800 lbm/h of H_2) and 2.8 kg/s of O_2 (22,000 lbm/h of O_2) are produced. The oxygen is released to the atmosphere, while the hydrogen is cooled and passed through a deoxygenator to remove residual oxygen. The hydrogen is then passed through a dryer to remove residual water vapor, and the final product hydrogen gas has a dew point of

$-60°$ C ($-76°$ F) or less. The dry gas then flows to the combined hydride bed–heat exchangers where it is adsorbed and stored. During charging, a valve downstream of the hydride beds is closed. Cooling water circulates through the heat transfer tubes buried in the metal hydride to remove the heat released as the hydrogen is adsorbed and the iron-titanium-hydride is formed. The sensible heat in the bed is also removed by this stream. The cooling duty is thus determined by the requirements that 0.35 kg/s (2800 lbm/h) of hydrogen be stored and 15.7 MJ be removed for each kilogram of hydrogen added.

During discharge, the valve between the hydrogen dryer and the hydride beds is closed, and the exit valve from the beds is opened. Heated water circulates through the same heat transfer tubes used during storage to supply the heat of dehydriding and sensible heat as the hydrogen is released. The sensible heat energy is neglected because it is small in comparison with the reaction energy required. The heating duty is thus determined by the requirements that 0.35 kg/s (2800 lbm/h) of hydrogen be released and 15.7 MJ be supplied for each kilogram of hydrogen released.

The general arrangement of a 26-MW(e) iron-titanium metal hydride storage (conductive type) electric power plant includes a 24-module, Lurgi high-pressure 3-MPa (440-psia) water-KOH electrolyzer and 10 iron-titanium hydride hydrogen storage heat-transfer beds.

The electric auxiliaries include an ac-dc rectifier to convert three-phase, 60-Hz power from the utility network to operate the electrolyzers. The cooling/heating water auxiliaries include: hydride cooling water-tower system, electrolyzer intercooling system, electrolyzer coolant cooler, and a dehydriding heating water system.

During the hydriding portion of the plant cycle, the gross electrical power input is 72.5 MW(e) three-phase ac to the solid state rectifiers. The rectifiers have an estimated conversion efficiency of about ~97%; therefore, 2.16 MW is rejected to the atmosphere as heat and ~70.0 MW(e) dc is distributed to the 24 electrolyzers. These electrolyzers convert the 70.0 MW(e) into 0.35 kg/s of H_2 (2800 lbm/h H_2) with a conversion efficiency of ~72%, based on the high heating value of hydrogen (140 MJ/kg). About 20.0 MW of heat is rejected directly through the circulating cooling water system to the atmosphere. The 0.35 kg/s of H_2 (2800 lbm/h of H_2) produced by the electrolyzers is stored in the hydride beds at a uniform rate with an assumed efficiency of 100% (i.e., with no leakage). The heat of hydriding, 5.54 MW (6750 Btu/lbm) is rejected to the atmosphere by means of the cooling water system. Hydrogen chemical energy is thus stored at the rate of 50.1 MW, based on the high heating value of hydrogen. The nominal hydriding effectiveness for the charging portion of the plant cycle is 69%. This effectiveness can be defined as the ratio of the rate of hydrogen chemical energy stored to the total plant power required.

During discharging or dehydriding, hydrogen is liberated from the hydride beds at a uniform rate of 0.35 kg/s of H_2 (2800 lbm/h of H_2).

3.5 Molten Salt and Hot Oil

3.5.1 System description

Figure 3.4 describes the molten salt and hot oil supply-delivery systems.

At the user end of the system, the required energy delivery rate is specified with the required use-temperature and the user's return fluid temperature. Heat exchangers provide for transfer of energy to the user-side fluid from the pipeline fluid. Once the required use temperature is set, the quality of the steam depends on the pressure chosen by the user.

The pipeline consists of a supply pipe and a return pipe, with pumping stations distributed equidistantly along both lines, which were assumed to be insulated and buried in a common or

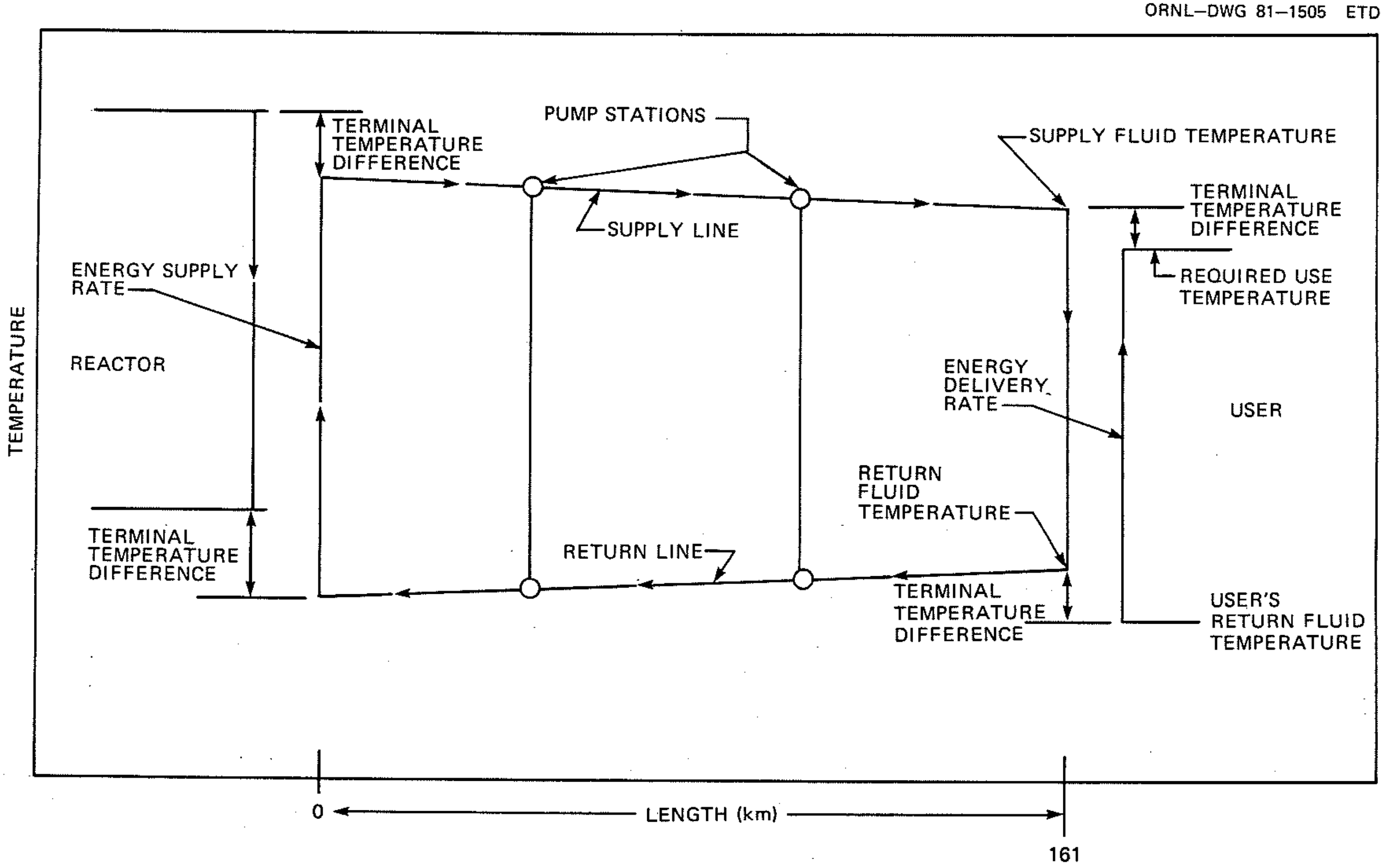

Fig. 3.4. Liquid supply-delivery system.

immediately adjacent trenches. At the pump stations, a supply-return pipe connection was assumed so that circulation could be maintained in subloops of the system while some sections of the line were out of service.

At the supply end of the system, energy is transferred from the source to the pipeline through heat exchangers.

Calculations of the quantities necessary for determination of capital and operating costs are shown in Appendix D. Fluid properties are given as a function of temperature in Fig. 3.5.

Molten salt was selected for consideration as a heat transport fluid because of its ability to transport energy at high temperature under low pressure without phase change. The salt selected, similar to HITEC,* is a eutectic mixture of KNO_3-$NaNO_2$-$NaNO_3$ with a melting point of 140°C (290°F). It is considered stable in the temperature range considered in this study [450 to 630°C (840 to 1170°F)], although some decomposition may occur at temperatures near 500°C (930°F).

In previous work,[8] corrosion of the containing pipe by the salt has not been considered a severe problem. However, recent data[9] indicate that corrosion of type-316 stainless steel (ss) may occur at intolerable rates at 550°C (1020°F) and above.

The heat transfer oil selected for study was Caloria HT-43.† This oil is representative of those available from various manufacturers. There is some variation in properties among these oils, but most have approximately the same thermal mass (density times specific heat). The primary difference among them is vapor pressure, which limits their (1) temperature range of application and (2) decomposition temperature limit. Note that some of the fluids [polychlorinated biphenyls (PCBs)] recently have been removed from the market because of environmental concerns.

Of the oils considered, Caloria HT-43 has a maximum allowable working temperature somewhat lower than others available [340 vs 400°C (640 vs 750°F)]. However, because it is approximately one-third as expensive as the competing oils and its properties are representative of the other oils, it was selected for this study.

3.5.2 Heat exchangers

Supply-end and user-end heat exchangers were sized using the following calculations:

$$\mathrm{LMTD} = (\mathrm{DTA} - \mathrm{DTB})/\ln(\mathrm{DTA}/\mathrm{DTB}) ,$$

and

$$A = \mathrm{QUSER}/(U \times \mathrm{LMTD}) ,$$

where

LMTD = log mean temperature difference,
DTA = terminal temperature difference on the hot end of the heat exchanger,
DTB = terminal temperature difference on the cold end of the heat exchanger,
A = heat exchanger surface area,
QUSER = energy delivery rate,
U = heat transfer coefficient [1140 $W/m^{2} \cdot {}^\circ C$ (200 $Btu/h \cdot ft^{2} \cdot {}^\circ F$) for salt, 490 $w/m^{2} \cdot {}^\circ C$ (87 $Btu/h \cdot ft^{2} \cdot {}^\circ F$) for oil].

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*Manufactured by Coastal Chemical under license from Dupont.

†Manufactured by Exxon Corporation.

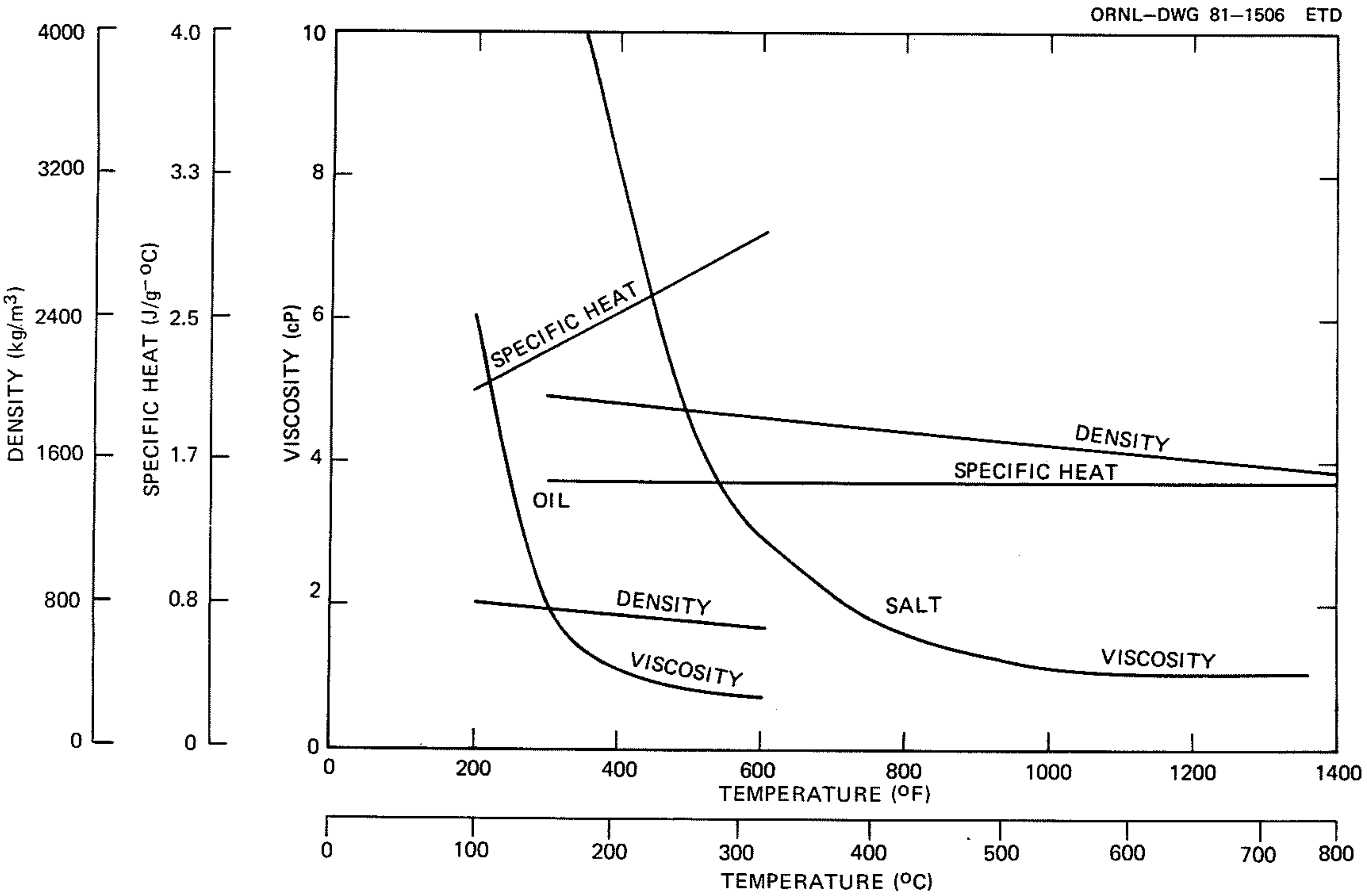

Fig. 3.5. Salt and oil fluid properties.

No pressure drop was included for losses in the heat exchangers. For conventional designs, these losses would be very small compared with the line losses.

3.5.3 Energy supply

The molten salt systems require an HTGR energy source to supply the necessary high temperatures. An oil system can be heated by either an HTGR or a pressurized-water reactor (PWR).

The rate of energy supplied from the source to the line was determined from the supply-end pipeline temperatures. The supply requirement was the sum of the delivery requirement and the pipeline loss minus the power input to the pipeline fluid by the pumps.

3.5.4 Piping and fluid requirements

Fluid inventory was determined from pipe volume. No allowance was made for filling heat exchangers.

Pipe material weight was determined from wall thickness, pipe diameter, line length, and material density. A factor of 1.07 was used as a multiplier to allow for fittings, turns, and elevation changes.

3.5.5 System optimization

Because of the large number of variables that influence the cost of pipeline transport of molten salt or hot oil, a preliminary analysis directed at determining the least-cost system configurations was required before generating cost figures for comparison and discussion. The primary system classifications were ss pipe with molten salt as the transport fluid, carbon steel (cs) pipe with molten salt, and cs pipe with heat transfer oil. Each classification was intended to serve a different temperature range.

The preliminary studies described in Appendix C consist of an informal optimization procedure in which basic cost data were fixed and the physical parameters were varied to discover their influence on cost and to allow selection of a set of variables that resulted in minimum delivered energy costs. This combination of variables is given in Table 3.4. The data that provided the basis of selection are given in a graph in Appendix C.

Table 3.4. Molten salt and hot oil system physical assumptions

System	User-end supply temperature [°C (°F)]	Insulation thickness supply/return [mm (in.)]	Pump station spacing [km (miles)]
Salt, ss [193°C (379°F) return]	675 (1250)	230/25 (9/1)	4.6 (2.85)
	621 (1150)	178/25 (7/1)	8.0 (5.0)
	565 (1050)	127/25 (5/1)	11.5 (7.14)
	510 (950)	102/25 (4/1)	12.4 (7.69)
	454 (850)	76/25 (3/1)	13.4 (8.33)
Salt, cs [193°C (379°F) return]	454 (850)	76/25 (3/1)	13.4 (8.33)
	399 (750)	51/25 (2/1)	20.1 (12.5)
	343 (650)	51/25 (2/1)	20.1 (12.5)
	288 (550)	51/25 (2/1)	20.1 (12.5)
Oil, cs [121°C (250°F) return]	327 (620)	51/25 (2/1)	20.1 (12.5)
	271 (520)	51/25 (2/1)	20.1 (12.5)
	216 (420)	51/25 (2/1)	20.1 (12.5)

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4. ECONOMIC ANALYSIS

4.1 Introduction

This section describes the economic analysis used to derive the cost of delivered energy for each energy transport option. Most cost data are based on information taken from a collection of Department of Energy (DOE) and industry reports. Data from these reports have been adjusted to reflect a chosen set of economic ground rules (Tables 4.1 through 4.6). These tables list general

Table 4.1. Economic factors

General inflation rate, %/year	6.0
Interest rate during construction, %	9.5
Federal income tax rate, %	46.0
State income tax rate, %	3.0
Property tax rate,[a] %	2.5
Insurance rate,[a] %	0.5
Interest rate on bonded debt, %	9.0
Return on equity, %	
Utility	13.0
Industry	16.0
Debt/equity capitalization ratio	
Utility	53/47
Industry	30/70
Average escalation rate for capital constuction, %	7.0
Fixed-charge rates, %	
Utility	17
Industry	24
Annual capacity factor, %	70

[a] Based on initial capitalized investment.

Table 4.2. Fuel prices

(1980 dollars)

Year	Cost [$/GJ ($/10⁶ Btu)]		
	Coal[a]	Oil[b]	Gas[c]
1980	1.23 (1.30)	4.90 (5.17)	2.37 (2.50)
1990	1.65 (1.74)	6.59 (6.95)	6.59 (6.95)
2000	2.01 (2.12)	8.85 (9.34)	8.85 (9.34)
2010	2.45 (2.58)	12.00 (12.55)	12.00 (12.55)
2020	2.99 (3.15)	15.98 (16.86)	15.98 (16.86)
2030	3.64 (3.84)	21.48 (22.66)	21.48 (22.66)

[a] Coal at 3.3¢/kg ($30/ton) in 1980, escalating to 4.4¢/kg ($40/ton) in 1990 and 2%/year above inflation thereafter.

[b] Oil at $30/bbl in 1980, escalating at 3%/year above inflation thereafter.

[c] Natural gas at $2.37/GJ ($2.50/MBtu) in 1980, escalating to price of oil in 1990 (10.8%/year) and maintaining parity with oil thereafter.

Table 4.3. Uranium fuel cycle services[a]

(1980 dollars)

Service	
Conversion, $/kg	5
Enrichment, $/SWU[b]	100
PWR fuel fabrication, $/kg U	130
HTGR fuel fabrication, $/kg U	800
PWR spent fuel disposal, $/kg U	150
HTGR spent fuel disposal, $/kg U	500

[a] Uranium price at $88/kg ($40/lb) U_3O_8 in 1980, escalating at 2%/year above inflation rate. Price of other fuel cycle services constant in 1980 dollars.

[b] SWU—separative work unit.

Table 4.4. Electric generation costs[a]

[mills/kWh(e)]

	System		
	PWR[b]	HTGR[b]	Coal
	1990 Start-up		
Capital	13.3	14.1	11.0
O&M	2.6	2.4	3.5
Fuel	8.4	7.7	22.6
Total	24.3	24.2	37.1
	2000 Start-up		
Capital	14.8	15.6	12.2
O&M	2.6	2.4	3.5
Fuel	9.4	8.5	27.5
Total	26.8	26.5	43.2

[a] Based on utility economic assumptions. Levelized to remain constant in dollars of 1980 purchasing power and to increase at the general rate of inflation in nominal (dollars in any given year) terms.

[b] Based on a 1200-MW(e) plant.

[c] Based on a 800-MW(e) plant.

economic assumptions, fuel prices, electrical generation costs, process heat reactor capital costs, and nuclear process heat costs.

Constant 1980 dollars were used for the evaluation. The costs cited in this study will escalate at the general rate of inflation. All capital costs were based on 1980 prices, inflated to the start-up date by a 7% escalation rate for capital construction and then deflated at a 6% rate for general inflation back to 1980.

All system construction times (excluding nuclear reactors) were assumed to be five years with the center-of-gravity of expenditures at two years and 9.5% interest during construction. This assumption led to an allowance for funds used during construction (AFDC) of 20% of the total capital requirement.

Table 4.5. Capital investment costs for 1000-MW(t) process heat reactors

(1980 dollars)

Plant or component	Cost ($10^6)	
	1990 start-up	2000 start-up
PWR[a]	520.0	570.0
HTGR[a]	650.0	710.0
VHTR[a]	720.0	790.0
Intermediate heat exchanger (IHX)[a] adder	120.0	130.0
Electric capacity[b] adder	30.0	35.0

[a] Scale factor to other sizes is 0.48:

$$C = C_0(S/S_0)^{0.48} ,$$

where C is cost at size S and C_0 is cost at size S_0 [S_0 is 1000 MW(t) in this table].

[b] 100 MW(e).

Table 4.6. Nuclear process heat costs[a]

($/GJ)[b]

	Reactor system		
	PWR[c]	HTGR[d]	VHTR[e]
1990 start-up			
Capital[f]	2.18	2.62	2.87
O&M[g]	0.40	0.39	0.38
Fuel[h]	0.85	0.95	1.05
Total	3.43	3.96	4.30
2000 start-up			
Capital	2.39	2.86	3.13
O&M	0.40	0.38	0.39
Fuel	0.96	1.06	1.16
Total	3.75	4.30	4.68

[a] Based on utility economic assumptions, 1000 MW(t). Levelized to remain in dollars of constant 1980 purchasing power and to increase at the general rate of inflation in nominal (dollars in any given year) terms.

[b] Includes IHX, but excludes electric generation.

[c] Process steam temperature 340° C (650° F).

[d] Process steam temperature 540° C (1000° F).

[e] Process steam temperature 760–870° C (1400–1600° F).

[f] Scale factor to other sizes is -0.52. For use of this factor, see footnote *a*, Table 4.5.

[g] Scale factor to other sizes = -0.7.

[h] Scale factor to other sizes = -0.1.

4.2 Economic Considerations of CHPs

4.2.1 Major equipment

Major equipment costs, including the reformer, methanator, and heat exchanger, were based on the costs reported in General Electric's report, *Closed Loop Chemical Systems for Energy Storage and Transmission.*[1] General Electric costs were escalated to 1980 dollars using the Chemical Engineering Plant Cost Index for fabricated equipment.[2] These costs were then escalated for a start-up in 2000 based on factors given in Sect. 4.1.

No scaling factor was used for the reformer and heat exchangers, because costs were based on materials plus fabrication. A scaling factor of 0.6 was used for the methanator as recommended by General Electric.[1] Table 4.7 shows the equipment costs and installed capacity. The gas-preheat heat exchangers were priced at $\$1200/m^2$ and the steam generator heat exchangers at $\$1800/m^2$. With a start-up in 2000, a 1000-MW methanator had a base price of $\$39{,}200/m^3$ and the materials cost of the reformer was $2600 per tube (1980 dollars).

Table 4.7. Major equipment costs and installed capacity

	Heat delivered at methanator (MW)	Installed capacity (MW)	Cost for 2000 start-up[a] (1980 $/MW)
Reformer	800	1,200	54,300
	1,000	1,500	54,300
	1,200	1,800	54,300
Methanator	800	1,200	66,100
	1,000	1,500	60,500
	1,200	1,800	56,200
Heat exchangers[b] reformer plant	800	800	40,000
	1,000	1,000	40,000
	1,200	1,200	40,000
Methanator plant[c] gas preheat	800	47	212,000
	1,000	59	212,000
	1,200	71	212,000
Steam generation[d]	800	800	66,000
	1,000	1,000	66,000
	1,200	1,200	66,000

[a] Based on installed capacity.
[b] Heat transfer coefficient of 1.6 $kW/m^2\cdot{}^\circ C$ (280 $Btu/h\cdot ft^2\cdot{}^\circ F$).[1]
[c] Heat transfer coefficient of 0.28 $kW/m^2\cdot{}^\circ C$ (50 $Btu/h\cdot ft^2\cdot{}^\circ F$).[1]
[d] Heat transfer coefficient of 0.45 $kW/m^2\cdot{}^\circ C$ (80 $Btu/h\cdot ft^2\cdot{}^\circ F$).[1]

4.2.2 Compressors

The CHP process compressor costs were estimated as a function of power requirements. Because no sitework or housing would be required for these process compressors, costs were estimated at $891,400/MW ($665/hp). When possible, the transmission line compressors were also located at the reformer plant to eliminate sitework and housing costs. The location of these compressors at the plant would also allow the use of steam-turbine-driven pumps. However, this alternative was not considered in this study.

Pumping power costs for both process and transmission pumps were charged at $26.50/MWh. This corresponds to electricity generated by an HTGR with a start-up in 2000 (Sect. 4.1).

4.2.3 Operating costs

Process heat costs were calculated using the base costs and scaling factors presented in Sect. 4.1. Table 4.8 lists these calculated costs.

Nonfuel O&M costs at the reformer and methanation plants were assumed to be 8% of the on-site capital investment (before AFDC). Transmission line O&M costs are discussed in Sect. 4.2.

Table 4.8. Process heat costs

Heat delivered at methanator (MW)	Process heat cost ($/GJ)
800	3.84
1000	3.51
1200	3.28

Working capital was assumed to be 5% of the total capital investment (before AFDC). Interest charges on the working capital were estimated at 10%/year.

4.2.4 Electrical generation equipment

The electric capacity adder cost given in Table 4.5 was used with a scale factor of 0.9 to calculate the cost of this equipment at the reformer plant. The equipment was oversized by 20% for conservatism. Table 4.9 shows the installed capacity and cost of this equipment.

Table 4.9. Electrical generation equipment

Heat delivered at methanator (MW)	Installed electrical capacity [MW(e)]	Cost[a] ($$10^6$)
800	300	94
1000	375	115
1200	450	136

[a] In 1980 dollars with a start-up date of 2000.

4.2.5 Delivered energy cost

Tables 4.10 through 4.13 give costs for open- and closed-loop CHP systems. System costs are broken down into process heat charges, O&M, interest on working capital, and fixed charges (17% of the total capital investment).

The power generation credit reflects the fact that a CHP must be coupled with an electrical generation plant for maximum cycle efficiency. If the combination of the power required to compress

Table 4.10. Cost in 1980 dollars of open-loop 800-MW(t)-delivery CHP

	Pipe distance (km)		
	80	160	320
Power input, 10^6 GJ year	31.165	31.165	31.165
Thermal energy delivered, 10^6 GJ year	17.660	17.660	17.660
Net electricity delivered, 10^6 GJ, year	1.368	0.652	−0.781
Overall efficiency, %	61.1	58.8	54.2
Capital investment, \$ 10^6			
On-site	513.127	546.598	546.598
Off-site	132.686	266.551	606.158
Total	645.814	813.149	1152.756
Working capital (WC), \$ 10^6	26.909	33.881	48.032
Operating costs, \$ 10^6 year (\$/GJ)			
Process heat[a]	119.674 (6.78)	119.674 (6.78)	119.674 (6.78)
O&M	38.760 (2.19)	44.906 (2.54)	52.736 (2.99)
Interest on WC	2.691 (0.15)	3.338 (0.19)	4.803 (0.27)
Fixed charges	109.788 (6.22)	138.235 (7.83)	195.969 (11.10)
Total cost	270.913 (15.34)	306.203 (17.34)	373.181 (21.13)
Power generation, \$ 10^6/year (\$/GJ)			
Electrical generation credit,[b]	32.500 (1.84)	32.500 (1.84)	32.500 (1.84)
Pumping power	17.238 (0.98)	22.551 (1.27)	33.057 (1.87)
Plant power	5.200 (0.29)	5.200 (0.29)	5.200 (0.29)
Net electrical credit	10.062 (0.57)	4.790 (0.27)	5.757 (-0.33)
Net cost	260.851 (14.77)	301.414 (17.07)	379.229 (21.47)

[a] At \$5.44/GJ.
[b] At \$26.5/MWh.

Table 4.11. Cost in 1980 dollars of closed-loop 800-MW(t)-delivery CHP

	Pipe distance (km)		
	80	160	320
Power input, 10^6 GJ/year	31.165	31.165	31.165
Thermal energy delivered, 10^6 GJ/year	17.66	17.66	17.66
Net electricity delivered, 10^6 GJ/year	1.271	0.451	−1.073
Overall efficiency, %	60.1	58.1	53.2
Capital investment, \$ 10^6			
On-site	518.917	559.272	565.661
Off-site	173.404	347.984	779.418
Total	692.321	907.256	1345.079
WC, \$ 10^6	28.847	37.802	56.045
Operating costs, \$ 10^6/year (\$/GJ)			
Process heat[a]	119.674 (6.78)	119.674 (6.78)	119.674 (6.78)
O&M	40.245 (2.28)	47.950 (2.72)	58.662 (3.32)
Interest on WC	2.885 (0.16)	3.780 (0.21)	5.605 (0.32)
Fixed charges	117.695 (6.66)	154.234 (8.73)	228.663 (12.95)
Total cost	280.499 (15.88)	325.638 (18.44)	412.604 (23.36)
Power generation, \$ 10^6/year (\$/GJ)			
Electrical generation credit,[b]	32.500 (1.84)	32.500 (1.84)	32.500 (1.84)
Pumping power	17.946 (1.02)	23.929 (1.35)	35.187 (1.99)
Plant power	5.200 (0.29)	5.200 (0.29)	5.200 (0.29)
Net electrical credit	9.354 (0.53)	3.371 (0.19)	−7.887 (−0.45)
Net cost	271.145 (15.35)	322.267 (18.25)	420.491 (23.81)

[a] At \$5.44/GJ.
[b] At \$26.5/MWh.

Table 4.12. Cost in 1980 dollars of open-loop 1000-MW(t)-delivery CHP

	Pipe distance (km)		
	80	160	320
Power input, 10^6 GJ/year	38.956	38.956	38.956
Thermal energy delivered, 10^6 GJ/year	22.075	22.075	22.075
Net electricity delivered, 10^6 GJ. year	1.192	0.137	−1.969
Overall efficiency, %	59.7	57.0	51.6
Capital investment, \$ 10^6			
On-site	654.377	680.693	680.693
Off-site	132.686	291.941	666.492
Total	787.063	972.634	1347.185
WC, \$ 10^6	32.794	40.526	56.133
Operating costs, \$ 10^6/year (\$/GJ)			
Process heat[a]	136.736 (6.20)	136.736 (6.20)	136.736 (6.20)
O&M	48.177 (2.18)	53.846 (2.44)	61.676 (2.79)
Interest on WC	3.279 (0.15)	4.053 (0.18)	5.613 (0.25)
Fixed charges	133.801 (6.06)	165.348 (7.49)	229.021 (10.37)
Total cost	321.586 (14.59)	359.983 (16.31)	433.046 (19.62)
Power generation, 10^6/year (\$/GJ)			
Electrical generation credit,[b]	40.625 (1.84)	40.625 (1.84)	40.625 (1.84)
Pumping power	25.359 (1.15)	33.119 (1.50)	48.636 (2.20)
Plant power	6.500 (0.29)	6.500 (0.29)	6.500 (0.29)
Net electrical credit	8.766 (0.40)	1.006 (0.05)	−14.511 (−0.66)
Net cost	313.227 (14.19)	358.977 (16.26)	447.557 (20.27)

[a] At \$4.68/GJ.
[b] At \$26.5/MWh.

Table 4.13. Cost in 1980 dollars of closed-loop 1000-MW(t)-delivery CHP

	Pipe distance (km)		
	80	160	320
Power input, 10^6 GJ/year	38.956	38.956	38.956
Thermal energy delivered, 10^6 GJ/year	22.075	22.075	22.075
Net electricity delivered, 10^6 GJ/year	1.051	−0.144	−2.304
Overall efficiency, %	59.4	56.3	50.8
Capital investment, \$ 10^6			
On-site	662.504	696.949	703.132
Off-site	176.002	378.570	861.643
Total	838.506	1075.519	1564.775
WC, \$ 10^6	34.938	44.813	65.199
Operating costs, \$ 10^6/year (\$/GJ)			
Process heat[a]	136.736 (6.20)	136.736 (6.20)	136.736 (6.20)
O&M	49.883 (2.26)	57.256 (2.59)	68.364 (3.10)
Interest on WC	3.494 (0.16)	4.481 (0.20)	6.520 (0.30)
Fixed charges	142.546 (6.46)	182.838 (8.28)	266.050 (12.05)
Total cost	332.659 (15.07)	381.311 (17.27)	477.632 (21.64)
Power generation, \$ 10^6/year (\$/GJ)			
Electrical generation credit,[b]	40.625 (1.84)	40.625 (1.84)	40.625 (1.84)
Pumping power	26.403 (1.20)	35.204 (1.59)	51.084 (2.31)
Plant power	6.500 (0.29)	6.500 (0.29)	6.500 (0.29)
Net electrical credit	7.722 (0.35)	−1.079 (−0.05)	−16.959 (−0.76)
Net cost	324.937 (14.72)	382.390 (17.32)	494.591 (22.41)

[a] At \$4.68/GJ.
[b] At \$26.5/MWh.

and transport the gas and the auxiliary power is greater than the power generated, the net electrical credit is negative.

The bottom line of each table (4.10 through 4.13) shows the net cost of transporting energy both on an annual and on a dollar per gigajoule basis. These costs vary from about $13 to $20/GJ. This energy is delivered to the customer in the form of steam or hot water. Cycle energy input, output, and overall efficiency are given at the top of each table. The efficiencies ranged from about 50 to 60%.

Figure 4.1 shows the net delivered cost as a function of distance for the four cases described in Tables 4.10 through 4.13.

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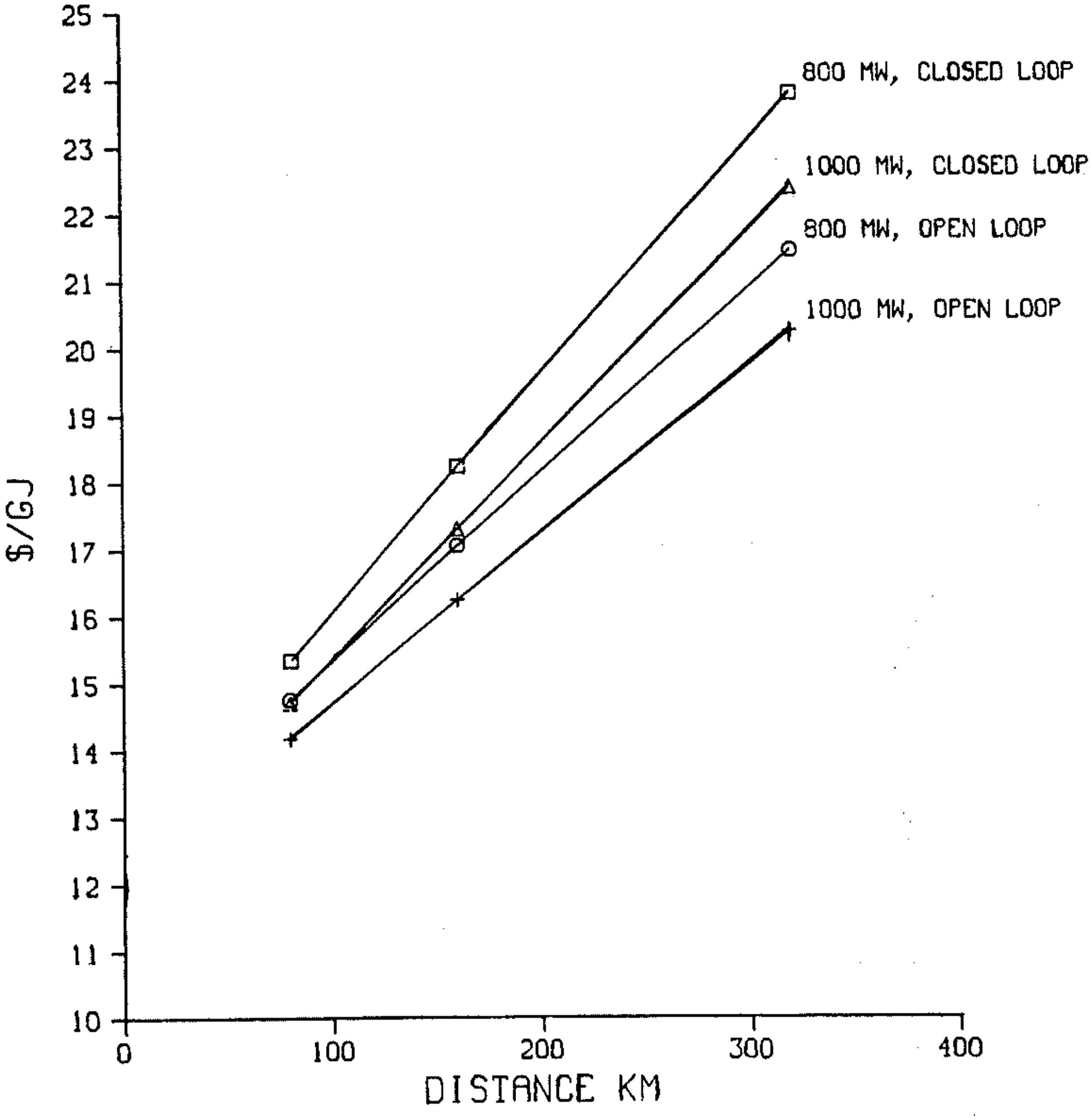

Fig. 4.1. Chemical heat pipes, 2000 start-up, 1980 dollars.

4.3 MBG from Coal

4.3.1 Capital investment

All MBG capital costs are based on plant costs reported in *Analysis of Industrial Markets for Low and Medium Btu Coal Gasification.*[3] These plant costs are reported in 1979 dollars and are based on cost information contained in non-site-specific quotations provided by process developers and licensees.

To make these cost estimates consistent with other components of this study, several adjustments were made. The load factor (or capacity factor) was changed to 70%. The plant size was scaled to correspond to energy outputs of 800, 1000, and 1200 MW. A scale factor of 0.8 was used in adjusting the capital costs to reflect the new plant sizes. The costs were escalated to 1980 dollars.

Start-up costs were scaled in direct proportion to plant size. Interest during construction was increased to reflect the additional capital cost of compression equipment (Sect. 4.3.3).

4.3.2 Coal cost levelization

As shown in Sect. 4.1, coal is estimated to escalate at a rate 2%/year above inflation. To determine a price in constant 1980 dollars that will increase at the rate of general inflation, making this cost uniform over the 30-year life of the plant, is necessary. This is done by making the present value of the projected costs equal to the present value of the projected revenues (where the price is defined in constant 1980 dollars that increase at the general inflation rate). The present value of the projected cost is

$$PV_c = \sum_{n=1}^{30} \frac{C_o(1 + i + 0.02)^n}{(1 + x)^n} ,$$

where

PV_c = present value of costs,
C_o = 1990 (for 1990 start-up) cost of coal in 1980 dollars (2000 cost for 2000 start-up),
i = inflation rate,
0.02 = cost rise above inflation (%/year),
x = effective discount rate,
n = index starting at end of first year of project and continuing through final year.

The present value of the projected revenue is:

$$PV_R = \sum_{n=1}^{30} \frac{\overline{P_o}(1 + i)^n}{(1 + x)^n} ,$$

where

PV_R = present value of the revenues,
$\overline{P_o}$ = price in constant 1980 dollars.

By setting $PV_c = PV_R$ and solving for $\overline{P_o}$:

$$\overline{P_o} = \left[C_o \sum_{n=1}^{30} \frac{(1 + i + 0.02)^n}{(1 + x)^n} \right] \div \left[\sum_{n=1}^{30} \frac{(1 + i)^n}{(1 + x)^n} \right] = C_o(1.29) .$$

This gives a levelized cost in constant 1980 dollars of \$51.70 and \$62.30/ton for a 1990 and a 2000 start-up, respectively, using the coal prices reported in Sect. 4.1. These prices correspond to \$2.13 and \$2.60/GJ, respectively.

Nuclear process heat was not evaluated as an energy source for coal gasification because its price was higher than that of coal over the time period considered (Tables 4.2 and 4.6).

4.3.3 Compression costs

The plant costs reported in *Analysis of Industrial Markets for Low and Medium Btu Coal Gasification*[3] did not allow for gas compression. Compression costs were added in two steps for this evaluation.

The compressor capital costs were calculated on a power basis as described in Sect. 4.2. Compressor site and housing costs were not added because the compressors would be located at the gasification plant.

The compressors were assumed to be gas-turbine driven with MBG as a fuel. This compression fuel was charged at \$6.60 and \$8.80/GJ (\$7 and \$9/10^6 Btu) for 1990 and 2000 start-ups, respectively. This approximates the cost of the gas before compression.

4.3.4 Operating costs

Annual O&M costs were assumed to equal 8% of the total plant costs. Working capital was estimated at 5% of the plant costs, excluding interest during construction and start-up costs. Interest on working capital was assumed to be 10%/year.

4.3.5 Delivered energy cost

Coal gasification plant characteristics are listed in Table 4.14 for three different energy delivery rates. Because a 70% capacity factor was assumed, the plant output is less than the plant capacity.

Tables 4.15 through 4.17 give costs for energy delivered from three different sized plants at three different pressures at three different distances. These costs range from about \$10 to \$13/GJ. Process gas compressors are included in the plant investment costs; compressors needed to transport the gas are included in the gas pumping costs.

Operating costs are broken down according to feedstock, compression fuel, O&M, interest on working capital, and fixed charges. The compression fuel cost represents the cost of the gas burned to power the initial gas compressors.

Figure 4.2 shows the cost of delivered MBG at a pressure of 0.9 MPa (1990 start-up) as a function of transport distance and energy delivery rate. Figure 4.3 shows the cost of delivered gas at a rate of 1200

Table 4.14. MBtu coal gasification plant characteristics

Energy delivery rate[a] [MW(t)]	Plant capacity (10^6 scm/d)	Plant output		Coal input capacity (tons/d)	Coal input[b] (tons/year)
		10^6 scm/year	10^6 GJ/year		
800	6.07	1,550	17.66	4,570	1,168,450
1,000	7.59	1,940	22.08	5,720	1,460,570
1,200	9.11	2,330	26.49	6,860	1,752,680

[a] Based on lower heating value of MBG.

[b] Coal heating value of 24 GJ/ton (11,600 Btu/lb).

Table 4.15. MBG[a] via coal gasification

(1980 dollars)

	Start-up date	
	1990	2000
800-MW(t) delivery capacity		
Investment, $ 10^6		
Plant	327.894	360.861
WC	13.051	18.043
Operating costs, $ 10^6/year ($/GJ)		
Coal[b]	60.759 (3.44)	72.444 (4.10)
Compression fuel[c]	16.657 (0.94)	21.416 (1.21)
O&M (8% plant)	26.232 (1.49)	28.869 (1.63)
Interest on WC (10%)	1.305 (0.07)	1.804 (0.10)
Fixed charges (17% plant)	55.742 (3.16)	61.346 (3.47)
Total operating costs	160.695 (9.10)	185.879 (10.52)
Transportation costs, $/GJ		
80 km	0.46	0.50
160 km	0.89	0.98
320 km	1.75	1.93
Total delivered cost, $/GJ		
80 km	9.56	11.02
160 km	9.99	11.50
320 km	10.85	12.45
1000-MW(t) delivery capacity		
Investment, $ 10^6		
Plant	395.060	433.448
WC	15.709	21.672
Operating costs, $ 10^6/year ($/GJ)		
Coal[b]	75.950 (3.44)	90.555 (4.10)
Compression fuel[c]	20,576 (0.93)	26.455 (1.20)
O&M (8% plant)	31.605 (1.43)	34.676 (1.57)
Interest on WC (10%)	1.571 (0.07)	2.167 (0.10)
Fixed charges (17% plant)	67.160 (3.04)	78.686 (3.34)
Total operating costs	196.862 (8.92)	227.539 (10.31)
Transportation costs, $/GJ		
80 km	0.41	0.45
160 km	0.80	0.88
320 km	1.58	1.75
Total delivered cost, $/GJ		
80 km	9.33	10.76
160 km	9.72	11.19
320 km	10.50	12.06
1200-MW(t) delivery capacity		
Investment, $ 10^6		
Plant	470.782	503.066
WC	18.717	25.153
Operating costs, $ 10^6/year ($/GJ)		
Coal[b]	91.139 (3.44)	108.666 (4.10)
Compression fuel[c]	24.589 (0.93)	31.615 (1.19)
O&M (8% plant)	37.663 (1.42)	40.245 (1.52)
Interest on WC (10%)	1.872 (0.07)	2.515 (0.09)
Fixed charges (17% plant)	80.033 (3.02)	85.521 (3.23)
Total operating costs	235.296 (8.88)	268.562 (10.14)

Table 4.15 (continued)

	Start-up date	
	1990	2000
Transportation costs, $/GJ		
80 km	0.38	0.41
160 km	0.74	0.81
320 km	1.46	1.61
Total delivered cost, $/GJ		
80 km	9.26	10.55
160 km	9.62	10.95
320 km	10.34	11.75

[a] Pressure at 0.9 MPa (130 psia).
[b] At $52/ton (1990), at $62/ton (2000), 11,600 Btu/lb.
[c] At $6.6/GJ (1990), at $8.5/GJ (2000).

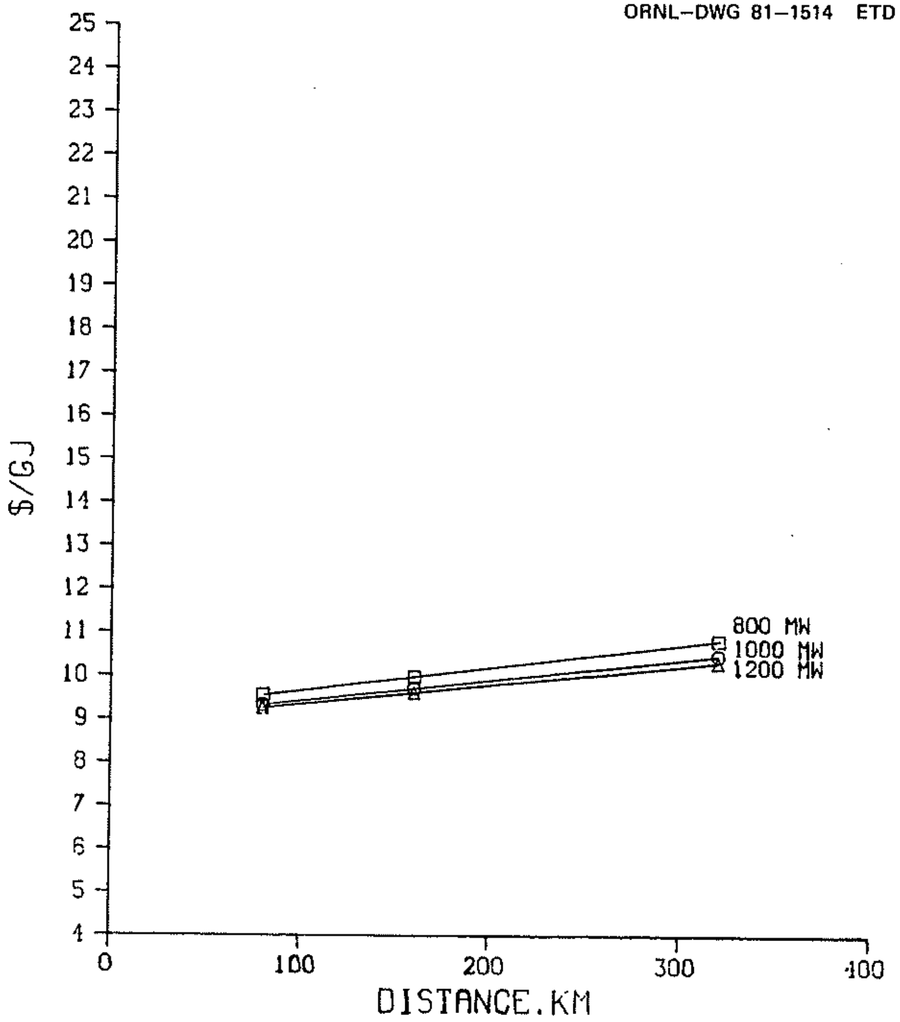

Fig. 4.2. MBG from coal, 0.9 MPa, 1990 start-up, 1980 dollars.

Table 4.16. MBG[a] via coal gasification

(1980 dollars)

	Start-up date	
	1990	2000
800-MW(t) delivery capacity		
Investment, \$ 10^6		
Plant	346.961	381.464
WC	13.829	19.073
Operating costs, \$ 10^6/year (\$/GJ)		
Coal[b]	60.759 (3.44)	72.444 (4.10)
Compression fuel[c]	25.650 (1.45)	32.978 (1.87)
O&M (8% plant)	27.757 (1.57)	30.517 (1.73)
Interest on WC (10%)	1.383 (0.08)	1.907 (0.11)
Fixed charges (17% plant)	58.983 (3.34)	64.849 (3.67)
Total operating costs	174.532 (9.88)	202.695 (11.48)
Transportation costs, \$/GJ		
80 km	0.41	0.45
160 km	0.79	0.87
320 km	1.56	1.72
Total delivered cost, \$/GJ		
80 km	10.29	11.93
160 km	10.67	12.35
320 km	11.44	13.20
1000-MW(t) delivery capacity		
Investment, \$ 10^6		
Plant	418.123	459.009
WC	16.650	22.950
Operating costs, \$ 10^6/year (\$/GJ)		
Coal[b]	75.950 (3.44)	90.555 (4.10)
Compression fuel[c]	31.639 (1.43)	40.785 (1.85)
O&M (8% plant)	33.450 (1.52)	36.721 (1.66)
Interest on WC (10%)	1.665 (0.08)	2.295 (0.10)
Fixed charges (17% plant)	71.081 (3.22)	78.031 (3.53)
Total operating costs	213.785 (9.68)	248.387 (11.25)
Transportation costs, \$/GJ		
80 km	0.37	0.40
160 km	0.71	0.78
320 km	1.40	1.55
Total delivered cost, \$/GJ		
80 km	10.05	11.65
160 km	10.39	12.03
320 km	11.08	12.80
1200-MW(t) delivery capacity		
Investment, \$ 10^6		
Plant	498.208	533.462
WC	19.835	26.673
Operating costs, \$ 10^6/year (\$/GJ)		
Coal[b]	91.139 (3.44)	108.666 (4.10)
Compression fuel[c]	37.765 (1.43)	48.556 (1.83)
O&M (8% plant)	39.857 (1.50)	42.677 (1.61)
Interest on WC (10%)	1.984 (0.07)	2.667 (0.10)
Fixed charges (17% plant)	84.695 (3.20)	90.688 (3.42)
Total operating costs	255.440 (9.64)	293.254 (11.07)

Table 4.16 (continued)

	Start-up date	
	1990	2000
Transportation costs, $/GJ		
80 km	0.33	0.37
160 km	0.65	0.72
320 km	1.29	1.42
Total delivered cost, $/GJ		
80 km	9.97	11.44
160 km	10.29	11.79
320 km	10.93	12.49

[a] Pressure at 1.8 MPa (260 psia).
[b] At $52/ton (1990), at $62/ton (2000), 11,600 Btu/lb.
[c] At $6.6/GJ (1990), at $8.5/GJ (2000).

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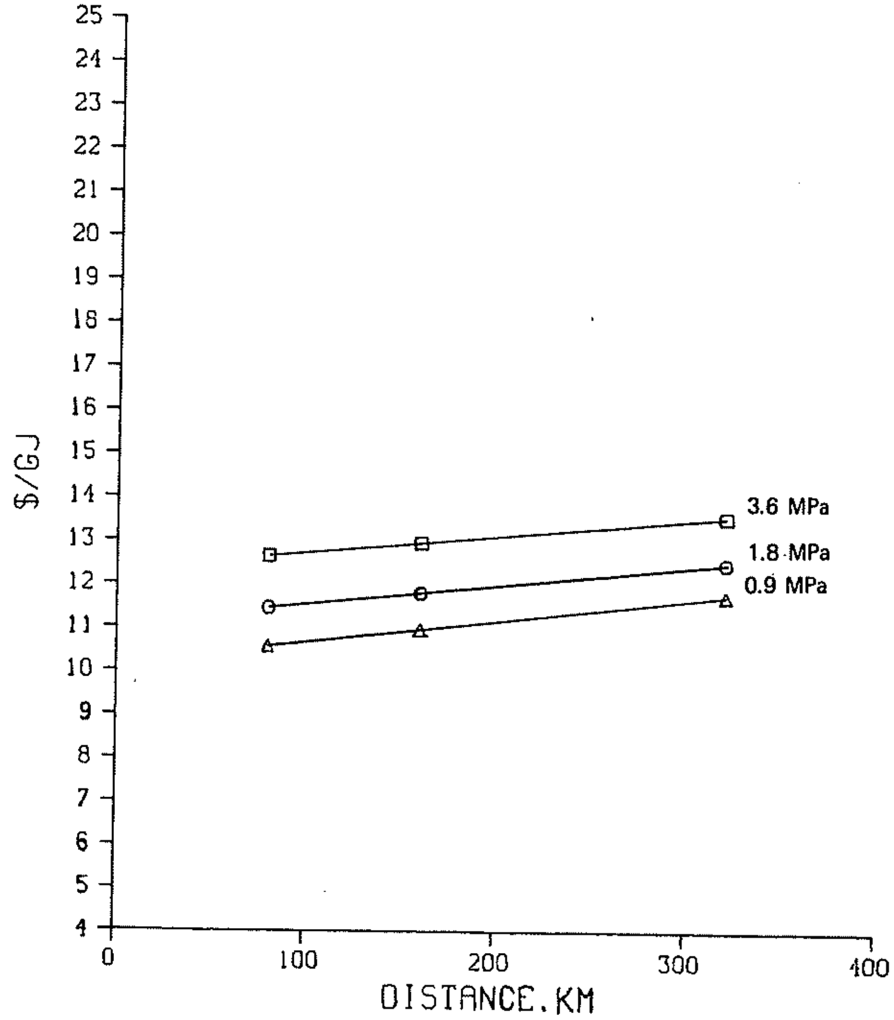

Fig. 4.3. MBG from coal, 1200 MW, 2000 start-up, 1980 dollars.

Table 4.17. MBG[a] via coal gasification

(1980 dollars)

	Start-up date	
	1990	2000
800-MW(t) delivery capacity		
Investment, \$ 10^6		
Plant	372.085	409.310
WC	14.853	20.466
Operating costs, \$ 10^6/year (\$/GJ)		
Coal[b]	60.759 (3.44)	72.444 (4.10)
Compression fuel[c]	37.721 (2.14)	48.498 (2.75)
O&M (8% plant)	29.767 (1.69)	32.745 (1.85)
Interest on WC (10%)	1.485 (0.08)	2.047 (0.12)
Fixed charges (17% plant)	63.254 (3.58)	69.583 (3.94)
Total operating costs	192.986 (10.93)	225.317 (12.76)
Transportation costs, \$/GJ		
80 km	0.37	0.41
160 km	0.71	0.79
320 km	1.40	1.55
Total delivered cost, \$/GJ		
80 km	11.30	13.17
160 km	11.64	13.55
320 km	12.33	14.31
1000-MW(t) delivery capacity		
Investment, \$ 10^6		
Plant	449.110	493.353
WC	17.913	24.668
Operating costs, \$ 10^6/year (\$/GJ)		
Coal[b]	75.950 (3.44)	90.555 (4.10)
Compression fuel[c]	46.554 (2.11)	59.855 (2.71)
O&M (8% plant)	35.929 (1.63)	39.468 (1.79)
Interest on WC (10%)	1.791 (0.08)	2.467 (0.11)
Fixed charges (17% plant)	76.345 (3.46)	83.870 (3.80)
Total operating costs	236.569 (10.72)	276.215 (12.51)
Transportation costs, \$/GJ		
80 km	0.33	0.36
160 km	0.63	0.70
320 km	1.25	1.38
Total delivered cost, \$/GJ		
80 km	11.05	12.87
160 km	11.35	13.21
320 km	11.97	13.89
1200-MW(t) delivery capacity		
Investment, \$ 10^6		
Plant	535.058	574.303
WC	21.338	28.715
Operating costs, \$ 10^6/year (\$/GJ)		
Coal[b]	91.139 (3.44)	108.666 (4.10)
Compression fuel[c]	55.307 (2.09)	71.109 (2.68)
O&M (8% plant)	42.805 (1.62)	45.944 (1.73)
Interest on WC (10%)	2.134 (0.08)	2.872 (0.11)
Fixed charges (17% plant)	90.960 (3.43)	97.632 (3.69)
Total operating costs	282.345 (10.66)	326.223 (12.31)

Table 4.17 (continued)

	Start-up date	
	1990	2000
Transportation costs, \$/GJ		
80 km	0.30	0.33
160 km	0.58	0.64
320 km	1.14	1.26
Total delivered cost, \$/GJ		
80 km	10.96	12.64
160 km	11.24	12.95
320 km	11.80	13.57

[a] Pressure at 3.6 MPa (520 psia).
[b] At \$52/ton (1990), at \$62/ton (2000), 11,600 Btu/lb.
[c] At \$6.6/GJ (1990), at \$8.5/GJ (2000).

MW (2000 start-up) as a function of distance and pressure. Increasing the pressure from 0.9 to 3.6 MPa increases the cost by 20% because the compression is an additional cost at the plant, and the additional transport costs are relatively small.

4.4 Hydrogen Economics[4]

4.4.1 General

Four hydrogen production processes were considered: (1) coal gasification, (2) KOH electrolysis, (3) SPE, and (4) SPE with storage. The hydrogen production cost estimates are based on information from *Production Economics for Hydrogen, Ammonia, and Methanol During the 1980–2000 Time Period,* by Exxon Research and Engineering Company.[4]

For this study, an assumption was made that any air separation plant required would be part of the on-site capital cost. The 1980 investment for oxygen plants was estimated as a function of capacity. The estimated utility requirements per 900 kg (1 ton) of oxygen required are

	99.5% O_2	
	At 3.4 MPa	At 0.3 MPa
Power, kWh	435	325
Cooling water, 10^3 m^3	138	138
Process water, 10^3 m^3	4.5	4.5

In the coal-based plants, larger power uses have been put on turbine drive, with the net required high-pressure steam generated in coal-fired boilers. Extraction turbines are used to provide any medium-pressure steam required. The heat rate used to convert high-pressure steam to electrical power is 10,300 Btu/kWh, and the boiler efficiency is 85% on higher heating value (HHV) for the large boilers required.

The electrolysis process for the production of hydrogen requires no steam.

For the coal-based plants, off-site costs have been estimated as the sum of the following:

1. 30% of on-site cost (includes costs such as storage, waste disposal facilities, and general and administrative facilities);
2. steam system capital cost (includes stack gas scrubbing equipment plus ash and dust collection equipment for coal-based plants);
3. cooling water system capital cost (provides a complete system including cooling tower, circulatory pumps, additive injection).

The amount of working capital usually depends on arrangements for payments on feedstock purchases and product sales and on storage volumes maintained for both. In the cost analysis for this study, one-month's storage of major feedstocks at cost (excluding gaseous feedstocks) and one-month's accounts receivables were used to estimate working capital. Working capital is considered as borrowed capital; only interest charges, at 10%, are included in the overall production costs.

Utility costs for coal gasification were estimated for three different areas in the United States, but only the Gulf Coast area data considered by this study are reported in Table 4.18.

Table 4.18. Utility costs for coal gasification, Gulf Coast area

(1980 dollars)

Montana/Wyoming coal (delivered), \$/GJ	1.46
Power (from mix of fuels), \$/kWh	0.027
Cooling water, \$/m^3	0.014
Steam at 4.1 MPa, 400°C, (600 psig, 750°F) from oil, \$/10^3 kg	11.00
Steam at 10.3 MPa and 480°C (1500 psig, 900°F) from oil, \$/10^3 kg	11.50
Boiler feed water, \$/m^3	0.22
Process water, \$/m^3	0.15

Annual salaries for operating labor were estimated for 1980 as $21,500 for shift operators, $24,200 for shift foremen, and $28,800 for supervisors. The total number of staff has been estimated by multiplying the men per shift by 4.6 and rounding to the next highest whole number. This assumes a 40-h average work week and 48 work weeks per year.

Production cost analysis has been divided into labor-related and capital-related categories to allow for easy comparison among alternate technologies. Labor-related costs, in addition to direct labor charges, include a part of maintenance and general plant overhead. These costs have been assumed at 100% of direct labor costs.

Assumed percentages for capital-related expenses are (1) 4.0% of on-site capital cost (without AFDC) for maintenance, (2) 2.6% of on-site capital cost (without AFDC) for general plant overhead, (3) 10% of working capital for interest, and (4) 17% of total investment (on-site plus off-site) for fixed-charge rate (consistent with Sect. 4.1).

Contingency factors that were applied to the capital cost (not including AFDC) estimates are 1.15 for coal gasification (K-T) and 1.30 for electrolysis (SPE).

Finding a market for the oxygen by-product from electrolysis plants is unlikely, particularly for plants operating with intermittent off-peak electricity. However, if the oxygen could be sold at a price of $24/ton, the oxygen credit would correspond to $1.46/GJ of hydrogen.

The Exxon 1980 cost estimates were adjusted for 1990 and 2000 start-ups using the escalation rates shown in Sect. 4.1. The original estimates for insurance, property taxes, depreciation, and return on investment were replaced with the comprehensive fixed-charge rate of 17% (Sect. 4.1) for consistency within this evaluation.

Outputs and capacity factors of the various processes were adjusted to achieve outputs of 800, 1000, and 1200 MW. The scaling factor for capital expenses (except for the SPE with storage process) was derived by dividing the original assumed output by the original capacity factor and multiplying this by a capacity factor of 0.7. The ratio of the desired output divided by this adjusted output was then raised to the 0.8 power to give an overall capital scaling factor. Noncapital expenses such as utilities and labor were assumed to be directly proportional to the energy output.

The SPE with storage capital scaling factor was determined by taking the ratio of desired output to original output to the 0.8 power [$(\text{desired}/\text{original})^{0.8}$]. No adjustment for capacity factor was made for this process.

4.4.2 Coal gasification

The Exxon report[4] presents cost estimates for a variety of locations and coal sources. This evaluation considered only the Gulf Coast location with a supply of Wyoming coal (Table 4.18). This Gulf Coast location was judged appropriate based on the market study in Sect. 2. The cost estimates would be lower for other possible locations. The Exxon report[4] also gives cost estimates for a "new" coal gasification process as well as for the proven K-T and Lurgi gasifiers. The estimates presented in this evaluation are for the K-T process as reported by Exxon.[4]

Coal prices were made uniform over the plant life as described in Sect. 4.3.2. The base 1980 price used was \$1.46/GJ (\$1.54/10^6 Btu) as reported by Exxon for Wyoming coal delivered to the Gulf Coast.

The coal gasification process under consideration used Wyoming coal at a Gulf Coast location, and the pricing was based on a K-T gasifier. Table 4.19 presents investment, operating, and pumping costs for three different sized plants, three gas transport distances, and two start-up dates—1990 and 2000. This hydrogen, delivered at a pressure of 3.6 MPa (520 psi), costs around \$12 to \$14/GJ. Figure 4.4 shows the variation in energy price as a function of distance and energy delivery rate for a 2000 start-up.

4.4.3 Electrolysis

Current state-of-the-art water electrolysis plants require very large investments. Although electricity is the highest cost of electrolytic hydrogen, costs associated with the plant investment are very significant. Costs included in this study for current state-of-the-art electrolysis facilities are indicative of the equipment supplied by Lurgi or Teledyne but are not intended to relate to a specific process.

The SPE process involves costs similar to those suggested for the General Electric Solid Polymer Electrolyte process, but such an improvement in costs possibly could be achieved by other groups conducting research and development (R&D) on electrolysis.

Hydrogen produced intermittently could be supplied continuously to a hydrogen-consuming operation if hydrogen storage were imposed between the production and consumption operations. Hydrogen storage using iron-titanium alloy has been studied extensively by Brookhaven National Laboratory (BNL).[5] The costs of storage used in this study were calculated on the basis of this BNL information.

Table 4.19. Hydrogen cost via coal gasification, Wyoming coal, Gulf Coast, K-T process[a]

	Start-up date	
	1990	2000
800-MW(t) delivery capacity		
Investment, \$ 10^6		
On-site	375.1	412.1
Off-site	255.0	280.1
Total[b]	630.1	692.2
WC	8.3	8.3
Operating cost, \$ 10^6/year (\$/GJ)		
Feedstock[c]	74.679 (4.23)	89.989 (5.10)
Utilities	4.353 (0.25)	4.353 (0.25)
Labor and supervision	2.506 (0.14)	2.506 (0.14)
Maintenance (4% on-site)	12.503 (0.71)	13.734 (0.78)
Plant overhead (2.6% on-site)	8.127 (0.46)	8.927 (0.51)
Interest on WC (10%)	0.830 (0.05)	0.830 (0.05)
Fixed charges (17% of total plant)	107.120 (6.07)	117.667 (6.66)
Total	210.112 (11.90)	238.002 (13.48)
Sulfur credit	[0.502 (0.03)]	[0.502 (0.03)]
Net cost	209.610 (11.87)	237.500 (13.45)
Transportation cost,[d] \$/GJ		
80 km	0.28	0.31
160 km	0.54	0.59
320 km	1.05	1.15
Total delivered cost, \$/GJ		
80 km	12.15	13.76
160 km	12.41	14.04
320 km	12.92	14.60
1000-MW(t) delivery capacity		
Investment, \$ 10^6		
On-site	448.2	492.4
Off-site	304.7	334.7
Total[b]	752.9	827.1
WC	10.4	10.4
Operating cost, \$ 10^6/year (\$/GJ)		
Feedstock[c]	93.349 (4.23)	112.486 (5.10)
Utilities	5.442 (0.25)	5.442 (0.25)
Labor and supervision	3.133 (0.14)	3.133 (0.14)
Maintenance (4% on-site)	14.941 (0.68)	16.412 (0.74)
Plant overhead (2.6% on-site)	9.712 (0.44)	10.668 (0.48)
Interest on WC (10%)	1.040 (0.05)	1.040 (0.05)
Fixed charges (17% of total plant)	127.990 (5.80)	140.597 (6.37)
Total	255.614 (11.58)	289.786 (13.13)
Sulfur credit	[0.628 (0.03)]	[0.628 (0.03)]
Net cost	254.986 (11.55)	289.158 (13.10)
Transportation cost,[d] \$/GJ		
80 km	0.24	0.26
160 km	0.46	0.51
320 km	0.91	1.00
Total delivered cost, \$/GJ		
80 km	11.79	13.36
160 km	12.01	13.61
320 km	12.46	14.10

Table 4.19 (continued)

	Start-up date	
	1990	2000
1200-MW(t) delivery capacity		
Investment, \$ 10^6		
On-site	518.6	569.6
Off-site	352.6	387.2
Total[b]	871.2	956.8
WC	12.5	12.5
Operating cost, \$ 10^6/year (\$/GJ)		
Feedstock[c]	112.018 (4.23)	134.983 (5.10)
Utilities	6.531 (0.25)	6.531 (0.25)
Labor and supervision	3.760 (0.14)	3.760 (0.14)
Maintenance (4% on-site)	17.287 (0.65)	18.989 (0.72)
Plant overhead (2.6% on-site)	11.236 (0.42)	12.343 (0.47)
Interest on WC (10%)	1.250 (0.05)	1.250 (0.05)
Fixed charges (17% of total plant)	148.104 (5.61)	162.670 (6.16)
Total	300.180 (11.37)	340.537 (12.90)
Sulfur credit	[0.753 (0.03)]	[0.753 (0.03)]
Net cost	299.427 (11.34)	339.784 (12.87)
Transportation cost,[d] \$/GJ		
80 km	0.22	0.24
160 km	0.42	0.46
320 km	0.82	0.90
Total delivered cost, \$/GJ		
80 km	11.56	13.11
160 km	11.76	13.33
320 km	12.16	13.77

[a] Plant specifications given in Sect. 3.4.

[b] 15% contingency added.

[c] Coal cost at plant in 1990, \$1.96/GJ; in 2000, \$2.39/GJ, coal heating value of 0.027 GJ/kg (11,600 Btu/lb).

[d] Pressure at 3 MPa (450 psia).

Hydrogen energy costs for an electrolysis (KOH) system are listed in Table 4.20 for three energy delivery rates, three transport distances, and two start-up dates. Figure 4.5 gives KOH hydrogen costs as a function of distance and energy delivery rate for a 1990 start-up.

SPE hydrogen costs are given in Table 4.21 for the same conditions. Figure 4.6 shows that SPE hydrogen costs are relatively insensitive to delivery rate and transport distance.

A process configuration that combines SPE hydrogen production with an iron titanium storage system is used to derive the hydrogen energy costs shown in Table 4.22 and Fig. 4.7. This system is slightly more sensitive to energy delivery rate than the straight SPE system because it is a more capital-intensive process.

Figures 4.8 and 4.9 contrast the various hydrogen production alternatives for 1990 and 2000 start-ups. The SPE system was not analyzed for a 1990 start-up because it is still a developing technology.

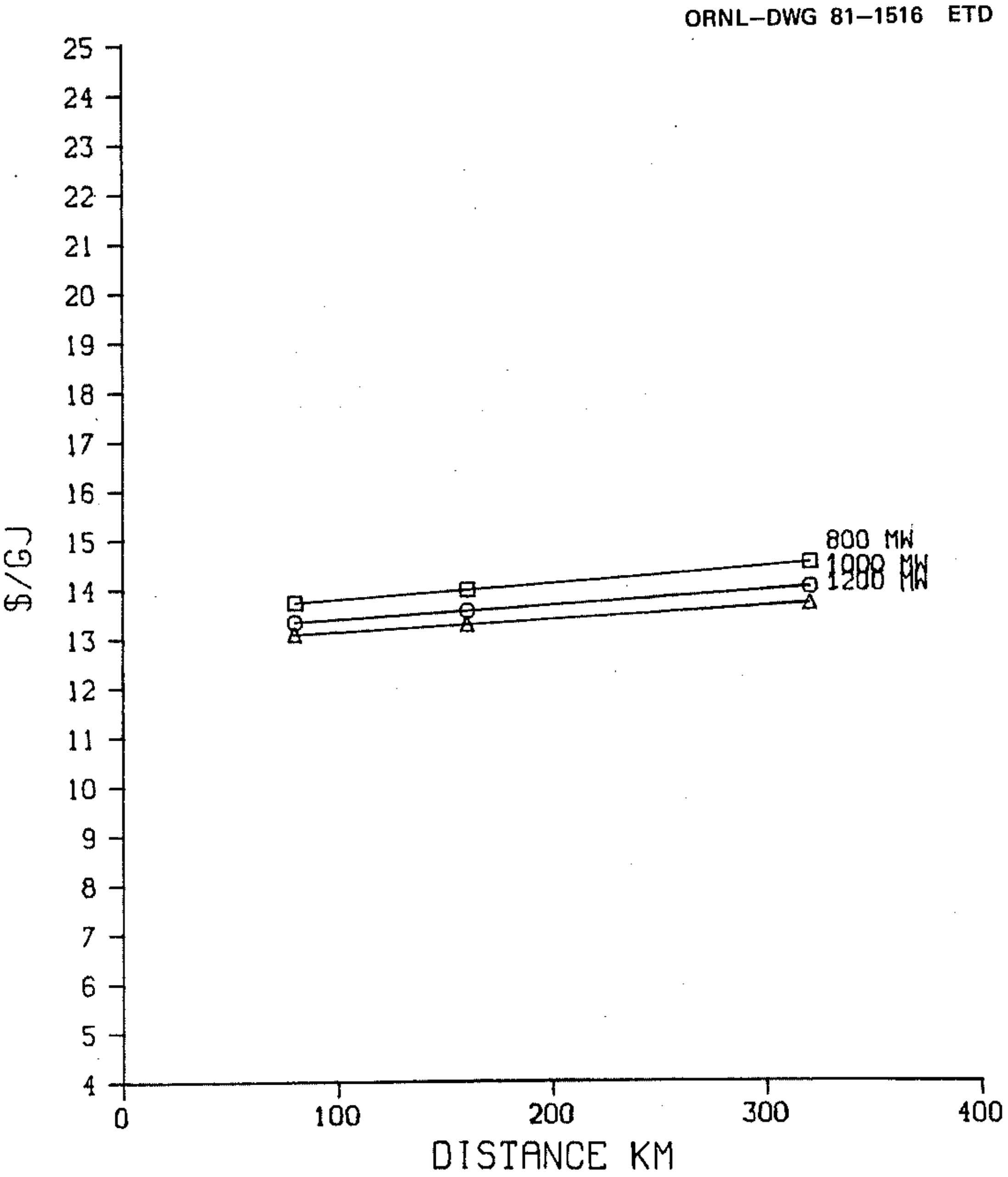

Fig. 4.4. Coal gasification hydrogen production, 3 MPa, 2000 start-up, 1980 dollars.

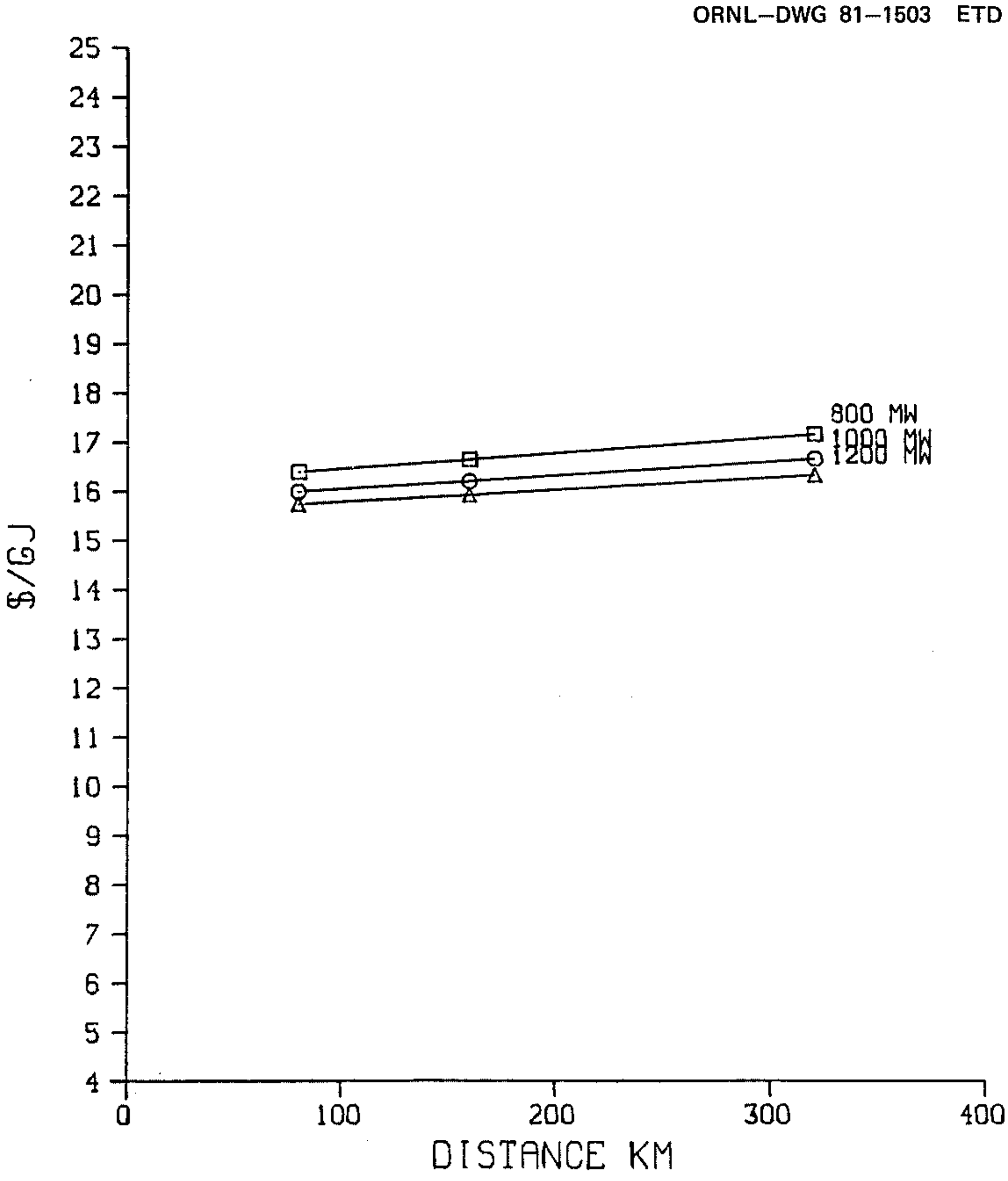

Fig. 4.5. KOH electrolytic hydrogen production, 3 MPa, 1990 start-up, 1980 dollars.

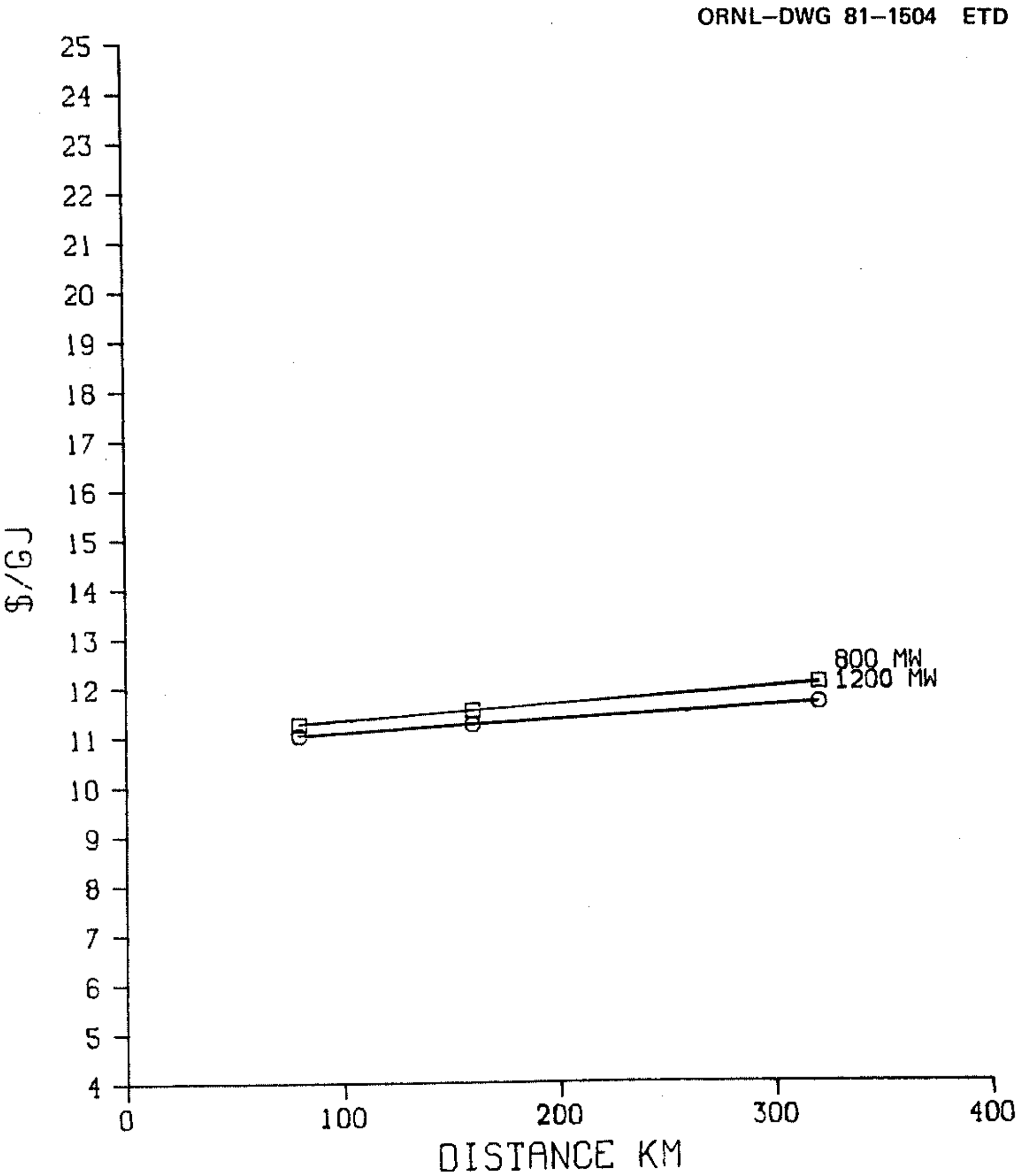

Fig. 4.6. SPE hydrogen production, 3 MPa, 2000 start-up, 1980 dollars.

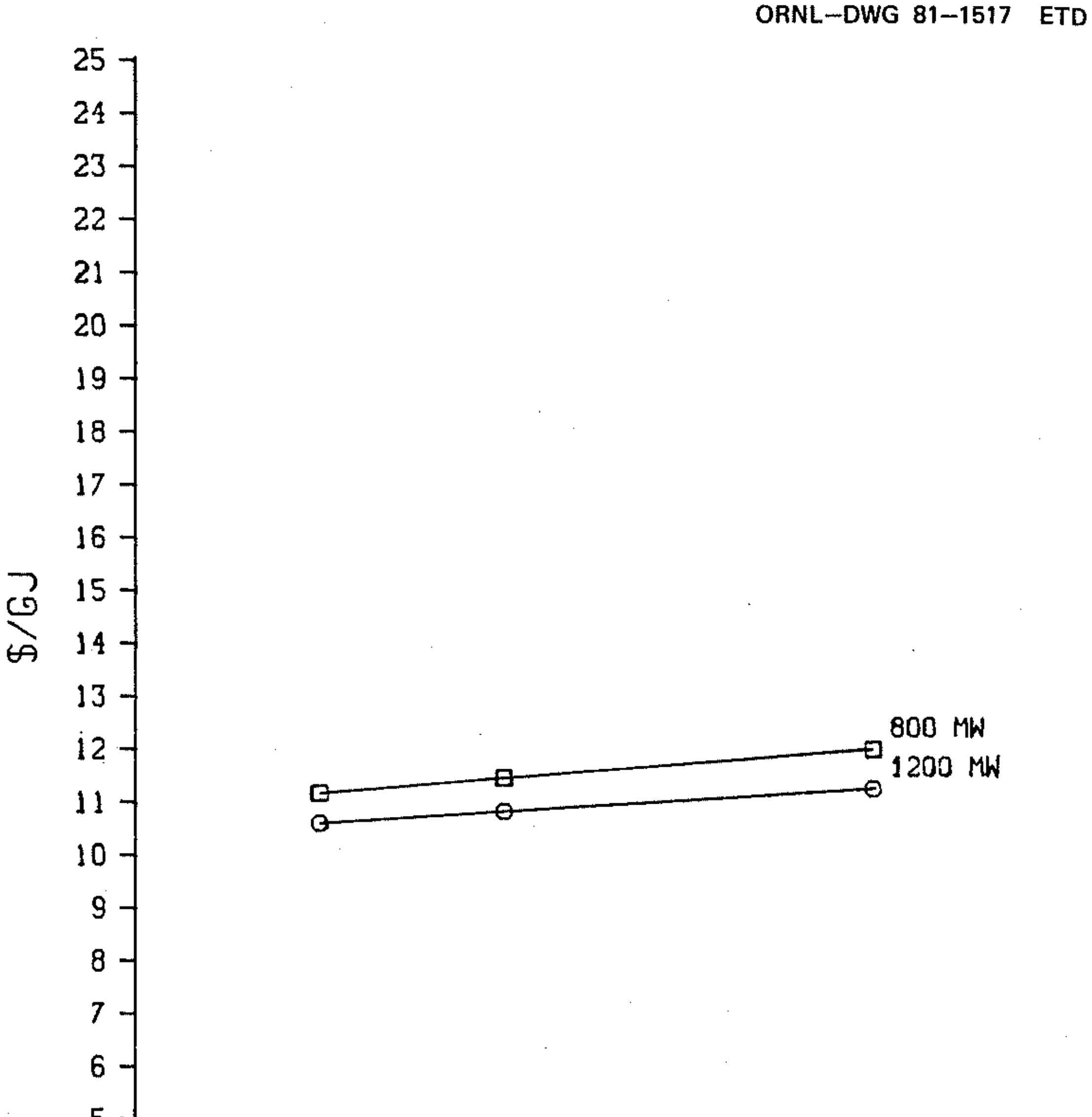

Fig. 4.7. SPE hydrogen production with storage, 3 MPa, 2000 start-up, 1980 dollars.

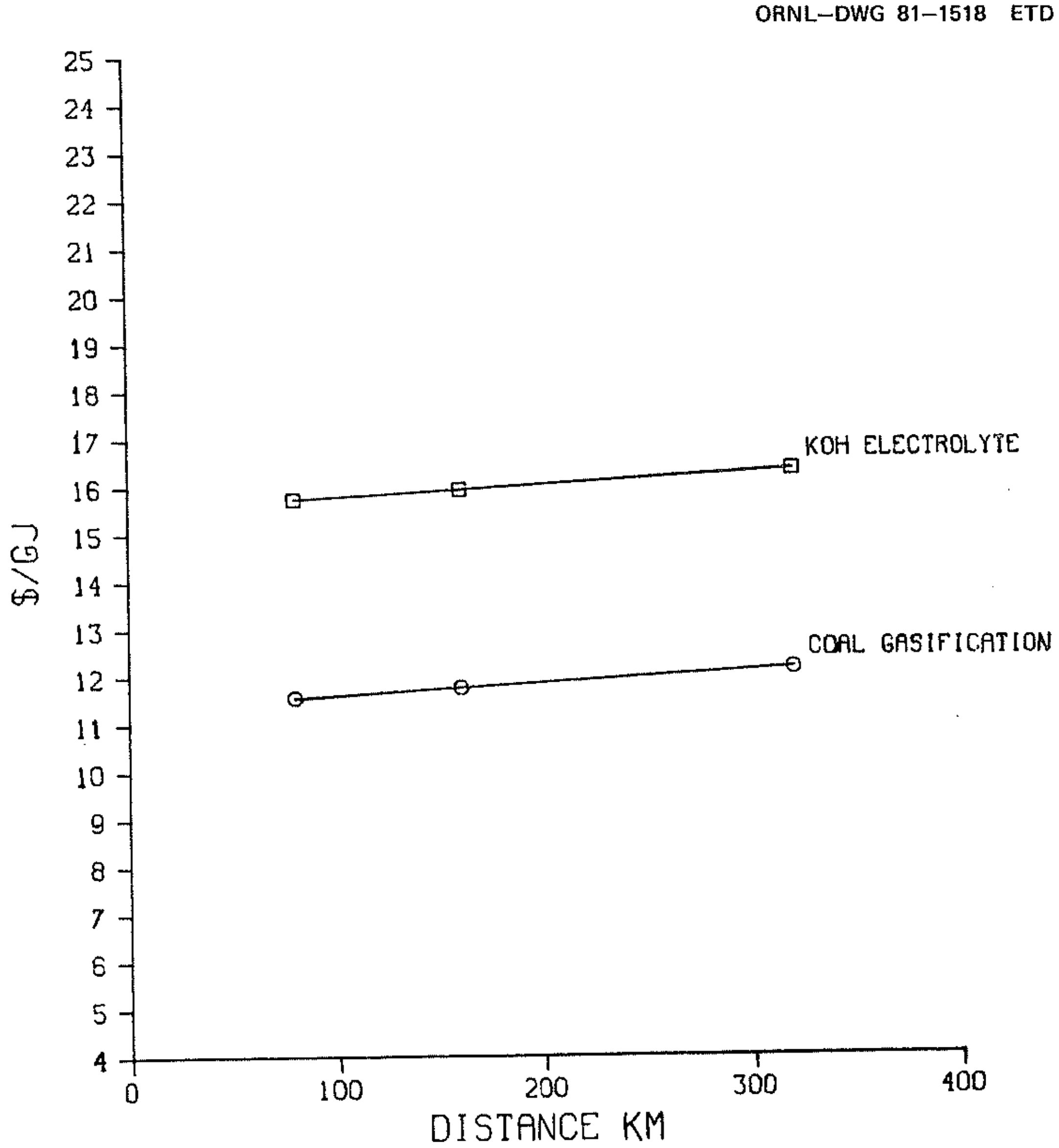

Fig. 4.8. Hydrogen energy transport options, 1200 MW, 3 MPa, 1990 start-up, 1980 dollars.

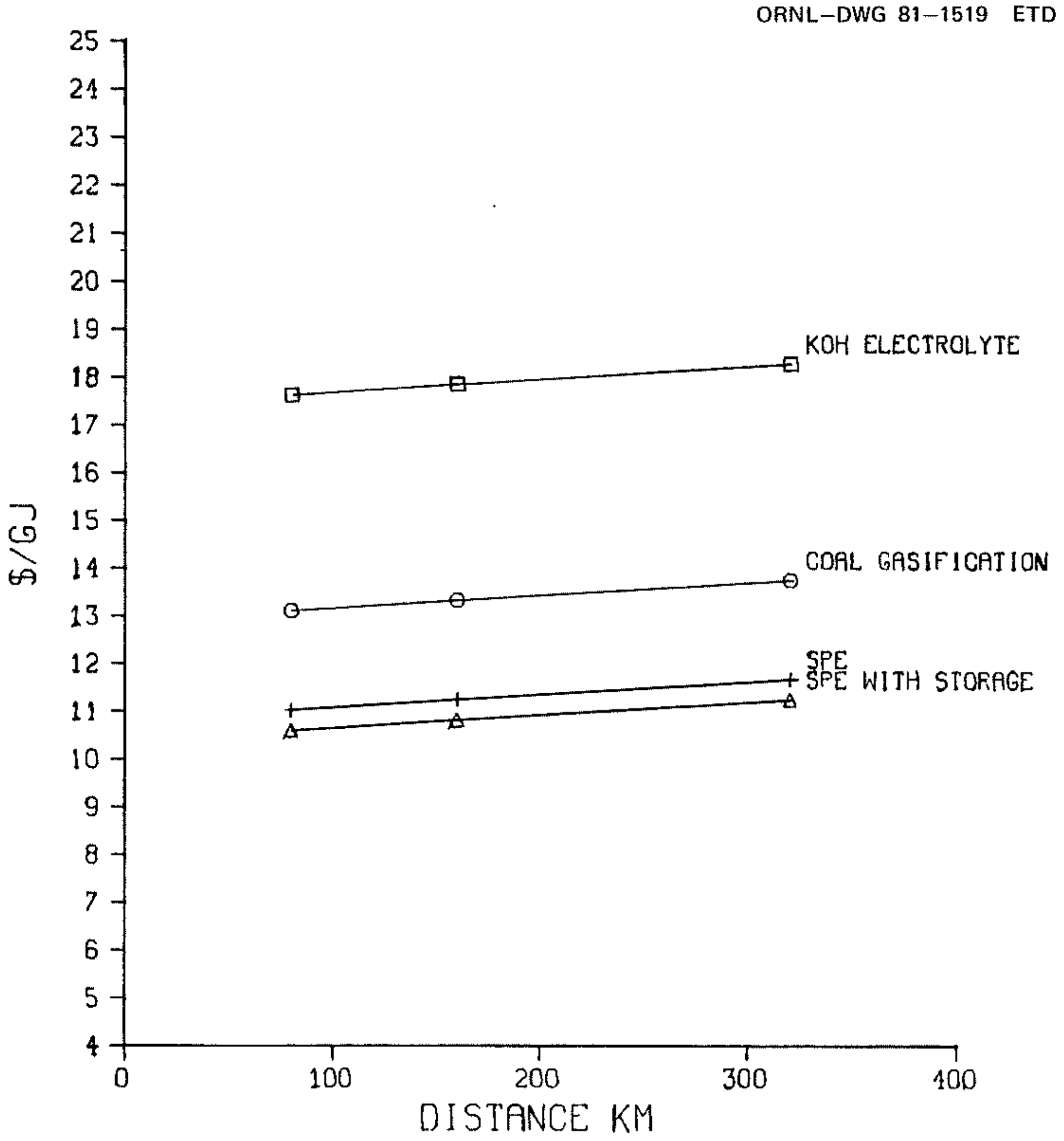

Fig. 4.9. Hydrogen energy transport options, 1200 MW, 3 MPa, 2000 start-up, 1980 dollars.

Table 4.20. Hydrogen cost via KOH electrolyte[a]

(1980 dollars)

	Start-up year	
	1990	2000
800-MW(t) delivery capacity		
Investment, $\$10^6$		
On-site	591.0	650.2
Off-site	92.9	101.3
Total[b]	683.9	751.5
WC	29.8	29.8
Operating cost, $\$10^6$/year ($/GJ)		
Electricity[c]	155.423 (8.80)	174.850 (9.90)
Water and chemicals	1.962 (0.11)	1.962 (0.11)
Labor and supervision	1.896 (0.11)	1.896 (0.11)
Maintenance (4% on-site)	19.700 (1.12)	21.672 (1.23)
Plant overhead (2.6% on-site)	12.805 (0.73)	14.087 (0.80)
Interest on WC (10%)	2.980 (0.17)	2.980 (0.17)
Fixed charges (17% total plant)	116.260 (6.58)	127.745 (7.23)
Total cost	311.026 (17.61)	345.192 (19.55)
Oxygen credit	[26.309 (1.49)]	[26.309 (1.49)]
Net cost	284.717 (16.12)	318.883 (18.06)
Transportation cost,[d] $/GJ		
80 km	0.28	0.31
160 km	0.54	0.60
320 km	1.06	1.17
Total delivered cost, $/GJ		
80 km	16.40	18.37
160 km	16.66	18.66
320 km	17.18	19.23
1000-MW(t) delivery capacity		
Investment, $\$10^6$		
On-site	706.3	776.9
Off-site	111.0	121.1
Total[b]	817.3	898.0
WC	37.21	37.21
Operating cost, $\$10^6$/year ($/GJ)		
Electricity[c]	194.303 (8.80)	218.591 (9.90)
Water and chemicals	2.453 (0.11)	2.453 (0.11)
Labor and supervision	2.371 (0.11)	2.371 (0.11)
Maintenance (4% on-site)	23.542 (1.07)	25.897 (1.17)
Plant overhead (2.6% on-site)	15.303 (0.69)	16.833 (0.76)
Interest on WC (10%)	3.721 (0.17)	3.721 (0.17)
Fixed charges (17% total plant)	138.934 (6.29)	152.657 (6.92)
Total cost	380.627 (17.24)	422.523 (19.14)
Oxygen credit	[32.890 (1.49)]	[32.890 (1.49)]
Net cost	347.737 (15.75)	389.633 (17.65)
Transportation cost,[d] $/GJ		
80 km	0.25	0.27
160 km	0.47	0.52
320 km	0.93	1.02
Total delivered cost, $/GJ		
80 km	16.00	17.92
160 km	16.22	18.17
320 km	16.68	18.67

Table 4.20 (continued)

	Start-up date	
	1990	2000
1200-MW(t) delivery capacity		
Investment, \$ 10^6		
On-site	817.2	898.9
Off-site	128.4	140.1
Total[b]	945.6	1039.0
WC	44.66	44.66
Operating cost, \$ 10^6/year (\$/GJ)		
Electricity[c]	233.183 (8.80)	262.331 (9.90)
Water and chemicals	2.944 (0.11)	2.944 (0.11)
Labor and supervision	2.845 (0.11)	2.845 (0.11)
Maintenance (4% on-site)	27.238 (1.03)	29.962 (1.13)
Plant overhead (2.6% on-site)	17.705 (0.67)	19.476 (0.74)
Interest on WC (10%)	4.466 (0.17)	4.466 (0.17)
Fixed charges (17% total plant)	160.746 (6.09)	176.623 (6.69)
Total cost	449.127 (17.01)	498.647 (18.89)
Oxygen credit	[39.472 (1.49)]	[39.472 (1.49)]
Net cost	409.655 (15.52)	459.175 (17.39)
Transportation cost,[d] \$/GJ		
80 km	0.22	0.24
160 km	0.42	0.47
320 km	0.83	0.92
Total delivered cost, \$/GJ		
80 km	15.74	17.63
160 km	15.94	17.86
320 km	16.35	18.31

[a] Plant specifications given in Sect. 3.4.

[b] 15% contingency included.

[c] Based on \$0.024 kWh in 1990 and \$0.027/kWh in 2000.

[d] Pressure at 3 MPa (450 psia).

4.5 Economic Considerations of Molten Salt and Hot Oil

4.5.1 Material costs

A cost of \$350/m^2 (\$33/ft^2) was used for heat exchangers in this study of molten salt and hot oil. Costs were found to vary widely, depending on temperature level, fluid, construction materials, and operating pressure. Cost estimates were found to range from \$15 to \$110/ft^2. The \$33/ft^2 figure represents a best judgement for the service conditions being considered and corresponds to the estimate used by General Electric[1] after escalation to 1980 dollars.

For salt, a cost of \$0.66/kg (\$0.30/lb) was used. Coastal Chemical, manufacturer of HITEC, indicated that salt components are used in varying percentages to produce mixtures for service at different temperature levels. The most expensive component, sodium nitrite, may cost as much as \$0.90/kg (\$0.41/lb).

However, production in the quantities implied by the pipeline diameters and lengths used in this study could result in lower prices because of bulk purchases of raw material.

Table 4.21. Hydrogen cost via SPE[a]

(1980 dollars, 2000 start-up)

	Delivery capacity [MW(t)]		
	800	1000	1200
Investment, $\$ 10^6$			
On-site	167.2	199.8	231.1
Off-site	39.2	46.9	54.2
Total[b]	206.4	246.7	285.3
WC	28.82	36.03	43.24
Operating cost, $\$ 10^6$/year ($/GJ)			
Electricity[c]	170.569 (9.66)	213.239 (9.66)	255.908 (9.66)
Water and chemicals	1.167 (0.07)	1.459 (0.07)	1.751 (0.07)
Labor and supervision	0.948 (0.05)	1.185 (0.05)	1.423 (0.05)
Maintenance (4% on-site)	5.573 (0.32)	6.659 (0.30)	7.705 (0.29)
Plant overhead (2.6% on-site)	3.622 (0.21)	4.328 (0.20)	5.008 (0.19)
Interest on WC (10%)	2.882 (0.16)	3.603 (0.16)	4.324 (0.16)
Fixed charges (17% total plant)	35.088 (1.99)	41.942 (1.90)	48.511 (1.84)
Total	219.849 (12.45)	272.411 (12.34)	324.638 (12.30)
Oxygen credit	[26.309 (1.49)]	[32.890 (1.49)]	39.472 (1.49)]
Net cost	193.540 (10.96)	239.521 (10.85)	285.166 (10.80)
Transportation cost,[d] $/GJ			
80 km	0.30	0.26	0.23
160 km	0.58	0.51	0.45
320 km	1.14	0.99	0.89
Total delivered cost, $/GJ			
80 km	11.26	11.11	11.03
160 km	11.54	11.36	11.25
320 km	12.10	11.84	11.69

[a] Plant specifications given in Sect. 3.4.
[b] 30% contingency included.
[c] Cost $0.027/kWh.
[d] Pressure at 3 MPa (450 psia).

4.5.2 Contingencies and construction costs

A project contingency of 20% of total pipeline costs (pipe material, insulation, construction, expansion joints, and pump stations) was assumed. This factor is based on judgment and is intended to cover site-specific problems such as terrain, river and road crossings, access roads, and extension of electrical service.

A process contingency of 5% of total pipeline costs was used. This factor is based on judgment and is intended to address the fact that the large-diameter high-temperature insulated lines may require heat tracing, additional engineering expense, field modification, and testing.

Engineering and inspection costs were estimated at 10% of total pipeline costs. This is somewhat lower than the usual assumption, but engineering costs are already partially included in the pipeline construction cost figures used.

The pipeline construction time was assumed to be five years, with the midpoint of expenditures at two years and 9.5% interest during construction. The AFDC was calculated as:

$$\text{AFDC} = [(1.095)^2 - 1] \times \text{Total Capital Requirement} .$$

Table 4.22. Hydrogen cost via SPE coupled to iron titanium hydrogen storage

(1980 dollars, 2000 start-up)

	Electrolysis	Storage	Total
800-MW(t) delivery capacity[a,b]			
Investment, $ 10^6			
On-site	275.6	197.3	472.9
Off-site	64.7	17.8	82.4
Total[c]	440.3	215.1	555.3
WC	11.102	0.627	11.729
Operating cost, $ 10^6/year ($/GJ)			
Offpeak electricity at $0.01/kWh	63.175		63.175 (3.58)
Chemicals, catalysts, utilities	1.168	3.761	4.929 (0.28)
Labor and supervision	2.276		2.276 (0.13)
Maintenance (4% on-site)	9.188	6.576	15.764 (0.89)
Plant overhead (2.6% on-site)	5.972	4.274	10.247 (0.58)
Interest on WC (10%)	1.110	0.063	1.173 (0.07)
Fixed charges	57.854	36.557	94.411 (5.35)
Total cost	140.743	51.231	191.965 (10.87)
Transportation cost,[d] $/GJ			
80 km			0.30
160 km			0.58
320 km			1.14
Total delivered cost, $/GJ			
80 km			11.17
160 km			11.45
320 km			12.01
1000-MW(t) delivery capacity[a,e]			
Investment, $ 10^6			
On-site	328.8	235.4	564.2
Off-site	77.3	21.1	98.4
Total[c]	406.1	256.5	662.6
WC	13.892	0.785	14.677
Operating cost, $ 10^6/year ($/GJ)			
Offpeak electricity at $0.01/kWh	79.052		79.052 (3.58)
Chemicals, catalysts, and utilities	1.462	4.706	6.168 (0.28)
Labor and supervision	2.847		2.847 (0.13)
Maintenance (4% on-site)	10.960	7.848	18.808 (0.85)
Plant overhead (2.6% on-site)	7.124	5.101	12.225 (0.55)
Interest on WC (10%)	1.389	0.079	1.468 (0.07)
Fixed charges	69.034	43.615	112.649 (5.10)
Total cost	171.868	61.349	233.217 (10.56)
Transportation cost,[d] $/GJ			
80 km			0.26
160 km			0.50
320 km			0.99
Total delivered cost, $/GJ			
80 km			10.82
160 km			11.06
320 km			11.55

Table 4.22 (continued)

	Electrolysis	Storage	Total
1200-MW(t) delivery capacity[a,f]			
Investment, $ 10^6			
On-site	381.1	272.9	654.0
Off-site	89.5	24.6	114.0
Total[c]	470.6	297.5	768.0
WC	16.652	0.941	17.594
Operating cost, $ 10^6/year ($/GJ)			
Offpeak electricity at $0.01/kWh	97.763		97.763 (3.58)
Chemicals, catalysts, utilities	1.752	5.642	7.394 (0.28)
Labor and supervision	3.413		3.413 (0.13)
Maintenance (4% on-site)	12.704	9.096	21.800 (0.82)
Plant overhead (2.6% on-site)	8.258	5.912	14.170 (0.53)
Interest on WC (10%)	1.665	0.094	1.759 (0.07)
Fixed charges	80.009	50.572	130.560 (4.95)
Total cost	202.564	71.299	273.859 (10.37)
Transportation cost,[a] $/GJ			
80 km			0.23
160 km			0.45
320 km			0.89
Total delivered cost, $/GJ			
80 km			10.60
160 km			10.82
320 km			11.26

[a] Electrolysis capacity based on 10 h/d, 330 d/year, other plant specifications given in Sect. 3.4.

[b] Fe-Ti hydrogen storage per cycle is 273,520 kg/h; overall load for 24 h/d is 186 × 10^6 scm/h (17.66 × 10^6 GJ/year).

[c] 30% contingency included.

[d] Pressure at 3 MPa (450 psia).

[e] Fe-Ti hydrogen storage per cycle is 342,262 kg/h; overall load for 24 h/d is 232 × 10^6 scm/h (22.075 × 10^6 GJ/year).

[f] Fe-Ti hydrogen storage per cycle is 410,280 kg/h; overall load for 24 h/d is 279 × 10^6 scm/h (26.49 × 10^6 GJ/year).

4.5.3 Delivered energy cost

Unit costs of delivered energy (transported 160 km) for the system assumptions described in Sect. 3.5 are shown in Fig. 4.10. The curves are grouped into three temperature ranges to be served by the three different systems. A constant energy supply cost of $3.96/GJ and a capacity factor of 0.9 were assumed in the following examination of cost as a function of temperature.

The lowest temperature range [up to 330° C (620° F)] is served by heat transfer oil in cs pipe. The temperature is limited because of thermal decomposition of the fluid at temperatures above 340° C (650° F). The intermediate temperature range [to 450° C (850° F)] is served by salt in cs pipe. The temperature limit is due to excessively large wall thicknesses above the allowable temperature range for the pipe material.

For both systems, minimum cost of delivered energy is achieved at the highest possible delivery temperature and the highest power level because (1) lower flow rates are required at the high delivery temperature (smaller pressure drop and resulting lower pumping costs and thinner wall pipe) and (2) more energy is delivered for the same capital investment.

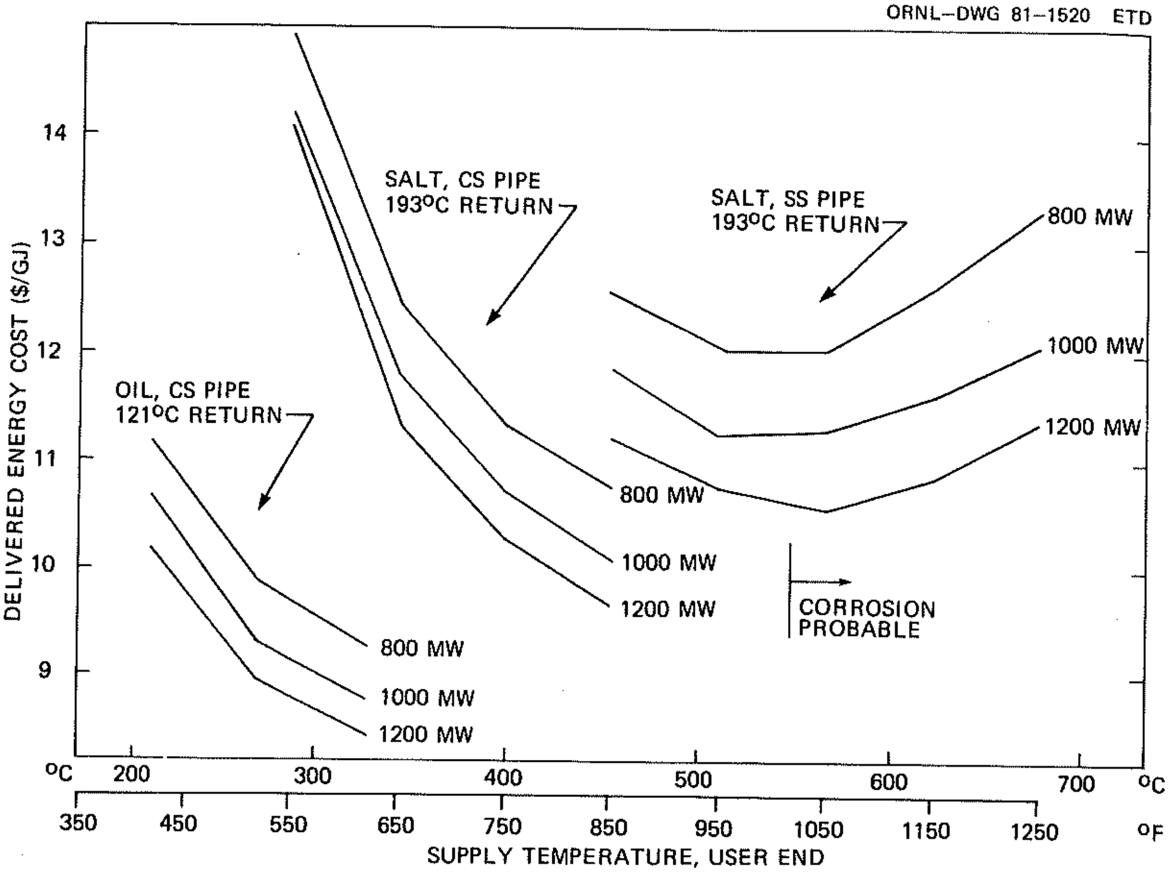

Fig. 4.10. Delivered energy cost as a function of supply temperature (160 km transport), 1990 start-up.

For the highest temperature range, minimum cost occurs near 565° C (1050° F) for each power level. The increased cost at higher temperatures is caused by the rapid decrease in allowable material stress, which results in the need for greater pipe wall thickness.

Recent studies[6] of stainless steel corrosion by molten salt of the HITEC type have established a practical temperature limit of 550° C (1020° F). For this reason, the data points presented here for temperatures in excess of 550° C (1020° F) may be of no practical interest. Results given in Table 4.23 indicate that the highest practical delivery temperature may be near 500° C (930° F), if the 550° C (1020° F) limit at the reactor end of the supply line cannot be exceeded.

Because the lowest unit cost of delivered energy occurs at the highest delivery temperature in each range, that cost is valid throughout the range. When servicing the lower temperatures in the range, the cost would decrease slightly because of the larger temperature difference across the heat exchangers. Consequently, a reduction occurs in the required heat exchanger surface area and cost. Heat exchanger costs, however, are not a large fraction of the total cost, so the assumption of constant cost throughout a temperature range is not greatly in error.

The result of assuming that each temperature range is to be serviced with the fluid at the highest delivery temperature for the range is shown in Fig. 4.11. Physical characteristics of these systems are given in Table 4.23.

Table 4.23. Pipeline temperature, size, energy quantities

(160-km transport)

System	Delivery (MW)	Reactor end, fluid temperature [°C (°F)]	Pipe diameter [mm (in.)]	Pump power (MW)	Line thermal loss (MW)	Reactor heat supply (MW)	Reactor size [MW(t)]
Oil, cs	800	348 (659)	914 (36)	53	197	944	1100
(330°C	1000	348 (658)	1016 (40)	58	209	1151	1400
delivery)	1200	341 (645)	1016 (40)	98	208	1310	1500
Salt, cs	800	495 (923)	711 (28)	70	238	968	1200
(450°C	1000	491 (916)	812 (32)	64	256	1192	1400
delivery)	1200	488 (911)	914 (36)	61	278	1417	1700
Salt, ss	800	640 (1184)	711 (28)	29	251	1022	1300
(565°C	1000	631 (1167)	812 (32)	26	271	1245	1500
delivery)	1200	616 (1140)	812 (32)	46	270	1424	1700
Salt, ss	800	567 (1053)	711 (28)	43	244	1001	1300
(510°C	1000	561 (1041)	812 (32)	40	264	1224	1500
delivery)	1200	556 (1033)	914 (36)	37	284	1447	1700
Salt, ss	800	507 (944)	812 (32)	35	261	1026	1300
(450°C	1000	499 (931)	914 (36)	36	279	1243	1500
delivery)	1200	488 (914)	914 (36)	61	278	1417	1700

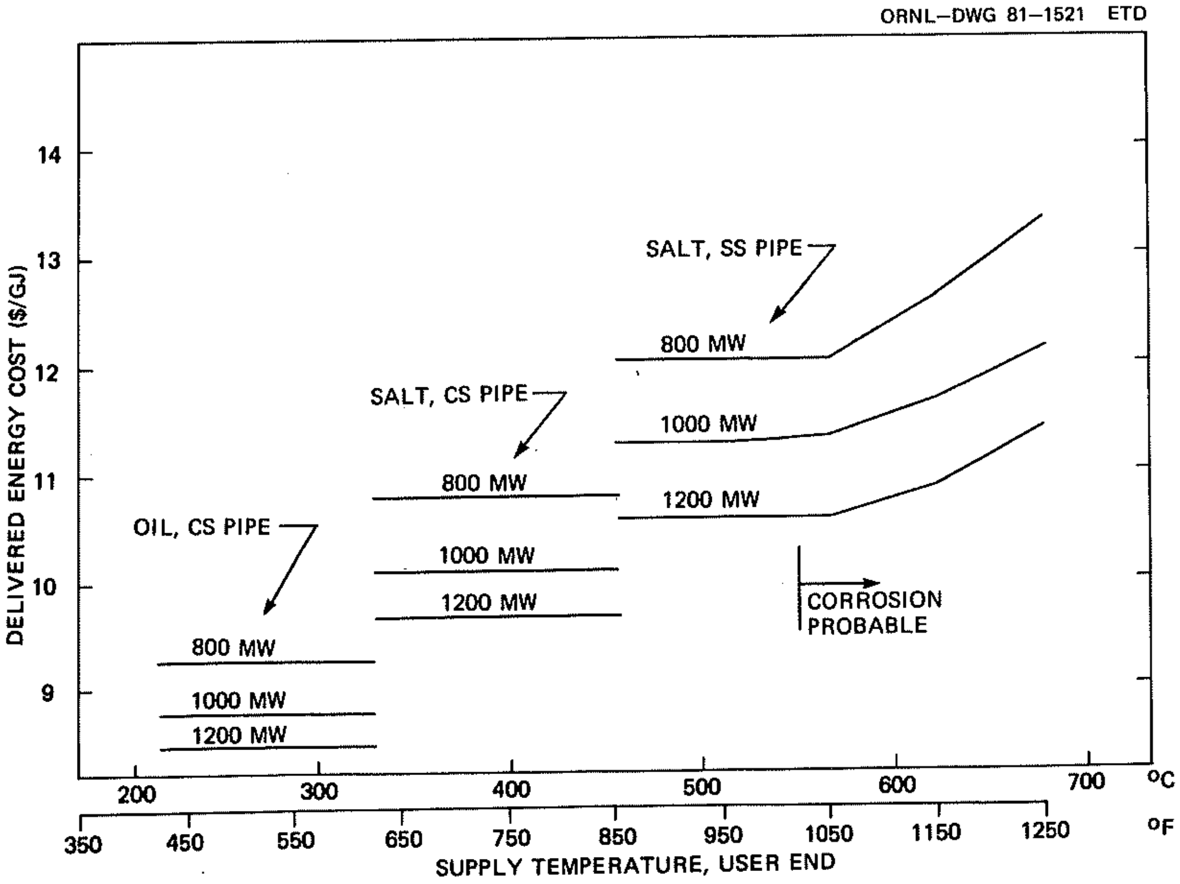

Fig. 4.11. Delivered energy cost at 160 km, 1990 start-up.

In Fig. 4.12 and Table 4.24 results are presented that describe the component costs at the transition temperatures between ranges. The reasons for cost differences between ranges are apparent from a comparison of component costs. At 330° C, where it is necessary to change from oil to molten salt, the cost increase is primarily due to the additional cost of the fluid. The increase in pump power costs is compensated for by a reduction in heat exchanger costs.

Comparison of the results at the high temperature (450° C) to those at the low temperature (330° C) for the intermediate range shows a cost decrease at the higher temperature because of the flow rate reduction possible with the higher temperature fluid. This flow rate reduction allows a decrease in pumping costs and a reduction in pipe size, which in turn reduces fluid inventory costs.

At transition to the highest temperature range, the cost increase is caused by increased pipe costs. At this transition temperature, it is necessary to change from cs to ss pipe. Although the higher allowable stress of ss permits smaller wall thicknesses, the cost ratio of 7.8:1 between the two materials results in a pipe cost increase ratio of 2.15:1.

To develop information on unit-delivered energy costs as a function of delivery distance, all physical variables except length were taken equal to their values for the 160-km system (Sect. 3.6). Optimization of the system for the particular transport distance being considered may result in different

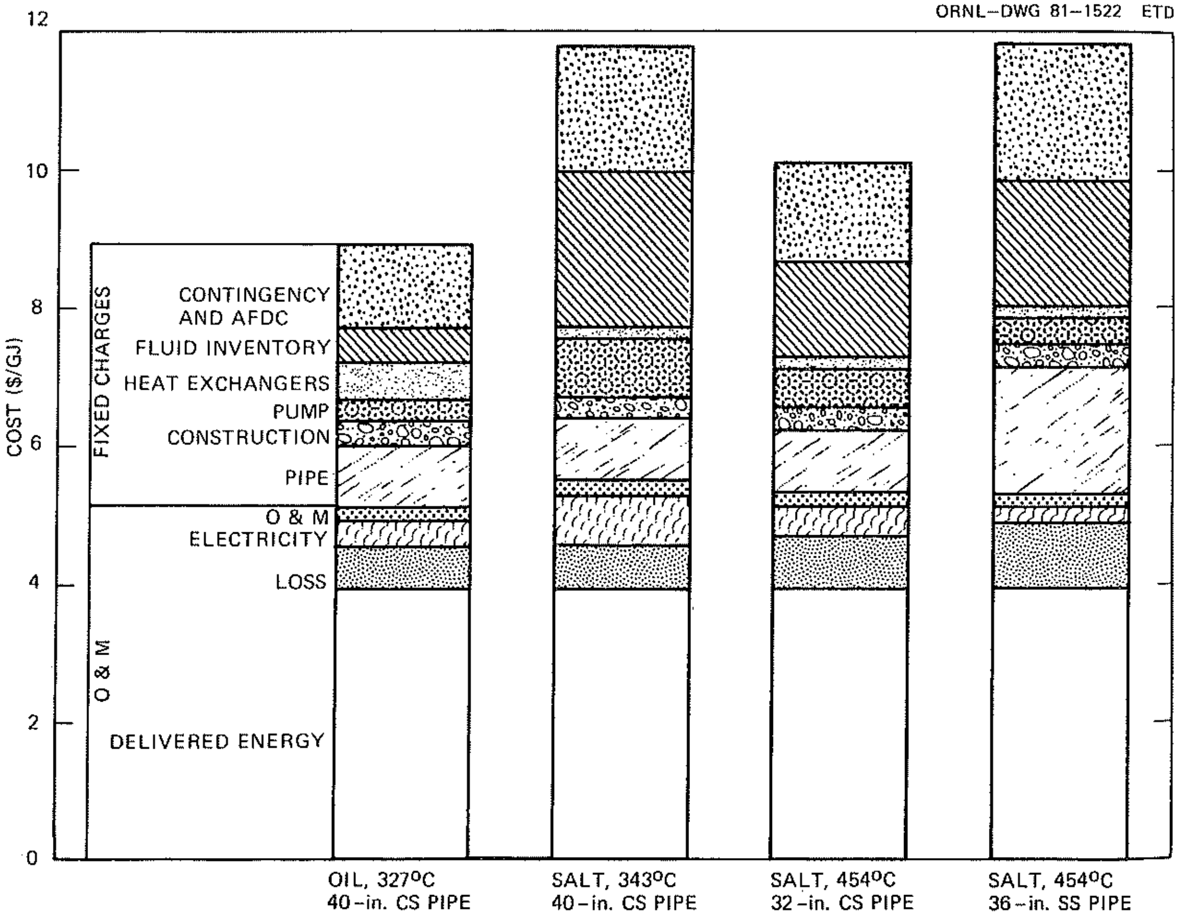

Fig. 4.12. Transitional costs, 1990 start-up.

Table 4.24. Salt and hot oil system component costs at transition temperatures, 1990 start-up

(1980 $/GJ)

Component	Low-temperature range, oil, cs, 330°C	Intermediate-temperature range, salt, cs		High-temperature range, salt, ss, 450°C
		340°C	450°C	
Pipe	1.09	1.20	1.13	2.43
Construction	0.46	0.46	0.41	0.44
Pump stations	0.37	1.05	0.73	0.48
Heat exchangers	0.72	0.15	0.15	0.17
Fluid inventory	0.60	2.92	1.83	2.37
Contingencies and AFDC	1.47	2.31	1.84	2.56
Total capital requirements	4.72	8.09	6.09	8.43
Electricity (pump power)	0.40	0.68	0.45	0.25
Pipeline thermal loss	0.52	0.55	0.66	0.83
Heat delivered	3.43	3.43	3.43	3.43
Other O&M	0.20	0.22	0.17	0.12
Total operational costs	4.55	4.88	4.71	4.63
Total costs	9.27	12.97	10.80	13.06

system configurations. Because the incremental costs as a function of changes in the physical variables were small in the sensitivity study (Appendix C), changes in these variables because of delivery distance are believed to be small. Such changes are considered as site-specific modifications that are adequately accommodated by the contingency allowances.

Cost variations as a function of transport distance are shown for the three temperature ranges in Figs. 4.13 through 4.15. In each case, the cost variation with distance is essentially linear.

At 30 km (20 miles) the unit-delivered energy cost for all systems is within ±10% of $5.10/GJ. From that point, costs increase most rapidly with length for systems consisting of the more expensive components—ss and salt.

The cost figures presented are based on a fixed charge rate of 0.17 and 1990 start-up date. Process heat costs were scaled according to the reactor sizes reported in Table 4.23. Scaling factors are given in Sect. 4.1. The components of resulting costs are shown in Tables 4.25 through 4.30 for the systems with the three temperature ranges shown in Fig. 4.11. Figures 4.16 and 4.17 show the delivered energy costs for the salt and hot oil systems as a function of distance for 1990 and 2000 start-up dates.

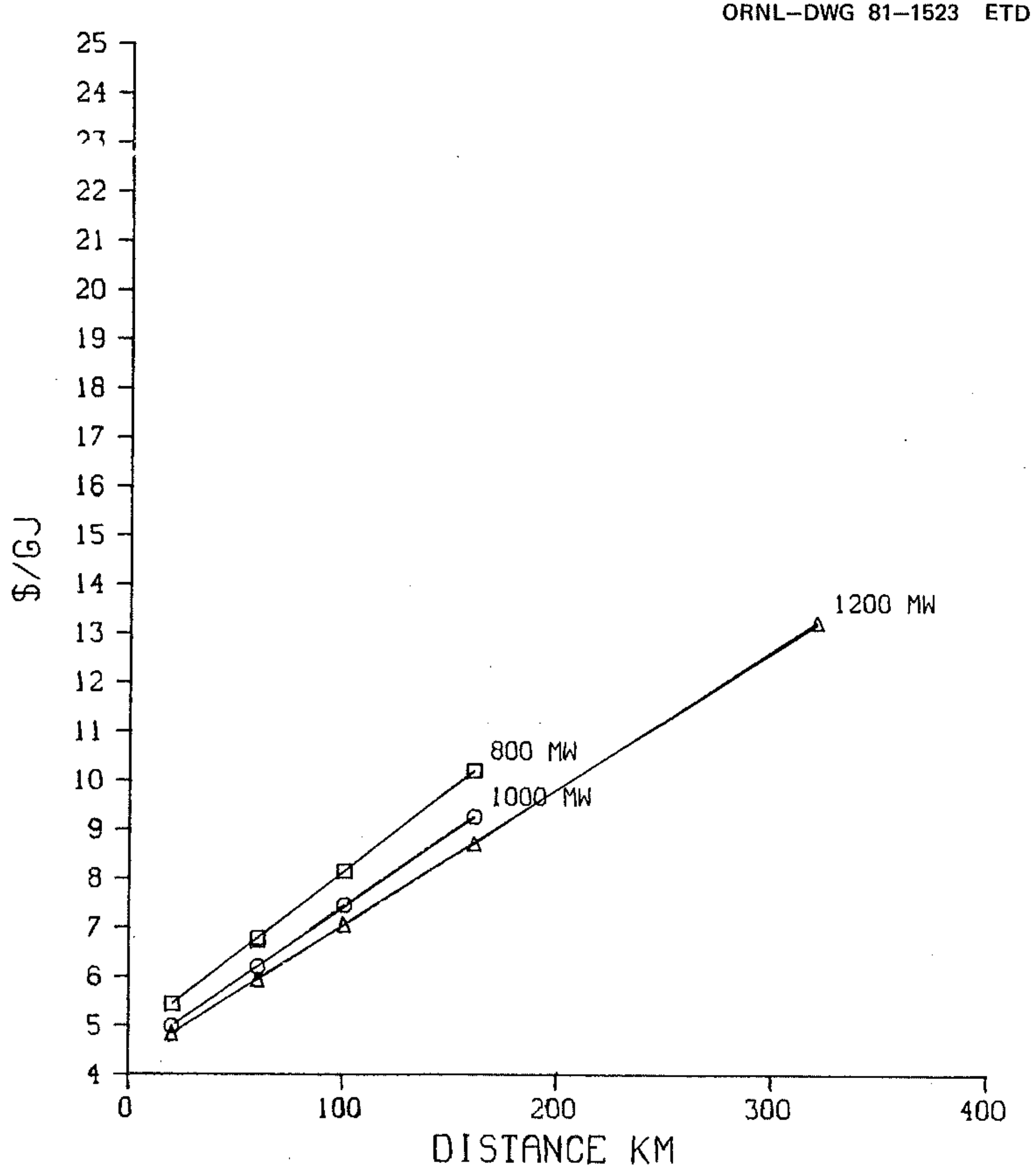

Fig. 4.13. Hot oil, 330° C and below, 1990 start-up, 1980 dollars.

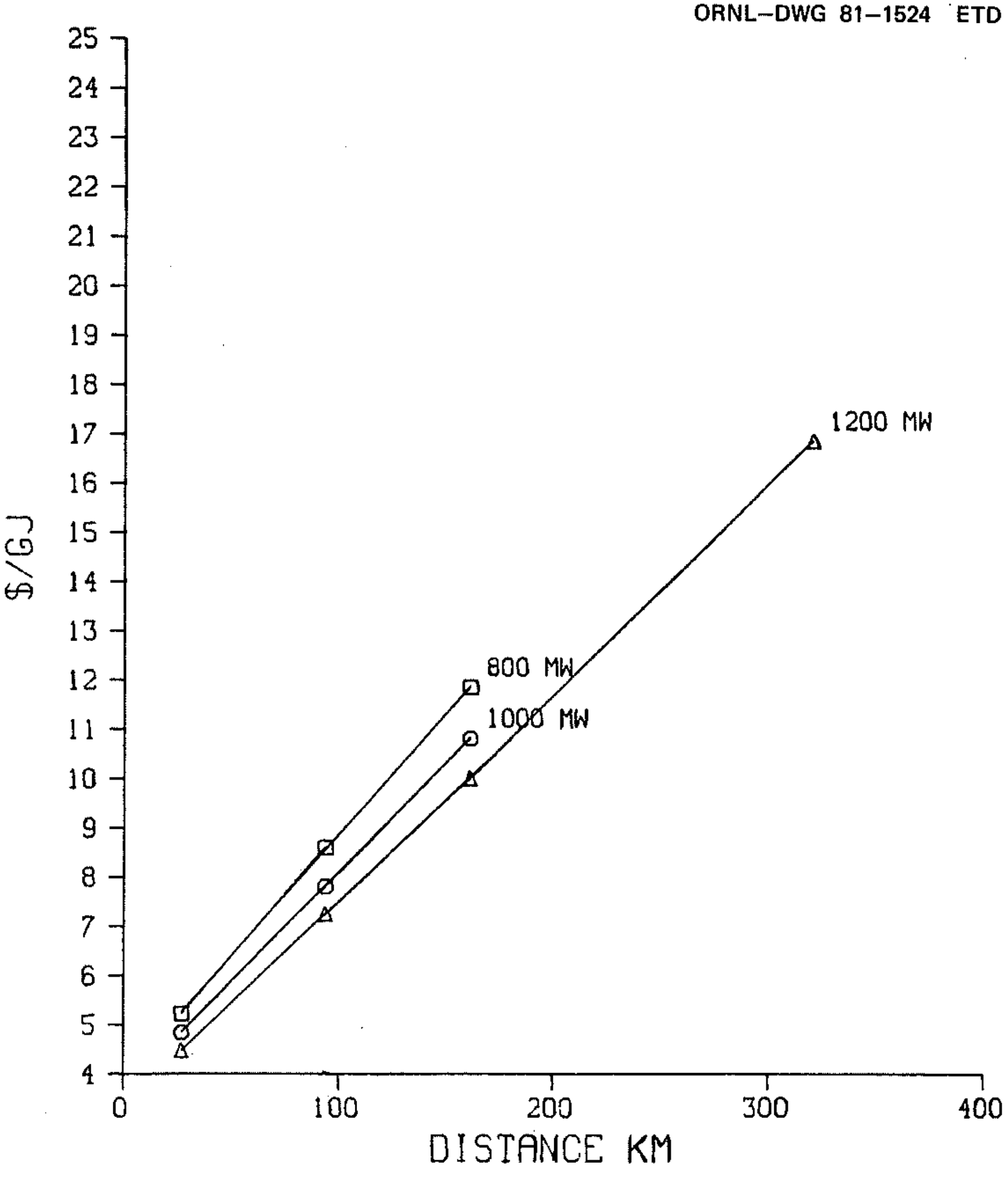

Fig. 4.14. Molten salt, cs pipe, 330 to 450°C, 1990 start-up, 1980 dollars.

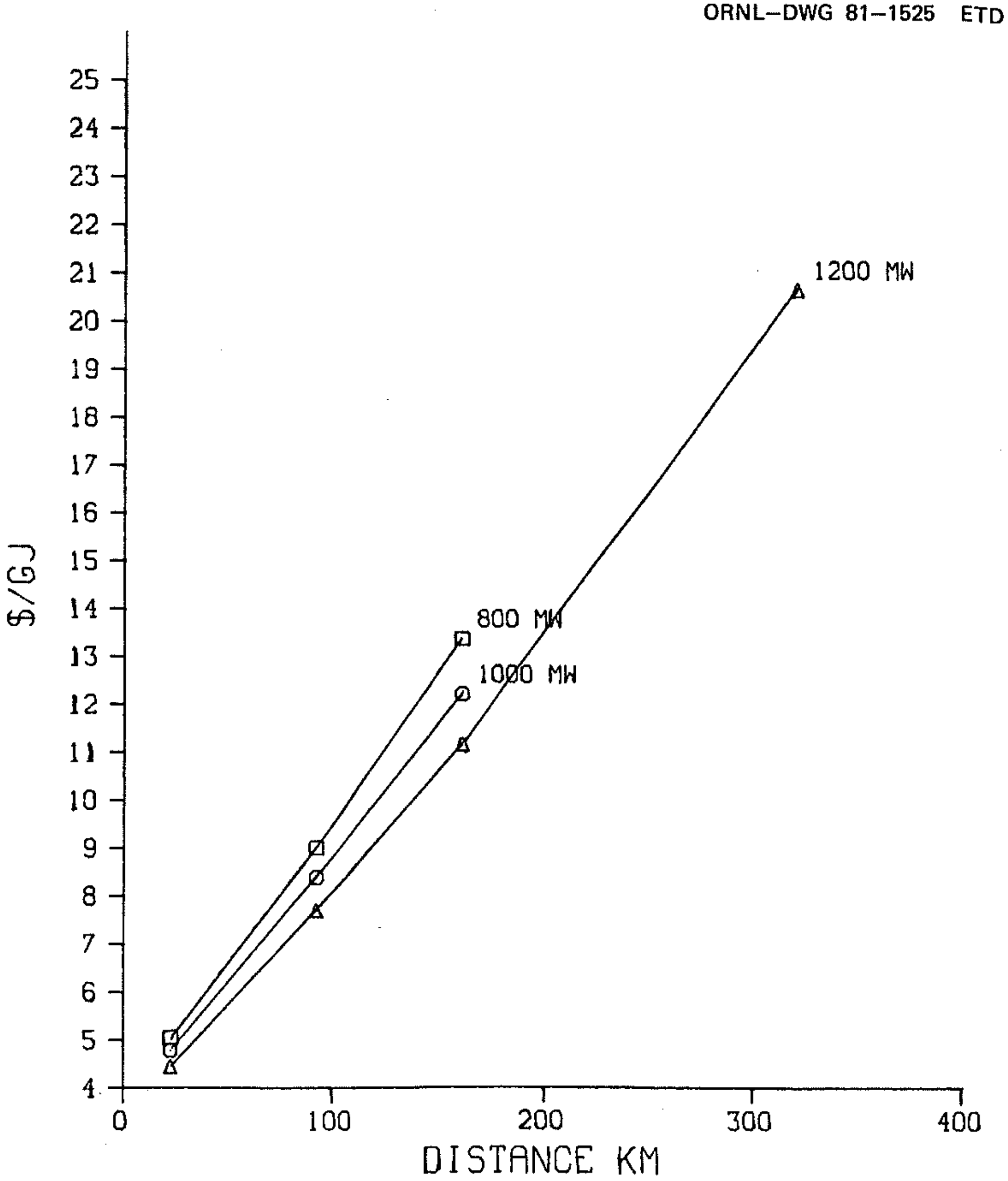

Fig. 4.15. Molten salt, ss pipe, 450 to 570° C, 1990 start-up, 1980 dollars.

Table 4.25. Salt, ss piping, 1990 start-up

(1980 $/GJ)

	Piping distance (km)			
	23	92	160	320
	800 MW			
Fixed charge[a]	1.29	4.68	8.49	
Process heat[b]	3.67	4.09	4.52	
Other O&M	0.09	0.24	0.38	
Total	5.05	9.01	13.39	
	1000 MW			
Fixed charge[a]	1.25	4.42	7.79	
Process heat[c]	3.44	3.78	4.15	
Other O&M	0.09	0.19	0.29	
Total	4.78	8.39	12.23	
	1200 MW			
Fixed-charge[a]	1.12	3.99	7.06	15.57
Process heat[d]	3.24	3.48	3.75	4.38
Other O&M	0.08	0.23	0.37	0.73
Total	4.44	7.70	11.18	20.68

[a] Fixed-charge rate of 17%.
[b] Process heat at $3.54/GJ.
[c] Process heat at $3.33/GJ.
[d] Process heat at $3.16/GJ.

Table 4.26. Salt, ss piping, 2000 start-up

(1980 $/GJ)

	Piping distance (km)			
	23	92	160	320
	800 MW			
Fixed charge[a]	1.41	5.14	9.32	
Process heat[b]	3.97	4.43	4.90	
Other O&M	0.09	0.24	0.38	
Total	5.47	9.81	14.60	
	1000 MW			
Fixed charge[a]	1.38	4.86	8.56	
Process heat[c]	3.74	4.12	4.50	
Other O&M	0.09	0.19	0.29	
Total	5.21	9.17	13.35	
	1200 MW			
Fixed-charge[a]	1.23	4.38	7.75	17.10
Process heat[d]	3.53	3.79	4.08	4.77
Other O&M	0.08	0.23	0.37	0.73
Total	4.84	8.40	12.20	22.60

[a] Fixed-charge rate of 17%.
[b] Process heat at $3.84/GJ.
[c] Process heat at $3.62/GJ.
[d] Process heat at $3.44/GJ.

Table 4.27. Salt, ss piping, 1990 start-up

(1980 $/GJ)

	Piping distance (km)			
	27	94	160	320
	800 MW			
Fixed charge[a]	1.30	4.08	6.63	
Process heat[b]	3.82	4.25	4.43	
Other O&M	0.11	0.27	0.80	
Total	5.23	8.60	11.86	
	1000 MW			
Fixed charge[a]	1.17	3.63	6.11	
Process heat[c]	3.53	3.80	4.09	
Other O&M	0.14	0.38	0.62	
Total	4.84	7.81	10.82	
	1200 MW			
Fixed charge[a]	1.12	3.45	5.80	11.64
Process heat[d]	3.25	3.50	3.73	4.33
Other O&M	0.11	0.30	0.48	0.93
Total	4.48	7.25	10.01	16.90

[a] Fixed-charge rate of 17%.
[b] Process heat at $3.66/GJ.
[c] Process heat at $3.43/GJ.
[d] Process heat at $3.16/GJ.

Table 4.28. Salt, ss piping, 2000 start-up

(1980 $/GJ)

	Piping distance (km)			
	27	94	160	320
	800 MW			
Fixed charge[a]	1.43	4.47	7.29	
Process heat[b]	4.15	4.62	4.81	
Other O&M	0.11	0.27	0.80	
Total	5.69	9.36	12.90	
	1000 MW			
Fixed charge[a]	1.29	3.99	6.71	
Process heat[c]	3.84	4.14	4.45	
Other O&M	0.14	0.38	0.62	
Total	5.27	8.51	11.78	
	1200 MW			
Fixed charge[a]	1.23	3.78	6.36	12.78
Process heat[d]	3.58	3.81	4.06	4.72
Other O&M	0.11	0.30	0.48	0.93
Total	4.92	7.89	10.90	18.43

[a] Fixed-charge rate of 17%.
[b] Process heat at $3.98/GJ.
[c] Process heat at $3.73/GJ.
[d] Process heat at $3.44/GJ.

Table 4.29. Hot oil, 1990 start-up

(1980 $/GJ)

	Delivery distance (km)				
	20	60	100	160	320
			800 MW		
Fixed charge[a]	1.35	2.40	3.46	5.07	
Process heat[b]	3.89	4.06	4.23	4.48	
Other O&M	0.20	0.33	0.46	0.66	
Total	5.44	6.79	8.15	10.21	
			1000 MW		
Fixed charge[a]	1.29	2.24	3.20	4.72	
Process heat[c]	3.46	3.55	3.64	3.95	
Other O&M	0.23	0.42	0.62	0.60	
Total	4.98	6.21	7.46	9.27	
			1200 MW		
Fixed charge[a]	1.25	2.13	3.01	4.32	7.88
Process heat[d]	3.37	3.44	3.52	3.63	3.95
Other O&M	0.21	0.37	0.52	0.77	1.42
Total	4.83	5.94	7.05	8.72	13.25

[a] Fixed-charge rate of 17%.
[b] Process heat at $3.80/GJ.
[c] Process heat at $3.43/GJ.
[d] Process heat at $3.33/GJ.

Table 4.30. Hot oil, 2000 start-up

(1980 $/GJ)

	Delivery distance (km)				
	20	60	100	160	320
			800 MW		
Fixed charge[a]	1.48	2.64	3.79	5.57	
Process heat[b]	4.23	4.41	4.60	4.87	
Other O&M	0.20	0.33	0.46	0.66	
Total	5.91	7.38	8.85	11.10	
			1000 MW		
Fixed charge[a]	1.41	2.46	3.52	5.18	
Process heat[c]	3.76	3.86	3.96	4.29	
Other O&M	0.23	0.42	0.62	0.60	
Total	5.40	6.74	8.10	10.07	
			1200 MW		
Fixed charge[a]	1.38	2.34	3.30	4.76	8.65
Process heat[d]	3.66	3.75	3.83	3.95	4.29
Other O&M	0.21	0.37	0.52	0.77	1.42
Total	5.25	6.46	7.65	9.48	14.36

[a] Fixed-charge rate of 17%.
[b] Process heat at $4.13/GJ.
[c] Process heat at $3.73/GJ.
[d] Process heat at $3.62/GJ.

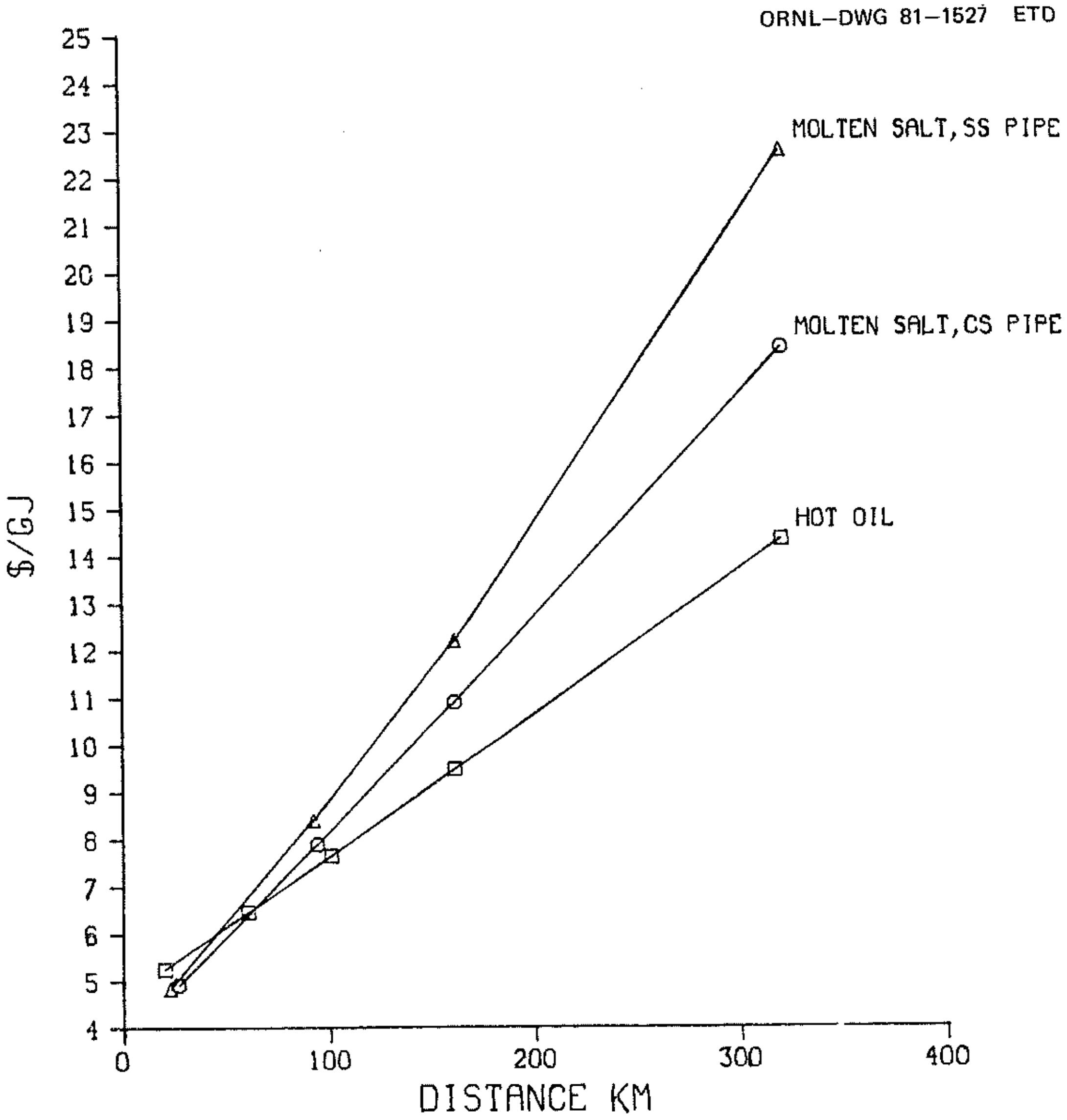

Fig. 4.16. Molten salt and hot oil delivered energy cost, 1200 MW, 1990 start-up, 1980 dollars.

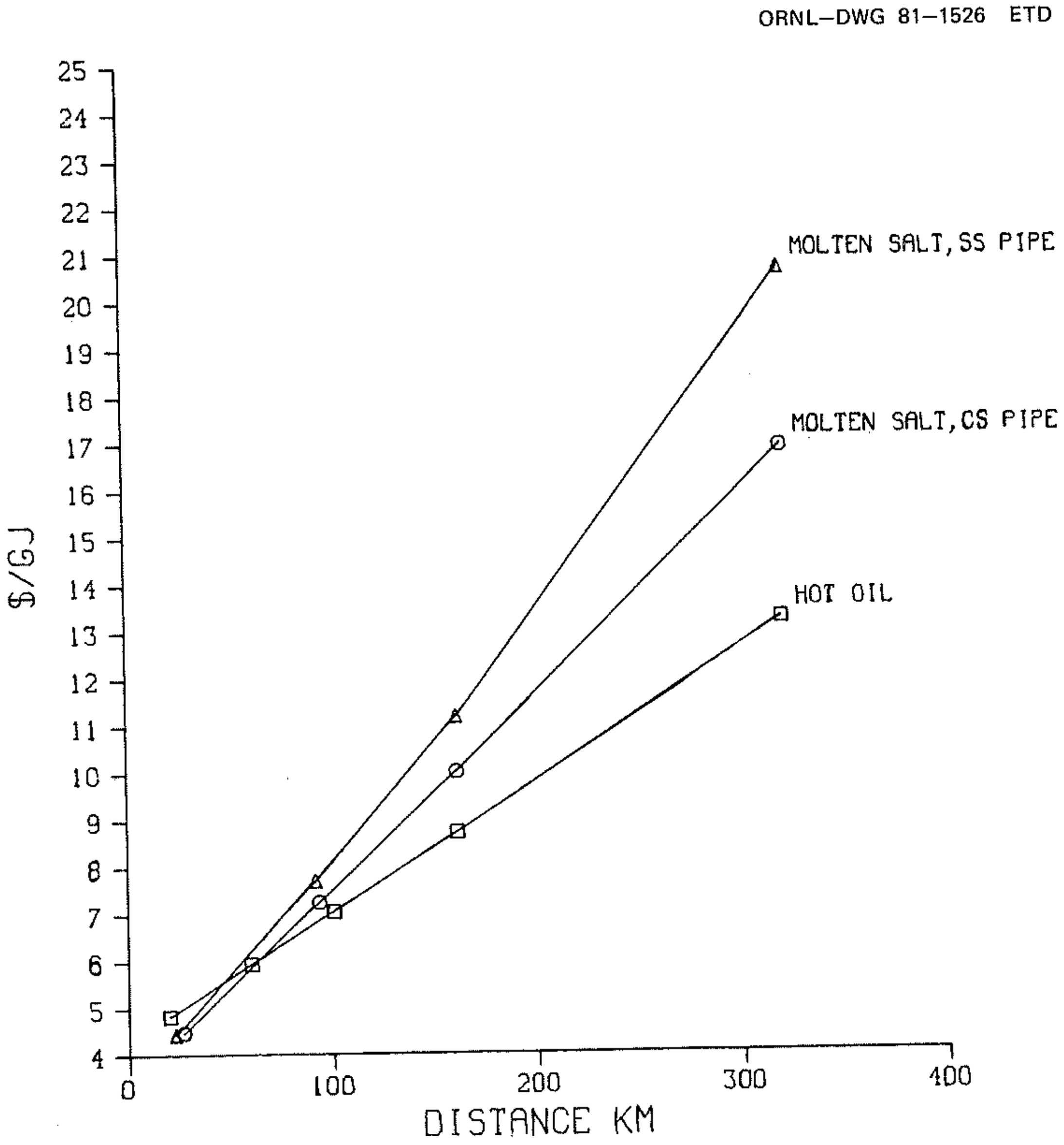

Fig. 4.17. Molten salt and hot oil delivered energy cost, 1200 MW, 2000 start-up, 1980 dollars.

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5. INSTITUTIONAL AND TECHNICAL CONCERNS

5.1 CHP

5.1.1 Institutional issues[1,2]

Acceptance of any long-distance transport system by industry will depend on questions of cost and reliability.

The additional probability of loss of service caused by the presence of the pipeline link in the energy supply chain is certain to cause considerable reluctance to accept the system by potential industrial energy consumers. Extensive reliability studies, pilot operation, and availability of back-up systems seem to be necessary steps in providing the necessary assurances.

Even with such assurances, demonstration of decided cost advantages or lack of alternative supplies will be necessary. In these systems, pipeline transport costs double or triple the bare cost of energy supplied to the consumer. Industry will examine every opportunity to avoid these transport charges by extensive consideration of all systems that allow local or on-site energy production.

The most likely candidate for supplying heat to the CHP is the VHTR with peak temperatures >800°C.

A number of institutional barriers must be overcome before commercial deployment of the VHTR.[1] The degree of difficulty associated with these issues depends on the specific application under consideration.

Design and operating experience must be obtained to justify the supplier entry into the commercial market and the utility/user willingness to commit to commercial orders.

The VHTR introduces many new concepts to the already complicated task of licensing nuclear power plants in the United States. Acceptance must be obtained for the following regulatory and institutional issues:

1. $CO\text{-}H_2/CH_4$ proximity to nuclear power plant (potential effects of explosion);
2. trace radioactive contamination (tritium) of the process gases;
3. $CO\text{-}H_2$ transmission pipelines; and
4. coupling with industrial processes.

The utilities have the financing, licensing, and operating experience necessary to construct and operate commercial nuclear power plants in the United States. Deployment of the VHTR for applications other than electrical generation necessitates the emergence of the utilities as energy companies rather than simply electricity and/or natural gas suppliers. In particular, new contractual and regulatory interfaces would have to be established. A further difficulty would result from the significant mismatch between lead times for industrial facilities and nuclear facilities.

The VHTR supply industry must be established to support commercial deployment of the VHTR, which includes the establishment of manufacturing facilities as well as plant engineering and construction experience.

While technical issues must be resolved in specific applications, institutional barriers offer, if anything, a more difficult challenge. Also included in these institutional issues are the following elements:

1. Deployment of a new reactor technology involves an enormous investment by both suppliers and utility users.
2. National energy policy, particularly with regard to nuclear energy, has been inconsistent, unstable, and to a large extent hostile to new development over the past decade.
3. The regulatory environment pertaining to energy, in general, and to nuclear energy, in particular, has become increasingly onerous during the past decade.
4. The flexibility that allows electric utilities to assume major development risks has been increasingly diminished by politically based public utility commissions.
5. The licensing basis for HTGRs is yet to be developed in the context of current requirements.

In view of these issues, the deployment risk of any HTGR (state of the art or advanced) could only be evaluated as open ended and unacceptable by any prudent business analysis. This view is further reinforced by the institutional impact on recent nuclear development programs both in the United States (Clinch River Breeder Reactor Project) and in the Federal Republic of Germany.

Another barrier that might arise is the placement of an unfamiliar technology, a chemical plant, in the process-steam user facility. Although the methanators might be considered analogous to process-steam boilers and are not complicated to operate, additional training for the operators would be required.

Who will market the CHP concept, and who will own the various components of the system? Traditionally, government-regulated utilities have supplied electricity. If a utility pursues the CHP concept, it will be a new type of utility—one that supplies heat for both process use and electricity.* The organizational difficulties of bringing the first large-scale thermal plant on-line to serve a variety of consumers with process steam will be enormous. Either the consumers must be lined up before start-up (possibly eight to ten years in advance) so that plans could be made for pipelines and geographical distribution of the heat, or a utility would have to build the plant on speculation, hoping to line up customers during and after the plant start-up.

If the latter route is followed, and multiple heat sources are required, each could be brought on-line as customers are brought under contract. Even this alternative might prove costly, because capital equipment would have to be purchased in advance and perhaps sit idle until consumers are located. A more likely scenario would be for the utility to bleed a small portion of the heat of existing plants into small reformer units and to set up a CHP pipeline grid (similar to existing electrical grids) into which a new plant's output could be fed upon its completion. Definitive answers to this problem of nucleation of a new technology must await further study on possible location sites, consumer geographical distribution, and needs (Sect. 2) in the areas surrounding those sites.

The question of responsiblity in case of an accident must be considered for this technology as well as for the others being developed now. A finite danger is associated with the transmission of hydrogen (fire) and carbon monoxide (fire and toxicity) on a large-scale through highly populated industrial areas. The situation is similar to that faced by the natural gas industry. Where does the utility's liability end and the user's start? Will the utility own all equipment and assume all liability?

The question of who owns the methanators in the CHP is important for reasons beyond safety or determination of responsibility. Currently, cogeneration of process steam and electricity is

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*Occasionally in this country and to a larger extent in some European countries, utilities have supplied both electricity and heat for district heating.

uneconomical for many industrial users; a difference exists in expected rate of return on capital investment between utilities and other industries. This difference results in the much higher capital charge in the cost of industry-generated electricity compared with the capital charge component in utility-generated electricity. As a result, the industrial user can frequently purchase power from the utility cheaper than producing it himself. This difference in rate of return has developed over many years as a result of government regulation of utilities. However, this difference clearly must be reevaluated if cogeneration (either within the CHP framework or otherwise) is to be pursued with the aim of conserving energy. If the industrial steam user produces electricity along with steam, it must be profitable to do so.

5.1.2 Technical issues[2,3]

In addition to these institutional issues, some technical issues that are common to all HTGR and VHTR options must be addressed.

Temperature fluctuations were observed in the primary coolant circuit at individual core-region outlets and at the steam generator module inlets during the rise to power of the Fort St. Vrain (FSV) reactor. The present HTGR core design has been modified to improve bypass control, and these modifications are expected to avert this problem in future HTGR designs.

Mechanical design of graphite components (core, reflector, and core support) includes consideration of irradiation, thermal and seismic stresses, and their long-term effects. Acceptance of a general approach to the combination of the stresses has not been achieved, but development of graphite design criteria for regulatory concurrence is in progress.

Development and characterization of metallic and ceramic materials for long-term operation in the HTGR or VHTR environment is required for applicable code qualification and design verification. This issue becomes more prominent as core outlet temperatures are increased.

Emphasis on nonproliferation has resulted in consideration of a low-enriched (20%) uranium/thorium (LEU/Th) fuel cycle. This change to LEU/Th fuel from high-enriched uranium/thorium (HEU/Th) fuel has led to the consideration of new fuel performance factors and a need for an improved data base to support the LEU/Th fuel design.

The technical barriers to using a nuclear reactor with the CHP lie in two areas: (1) any such VHTR is, at best, many years from commercial implementation; (2) the technology to couple the two systems into one capable of containing accidental radioactive emissions for safety reasons is not yet available. Work in Germany and the United States is currently addressing both problems. Developmental work in Germany has progressed on the pebble bed reactor (PBR) to the extent that their system has heated helium gas for an extended period to temperatures capable of supplying heat to a methane reformer.[3] General Atomic has had an extended development program for an HTGR.

KFA[3] is investigating the reactor coupling problems in detail; at the Eva-Adam facility, the ultimate goal is mating a methane reformer with a PBR. The potential of the CHP system for saving premium fuels should stimulate additional nuclear reactor development.

Coupling of the reactor to a reformer should be considered in developing source designs. In current reformers, cycling the reformer tubes over an extended temperature range in a short time interval is catastrophic because of extreme metal expansion/contraction stresses placed on the tubes and the structural degradation of catalyst pellets under such cycling. Twice a year is considered too often to bring current reformers down to ambient temperatures. Current industrial practice is to feed high-temperature steam to the unit to limit thermal shock in case of fuel outage to the system.

High-temperature methanators appear to be within the state of the art. However, no large industrial-scale installation exists; therefore no long-term operating experience on a high-temperature process exists. This barrier may well be eliminated in the near future if new coal-gasification plants come on stream and use high-temperature methane as part of their process design.

Thermal cycling might present a problem in the methanator analogous to that in the reformer, if used on a one- or two-shift basis. However, current indications are that this thermal cycling problem is not a significant barrier. This is particularly true for designs utilizing adiabatic reactor beds and external heat exchangers. Cycling methanator duty may be less of a problem than that for normal boilers.

Because all CHP-transmitted heat users would not be clustered at one geographical site, actual transmission philosophies will differ from the ones used in this study, and complicated grid arrangements to several load centers may in fact be required. Because existing industries may well be interested in coupling into CHP lines, the technical feasibility of installing the required pipelines in industrially congested areas may present a problem. The use of existing natural gas lines is a possibility; an analysis of this opportunity should await the selection of a specific user/geographical distribution.

5.1.3 Summary

It appears unlikely that the CHP can compete with MBG or SPE hydrogen. Costs are prohibitively high, and the possibility of VHTR commercialization in the United States is exceptionally poor.

5.2 MBG[4]

5.2.1 Institutional issues

Standards affecting coal gasification technologies are issued or proposed under the Clean Air Act, the "Nonattainment" and "Significant Deterioration" provisions of the Clean Air Act amendments, the Clean Water Act, the Toxic Substances Control Act, and the Resource Conservation and Recovery Act. These standards pose no insurmountable barriers to the commercial use of coal gasification processes although uncertainties remain in regard to final regulation of these laws.

5.2.2 Technical issues

When used with available control technology, coal gasification is one of the cleanest ways of using coal today. The exact composition of coal gasification by-products and emissions, however, depends primarily on feedstock characteristics and operating conditions.

The main environmental concerns regarding coal gasification include particulates, ash, nitrogen, sulfur, and organic condensates.

Coal fines disposal may present a problem for small installations where it may not be economical to use briquetting. These fines may be burned if a coal-fired boiler is used; otherwise they may have to be returned to the mine.

Ash or slag can be treated and handled the same way as ash from a boiler: (1) sluiced to an ash pond, (2) dumped directly into a container and trucked away, or (3) disposed of by landfill, particularly when a high percentage of carbon is unconverted, thus presenting a potential fire hazard.

The sulfur content of the coal will appear in the gas mainly as hydrogen sulfide and as complex organic sulfur compounds. Because gasification takes place in a reducing atmosphere, no sulfur dioxide is formed. However, if sulfur-carrying gas is not treated, its combustion is likely to produce sulfur dioxide. To produce a fuel gas with sulfur oxide contents below Environmental Protection Agency (EPA) standards, the sulfur content in the coal must be no greater than 0.72%. Gas desulfurization is usually done in a costly Stretford unit after removing hydrogen cyanide. The main advantage in using a sulfur removal unit is that lower cost, high-sulfur coals may be used while meeting EPA SO_x standards. Additionally, products that may be affected by sulfur oxides can be heated directly by the clean gas. In addition to cost, a Stretford unit adds complexity to what may have been otherwise a simple system.

Organic condensates (tars and oils) are present when gasifiers operate at low temperatures, such as in Lurgi fixed-bed gasifiers. These organics are removed by condensation and separation through a water wash and, if needed, an electrostatic precipitator. Tars and oils are profitably separated from the water and can be used as a fuel or as chemical feedstock. The water, however, must be treated before discharge because it contains dissolved phenols and ammonia; it would probably be stream stripped and recycled to the quench and water-wash sections of the plant. Lurgi gasification, in particular, produces a large amount of condensate, and the required detarring and deoiling results in a significant drop in process thermal efficiency.

MBG pipelines present the same problems and hazards as CHP pipelines. Because combustible gas pipeline systems are currently in use, this should not pose a significant barrier. Piping standards and installation practices now used for natural gas pipelines could be applied to MBG pipelines.

5.2.3 Summary

The MBG seems likely to compete with oil from ~1990 onward. No significant technical or institutional barriers exist, and the cost is lower than most of the options considered in this study.

5.3 Hydrogen

5.3.1 Institutional issues

Hydrogen produced through coal gasification will be subject to the same standards and regulations discussed in Sect. 5.2.1 for MBG.

5.3.2 Technical issues

Technologies for producing hydrogen via KOH electrolysis or coal gasification are well established. The SPE electrolysis technology is still under development, although small SPE systems have been used for aerospace and military applications. General Electric is, however, planning a large scale pilot plant operation for ~1984 to 1986 (Ref. 5).

The major hazard associated with pipeline transmission of hydrogen gas is leakage followed by combustion. Because of its small molecular size, hydrogen is more likely to leak than any other combustible gas. Flammability limits are unusually broad (4 to 75% by volume with air) at normal temperature and pressure. The ignition temperature [580°C (1075°F)] is higher than that of most hydrocarbons [200 to 370°C (400 to 700°F)], but its ignition energy is an order of magnitude lower. Once ignited, flame speeds are of the order of tens of feet per second.

Hydrogen flames are nearly invisible and are characterized by a low level of thermal radiation to surrounding regions. Thus, a hydrogen fire is less likely to ignite other combustible materials by radiation than is a hydrocarbon fire. However, because a hydrogen flame is nearly invisible, it is difficult to detect; thus it is more of a personnel hazard than are hydrocarbon flames.

An explosion of hydrogen-air mixtures is very unlikely to occur as a result of a pipeline leak. Detonation limits of hydrogen-air mixtures range from 18 to 59% by volume. Because the hydrogen molecule has a very low mass, it would rise rapidly in air. Unless the leak occurred in a confined region, a concentration of 18% would probably not be reached. Furthermore, the ignition energy for detonation is very large compared with that required for combustion.

In the United States, a 0.2-m (8-in.) hydrogen pipeline has been operating for several years from Bayport to Baytown, Texas. Very little information about this pipeline appears in the literature. Most of the reported hydrogen pipeline operating experience is from Germany.[6] A hydrogen pipeline network 208 km in length is in use in the Rhine-Ruhr area. The original segments of this network have been operating since 1938. Pipe size varies, but much of the system appears to be 300 mm in diameter. Maximum pressure is 1.5 MPa gage; in 1973, the integrated network handled some 206×10^6 scm of hydrogen.

German operating experience has been quite good over a period of 30 years. Some fires have occurred, but there have been no explosions or problems with hydrogen embrittlement. Chemische Werke Hüls AG, the German operating organization, has expressed a high degree of confidence in their technical ability to design, construct, and operate an integrated hydrogen pipeline network.

5.3.3 Summary

The SPE appears to be the most promising way of supplying nonelectrical energy based on nuclear power to meet industrial energy needs. The technology development program is progressing according to schedule and shows promising results.

Utilities might invest in SPE hydrogen as a storage medium as well as an energy transport alternative. The oxygen by-product could be sold to synfuel plants or used for MBG production.

5.4 Molten Salt and Hot Oil[7,8]

5.4.1 Institutional issues[1,2]

Both molten salt and hot oil systems require an HTGR as an energy source. A hot oil system could conceivably be coupled to an LWR but could not then deliver the temperatures considered in this study. There are many barriers to HTGR commercialization in the United States. Barriers for the VHTR are discussed in Sect. 5.1.1 and may be summarized as: (1) licensing and regulatory problems, (2) supplier industry problems, (3) high capital investment requirements, and (4) rigid utility role definitions.

Because of the distances and the temperatures and fluids involved, acceptance of the systems by the general public is questionable. Questions of transport at high temperature through relatively high population density areas will add features to the controversy that were not included in the debate over the Alaskan oil pipeline.

An additional problem may arise in locating sources of capital for the project. The project is unlikely to be financed by venture capital because it is a new type of high-risk utility venture. Because of the limited potential for return on investment, it may be necessary to seek government cooperation in the form of low interest rate loans or tax-free bond status.

5.4.2 Technical issues

Technical issues common to all HTGR and VHTR options are discussed in Sect. 5.1.2: design modifications, graphite design criteria, metallic and ceramic materials development, and fuel cycle modifications.

Although heat-transfer salt does not have properties that inherently threaten either human life or the environment, operation with the salt at high temperature is a cause for concern.

Pipeline rupture and a fluid spill would result in property and environmental damage and create a potential threat to persons in the area. However, if a buried pipeline is considered, the potential danger is diminished except at pump stations.

Because of the line's segmented loop design, a broken section could be isolated and the volume of spilled fluid controlled. Further, because of its high melting point, spilled salt would freeze and its spread would be limited. Cleanup of the solid salt would be relatively easy, and no long-lasting effects on the environment should occur.

For oil, the potential exists for combustion of the pipeline fluid in the event of a pipe break. However, this problem should be accommodated by normal safety procedures and systems at the pump stations.

The consequences of spill from the line would be more extensive for oil than for salt because the oil spread is not self-limited by freezing. Neither is cleanup of an oil spill so direct a procedure as for salt, where gathering the solid phase material is rather easily accomplished. The probability of occurrence of such an accident should be only slightly greater than for normal liquid pipeline systems. The hot oil may, however, be more likely to burn immediately.

Two recognized technical concerns are the reliability of high-temperature pumps and the problem of system start-up or restart after a shutdown.

Pumps suitable for high-temperature service with molten salt are available. However, valid concerns exist regarding bearing life, seal life, and maintainability. Because of these concerns, redundancy would be a requirement rather than a design option.

The problem of start-up or remelt of the salt system may not be so significant as commonly considered. With the segmented loop design of the system, these operations could be accomplished section by section. Preheat of the line by steam flow is possible. Water dilution of the salt mixture lowers its melting point and can be used to ease the task of remelt.

Resistance heating by passing current from a low-voltage source through the pipe wall has apparent potential application to the remelt problem, particularly in short line sections. These procedures will be expensive and will require more study and refinement before application to large-scale systems.

A most discouraging technical feature for molten salt systems is the fact that corrosion rates at high temperature may be unacceptably high. In earlier studies consulted, corrosion was not considered a significant problem. However, other results[9] indicate that corrosion rates may limit systems to operation below 550°C. This would imply energy supply at temperatures of the 500°C level, which excludes a significant fraction of the potential market for high-temperature energy.

Operation of the oil system heat exchangers at high temperatures may cause deterioration of the fluid and fouling of the heat exchanger surface. Fluid deterioration is a problem in high-temperature systems, and the system would probably need to operate with continuous fluid regeneration. Careful temperature control can alleviate the problem, and regeneration procedures are well-developed standard items of technology.

5.4.3 Summary

The molten salt and hot oil systems are economically competitive over limited distances (~120 km). The basic weakness of these systems is their reliance on HTGR energy sources. Unless some of the substantial institutional barriers faced by HTGR commercialization are removed, it is unlikely that one of these systems will ever be built.

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6. RESULTS AND CONCLUSIONS

6.1 Introduction

This study examined the possibility of substituting nonelectrical energy forms, generated at central nuclear plants, for the industrial energy consumption of oil and natural gas. The purpose of the study was to examine prior work in this area and place the various alternatives on an equal basis for comparison. The comparison was then used to understand the relative benefits of HTGRs and LWRs and to indicate possible directions for future research and development.

6.2 CHP

The CHP was the most expensive energy transport alternative considered, which is not surprising, because it delivers only ~1.7 MJ for every standard cubic meter of gas transported. This output can be compared with ~33.5 MJ/scm for natural gas and ~10 MJ/scm for MBG. Unless some major advance is made in CHP technology, this system does not merit further consideration.

Compounding the economic problems inherent in the CHP system is its reliance on the VHTR. The significant barriers to VHTR commercialization might alone rule out the CHP even if the CHP were not so expensive.

6.3 MBG and Hydrogen

The results for MBG and SPE and coal-derived hydrogen are very similar. They both show competitive costs and can be economically transported over long distances. The KOH hydrogen was much more expensive and should not be considered for future industrial fuel substitution.

In this study, each system was sized to produce the same energy output to allow easy system comparison. This sizing may not be ultimately correct, because the longer distance alternatives, hydrogen and MBG, are likely to gather in more users. The extension of the hydrogen and MBG systems to larger sizes was beyond the scope of this project. The beneficial effects of size will have been largely determined at the 1200-MW delivered-energy level. Because the transport costs for these systems are a very small portion of their total cost, pipeline size changes should have a negligible effect on the delivered-energy cost.

Both MBG and hydrogen could fill fuel and feedstock needs in the petroleum and chemical industries. Both of these industries meet ~70% of their steam requirements by firing oil and natural gas boilers. The use of MBG or hydrogen could substitute nuclear and coal resources for these prime fuels.

The SPE hydrogen pricing was based on PWR-generated electricity. The only impediment to the immediate application of SPE systems is the developmental nature of the SPE technology. Major pilot plants are needed to further test and prove the concept. An SPE system could be placed on a user's site and supplied with nuclear-based electricity. This option was not evaluated here.

The improved MBG and coal-based hydrogen systems are in a similar situation. Again, large-scale plants are needed to test the concept and to provide operating experience.

The next step in any evaluation of these systems should be a site-specific study with a defined set of industrial customers.

6.4 Molten Salt and Hot Oil

Molten salt and hot oil systems are shown to be competitive for short and intermediate distances ($\lesssim$120 km). The high temperatures available with these sensible heat carriers are appropriate for a large portion of the industrial energy market including the chemical, petroleum, metals, and stone/clay/glass industries. The chemical and petroleum industries in particular consume large amounts of oil and natural gas to raise steam of varying pressures. Some of this steam is required at the high pressures that would require the high-temperature energy available with a molten salt system. These industries are concentrated in the central southwest United States.

Based on their economics, these systems deserve immediate further consideration. However, their dependence on an HTGR energy source and the significant institutional barriers to HTGR commercialization raise serious doubts about the practicality of these energy transport systems. Unless these barriers are lessened or removed, site-specific evaluations are not merited.

6.5 Summary

Figures 6.1 and 6.2 show the costs of each alternative for 1990 and 2000 start-ups. Oil and electricity costs are given for comparison. These results indicate that centralized hydrogen distribution systems supplied by nuclear power would be competitive with synthetic gas systems for long-distance transmission for 2000 start-up. If the barriers to HTGR commercialization are overcome, the molten salt and hot oil systems can be considered for intermediate distances.

The CHP systems are not economically competitive in this study. Any future work on the CHP should be directed at finding new technological methods that could improve the economics of the system.

The industrial market review included in this study shows that there are several geographical areas where groups of industrial energy users are clustered. While this study is focused on the industrial market, the concept could also be applied to commercial or residential district heating systems.

The next logical step is the detailed examination of candidate distribution systems (molten salt, hot oil, MBG, and hydrogen) laid out to serve existing or potential user groups. This examination would extend previous studies[1] of isolated groups of users.

Both PWRs and HTGRs appear to have significant market potential. The PWR is used in the SPE hydrogen system and the HTGR in the hot oil and molten salt systems.

Although this study was directed at nonelectrical energy forms, the results show that the direct use of electricity may be a good choice. The concept of electrification deserves serious consideration and further study.

Reference

1. Dow Chemical Company, *The Potential Industrial Market for Process Heat from Nuclear Reactors,* Dow Chemical Company for Oak Ridge National Laboratory, ORNL/TM-5516 (July 1976).

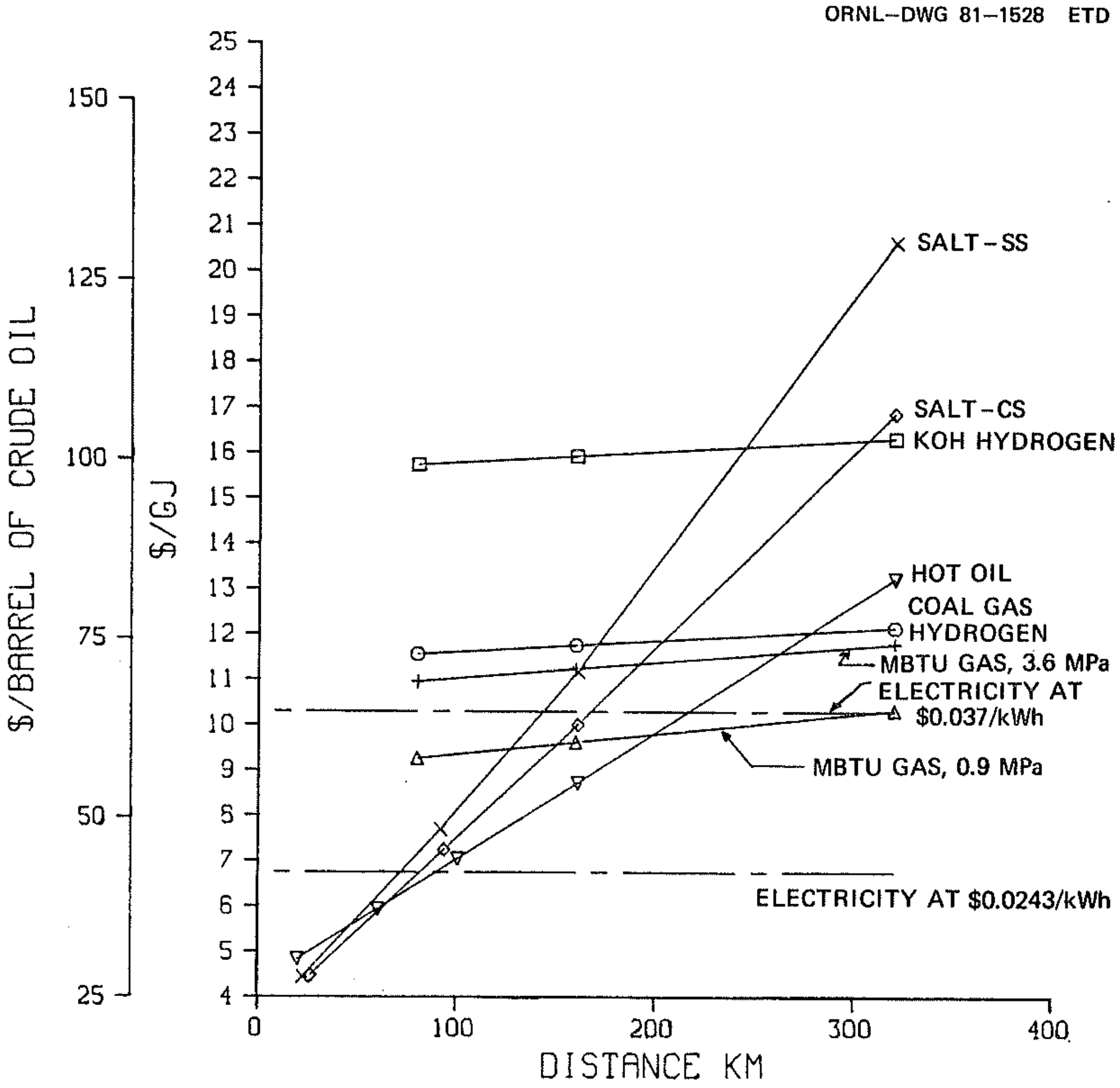

Fig. 6.1. Thermal energy transport options, 1200 MW, 1990 start-up, 1980 dollars.

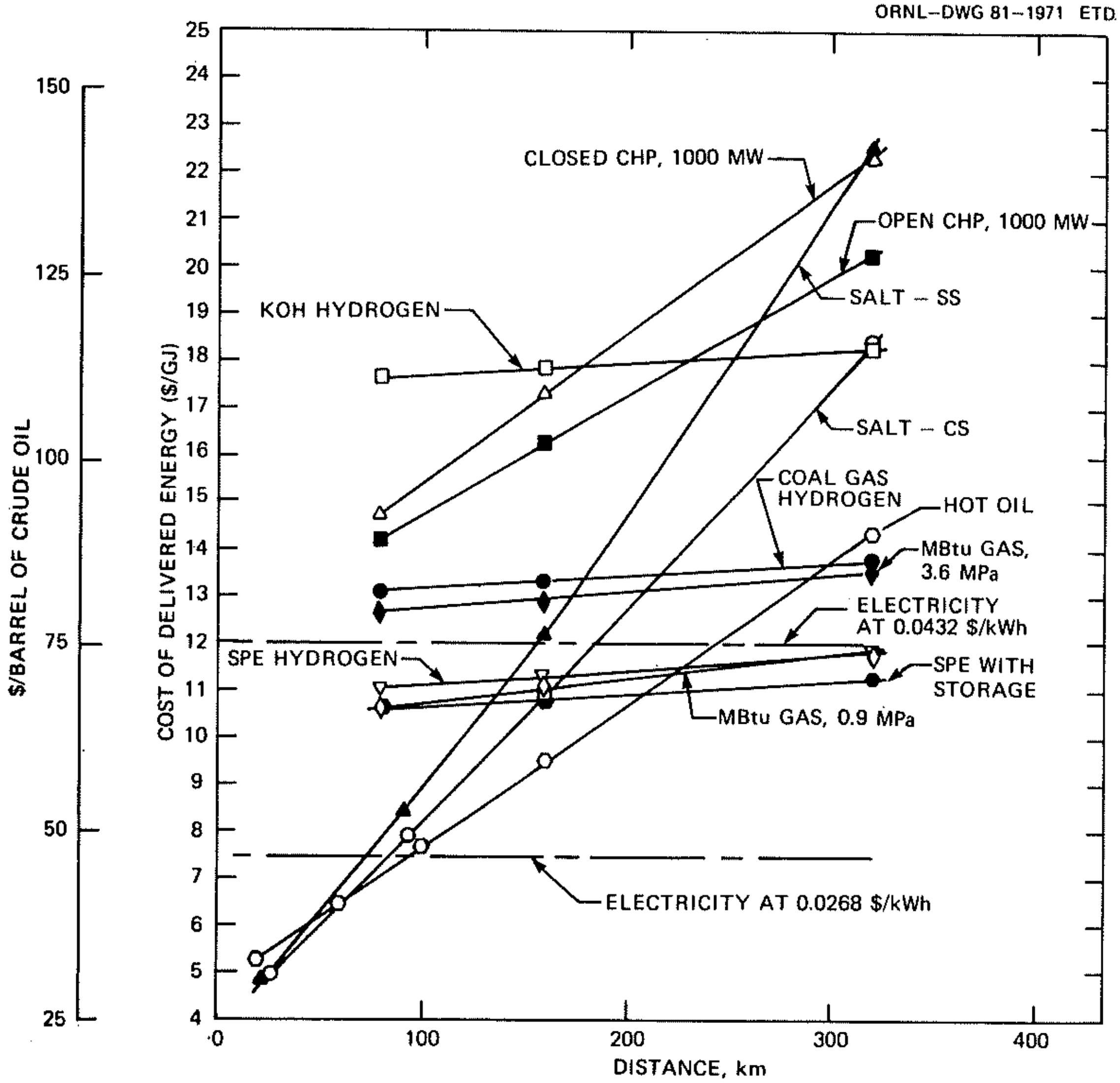

Fig. 6.2. Thermal energy transport options, 1200 MW, 2000 start-up, 1980 dollars.

Appendix A

INDUSTRIAL PROCESS HEAT APPLICATIONS AND REQUIREMENTS

Table A.1. Process heat requirements by two-digit SIC code, 1974

(10^{15} J)

Industry	SIC	Hot water <100° C	Steam		Direct heat or heated gas						Total fuels consumed
			100–200° C	200–350° C	<100° C	100–200° C	200–350° C	350–550° C	550–1000° C	>1000° C	
Food	20	118.1	38.1		205.8	309.4	124.9	20.2			817.2
Tobacco	21		2.9			7.6					10.5
Textile	22	127.0	34.7			68.8					230.5
Lumber, wood	24	3.0	13.4			147.2					163.6
Furniture	25				43.1						43.1
Paper	26	175.4	727.3	77.6		157.2				95.7	1,233.2
Chemicals	28	31.5	1,105.2	429.7	168.9	265.7	205.3		276.9	63.9	2,549.1
Petroleum	29		56.4	178.5		99.2	128.6	614.8	378.4		1,455.9
Rubber, plastics	30			68.3		119.8					188.1
Leather	31	2.3			14.7						17.0
Stone, clay, glass	32	18.8	30.6	24.6		46.9	12.9	28.6	247.6	824.5	1,234.5
Primary metals	33	86.1	40.8		3.0		46.0		0.7	1,968.3	2,144.9
Fabricated metal	34				141.6			180.2			321.8
Machinery	35	44.6	4.8	73.1		35.3	37.4			78.1	273.2
Electric	36				4.5	14.0		146.6			165.1
Transportation	37	46.6	5.0	76.4		36.9	39.1			81.6	285.6
Total including SIC 29		654.1	2,059.2	928.2	581.6	1,308.0	594.2	990.4	903.6	3,112.1	11,131.4
Percent of total fuels		(5.9%)	(18.5%)	(8.3%)	(5.2%)	(11.8%)	(5.3%)	(8.9%)	(8.1%)	(28.0%)	(100%)
Total excluding SIC 29		654.1	2,002.8	749.7	581.6	1,208.8	456.6	375.6	525.2	3,112.1	9,675.5
Percent of total fuels		(6.8%)	(20.7%)	(7.7%)	(6.0%)	(12.5%)	(4.8%)	(3.9%)	(5.4%)	(32.2%)	(100%)

Source: Sara Wood Boercker, *Characterization of Industrial Process Energy Services* (Oak Ridge, Tennessee: Oak Ridge Associated Universities, Institute for Energy Analysis, May 1979, ORAU IEA-79-9-[R], p. 11.

Table A.2. Purchased fuels and electric energy used by major industry groups, 1977

SIC	Major industry group	Total fuels and electric energy (10^{12} Btu)	Fuels (10^{12} Btu)	Fuel oil		Bituminous coal, lignite, and anthracite (10^3 short tons)	Coke and coke breeze (10^3 short tons)	Natural gas (10^9 ft^3)	Electric energy	
				Distillate (10^3 bbl)	Residual (10^3 bbl)				Purchased (10^6 kWh)	Generated less sold (10^6 kWh)
28	Chemicals and allied products	2,939.5	2,418.8	21,489.8	44,244.5	13,045.8	49.5	1,451.8	152,604.3	13,205.3
33	Primary metals industry	2,370.3	1,839.7	11,628.8	41,702.2	4,815.0	13,840.9	889.0	155,482.7	13,292.9
26	Paper and allied products	1,293.3	1,145.5	6,707.0	76,126.7	8,152.2	D[a]	348.3	43,297.7	27,368.1
29	Petroleum and coal products	1,288.8	1,183.4	4,179.1	7,542.4	183.0	D	1,027.8	30,917.5	4,432.2
32	Stone, clay, and glass products	1,263.3	1,153.9	14,786.5	12,202.6	13,011.0	257.6	523.2	32,081.6	441.2
20	Food and kindred products	980.5	834.9	13,578.6	18,573.0	3,245.4	55.0	429.6	42,686.0	2,343.9
34	Fabricated metals products	404.3	307.8	4,642.5	3,471.6	388.7	53.7	193.5	28,283.0	S[b]
37	Transportation equipment	389.1	281.7	3,030.3	5,508.4	1,728.5	D	136.9	31,489.9	S
22	Textile mill products	337.6	239.5	4,680.8	11,737.6	1,208.7	Z[c]	74.4	28,623.0	442.5
35	Machinery except electric	336.5	237.9	3,342.0	3,536.4	709.1	107.6	126.9	28,912.2	265.7

[a] D = Withheld to avoid disclosing figures for individual companies.

[b] S = Withheld because estimate did not meet publication standards.

[c] Z = Less than $50,000, or less than half the unit of measure specified for quantities.

Source: U.S. Department of Commerce, Bureau of the Census, *1977 Census of Manufacturers, Preliminary Report, Fuels and Electric Energy Consumed* (Washington, D.C., May 1979, MC-77-SR-4(P)), pp. 4 and 5.

Table A.3. List of industrial process heat applications and annual requirements (1974)

Industry—SIC group	Application temperature requirement [°C (°F)]	Process heat used for application [10^{12} Btu/year (10^{12} kJ/year)]
Iron ore—1011		
Pelletizing of concentrates	1288-1371 (2350-2500)	37.2 (39.2)
Copper concentrate—1021		
Drying	121 (250[a])	1.7 (1.8)
Bituminous coal—1211		
Drying (including lignite)	66-104 (150-220[a])	18 (19)
Sand and gravel—1442		None
Potash—1474		
Drying filter cake	121 (250[a])	1.03 (1.09)
Phosphate rock—1475		
Calcining	760-871 (1400-1600)	0.71 (0.75)
Drying	232 (450[a])	10.5 (11.1)
Sulfur—1477		
Frasch mining	163-171 (325-340)	60 (63)
Meat Packing—2011		
Sausages and prepared meats—2013		
Scalding, carcass wash, and cleanup	60 (140)	43.7 (46.1)
Singeing flame	260 (500)	1.06 (1.12)
Edible rendering	93 (200)	0.52 (0.55)
Smoking/cooking	68 (155)	1.16 (1.22)
Poultry dressing—2016		
Scalding	60 (140)	3.16 (3.33)
Natural cheese—2022		
Pasteurization	77 (170)	1.28 (1.35)
Starter vat	57 (135)	0.02 (0.02)
Make vat	41 (105)	0.47 (0.50)
Finish vat	38 (100)	0.02 (0.02)
Whey condensing	71-93 (160-200)	10.2 (10.8)
Whey drying	49 (120[a])	2.94 (3.10)
Process cheese blending	74 (165)	0.07 (0.07)
Condensed and evaporated milk—2023		
Stabilization	93-100 (200-212)	2.93 (3.09)
Evaporation	71 (160)	5.20 (5.48)
Spray drying	177-204 (350-400)	3.58 (3.78)
Sterilization	121 (250)	0.54 (0.57)
Fluid milk—2026		
Pasteurization	72-77 (162-170)	1.44 (1.52)
Canned specialities—2032		
Beans		
Precook (blanch)	82-100 (180-212)	0.40 (0.42)
Simmer blend	77-100 (170-212)	0.24 (0.25)
Sauce heating	88 (190)	0.20 (0.21)
Processing	121 (250)	0.38 (0.40)
Canned fruits and vegetables—2033		
Blanching/peeling	82-100 (180-212)	1.88 (1.98)
Pasteurization	93 (200)	0.15 (0.16)
Brine syrup heating	93 (200)	1.02 (1.08)
Commercial sterilization	100-121 (212-250)	1.67 (1.76)
Sauce concentration	100 (212)	0.44 (0.46)
Dehydrated fruits and vegetables—2034		
Fruit and vegetable drying	74-85 (165-185)	5.84 (6.16)
Potatoes		
Peeling	100 (212)	0.33 (0.35)
Precooking	71 (160)	0.47 (0.50)
Cooking	100 (212)	0.47 (0.50)
Flake dryer	177 (350)	1.09 (1.15)
Granule flash dryer	288 (550)	1.09 (1.15)

Table A.3 (continued)

Industry—SIC group	Application temperature requirement [°C (°F)]	Process heat used for application [10^{12} Btu/year (10^{12} kJ/year)]
Frozen fruits and vegetables—2037		
Citrus juice concentration	88 (190)	1.33 (1.40)
Juice pasteurization	93 (200)	0.27 (0.28)
Blanching	82–100 (180–212)	2.26 (2.38)
Cooking	77–100 (170–212)	1.41 (1.49)
Wet corn milling—2046		
Steep water evaporator	177 (350)	3.66 (3.86)
Starch dryer	49 (120^{a})	3.03 (3.20)
Germ dryer	177 (350)	1.92 (2.03)
Fiber dryer	538 (1000)	2.93 (3.09)
Gluten dryer	177 (350)	1.32 (1.39)
Steepwater heater	49 (120)	0.77 (0.81)
Sugar hydrolysis	132 (270)	1.89 (1.99)
Sugar evaporator	121 (250)	2.74 (2.89)
Sugar dryer	49 (120^{a})	0.16 (0.17)
Prepared feeds—2048		
Pellet conditioning	82–88 (180–190)	2.28 (2.40)
Alfalfa drying	204 (400^{a})	16.8 (17.7)
Bread and baked goods—2051		
Proofing	38 (100)	0.84 (0.89)
Baking	216–238 (420–460)	6.40 (6.75)
Cane sugar refining—2062		
Mingler	52–74 (125–165)	0.59 (0.62)
Melter	85–91 (185–195)	3.30 (3.48)
Defecation	71–85 (160–185)	0.44 (0.46)
Revivification	399–599 (750–1110)	3.96 (4.18)
Granulator	43–54 (110–130)	0.44 (0.46)
Evaporator	129 (265)	26.39 (27.84)
Beet sugar—2063		
Extraction	60–85 (140–185)	4.63 (4.88)
Thin juice heating	85 (185)	3.08 (3.25)
Lime calcining	538 (1000)	2.98 (3.14)
Thin syrup heating	100 (212)	6.68 (7.05)
Evaporation	132–138 (270–280^{a})	30.8 (32.5)
Granulator	66–93 (150–200)	0.15 (0.16)
Pulp dryer	110–138 (230–280^{a})	16.5 (17.4)
Soybean oil mills—2075		
Bean drying	71 (160)	4.05 (4.27)
Toaster desolventizer	102 (215)	6.08 (6.41)
Mean dryer	177 (350^{a})	4.36 (4.60)
Evaporator	107 (225)	1.62 (1.71)
Stripper	100 (212)	0.30 (0.32)
Animal and marine fats—2077		
Continuous rendering of inedible fat	166–177 (330–350)	16.5 (17.4)
Shortening and cooking oil—2079		
Oil heater	71–82 (160–180)	0.72 (0.76)
Wash water	71–82 (160–180)	0.12 (0.13)
Dryer preheat	93–132 (200–270)	0.60 (0.63)
Cooking oil reheat	93 (200)	0.32 (0.34)
Hydrogenation preheat	149 (300)	0.37 (0.39)
Vacuum deodorizer	149–204 (300–400)	0.35 (0.37)
Malt beverages—2082		
Cooker	100 (212)	1.53 (1.61)
Water heater	82 (180)	0.53 (0.56)
Mash tub	77 (170)	0.60 (0.63)
Grain dryer	204 (400^{a})	9.18 (9.68)
Brew kettle	100 (212)	3.98 (4.20)

Table A.3 (continued)

Industry SIC group	Application temperature requirement [°C (°F)]	Process heat used for application [10^{12} Btu/year (10^{12} kJ/year)]
Distilled liquor—2085		
Cooking (whiskey)	100 (212)	3.16 (3.33)
Cooking (spirits)	160 (320)	6.27 (6.61)
Evaporation	121-143 (250-290[a])	2.32 (2.45)
Dryer (grain)	149-204 (300-400)	1.94 (2.05)
Distillation	110-121 (230-250)	7.69 (8.11)
Soft drinks—2086		
Bulk container washing	77 (170)	0.21 (0.22)
Returnable bottle washing	77 (170)	1.27 (1.34)
Nonreturnable bottle warming	24-29 (75-85)	0.43 (0.45)
Can warming	24-29 (75-85)	0.52 (0.55)
Cigarettes—2111		
Drying	104 (220[a])	0.43 (0.45)
Rehumidification	104 (220[a])	0.43 (0.45)
Tobacco stemming and redrying—2141		
Drying	104 (220[a])	0.50 (0.26)
Finishing plants, cotton—2261		
Washing	100 (200)	15.4 (16.2)
Dyeing	100 (200)	4.5 (4.7)
Drying	135 (275)	22.2 (23.4)
Finishing plants, synthetic—2262		
Washing	93 (200)	35.9 (37.9)
Dyeing	100 (212)	15.2 (16.0)
Drying and heat setting	135 (275)	23.2 (24.5)
Logging camps—2411		None
Sawmills and planing mills—2421		
Kiln drying of lumber	149 (300)	63.4 (66.9)
Plywood—2435		
Plywood drying	121 (250)	50.6 (53.4)
Veneer—2436		
Veneer drying	100 (212)	57.8 (61.0)
Wooden furniture—2511		
Makeup air and ventilation	21 (70)	5.7 (6.0)
Kiln dryer and drying oven	66 (150)	3.8 (4.0)
Upholstered furniture—2512		
Makeup air and ventilation	21 (70)	1.4 (1.5)
Kiln dryer and drying oven	66 (150)	0.9 (0.9)
Pulp mill—2611		
Paper mills—2621		
Paperboard mills—2631		
Building paper—2661		
Pulp digestion	188 (370)	253 (267)
Pulp refining	66 (150)	175 (185)
Black liquor treatment	138 (280)	164 (173)
Chemicals recovery-calcining	1038 (1900)	96 (101)
Pulp and paper drying	143 (290)	383 (404)
Solid and corrugated fiber boxes—2653		
Corrugating and glue setting	149-177 (300-350)	21.6 (22.8)
Alkalies and chlorine—2812		
Mercury cell (to be phased out by 1983)		6.4 (6.8)
Diaphragm cell	177 (350)	82.1 (86.6)
Cyclic intermediates—2865		
Ethylbenzene	177 (350)	3 (3)
Styrene	121-177 (250-350)	35 (37)
Phenol	121 (250)	0.45 (0.47)

Table A.3 (continued)

Industry SIC group	Application temperature requirement [°C (°F)]	Process heat used for application [10^{12} Btu/year (10^{12} kJ/year)]
Alumina—28195		
Digesting, drying, heating	138 (280)	113.2 (119.4)
Calcining	1204 (2200)	35.3 (37.2)
Plastic materials and resins—2821		
Polystyrene, suspension process		
Polymerizer preheat	93–102 (200–215)	0.102 (0.107)
Heating wash water	88–93 (190–200)	0.064 (0.068)
Drying	93 (200)	0.034 (0.036)
Synthetic rubber—2822		
Cold SBR latex crumb		
Bulk storage	27–38 (80–100)	0.179 (0.189)
Emulsification	27–38 (80–100)	0.086 (0.091)
Blowdown vessels	54–63 (130–145)	0.865 (0.912)
Monomer recovery by flashing and stripping	49–60 (120–140)	4.095 (4.319)
Dryer air temperature	66–93 (150–200)	3.663 (3.864)
Cold SBR, oil-carbon black masterbatch		
Dryer air temperature	66–93 (150–200)	0.506 (0.534)
Oil emulsion holding tank	27–38 (80–100)	0.028 (0.030)
Cold SBR, oil masterbatch		
Dryer air temperature	66–93 (150–200)	1.09 (1.15)
Oil emulsion holding tank	27–38 (80–100)	0.090 (0.095)
Cellulosic man-made fibers—2823		
Polyester	<288 (<550)	48.9 (51.6)
Nylon	<279 (<535)	41.7 (44.0)
Acrylic	<121 (<250)	23.5 (24.8)
Polypropylene	<282 (<540)	3.9 (4.1)
Noncellulosic fibers—2824		
Rayon	<100 (<212)	37.8 (39.9)
Acetate	<100 (<212)	37.6 (39.7)
Pharmaceutical preparations—2834		
Autoclaving and cleanup	121 (250)	18.85 (19.88)
Tablet and dry-capsule drying	121 (250)	1.00 (1.05)
Wet capsule formation	66 (150)	0.05 (0.05)
Soaps and detergents—2841		
Soaps		
Various processes in soap manufacture	82 (180)	0.50 (0.53)
High-temperature processes	254 (490)	0.002 (0.002)
Spray drying	260 (500[a])	0.001 (0.001)
Detergents		
Various low-temperature processes	82 (180)	0.36 (0.38)
High-temperature processes	260 (500)	0.001 (0.001)
Drum-dried detergents	177 (350[a])	0.31 (0.33)
Spray-dried detergents	260 (500[a])	0.019 (0.020)
Organic chemicals, n.e.c.—2869		
Ethanol	93–121 (200–250)	6 (6)
Isopropanol	93–177 (200–350)	11 (12)
Cumene	121 (250)	1 (1)
Vinyl chloride monomer	121–177 (250–350)	9 (9)
Urea—2873215		
High-pressure steam-heated stripper	191 (375)	5.07 (5.35)
Low-pressure steam-heated stripper	143 (290)	0.89 (0.94)
Explosives—2892		
Dope (inert ingredients)		
Drying	149 (300)	0.006 (0.006)
Wax melting	93 (200)	0.118 (0.124)
Nitric acid concentrator	121 (250)	0.070 (0.074)
Sulfuric acid concentrator	93 (200)	0.027 (0.028)
Nitric acid plant	93 (200)	0.223 (0.235)
Blasting cap manufacture	93 (200)	0.016 (0.017)

Table A.3 (continued)

Industry SIC group	Application temperature requirement [°C (°F)]	Process heat used for application [10^{12} Btu/year (10^{12} kJ/year)]
Petroleum refining—2991		
Crude distillation		
Atmospheric topping	343 (650)	275 (290)
Vacuum distillation	227–427 (440–800)	183 (193)
Thermal operations	291–543 (555–1010)	154 (162)
Catalytic cracking	607 (1125)	447 (471)
Delayed coking	482 (900)	225 (237)
Hydrocracking	268–432 (515–810)	91 (96)
Catalytic reforming	496 (925)	498 (525)
Catalytic hydrorefining	371 (700)	52 (55)
Hydrotreating	371 (700)	124 (131)
Alkylation	7–171 (45–340)	59 (62)
Hydrogen plant	871 (1600)	124 (131)
Olefins and aromatics	649 (1200)	124 (131)
Lubricants	Unavailable	25 (26)
Asphalt	Unavailable	96 (101)
Butadiene	121–177 (250–350)	60 (63)
Paving mixtures—2951		
Aggregate drying	135–163 (275–325[a])	88.1 (92.9)
Heating asphalt	163 (325)	4.93 (5.20)
Asphalt felts and coatings—2952		
Saturator	204–260 (400–500)	1.52 (1.60)
Asphalt coating	149–204 (300–400)	1.23 (1.30)
Drying (steam)	177 (350)	3.32 (3.50)
Sealant	149–204 (300–400)	0.57 (0.60)
Tires and inner tubes—3011		
Vulcanization	121–171 (250–340)	6.18 (6.52)
Plastics products—3079		
Blow-molded bottles		
High-density polyethylene	218 (425)	3.52 (3.71)
Leather tanning and finishing—3111		
Bating	32 (90)	0.094 (0.099)
Chrome tanning	29–54 (85–130)	0.060 (0.063)
Retan, dyeing, fat liquor	49–60 (120–140)	0.15 (0.16)
Wash	49 (120)	0.034 (0.036)
Drying	43 (110[a])	2.05 (2.16)
Finishing drying	43 (110[a])	0.13 (0.14)
Flat glass—3211		
Melting	1260–1482 (2300–2700)	50.1 (52.8)
Fabrication (including tempering and laminating)	799–1093 (1470–2000)	3.5 (3.7)
Annealing	499 (930)	5.9 (6.2)
Glass containers—3221		
Melting-firing	1482–1593 (2700–2900)	98.60 (104.0)
Conditioning	816–1093 (1500–2000)	42.25 (44.56)
Annealing	649 (1200)	12.81 (13.51)
Post forming	649 (1200)	1.42 (1.50)
Hydraulic cement—3241		
Drying	135–163 (275–325[a])	8.0 (8)
Calcining	1260–1482 (2300–2700)	468.0 (494)
Brick and structural tile—3251		
Brick kiln	1371 (2500)	70.4 (74.2)
Clay refractories—3255		
Refractories firing	1816 (3300)	9.0 (9.5)

Table A.3 (continued)

Industry/SIC group	Application temperature requirement [°C (°F)]	Process heat used for application [10^{12} Btu/year (10^{12} kJ/year)]
Concrete block—3271		
Low-pressure curing	74 (165^a)	12.29 (12.96)
Autoclaving	182 (360)	5.42 (5.72)
Ready-mix concrete—3273		
Hot water for mixing concrete	49–88 (120–190)	0.34 (0.36)
Lime—3274		
Calcining	982 (1800)	129.9 (137.0)
Gypsum—3275		
Kettle calcining	166 (330)	10.0 (10.5)
Wall board drying	149 (300)	11.18 (11.79)
Treated minerals—3295		
Expanded clay and shale		
Bloating process	982 (1800)	29.1 (30.7)
Fuller's earth		
Drying and calcining	593 (1100)	6.37 (6.72)
Kaolin		
Calcining	1040 (1900)	1.4 (1.5)
Drying	110 (230^a)	12.7 (13.4)
Expanded perlite		
Drying	71 (160^a)	0.22 (0.23)
Expansion process	871 (1600)	1.7 (1.8)
Barium		
Drying	110 (230^a)	0.34 (0.36)
Blast furnaces and steel mills—3312		
High-temperature uses	1482 (2700)	3300 (3480)
Ferrous castings		
Gray iron foundries—3321 (73% of heat)		
Malleable iron foundries—3322 (10% of heat)		
Steel foundries—3323 (17% of heat)		
Melting in cupola furnaces	1482 (2700)	146 (154)
Mold and core preparation	149–246 (300–475)	117.7 (124.1)
Heat treatment and finishing	482–982 (900–1800)	16 (17)
Pickling	38–100 (100–212)	151 (160)
Primary copper—3331		
Smelting and fire-refining	1095–1371 (2000–2500)	32.58 (34.37)
Primary zinc—3333		
Pyrolytic reduction	1300 (2400)	1.0 (1.1)
Primary aluminum—3334		
Prebaking anodes	1093 (2000)	8.14 (8.59)
Galvanizing—3479		
Cleaning, pickling	54–88 (130–190)	0.011 (0.012)
Galvanizing (melting zinc)	454 (850)	0.014 (0.015)
Motors and generators—3621		
Drying and preheat	66 (150)	0.043 (0.045)
Baking	177 (350)	0.133 (0.140)
Oxide coat laminations	816–927 (1500–1700)	0.72 (0.76)
Annealing	816 (1500)	0.67 (0.71)
Motor vehicles—3711		
Baking-prime and paint ovens	121–149 (250–300)	0.29 (0.31)
Casting foundry	1454 (2650)	23 (24)
Inorganic pigments—2816		
Drying chrome yellow	93 (200)	0.075 (0.079)

[a] No special temperature required; requirement is simply to evaporate water or to dry the material.

Appendix B

CURRENT HYDROGEN PRODUCTION PRACTICES

While the bulk of hydrogen produced is manufactured by steam-reforming methane, other important processes are used, particularly for captive hydrogen production in refining and petrochemical plants.

B.1 Steam Reformer Process (Fig. B.1)

The largest quantities of hydrogen are manufactured by the catalytic reaction of hydrocarbons and steam (yielding H_2 and CO_x) followed by the water-gas shift reaction. The most commonly used raw material for this process is natural gas, although other natural and refinery hydrocarbons (including naphtha cuts) may be used. The following description is based on propane as a raw material.

Commercial propane, obtained from either natural-gasoline plants or oil refineries, contains small amounts of organic sulfur compounds that are removed before processing. The propane from storage passes (in the form of vapors) through a heater at a temperature of ~370° C. The hot gases then pass over a bauxite or metallic oxide catalyst, which converts the sulfur compounds (mercaptans, organic sulfides, and carbonyl sulfide) to hydrogen sulfide. After cooling, the gases are scrubbed with aqueous sodium hydroxide and water to remove the soluble sulfides. When natural gas is used as the process feed, sulfur is removed by passage of the gas through drums containing activated carbon.

The sulfur-free propane vapors are then mixed with steam and passed into the top of a reforming furnace. One type of furnace consists of a number of vertical nickel-chromium-iron alloy tubes (7.5 to 20 cm in diameter and about 7.5 m long) mounted in a refractory furnace. Heat for the endothermic reaction is supplied by multiple horizontal burners located at various levels, with the flue gases passing upward countercurrent to the process gas. At temperatures of 760 to 980° C, the propane gas passes

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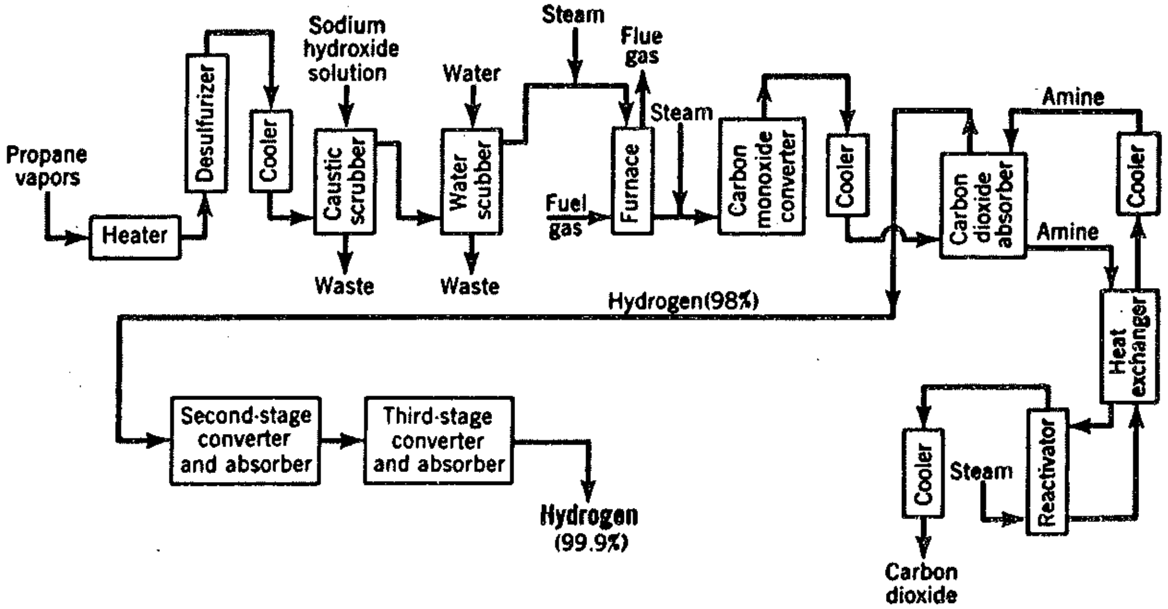

Fig. B.1. Hydrogen from light hydrocarbons and steam.

down the tubes over a supported nickel catalyst at a space velocity of about 600 volumes per hour per volume of catalyst. The propane is converted to H_2, CO, and CO_2, with a trace of methane remaining in the mixture.

Operating pressure in the reformer may be as high as 4.2 MPa (600 psi). Although the steam-hydrocarbon reaction is benefited by low pressure, there are compensating economic advantages in operating at high pressure. Accordingly, new plants usually are designed for high-pressure operation.

In any event, the reformed gases are cooled to ~370° C when mixed with steam and then passed into the first-stage CO converter containing an iron oxide catalyst promoted with chromium oxide. The exothermic conversion reaction (water-gas shift) takes place at a temperature of ~425° C and a space velocity of 100 (or greater) volumes of gas per volume of catalyst per hour. Both this catalyst and the nickel-reforming catalyst are rugged and have a normal life of one year or more.

From the converter, the gases containing a small amount of CO are cooled to ~38° C and passed into a packed (or bubble-tray) tower. Aqueous 15 to 20% monoethanolamine is circulated down through the countercurrently blowing gas (Girbotol process). The amine solution absorbs the CO_2 and, after passing through heat exchangers, is run to the top of a reactivating tower. The CO_2 is desorbed by steam generated by heating the solution in a reboiler at the bottom of the tower. The CO_2 removed amounts to about 30 volumes per 100 volumes of hydrogen, and since it is recovered at a purity of 99.8%, the CO_2 is available as a useful by-product. The regenerated amine solution is then returned to the system. At atmospheric pressure and 38° C, hydrogen gas containing 20% CO_2 may be purified to 0.1% CO_2 by scrubbing, with the monoethanolamine absorbing 15 to 30 m^3 of CO_2 per m^3 of solution circulated. Approximately 0.12 kg of steam per liter of solution is required for regeneration. The carbon-dioxide-free hydrogen coming from the absorber still contains ~1% CO. This CO is removed by passing H_2 through two more stages of CO conversion, followed by CO_2 removal. From the last absorber, the purified H_2 analyzes better than 99.9% pure. The gas may be compressed to ~1 MPa (150 psi) and placed in storage tanks. Many new plants require only one stage of shift conversion in which a highly active low-temperature catalyst is used.

Shift converter gas may also be purified by absorption of CO_2 in hot carbonate solutions. In one modification (Vetrocoke process), the potassium carbonate solution is activated with arsenic trioxide. Small amounts of carbon oxides may be removed by conversion to methane by passing the mixture over a nickel catalyst at 315 to 425° C.

B.2 From Hydrocarbons by Partial Oxidation (Fig. B.2)

The following reactions take place in this process:

$$C_nH_{2n-2} + mO_2 \rightarrow xCO_2 + (n-x)CO + (n-1-y)H_2 + yH_2O\ ,$$

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

The material and utility requirements to produce 28 scm (1000 scf) H_2 are

Fuel	10 kg
Oxygen (95%)	10 kg
Electricity	7.2 MJ (2 kWh)
Refrigeration	4100 kg/h

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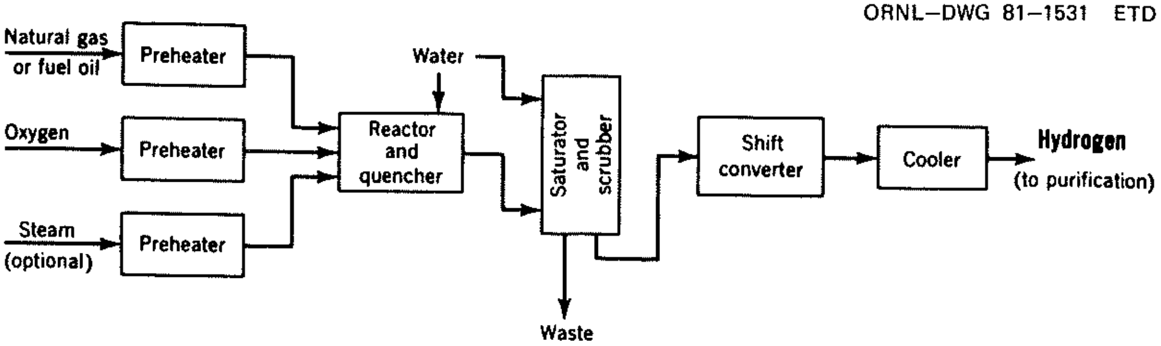

Fig. B.2. Hydrogen from hydrocarbons by partial oxidation.

Process

Synthesis gas, a mixture of H_2 and CO, may be made by partially burning any hydrocarbon feed (natural gas through fuel oil) with O_2. The CO then reacts with water vapor in a shift converter to form more H_2.

Natural gas, for example, is preheated to 650°C and sent to a refractory-lined pressure vessel where it reacts with preheated oxygen. Hydrocarbon feed rate varies from 1000 to 3000 volumes per unit volume of reactor space. The O_2/C ratio is held between 1.0 and 1.2 to hold C formation below 0.5 kg/1000 m^3 (14 kg/million ft^3) of gas. Customary operating pressure is 1.4 to 3.5 MPa (200 to 500 psi).

Reaction products are removed continuously from the reaction zone. In addition to the major products, CO and H_2, the burner effluent contains 5 mole % water, <2 mole % CO_2, <0.1 mole % O_2 and ~0.5% unreacted methane. The effluent gases are quenched rapidly (<1 s) from 1425°C to 530 or 815°C. In a similar process used largely in Europe, partial combustion is moderated by the presence of N_2, steam, or CO_2; the quench is carried out in a waste-heat boiler.

In either case the gases may be scrubbed (to remove carbon particles), saturated, and then passed to a shift converter, where the CO reacts with steam to produce CO_2 and H_2. The H_2 may be purified by conventional means.

B.3 From Water Gas and Steam (Fig. B.3)

The following reactions were used in this process:

$$\text{C (amorph)} + H_2O \rightarrow CO + H_2 \text{ (water gas)},$$

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

The conversion of CO is 94 to 96% complete.

The material and utility requirements to produce 28 scm (1000 scf) H_2 (97% pure) are

Coke	19 kg
Steam	200 kg
Cooling water	7.6 m^3
Electricity	11 MJ (3 kWh)

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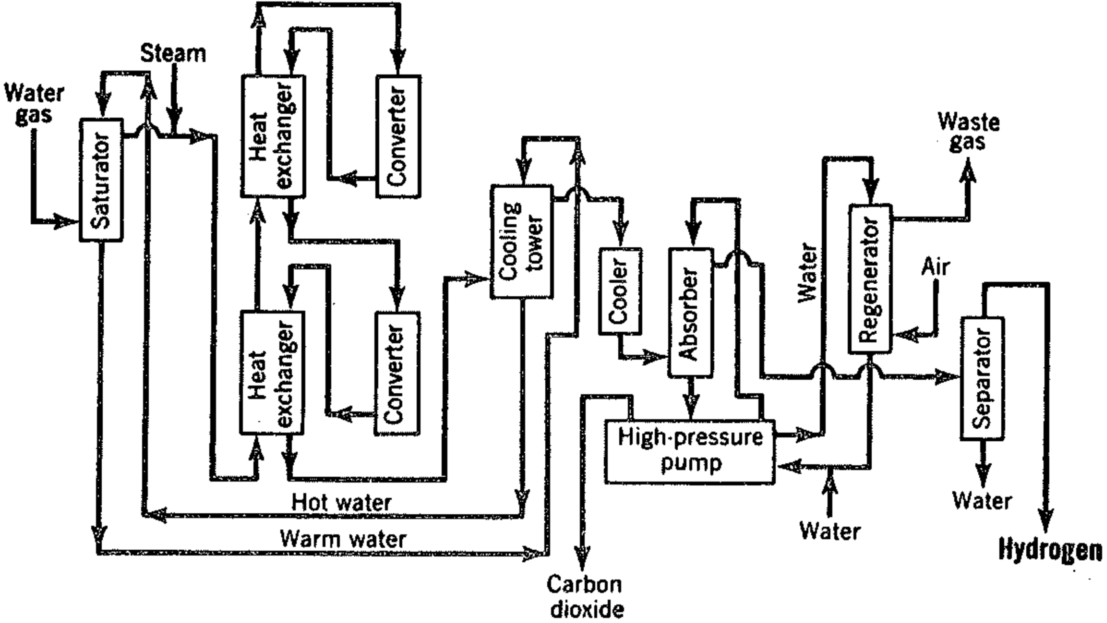


Fig. B.3. Hydrogen from water gas and steam.

Process

Hydrogen is produced along with CO_2 by catalytic reaction of water gas (40% CO–50% H_2) with steam at elevated temperatures. The CO_2 is removed by scrubbing the gas; relatively pure H_2 remains. The process is particularly adaptable to low-purity H_2 containing N and CO for the synthesis of ammonia and methanol, respectively.

Water gas, also called blue gas, is produced by the reaction of steam with incandescent coke or coal at a temperature of 1000° C or higher. Analysis of the gas yields ~40% CO, 50% H_2, 5% CO_2, and 5% N and methane. The gas from a holder is run into a saturator, where it contacts hot water and is heated to between 75 and 85° C. The saturated water gas (1 volume) is then mixed with steam (3 volumes) and passed into a two-stage catalytic converter, where CO reacts with water vapor. The first stage operates at a temperature of 425 to 480° C and the second at 370 to 400° C, with heat exchangers between the stages. Two stages are employed because of the exothermic character of the reaction and the decreased conversion at higher temperatures. By using this two-stage procedure on the water-gas shift reaction, the major part of the conversion takes place with a relatively small amount of catalyst, whereas the balance is effected at a lower temperature conducive to high overall yield. The catalyst commonly used is FeO_2 promoted with CrO_2 and is required in amounts of 6425 to 9600 kg/ 1000 m^3 (180 to 270 kg/ 1000 ft^3) of water gas converter per hour. The life of the catalyst is long; removal of sulfur compounds from the water gas is ordinarily unnecessary, because the catalyst is sulfur-resistant and converts sulfur compounds into H_2S that is removed with the CO_2.

The exit gases, containing about 64% H_2, 31% CO_2, 4% N and methane, and 1% CO, are cooled in water towers and passed to purification units.

The method of purifying depends on the ultimate use of the hydrogen gas. For uses at high pressures, the reaction gases may be compressed to 1.4 to 2.8 MPa (200 to 400 psi) and passed through water scrubbers, where most of the CO_2 is removed. Because of its solubility in water at these pressures, H_2 is lost in amounts of ~3 to 5%. The resulting gas contains 0.5 to 1.5% residual CO_2 and 1.5 to 2.0% CO. These impurities may be removed by compressing the gas to 14 to 21 MPa (2000 to 3000 psi) and scrubbing with an ammoniacal cuprous formate or cuprous chloride solution.

The gaseous mixture from the converters may be purified at atmospheric pressure by scrubbing with amine solutions. This method, known as the Girbotol process, saves compression costs and avoids hydrogen loss, owing to the water-scrubbing operation. After CO_2 removal by this method, the gases analyze about 97.2 mole % H_2, 1.2% CO, 1.2% N, 0.3% methane, and 0.1% CO_2. Purer hydrogen may be obtained by starting with water gas having a lower nitrogen content and catalytically removing the CO. The hydrocarbon–steam process section provides details on the amine purification process.

The composition of the starting water gas is varied to suit the end use destined for the hydrogen. For ammonia synthesis, a synthesis gas is generated so that after conversion the gas contains the proper ratio of nitrogen to hydrogen. A low-nitrogen water gas is produced for methanol synthesis, which requires a ratio of 2 moles H_2 to 1 mole CO in the starting gases.

Carbon monoxide may also be liquefied by refrigeration and thus separated from the water gas produced by the first reaction shown, leaving fairly pure hydrogen. In one such process, the refrigerant used is liquid air after preliminary cooling by an ammonia refrigeration system; in another, the cooling is obtained by expansion of the hydrogen from a pressure of 2.1 MPa (20 atm) to a lower one by working against a piston. The liquid CO is then easily separated from the still gaseous H_2.

The steam–water gas process is practically obsolete but is described here for its historical interest as the dominant process for hydrogen production at one time.

B.4 Steam-Iron Process

This process produces H_2 by the reaction of steam at high temperature with reduced FeO_2 to yield H_2, and then the rereduction of the FeO_2 with a reducing gas such as water gas or producer gas in a cyclic operation. The steam-iron process is obsolete.

B.5 From Water by Electrolysis

Direct current is passed between iron or nickel-plated iron electrodes in a solution consisting of 10 to 25% caustic soda or potash. Only distilled water is added to the electrolyte, because the solute is not consumed. Operating voltage is ~2 V; current density is ~650 A/m^2. Electricity required is 4950 kWh or 18 GJ/100 smc of H_2 (140 or more kWh/1000 scf) plus one-half that volume of O_2 (if the O_2 is recovered). Where low-cost electricity is available, as in Canada and Norway, this process is an attractive way of producing heavy water.

Hydrogen is also obtained as a by-product in other electrochemical processes, although recovery is not always economical. Typical examples are the electrolysis of sodium chloride or potassium chloride solutions for the manufacture of chlorine, sodium or potassium hydroxide, sodium chlorate, and sodium perchlorate.

B.6 Thermal Decomposition of Hydrocarbons

Hydrogen is obtained as a by-product in the manufacture of carbon black by the thermal decomposition of natural gas. A brick checker-work chamber is heated to between 1090 and 1370°C, and natural gas is passed through it. After the carbon black is separated, the gases contain 70 to 85% H_2.

By scrubbing to remove carbon oxides, hydrogen pure enough for ammonia synthesis is obtained. This process is one of the cheapest sources of low-purity hydrogen.

B.7 Catalytic Reforming of Petroleum Stocks

Considerable hydrogen [11 to 22 scm (400 to 800 scf) per barrel of feedstock] is produced by petroleum refiners in catalytic reforming operations. Hydrogen-rich gas from the reforming operation contains hydrogen sulfide and hydrocarbons of low molecular weight. Purification involves only a caustic wash (to remove hydrogen sulfide) followed by refrigeration to condense most of the water and higher hydrocarbons. The resulting gas is usually further dried by passing it over a solid adsorbent (e.g., activated alumina) and may be washed with liquid nitrogen to remove more hydrocarbons. Although the product (95% H_2) is suitable for ammonia synthesis and was formerly so used, present refining processes have such large requirements for hydrogen (for hydrodesulfurization and other uses) that all or practically all by-product hydrogen is used on-site and does not reach the market.

B.8 From Methanol and Steam

Hydrogen of higher than 98% purity may be manufactured by the catalytic reaction of methanol and steam at 260°C, according to the following equation:

$$CH_3OH + H_2O \rightarrow CO_2 + 3H_2 .$$

The exothermic reaction at moderately low temperatures may be conducted at up to 2.1 MPa (300 psi), so that the hydrogen can be delivered at elevated pressures. For 28 scm (1000 scf) of H_2 plus 18 kg of pure by-product CO_2, ~20 L of methanol and 15 L of fuel are required. The ease and simplicity of fabricating the plant, handling raw materials, and purifying the hydrogen make this process particularly practical for portable plants. The present cost of methanol, however, precludes the widespread or large-scale use of this process.

B.9 From Ammonia by Dissociation

For metal treating and hydrogen welding, a source of mixed gas (75% H_2, 25% N) is the catalytic cracking of ammonia. Economies in transportation and storage make this process feasible for production of less than 565 scm (20,000 scf) gas mixture per day. In the process, liquid ammonia is vaporized, preheated by hot gases from the reaction chamber, and then passed into an electrically heated reactor. In the presence of a supported nickel oxide catalyst, the ammonia may be cracked at pressures up to 350 kPa (20 psig) and temperatures ranging from 900 to 980°C. The product gases may be scrubbed with water to remove traces of unconverted ammonia and then dehydrated to yield a dry gas.

Because each kilogram of liquid ammonia yields 2.8 scm of dissociated gas containing ~2.1 scm of H_2 and 0.7 scm of N, a single 68-kg (150-lb) cylinder of liquid ammonia can replace 33 cylinders of H_2. Pure nitrogen may also be obtained by burning out the hydrogen at a cost comparable with or perhaps lower than cylinder nitrogen. Although production of pure hydrogen by this process is not economically feasible, the cracked gas is quite satisfactory for many uses, particularly in the steel industry's annealing process.

B.10 High-Purity Hydrogen

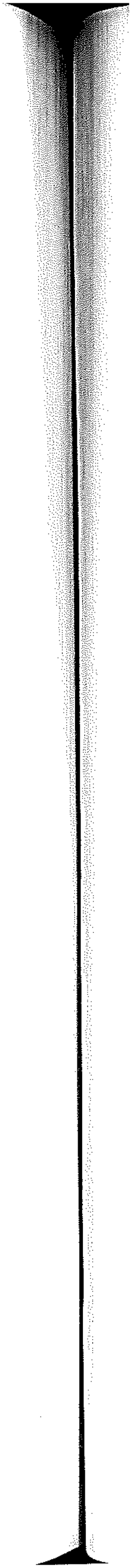
Extremely pure hydrogen (99.9%) can be produced by diffusing less pure gas through palladium films. At about 315° C, molecular hydrogen dissociates to atoms on palladium; the atoms then diffuse through the metal and recombine on the other side. The liquefaction process, described in the following section, also yields extremely pure hydrogen.

B.11 Liquid Hydrogen

The steam reformer process is usually the source of hydrogen in liquefied hydrogen plants. After the removal of impurities, further purification is obtained by refrigeration and adsorption; feed to the liquefier must be quite pure to prevent blocking of the lines and valves by condensable impurities.

The final liquefier section depends on the ortho-para shift of the hydrogen molecule. Hydrogen can exist in two forms, depending on the rotation of the two atoms in the molecule. In orthohydrogen, the atoms spin in the same direction; in parahydrogen they spin in opposite directions. The conversion temperature is also very low, –70° C, above which Joule-Thompson expansion raises rather than lowers the temperature.

Normal hydrogen is only 25% para at atmospheric temperature; at liquid hydrogen temperature it is 99.7% para, and the heat generated by the conversion is $\gtrsim$11% of the total heat removed during liquefaction. If normal liquid hydrogen were stored, as much as two-thirds of it would be vaporized as conversion took place. Therefore, the liquefier cycle includes rapid conversion over a ferric hydroxide catalyst. This conversion takes place in two steps, at liquid nitrogen and liquid hydrogen temperatures, in the presence of liquid refrigeration. Thus liquid hydrogen is over 95% para. Purity may exceed 99.99%.



Appendix C

SENSITIVITY STUDY FOR MOLTEN SALT AND HOT OIL TRANSPORT

Delivered-energy costs are largely determined by the cost of pipeline transport. Pipeline costs are determined by a number of physical variables that may be selected in proper combinations to produce a least-cost transport system.

This section describes results of a sensitivity study to determine the influence of each variable on delivered-energy costs and allow selection of that set of variables that produces the least-cost system.

Variables selected for study were insulation thickness, pipe diameter, pump station spacing (distance between stations), heat-exchanger surface area, delivered-energy temperature, and delivered-energy rate.

For these results, fixed parameters were fixed-charge rate (0.17); process heat costs ($3.96/GJ for hot oil, $4.18/GJ for salt); electrical costs ($0.0243/kWh); cost factors for all construction, materials, and operation; pipeline length [160 km (100 miles)]; and start-up date (1990).

C.1 Molten Salt, cs Piping

Figure C.1 shows typical results, using salt as a fluid and A106B steel as the pipe material, for 1000-MW delivery at 400°C (750°F) at the user-end supply pipe and 195°C (380°F) at the user-end return pipe.

Points marked 1 through 4 and 5 through 8 show cost variation as a function of insulation thickness for two pipe diameters. The combinations of supply/return line insulation thicknesses for each set of points were 25 mm/25 mm, 51 mm/25 mm, 76 mm/51 mm, and 102 mm/51 mm, respectively. The unit energy cost is very insensitive to insulation thickness. In subsequent computer runs, a combination of 51 mm/25 mm was used.

Points 9 through 12 in combination with those previously noted define the results with an assumption of four pump stations (54-km station spacing). Points 15, 17, and 19 define the curve for six stations (32-km spacing). Points 16, 18, 20 and 21 through 24 define 11 (16-km spacing) and 21 (8-km spacing) station results, respectively. The unit energy cost decreases with addition of pump stations. However, point 25 is for 16 stations, and the lower unit cost indicates that the probable optimum number of stations is between 11 and 21. Note that near the optimum, the cost becomes relatively insensitive to number of stations or station spacing. The apparent reason for this insensitivity is that the maximum pressure in each section has dropped to a level where the minimum pipe wall thickness is allowed in most of the line so pipe material costs have become constant. Additional pump stations then begin to increase total system costs.

Points 26, 28, 29, and 31 represent results for a user-end supply temperature of 450°C (850°F) [rather than 400°C (750°F), as previously], but the user still receives heat at 370°C (700°F). Heat exchanger costs decrease only a small amount, but the lower salt flow rate made possible by the higher temperature decrease across the exchangers produces significant reductions in pump station horsepower and, therefore, in capital and operating costs.

Point 30 is the same as point 26 except that the user's temperature at the heat exchanger exit is 430°C (800°F) rather than 370°C (700°F). The resulting influence on unit energy cost of heat exchanger size is negligible.

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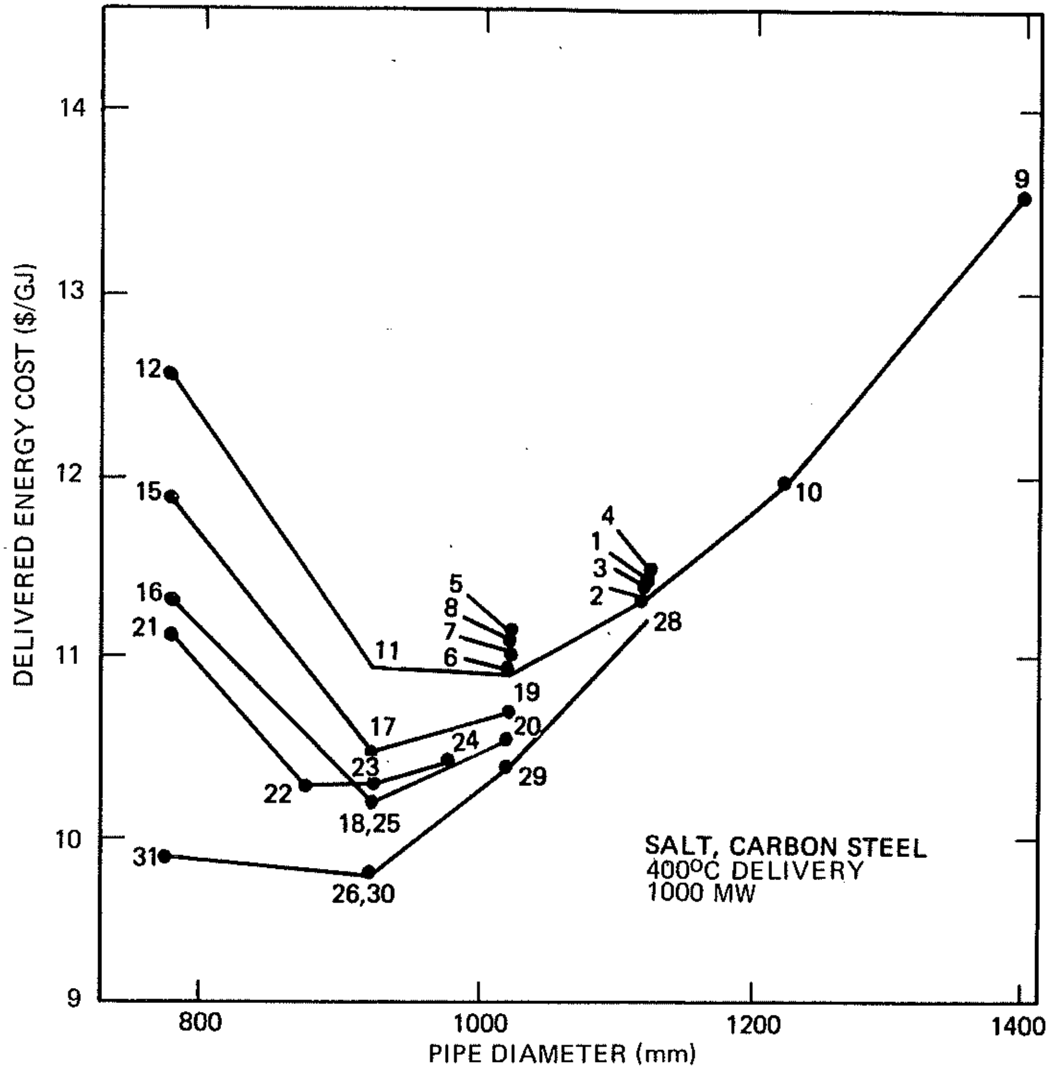

Fig. C.1. Salt, cs pipe, 400°C delivery, 1000 MW.

These results indicate that minimum cost results from using the highest possible supply temperature and the lowest possible return temperature at the user's end, independent of the temperature level desired on the user's side of the heat exchanger. The return temperature was fixed at 195°C (380°F) because this allowed minimum line temperature of ~175°C (350°F) [35°C (65°F) above the salt freezing point].

The unit energy cost is sensitive to pipe diameter; however, near the optimum diameter the cost is fairly insensitive to all combinations of the other variables. The variation shown, at 910-mm (36-in.) diameter, is ~10%. This low variation indicates that the costs will not be greatly in error if other than the precise optimum line configuration is selected.

The procedure for determining the configuration for other power levels, temperature levels, and pipe materials was to first check the influence of changes in supply-line insulation thickness and pump station spacing. When the results became relatively insensitive to further changes in these variables, a standard configuration was defined and used for checking minimum-cost pipe diameter.

C.2 Molten Salt, ss Piping

Figure C.2 is a display of the results for 1000-MW delivery at 675°C (1250°F) supply line temperature with ss supply pipe.

Points 74, 73, 72, and 75 show changes in cost because of, respectively, 150, 180, 200, and 230 mm (6, 7, 8, and 9 in.) of insulation on the supply line (25 mm on the return line for all cases). The incremental cost reduction because of additional insulation is decreasing rapidly, so 230 mm of insulation was selected as standard for this temperature level.

The sensitivity of cost to decreasing length of line segment between pump stations is shown by points 75 through 79, which are for segment lengths of 16, 11, 8, 4.6, and 5.4 km, respectively. The cost

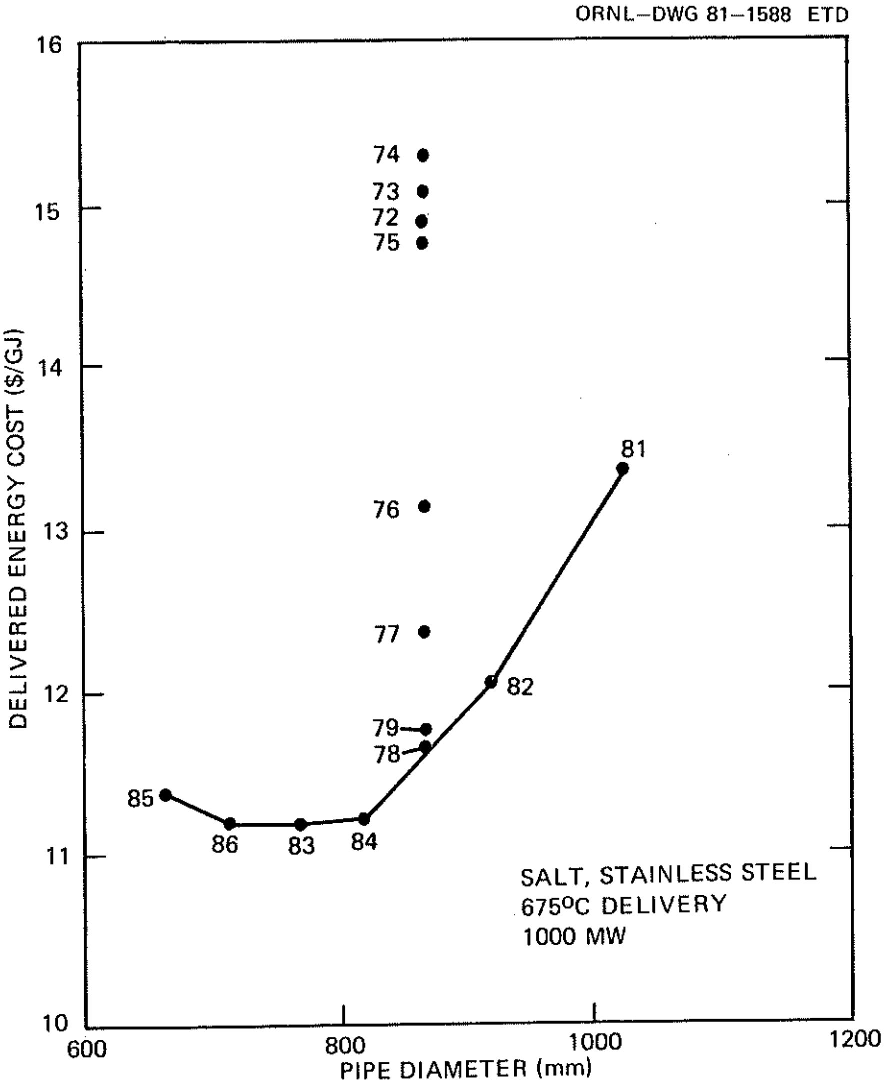

Fig. C.2. Salt, ss pipe, 675°C delivery, 1000 MW.

becomes relatively insensitive to the segment length near lengths of 5 km. The 4.6-km station spacing (36 stations per 161 km) was selected as standard for this temperature level.

Changing from 16- to 4.6-km spacing (points 75 and 78) allowed reduction of necessary pump pressure per section from near 690 kPa to ~210 kPa (100 to 30 psi). Pipe wall thicknesses that ranged from 8.6 to 262 mm (0.34 to 1.03 in.) for point 75 were reduced to a range of 6.4 to 135 mm (0.25 to 0.53 in.) for point 78. Supply pipe material cost dropped from $\$527.4 \times 10^6$ to $\$172.5 \times 10^6$, while the cost of pump stations (36 vs 11) increased only from $\$29.1 \times 10^6$ to $\$63.1 \times 10^6$.

Studies similar to those previously described were also conducted for user-supply temperatures of 620, 565, and 510° C (1150, 1050, and 950° F) for ss pipe and 1000-MW delivery to establish a standard supply-line insulation thickness and pump station spacing at those temperatures. The results are not included, but the standard values selected are shown in Table C.1.

Table C.1. Optimal pipeline parameters determined by sensitivity study (salt)

User-end supply temperature [°C (°F)]	Station spacing [km (miles)]	Insulation supply/return [mm/mm (in./in.)]	User-end return temperature [°C (°F)]
675 (1250)	4.6 (2.85)	225/25 (9/1)	195 (380)
620 (1150)	8.0 (5.0)	175/25 (7/1)	195 (380)
565 (1050)	11.5 (7.14)	125/25 (5/1)	195 (380)
510 (950)	12.4 (7.69)	100/25 (4/1)	195 (380)
455 (850)	13.4 (8.33)	75/25 (3/1)	195 (380)
400 (750)	16.1 (10.0)	75/25 (3/1)	195 (380)
345 (650)	16.1 (10.0)	50/25 (2/1)	195 (380)
290 (550)	16.1 (10.0)	50/25 (2/1)	195 (380)

These values were used to generate information for all subsequent cost comparisons, independent of delivered power, line length, or pipe material. These variables clearly will influence the true least-cost configuration, as will different assumptions for process heat costs, electrical costs, and fixed-charge rate. However, the results discussed in Sects. C.1 and C.2 should establish that the values of Table C.1 provide cost results consistent with the estimated accuracy of a study of this type.

One set of points was developed demonstrating that, for 565°C (1050°F) supply, the optimum diameter was the same at 80- and 160-km (50- and 100-mile) lengths. Therefore, for calculations involving different line lengths, the proper diameter at any length was assumed to be the same as the diameter at the 160-km length.

C.3 Hot Oil, cs Piping

The results of the sensitivity study for the cases using oil as a thermal transport medium are shown in Fig. C.3. Basic assumptions included a fixed-charge rate of 17%, process heat at $3.96/GJ ($4.18/10^6 Btu), electricity at $0.0243/kWh, and 1000-MW delivery at 315°C (600°F) to the user. Terminal temperature differences were 10°C (20°F) at the high-temperature ends and 30°C (50°F) at the low-temperature ends of the exchangers. Return from the user was at 90°C (200°F). User-end pipeline temperatures were 325°C (620°F), supply line, and 120°C (250°F), return line.

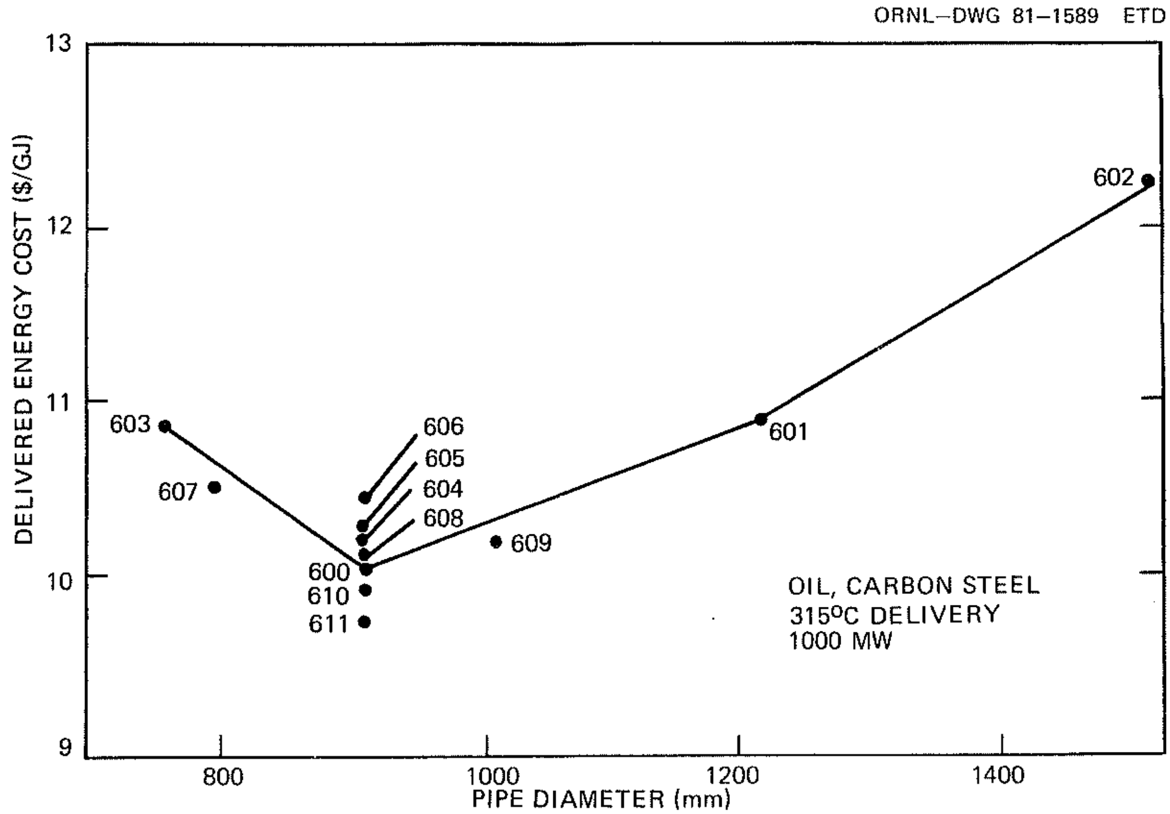

Fig. C.3. Oil, cs pipe, 315° C delivery, 1000 MW.

In Fig. C.3, the line connects points calculated using 16-km (10-mile) pump station spacing, supply/return insulation thicknesses of 75/25 mm (3/1 in.), and heat exchanger U-values of 494 $W/m^2 \cdot {}^\circ C$ (87 $Btu/h \cdot ft^2 \cdot {}^\circ F$). Point 604 was computed using 26.7-km (16.6-mile) station spacing, which increased the cost ~2%.

Changing supply line insulation thicknesses to 130 mm (5 in.) at point 605 and 180 mm (7 in.) at point 606 also caused cost increases. For points 607 through 609, the insulation thickness was reduced to 50 mm (2 in.), and the station spacing was increased to 20 km (12.5 miles). Point 608 is less than 1% above point 600 (75-mm, 16-km spacing), indicating that cost is relatively insensitive to insulation thickness and station spacing near these values.

The effects of changes in U-value are illustrated by comparison of points 608 and 610. For 608, the U-value was 494 $W/m^2 \cdot {}^\circ C$, while for 610 it was 625 $W/m^2 \cdot {}^\circ C$. This 26% increase in U represents the range of values reported in the literature at the 315°C (600°F) temperature level. The resulting cost difference is ~2%.

The influence of user-return temperature is shown by comparison of points 608 and 611. Point 611 was computed using a temperature of 65°C (150°F) rather than 90°C (200°F). Although the lower return temperature allows an oil flow rate reduction of ~11%, the cost decrease is less than 4%, and cost is shown to be relatively insensitive to these return temperature levels at 315°C (600°F) supply temperature.

One set of values was selected for the oil pipeline configuration, independent of supply temperature or energy delivery rate. Because return of steam condensate from the user at near atmospheric pressure would be a common situation, a return temperature of 90° C (200° F) was selected as standard. Based on results of the comparisons discussed previously, other values chosen were 50-mm (2-in.) supply line insulation, 20-km (12.5-mile) station spacing, and a U-value of 494 W/m^2·°C.

Terminal temperature differences were 11°C (20°F) for the high-temperature end, and 28°C (50°F) for the low-temperature end of the exchangers. These values represent acknowledgment that oil may thermally decompose in contact with surfaces much above 345°C (650°F). Low-temperature differences and large heat exchangers will be necessary to avoid decomposition.

C.4 Cost Variation with Pipe Diameter, Delivery Temperature, and Energy Delivery Rate

Having determined the appropriate combinations of pump station spacing and insulation thickness for the range of delivery temperatures, the cost variation as a function of diameter and energy delivery rate was investigated (Figs. C.4 through C.12). A process heat cost of $4.18/GJ was used for calculation with salt, and Fig. C.4 illustrates that the lowest cost for delivery of 1200 MW, using molten salt in ss pipe, is achieved in an 810-mm (32-in.) diameter pipe with 565°C (1050°F) user-end salt temperature. The cost variation with diameter is negligible; diameters in the range of 710 to 915 mm (28 to 36 in.) yield costs that differ by only a few percent.

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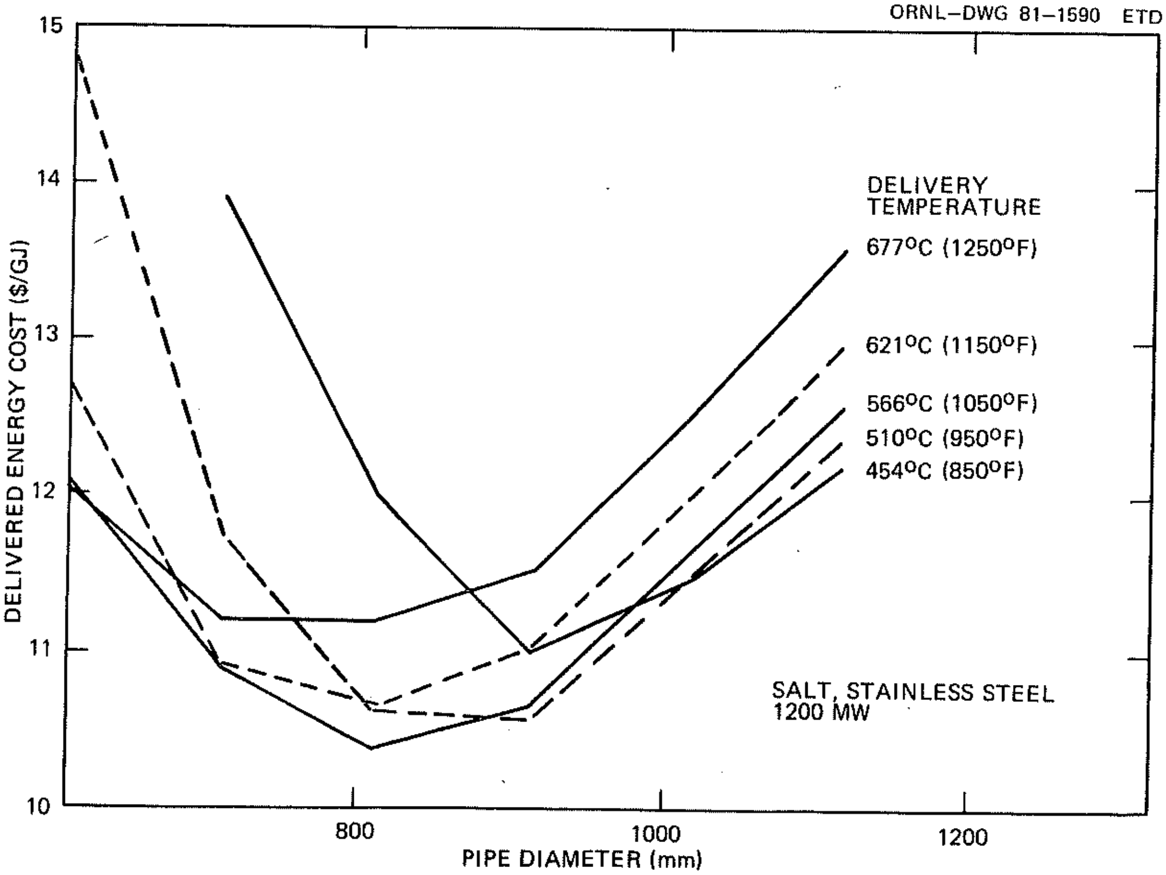

Fig. C.4. Salt, ss pipe, 1200 MW.

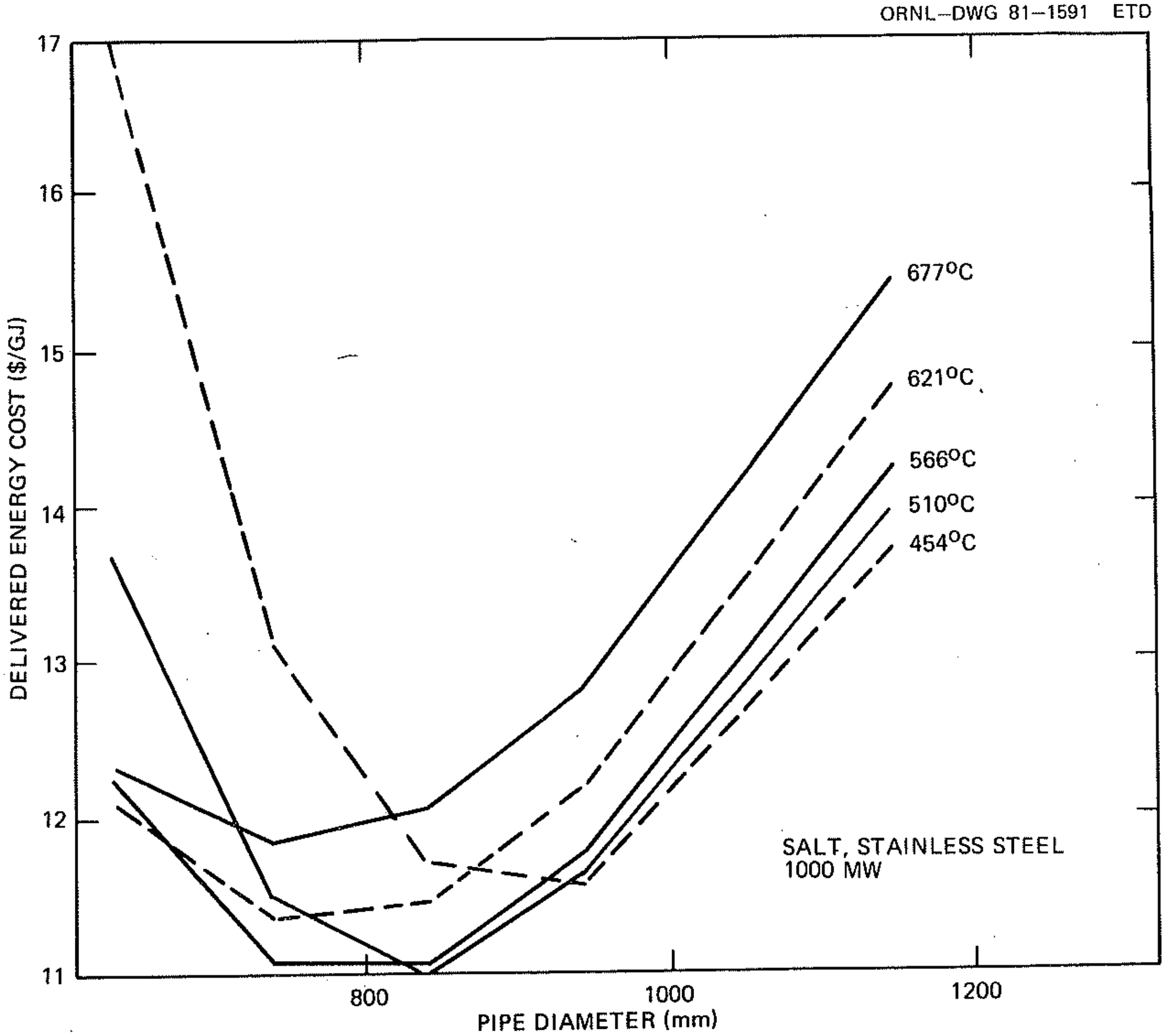

Fig. C.5. Salt, ss pipe, 1000 MW.

The cost variation with supply pipeline terminal temperature is also weak, because of the sharp decrease in pipe allowable stress at higher temperatures. Although a lower mass flow rate (therefore smaller diameter and lower operating cost) is required at higher temperatures for a given delivered power, the greater wall thickness required offsets the saving due to diameter reduction.

A similar result is depicted in Fig. C.5 for 1000-MW delivery and in Fig. C.6 for 800-MW delivery. These figures also demonstrate that unit-delivered energy costs increase as power level is reduced. This relationship is typical of capital-intensive projects.

The results shown in Figs. C.7 through C.9 are for salt in cs pipe. Computer runs were made for terminal temperatures up to 680°C (1250°F), but all the results above 400°C (750°F) were discarded because the allowable maximum temperature for the pipe material was exceeded or because pipe wall thicknesses in excess of 50 mm (2 in.) were required.

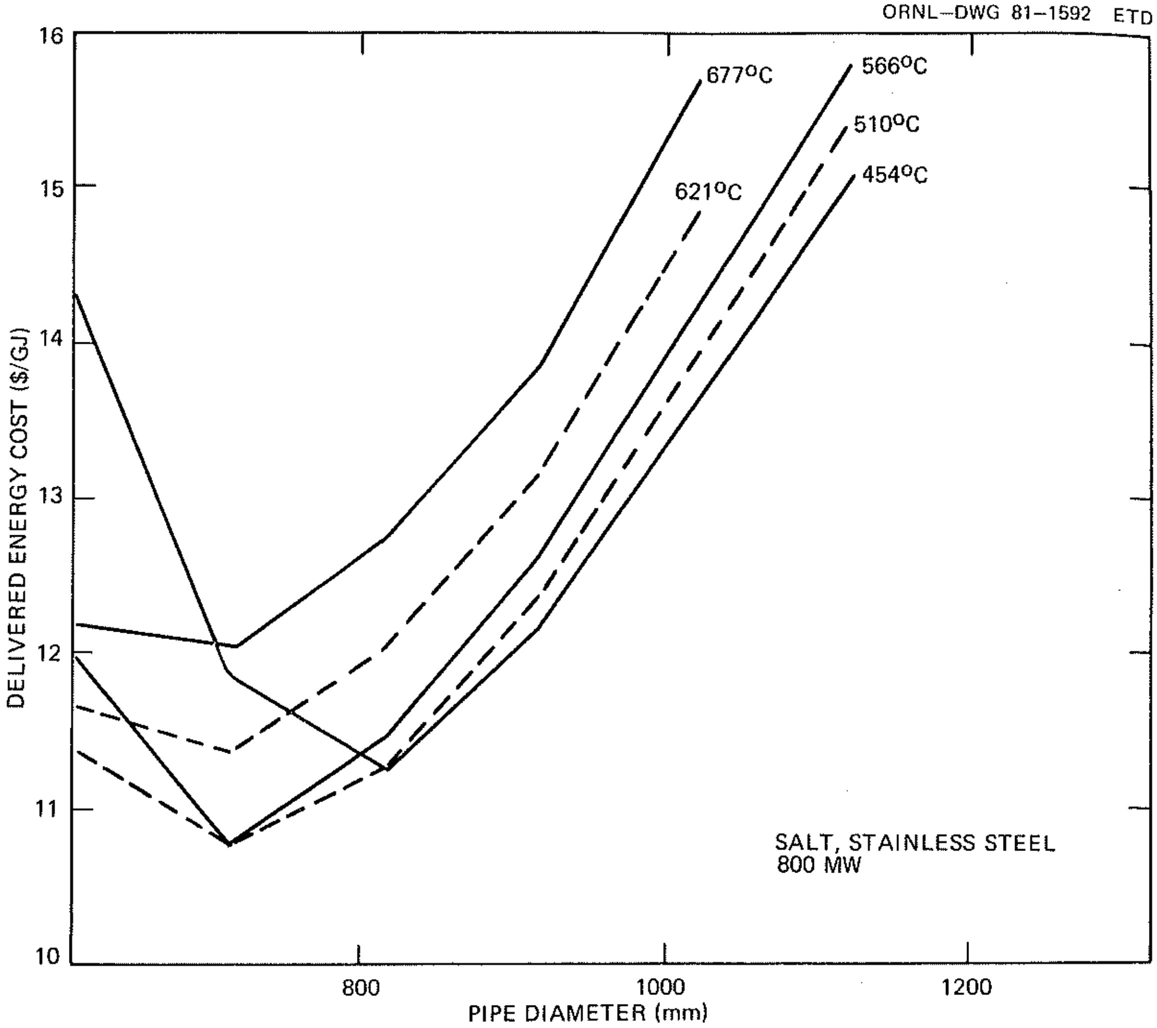

Fig. C.6. Salt, ss pipe, 800 MW.

Results indicate that significant savings occur from use of high-temperature transport, which causes lower salt flow rates and thus smaller pipe diameter and/or lower pressure drop (thinner wall and lower operating cost). Low-temperature systems require large diameter pipe because of the necessary higher flow rates for a given power delivery.

Results shown in Figs. C.10 through C.12 are for oil in cs pipe. The trends are similar to those for the case of salt with cs pipe, but the influence of diameter on cost is less. The minimum point on each curve of these figures was used to define the least-cost system for a given fluid-material-temperature-power combination.

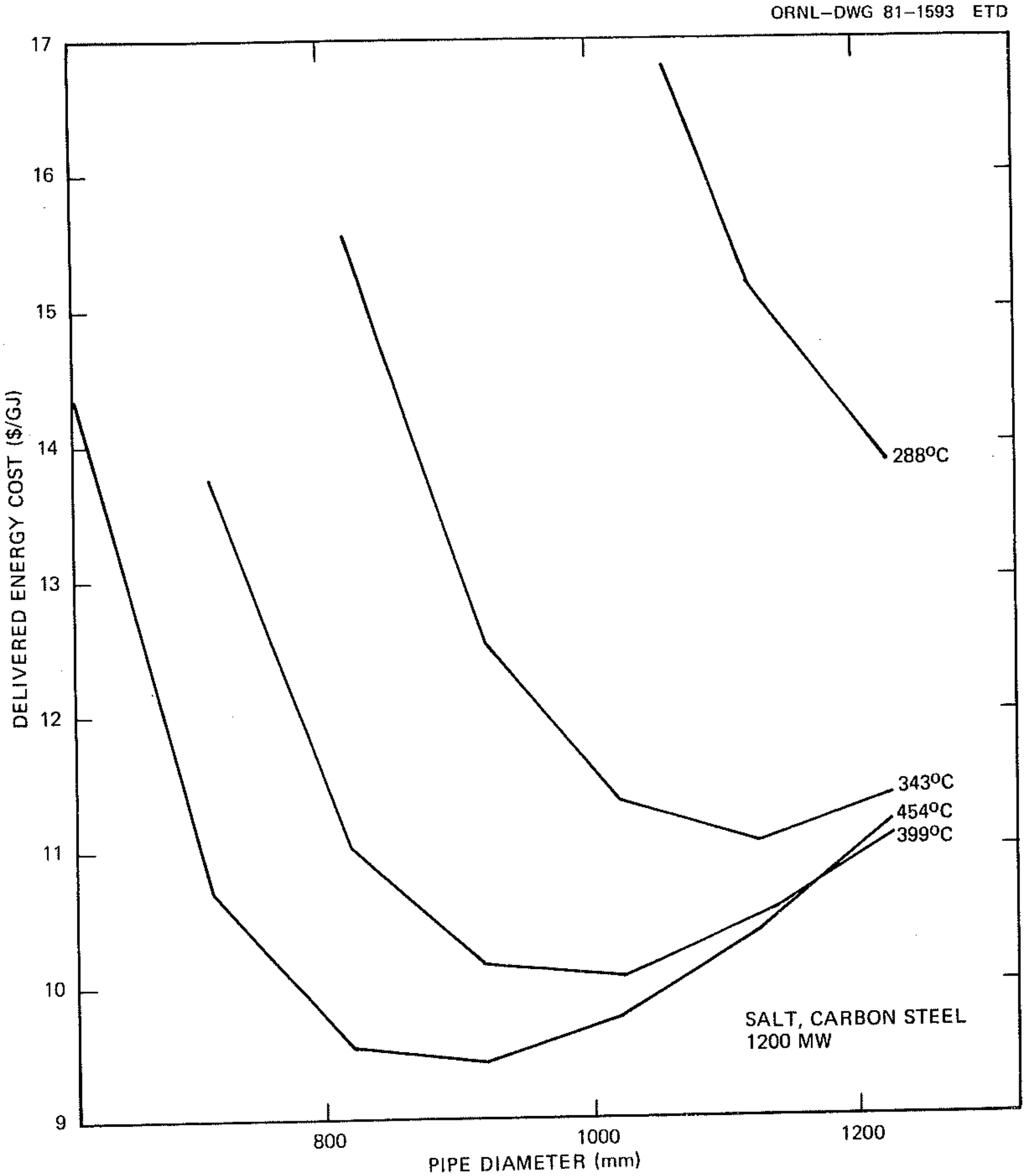

Fig. C.7. Salt, cs pipe, 1200 MW.

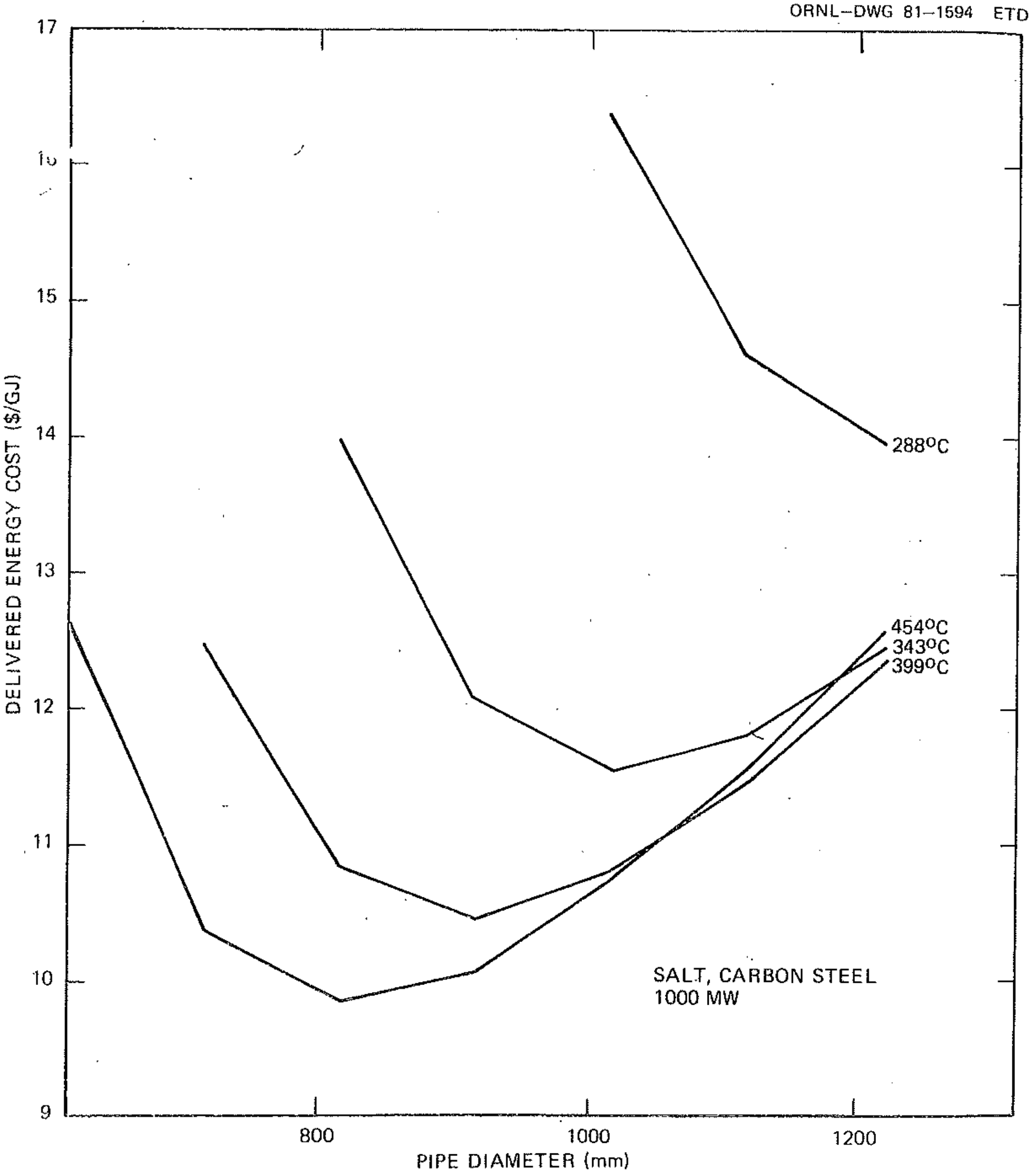

Fig. C.8. Salt, cs pipe, 1000 MW.

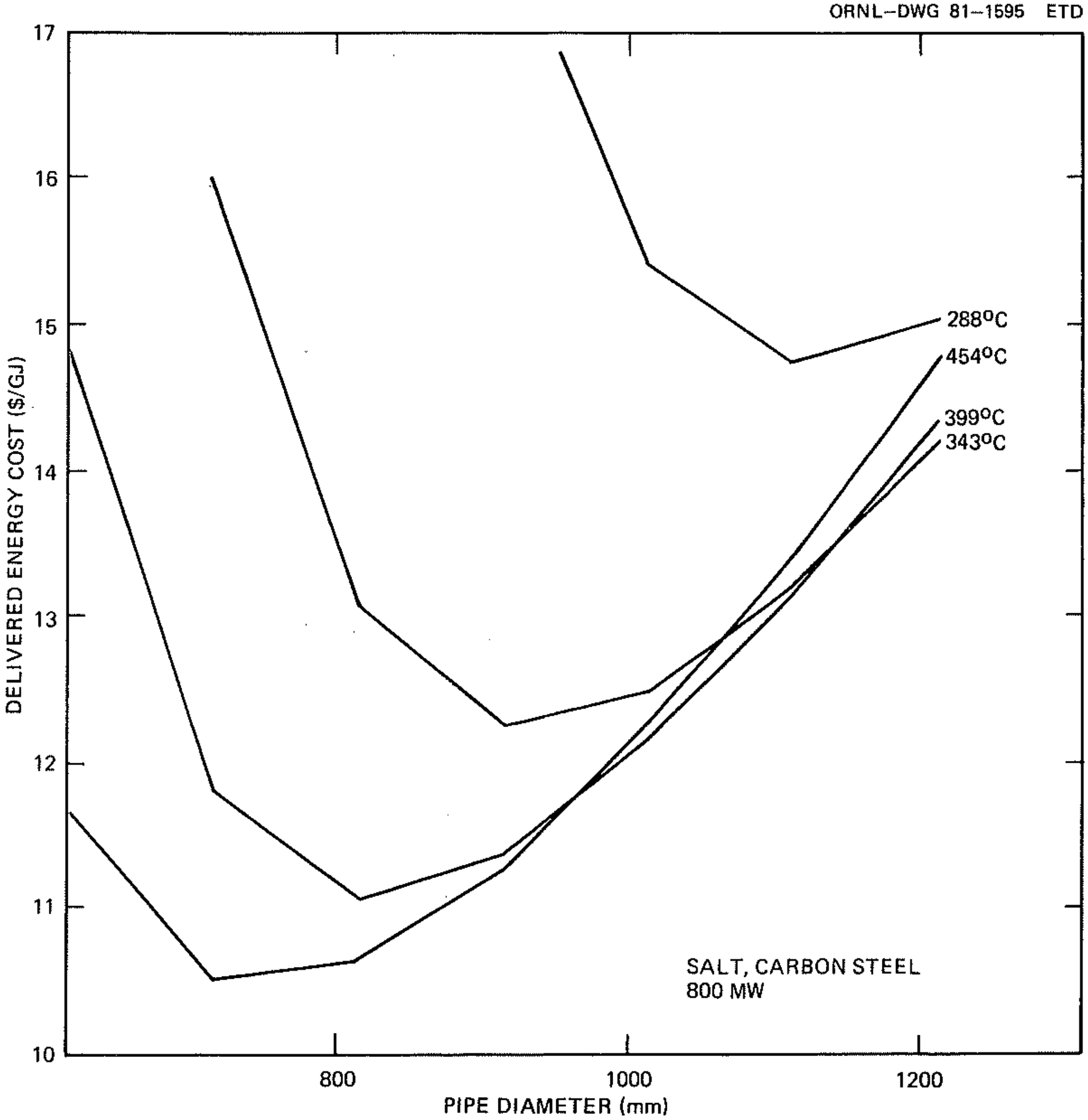

Fig. C.9. Salt, cs pipe, 800 MW.

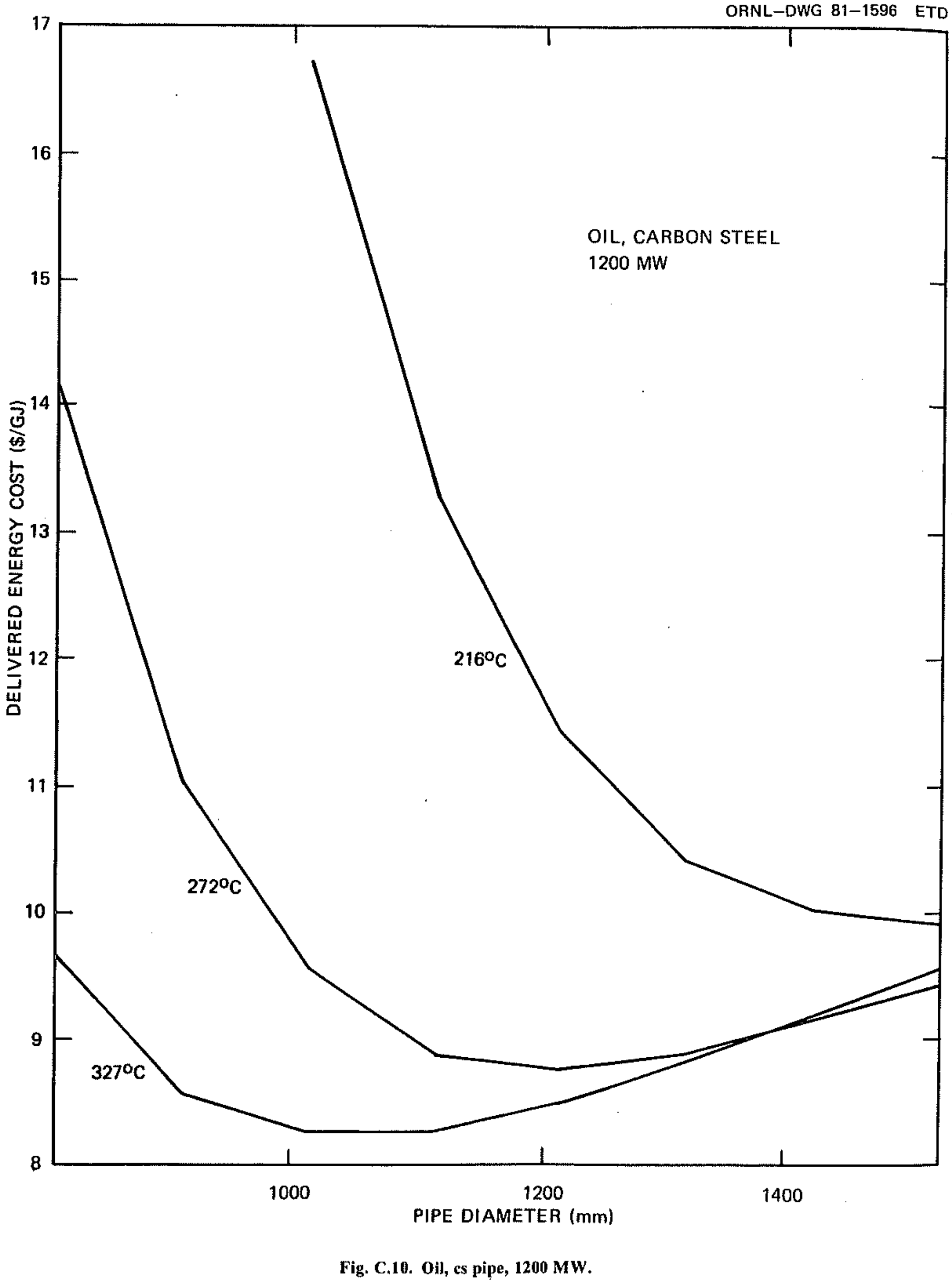

Fig. C.10. Oil, cs pipe, 1200 MW.

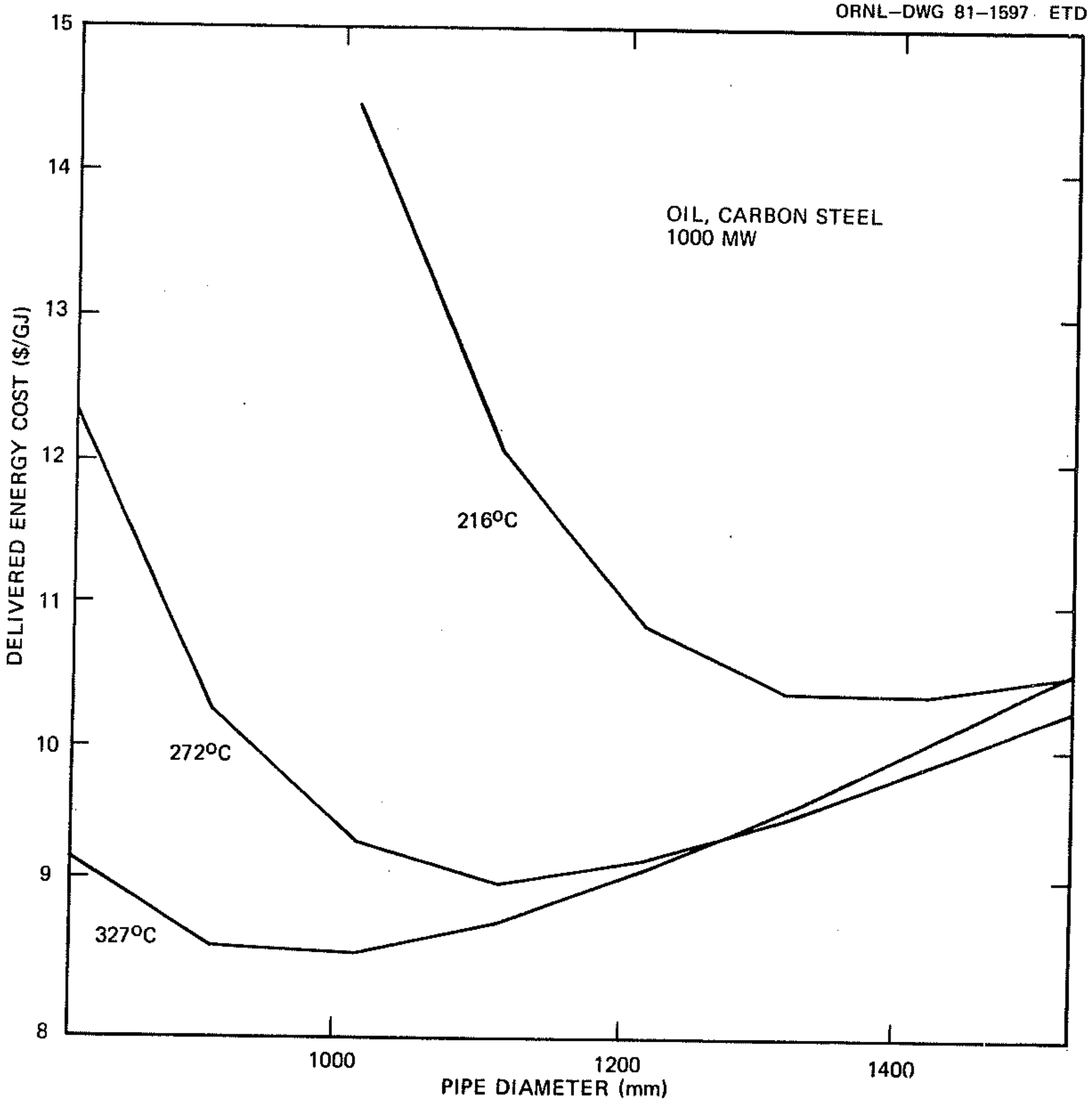

Fig. C.11. Oil, cs pipe, 1000 MW.

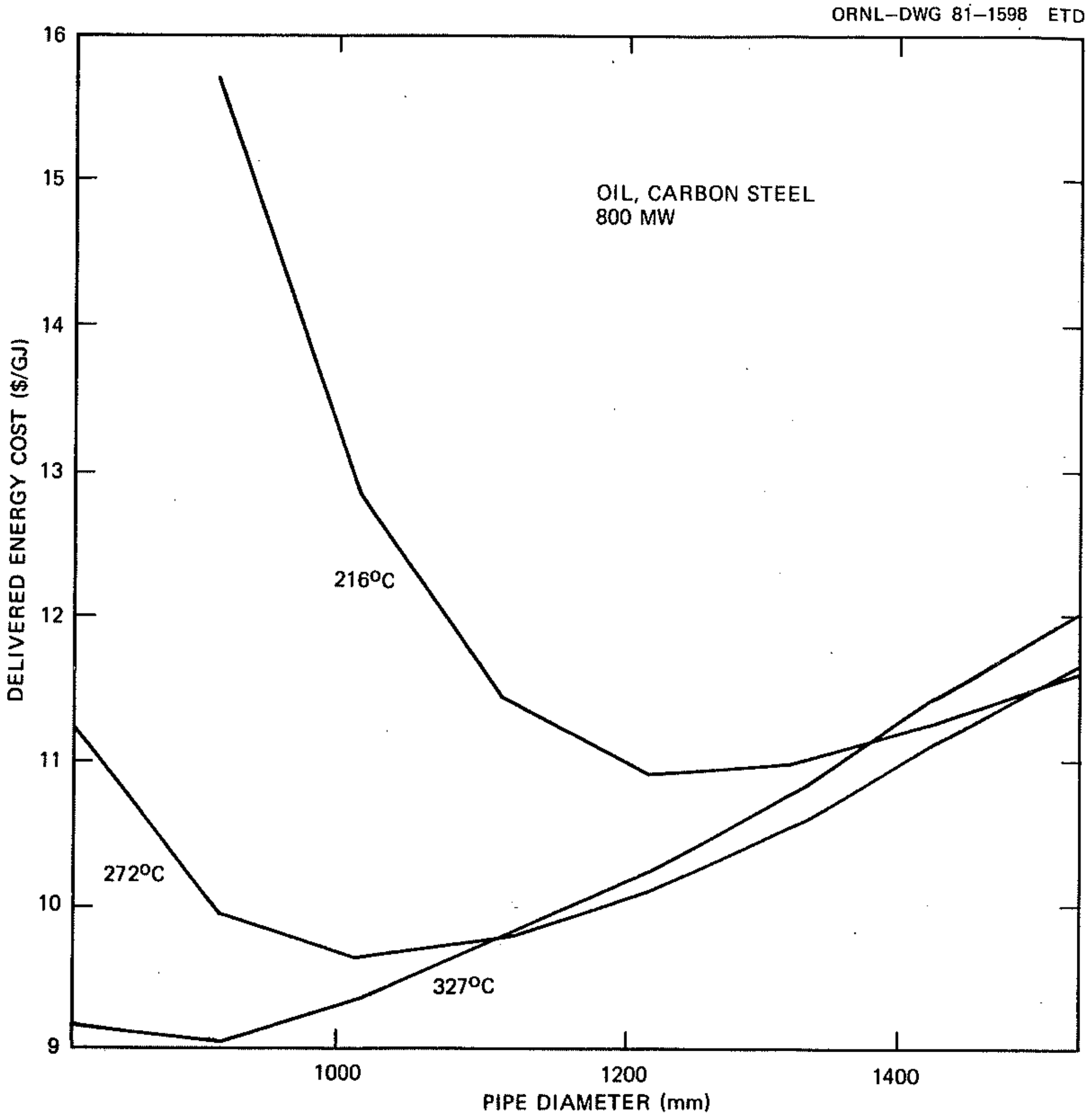

Fig. C.12. Oil, cs pipe, 800 MW.

Appendix D

PIPING SYSTEMS

D.1 Gas Piping and Pumping

The gas pipeline model is based on the equations shown in Table D.1 for the isothermal flow of a compressible fluid. These equations are solved through the specification of a mass flow rate, gas temperature, maximum and minimum gas pressures, and pipe diameter. These correlations are based on equations given in Refs. 1 and 2. Gas properties including molecular weight, lower heating value, and density must be known. The variables used in Table D.1 are defined in Table D.2.

Table D.1. Gas flow equations[a]

$$Q = \omega \cdot LHV \cdot C_1$$

$$M_1 = \left(\frac{\omega}{g}\sqrt{gRZT/k}\right)/(C_2\pi r^2 P_1)$$

$$L = \frac{d}{4f}\left[\frac{1-(P_2/P_1)^2}{kM_1^{\,2}} - \ln\left(\frac{P_1}{P_2}\right)^2\right] C_3$$

$$HP = \frac{\omega RT}{\eta_1 C_4}\left(\frac{k}{k-1}\right)\left[\left(\frac{P_1}{P_2}\right)\left(\frac{k-1}{k}\right) - 1\right]$$

[a] Variables are defined in Table D.2.

Table D.2. Variable definitions

Variable	Units	Definition
C_1	1 (3600 s/h)	Conversion constant
C_2	10^5 nt/m^2b (1)	Conversion constant
C_3	1 (5279 ft/mile)	Conversion constant
C_4	1 (550 ft lbf/hp s)	Conversion constant
d	m (in.)	Pipe diameter
f	Dimensionless, 0.003	Pipe friction factor, function of Reynolds numbers and pipe roughness
g	1.0 kg· m/N·s^2 (32.174 lb_m·ft/lb_f·s^2)	Gravitational acceleration
HP	Watt (hp)	Compressor power per pumping station
k	Dimensionless	Gas constant = Cp/Cv
L	m (miles)	Distance between pumping stations
LHV	J/kg (Btu/lb)	Lower heating value
M_1	Dimensionless	Mach number at P_1
P_1	bars (psia)	Maximum gas pressure
P_2	bars (psia)	Minimum gas pressure
Q	J/s (Btu/h)	Energy flow rate
r	m (in.)	Radius of pipe = d/2
R	Newton-m/kg K [(lb_f·ft)/(lb_m·°R)]	Gas constant = Ru/MW
T	K (°R)	Temperature
Z	Dimensionless	Gas compressibility = $P/\rho RT$
η_1	0.85 (0.85)	Compressor efficiency
MW	g/g-mole (lb/lb-mole)	Molecular weight
ω	kg/s (lb/s)	Mass flow rate
ρ	10^5 kg/m^3 (lb/ft^3)	Gas density

D.2 Liquid Piping and Pumping

Liquid piping calculations (shown in Tables D.3 and D.4) begin with the determination of user-end pipeline fluid temperatures and then proceed with the determination of the enthalpy change, mass flow rate, pipe cross section, velocity, Reynolds number, friction factor, and pressure drop in a segment of pipe between two pumping stations. The magnitude of the pressure drop was also used as the maximum pressure to which that section of line would be subjected. Energy input from the pump to the fluid was determined by dividing the frictional energy dissipation by the pump efficiency. Heat loss per unit length of pipe was determined as a function of soil and insulation thermal resistances and pipe and ground temperatures.[3] An energy balance on each line segment was then used to determine the fluid enthalpy and, therefore, temperature at the next pump station.

Table D.3. Liquid pumping calculation[a]

$$\mathrm{TS(I)} = \mathrm{TSU} + \mathrm{DTAU}$$

$$\mathrm{TR(I)} = \mathrm{TRU} + \mathrm{DTBU}$$

$$\Delta h = \int_{\mathrm{TR(I)}}^{\mathrm{TS(I)}} (C_p)\, dT$$

$$\dot{m} = \mathrm{QUSER}/\Delta h$$

$$A = \pi D^2/4$$

$$V = \dot{m}/(\rho \times A)$$

$$\mathrm{Re} = VD\, \rho/\mu)$$

$$f = 0.0056 \left[1.0 + \left(\frac{3}{D} + \frac{10^6}{\mathrm{Re}}\right)^{1/3}\right]$$

$$P = \frac{fL}{D}\left(\frac{\rho V^2}{2}\right)$$

$$\mathrm{PMAX} = \Delta P$$

$$\mathrm{Fric} = \dot{m}\, \frac{fL}{D}\, \frac{\rho V^2}{2\mathrm{eff}}$$

$$\mathrm{Rth} = \frac{1}{2\pi \mathrm{Ks}} \times \frac{K_S}{K_I} \ln \left(\frac{D/2 + \mathrm{TKINS}}{D/2}\right) + \ln \left[\frac{3 + D/2}{D/2} + \sqrt{\left(\frac{3 + D/2}{D/2}\right)^2 - 1}\right]$$

$$\mathrm{Loss} = (T_{\mathrm{pipe}} - T_{\mathrm{ground}})/\mathrm{Rth}$$

$$h_{I+1} = h_I + \frac{\mathrm{Loss} - \mathrm{Fric}}{\dot{m}}\text{, supply line}$$

$$h_I + \frac{\mathrm{Loss} - \mathrm{Fric}}{\dot{m}}\text{, return line}$$

$$t_{\mathrm{wall}} = (\mathrm{PMAX})\,(D)/(2S + 0.4 \times 2 \times \mathrm{PMAX})$$

[a] Variables are defined in Table D.4.

Having determined the maximum temperature in the pipeline segment and its maximum pressure requirement, the required pipe wall thickness was determined as a function of S, the allowable stress for the pipe material as given in the American National Standards Institute (ANSI) B 31.3 Code.[4] These values are shown in Fig. D.1 for type 304 ss and cs (A-106B) pipe. Minimum wall thicknesses were

Table D.4. Variable definitions

TS(I)	Supply fluid temperature at user's site
TR(I)	Return fluid temperature at user's site
TSU	Required use temperature
DTAU, DTBU	Heat exchanger terminal temperature differences
TRU	User's return fluid temperature
Δh	Enthalpy change in supply fluid at user's site
C_p	Fluid specific heat
m	Fluid mass flow rate
QUSER	Energy delivery rate
A	Pipe cross section
D	Pipe diameter
V	Fluid velocity
ρ	Fluid density
Re	Reynolds number
μ	Fluid viscosity
f	Friction factor
ΔP	Pressure drop between stations spaced "L" apart
PMAX	Maximum pressure in any line section
Fric	Energy input to the fluid, equal to the frictional energy dissipation divided by the pump efficiency
L	Distance between pumping stations
eff	Pump efficiency (assumed = 0.8)
Rth	Thermal resistance[a]
K_S	Soil thermal conductivity
K_I	Insulation thermal conductivity
TKINS	Insulation thickness
Loss	Heat loss per unit length of pipe through the insulation and soil
T_{pipe}	Average temperature of pipe in the segment being considered
T_{ground}	Average groundwater temperature [assumed = 16° C (60° F)]
h_{I+1}	Fluid enthalpy at next pump station (numbers proceeding from user end to supply end)
h_I	Fluid enthalpy at current pump station
t_{wall}	Required pipe wall thickness
S	Pipe material allowable stress, given in ANSI B 31.3 code[b]

[a] See Ref. 3.

[b] See Ref. 4.

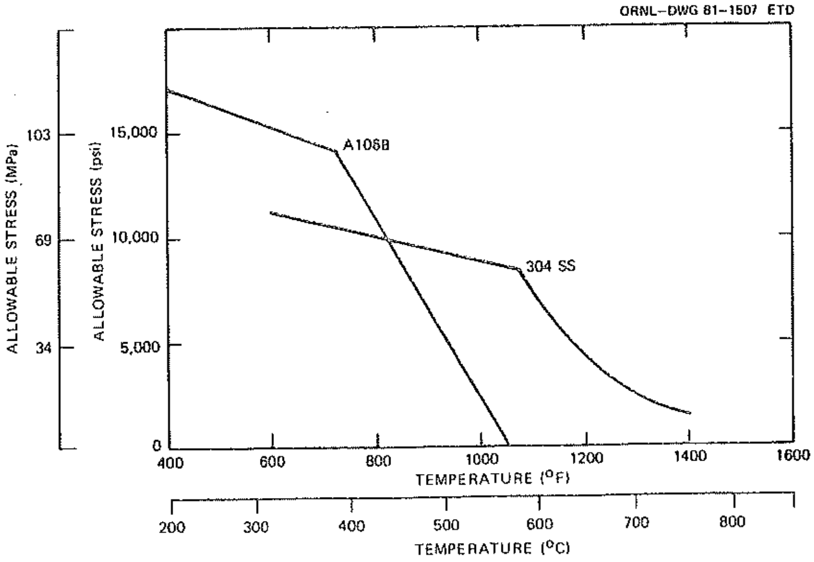

Fig. D.1. Allowable stress for A106B steel and 304 ss pipe.

established for each diameter,[5] and thickness increments of 0.8 mm (1/32 in.) were considered with the next increment above the calculated thickness requirement being used. This procedure was repeated for each pipe segment of the total line, with pipe wall thickness changes being possible from one segment to the next.

Pipe thermal expansion was computed from the average temperatures of both the entire supply and return lines. The expansion coefficients were 4.3×10^{-5} mm/m·°C for ss pipe and 3.25×10^{-5} for cs.[6]

D.3 Pumping and Piping Economics

D.3.1 O&M

A pumping/piping system was assumed to require a four-person staff per shift working 24 h/d for operational duties at a cost of $14/man-hour. This was assumed to remain constant within the range of pumping distances considered in this study.

Maintenance costs were estimated at 3%/year of the total initial piping cost for gas pipelines. Maintenance costs were also at 3%/year of the total initial cost of heat exchangers and pumping stations for liquid pipelines. Maintenance labor was estimated at 60% of this 3% or 1.8% of the initial investment. The remainder of the maintenance costs were assumed to be materials cost.

Overhead was assumed to be 30% of the total labor charge, which included both operating and maintenance labor.

D.3.2 Gas pipeline economics

Cost model equations for gas pipelines are shown in Table D.5. The variables used in these equations are defined in Table D.6. The first two equations give the pipe weight per mile based on 410 MPa (60,000 psi) minimum yield strength pipe. A minimum pipe wall thickness was calculated based on a gas pressure of 5.2 MPa (750 psi).[7]

Table D.5. Gas pipeline cost equations[a]

$$
\begin{aligned}
t &= 0.7P_1d/C_5 \\
W &= C_6t(d - t) \\
C_{pipe} &= (A_1 + A_2d + A_4)C_7 \text{ if } d \leqslant 0.4 \text{ m (16 in.)} \\
C_{pipe} &= (A_1W + A_3d^{C_8} + A_4)C_7 \text{ if } d > 0.4 \text{ m (16 in.)} \\
C_{pump} &= [B_1\,(HP) + B_2d]/L \\
C_{cap} &= FCR(C_{pipe} + C_{pump}) \\
C_{fuel} &= (CF \times HP \times CG \times C_9 \times C_{10})/(L \times \eta_2 \times C_{11}) \\
C_{total} &= C_{cap} + C_{fuel} + C_{O\&M} \\
Cost &= [(C_{cap}/CF) + C_{fuel} + C_{O\&M}]/Q/C_{12}
\end{aligned}
$$

[a] The variables are defined in Table D.6.

Pipe cost C_{pipe} consists of three factors. Pipe cost per unit of weight is represented by A_1. Right-of-way (ROW) costs are represented by A_4. Installation costs are a linear function of pipe diameter for diameters <0.41 m (16 in.), and A_2 is the proportional cost. For diameters >0.41 m, the installation cost is nonlinear and is represented by $A_3d^{0.48}$; this expression is derived in Sect. D.3.3.

Table D.6. Variable definitions

Variable	Unit	Definition
t	m (in.)	Pipe wall thickness
C_5	414 MPa (60,000 psi)	Constant
C_6	24,640 kg/m^3 (28.2 ton/in.^{2}-mile)	Constant
P_1	b (psia)	Maximum gas pressure
d	m (in.)	Pipe diameter
W	kg/m (tons/mile)	Pipe weight per unit length
C_{pipe}	\$/m (\$/mile)	Pipe cost per unit length
A_1	\$/kg (\$/ton)	Specific cost of pipe
A_2	\$/m^2 (\$/in.-mile)	Installation cost, including anti-corrosion wrapping and coating
A_3	\$/m$^{1.48}$ [\$/(in.)$^{0.48}$-mile]	Installation cost, including anticorrosion wrapping and coating
A_4	\$/m (\$/mile)	ROW cost per unit length
C_7	1.3	Dimensionless contingency factor
C_8	0.48	See Sect. D.3.3 for derivation
C_{pump}	\$/m (\$/mile)	Pump cost per unit length of pipe
B_1	\$/W (\$/hp)	Pump cost per unit of power
B_2	\$/m (\$/in.)	Pump station cost, including land, buildings, and controls
L	m (mile)	Distance between pump stations
C_{cap}	\$/m (\$/mile)	Annual fixed-charge per unit distance
FCR	0.17	Fixed-charge rate (Sect. 4.1)
CF	0.70	Capacity factor
C_{fuel}	\$/m (\$/mile)	Annual fuel charge per unit distance
HP	W (hp)	Compressor power per station (or per distance L)
CG	\$/GJ (\$/10^6 Btu)	Gas cost
C_9	8760 h/year (8760 h/year)	Conversion constant
C_{10}	3600 s/h (2545 Btu/hp·h)	Conversion constant
η_2	0.31	Combustion efficiency for gas burned at pumping stations
C_{11}	10^9 J/GJ (10^6 Btu/10^6 Btu)	Conversion constant
C_{total}	\$/m (\$/mile)	Total annual cost per unit length
$C_{O\&M}$	\$/m (\$/mile)	Annual O&M costs per unit length (Sect. D.3)
Cost	\$/Jm (\$/Btu·mile)	Total annual cost per unit distance per unit energy
Q	J/s (Btu/h)	Energy flow rate
C_{12}	3.1536 × 10^7 s/year (8760 h/year)	Conversion constant

Compressor costs consist of two terms, B_1 and B_2. The first term, B_1, represents the cost proportional to installed power. The second term, B_2, reflects the cost of the land, buildings, controls, and other cost factors, which are associated with each pumping station. Station costs are correlated with pipe diameter. Total pump costs are divided by the distance L between stations to give pump and station costs on a per unit distance basis.

The piping cost factors A_1 through B_2 are listed in Table D.7. These costs are based on 1980 costs escalated at a capital construction rate of 7%/year to the start-up year and then de-escalated at the general inflation rate of 6%/year to the 1980 dollar basis.

A contingency of 30% was added to the pipe costs. This contingency provides for site-specific accommodations and any necessary additional engineering and inspection.

Table D.7. Gas piping cost factors

Variable	1990 start-up	2000 start-up
A_1, \$/kg (\$/ton)	0.73 (660)	0.80 (725)
A_2, \$/m^2 (\$/in.-mile)	295 (12,080)	325 (13,270)
A_3, \$/m$^{1.48}$ (\$/in.$^{0.48}$-mile)	185 (51,090)	205 (56,120)
A_4, \$/m (\$/mile)	6.80 (10,980)	7.50 (12,070)
B_1, \$/W (\$/hp)	0.80 (600)	0.90 (665)
B_2, \$/m (\$/in.)	1,564,000 (39,720)	1,718,000 (43,630)

D.3.3 Liquid pipeline economics

Pipe costs were found to be proportional to pipe weight, to be independent of diameter and wall thickness, and to vary only slightly with specification or grade (except for ss). The values used, \$0.71 and \$5.51/kg (\$0.32 and \$2.50/lb) for cs and ss, respectively, represent a best judgement based on discussions with representatives of both Shell Pipeline and Black and Veatch Consulting Engineers and on catalog information provided by U.S. Steel.

Straight line pipe costs were computed and increased by a 2% allowance for fittings and a 5% allowance for detours, elevation changes, and other factors.

Based on information from Johns-Manville Contracting Division, pipe insulation was assumed to cost \$810/m^3 (\$23/ft^3) for calcium silicate, installed and protected by a waterpoof membrane. The costs were quoted on a dollars-per-inch-square-foot basis and were converted to the dollars-per-cubic-meter basis used in this study.

The \$810 figure is higher than the assumed costs found in other studies of this type, which ranged from \$280 to \$640/m^3 of insulation. No attempt was made to optimize the insulation system by use of less expensive insulation types in the outer layers where temperatures would be lower.

The cost expression used for expansion joints was cost = $15.781D^{1.797}$, where D is piped diameter per joint. This expression was based on information, supplied by the Kansas City representative of Pathway Bellows, Inc., for high-temperature corrosion-resistant units with an expansion capability of 203 mm (8 in.) per unit. The basic cost data and the cost expression given in this section are shown in Fig. D.2.

Contacts with Koch Oil Company and Shell Oil Pipeline indicated that 1980 pipeline construction costs average in the range of \$10,000 to \$12,000/in. diam-mile length. These costs are highly variable and very much site-specific.

Oil and Gas Journal[8] (Aug. 13, 1979) gives a table of 1978 pipeline costs. Lines, ranging from 7.9 to 346 km (4.9 to 215 miles) in length in sizes from 0.2 to 1.1 m (8 to 42 in.), are located in a variety of states (and, therefore, terrain), for example, Louisiana, Wyoming, Oklahoma, Texas, and West Virginia.

Labor and miscellaneous costs ranged from \$3,235 to \$15,638/in.-mile; the average for the 14 lines described was \$10,890 (scaled from 1978 at 15% escalation). These miscellaneous costs include surveying, mapping, hydrotesting, building road and canal crossings, making X rays, engineering, inspection, and cathodic protection.

This \$10,890 figure is consistent with the \$10,000 to \$12,000 figure given by both Koch Oil and Shell. The Koch Oil and Shell sources indicated that the "dollars per inch-mile" rule was valid up to about 0.4-m (16-in.) lines but costs per inch-mile are lower for larger diameters. This was confirmed by data[9] plotted on the attached Fig. D.3. The average lay varies strongly with diameter but is only a weak

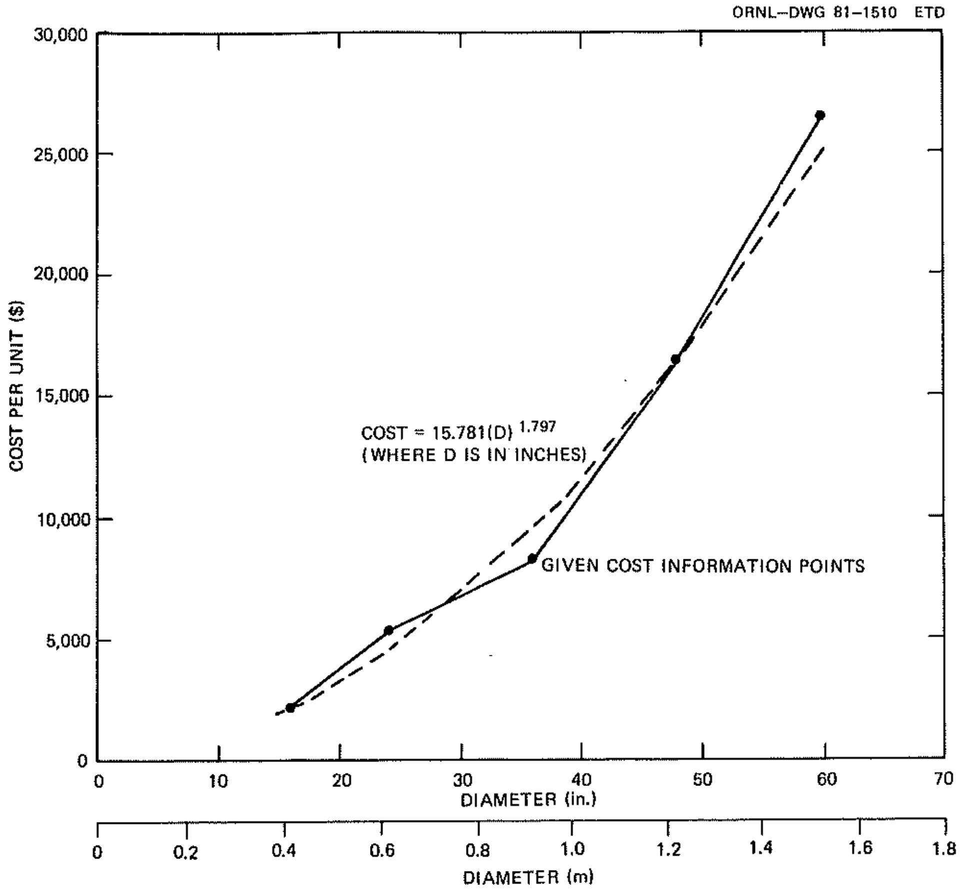

Fig. D.2. Expansion joint costs.

function of wall thickness. To represent the average lay in the range of 16- to 42-in. pipe, the curve $D^{0.48}L = 28381$ (Fig. D.3) appears adequate.

The expression used in this study to evaluate installation cost is cost = \$46,510 ($D^{0.48}L$), where D is in inches and L is in miles. This expression evaluates the cost of a 16-in. line to be \$11,000/in.-mile (as justified above) and scales the diameter dependence to the average-lay data of Fig. D.3.

Installation costs were included separately in the cost analysis for both the supply and return lines. For adjacent pipes, not all operations would double in scale, so this procedure overestimates installation costs.

Right-of-way costs were found to be highly variable, ranging from \$4,186 to \$29,000/mile in the 1980 *Oil and Gas Journal* survey.[8] Koch Oil indicated a range of \$3,200 to \$8,000/mile for ROW only for 16-in. and smaller lines.

A value of \$10,000/mile was selected for this study. This cost is about average and refinement is not warranted because it represents a relatively minor cost.

The *Oil and Gas Journal*[8] gives 1978 compressor station average costs of \$511/hp, installed. (This inflates to \$676 at the 15% annual rate quoted to 1980.)

Shell Oil Pipeline suggested \$800/hp for electric motor drive pumping stations, including metering and controls.

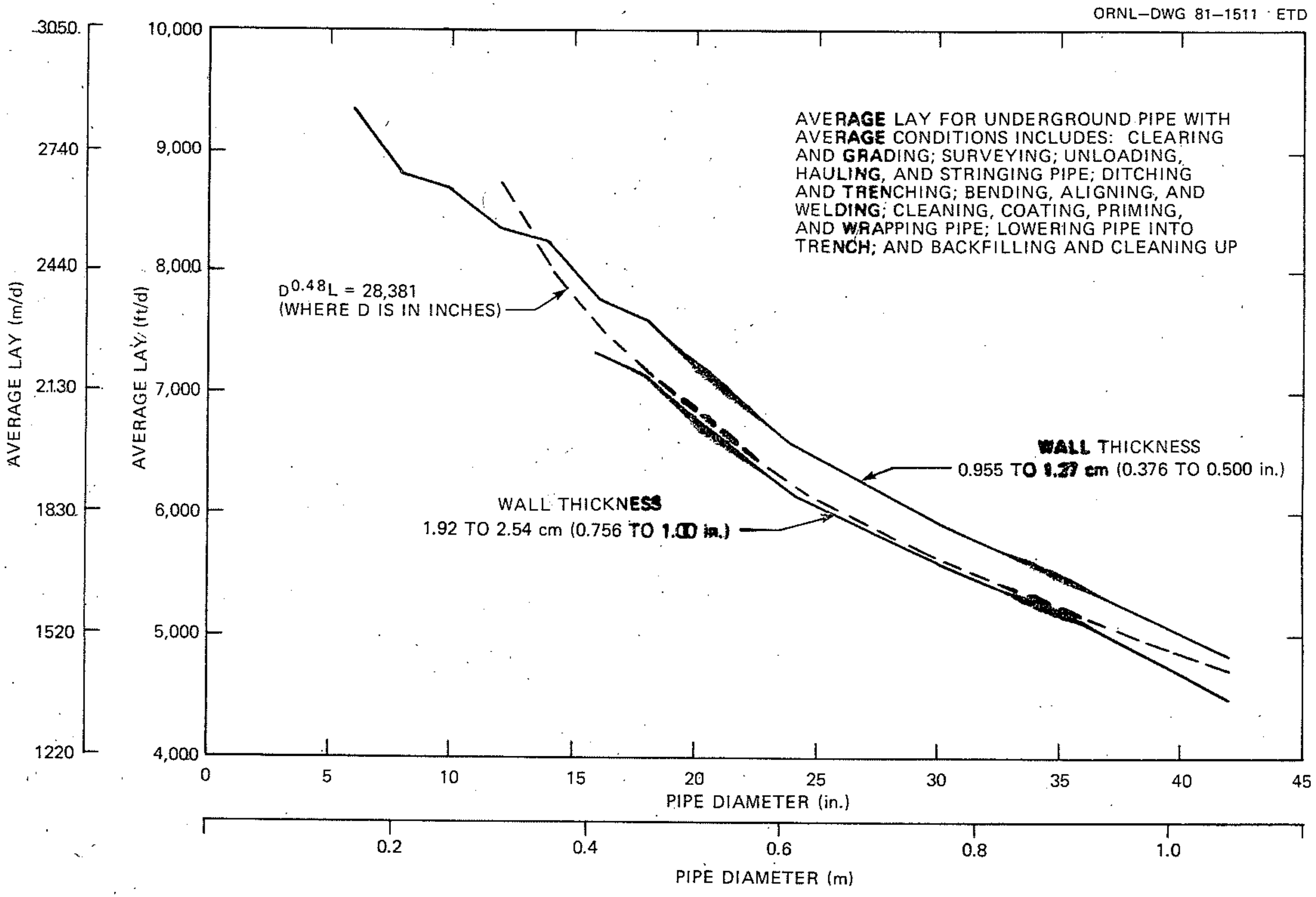

Fig. D.3. Average lay of underground pipe vs pipe diameter. Data were taken from *Cost Estimating Manual for Pipelines and Marine Structures*, by John S. Page, Gulf Publishing, Houston (1977).

Some authors[7] estimate costs of compressor stations using a cost per horsepower factor plus a cost per inch-diameter factor to represent costs of valves, fittings, and structure necessary at the station. The *Oil and Gas Journal* information is plotted in Fig. D.4, and a linear fit of the described form is shown.

A linear fit of the described form is

$$\text{cost} = 377 \times \text{hp} + 1{,}025{,}000\ ,$$

or, inflated at 15%/year from 1978 to 1980, the form is

$$\text{cost} = 499 \times \text{hp} + 1{,}355{,}000\ .$$

With a 36-in. average diameter assumption:

$$\text{cost} = 499 \times \text{hp} + 37{,}639 \times \text{diam (in)}\ .$$

The expression given in Ref. 7, when escalated to 1980 dollars, is

$$\text{cost} = 550 \times \text{hp} + 36{,}200 \times \text{diam}\ .$$

Agreement between the two sources in magnitude and form indicates that this is a reasonable estimate for compressor station costs.

Although pumps are generally less expensive than compressors, because of the expected extra cost of high-temperature salt pumps, a factor of $800 rather than $550/hp was selected for use in this study. For oil, $400/hp was used. The cost expressions used in this study are for salt:

$$\$\ \text{cost} = 800 \times \text{hp} + 36{,}200 \times \text{diam}\ ,$$

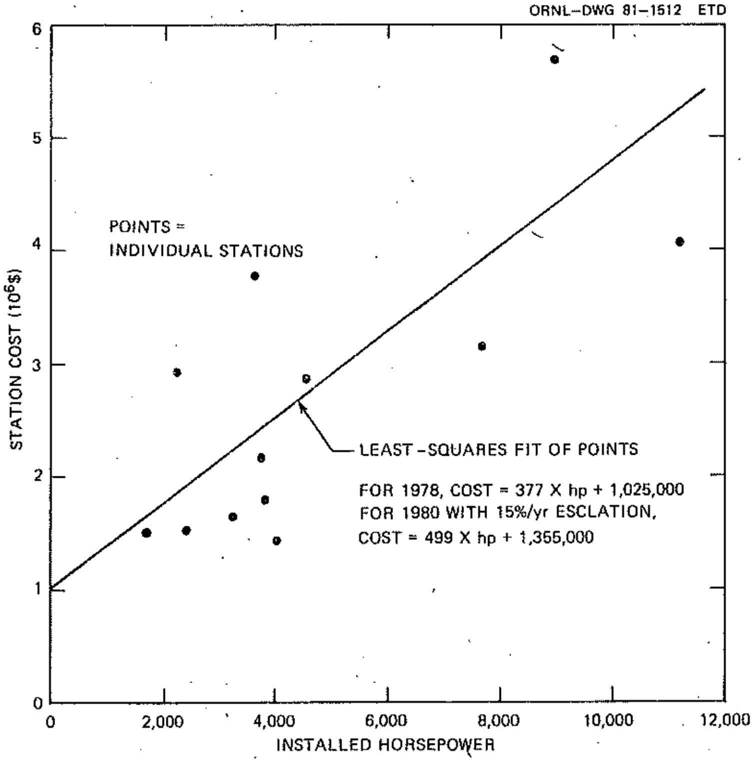

Fig. D.4. Compressor station costs. Data were taken from the *Oil and Gas Journal*, Aug. 13, 1979.

and for oil:

$$\$ \text{ cost} = 400 \times \text{hp} + 36{,}200 \times \text{diam} ,$$

where hp = station bhp (80% efficiency assumed) and diam = pipe diameter, in.

As discussed in Sect. 5.5.2, redundant pumps will be required. This is not considered in the cost estimation presented in this section.

A summary of all liquid pipeline cost equations, variable definitions, and cost factors is given in Tables D.8 through D.10.

Table D.8. Liquid pipeline cost equations[a]

$$t = p\,do/[2S + 2\,(0.4p)]$$

$$di = do - 2t$$

$$\text{CPIPE} = [\pi(do^2 - di^2)/4]F_1 \cdot \text{CC1} \cdot \text{G1}$$

$$\text{CINS} = \left\{\left[\pi\left(do + \frac{t_{ins}}{2}\right)^2 - do^2\right]/4\right\} \; F_2$$

$$\text{CROWD} = F_3$$

$$\text{CLAY} = F_4 do^{0.48}$$

$$\Delta L = \alpha(T_{avg} - T_{amb})\; L$$

$$\text{CJNTS} = \Delta L\; F_5\; do^{1.797}$$

$$\text{BHP} = (\Delta p \cdot Q/\eta_p)\; \text{CC2}$$

$$\text{CPUMP} = F_6 \text{BHP} + F_7 do$$

$$\text{CTLPIP} = [(\text{CPIPE} + \text{CINS} + \text{CROWD} + \text{CLAY})(L) + (\text{JNTS} + \text{CPUMP})](\text{CCC})$$

$$\text{CFINV} = (\pi\; di^2/4)\rho_F \cdot F_8$$

$$\text{LMTD} = (\text{DTA} - \text{DTB})/(\ln \text{DTA}/\text{DTB})$$

$$\text{ASX} = \text{QXFR}/(\text{U} \cdot \text{LMTD})$$

$$\text{CHX} = F_9 \cdot \text{ASX} \cdot \text{CCC}$$

$$\text{CPCS} = 0.05\; \text{CTLPIP}$$

$$\text{CPRJ} = 0.2\; \text{CTLPIP}$$

$$\text{CEI} = 0.1\; \text{CTLPIP}$$

$$\text{COPLBR} = 2 \times 2 \times 14.00 \times 8760$$

$$\text{CMLBR} = 0.03\;(\text{CHX} + \text{CPUMP})(0.6)$$

$$\text{CMMAT} = 0.03\;(\text{CHX} + \text{CPUMP})(0.4)$$

$$\text{ADOH} = 0.3\;(\text{COPLBR} + \text{CMLBR})$$

$$\text{CELEC} = F_{10} \times \text{BHP} \times \text{CC3} \times \text{CAPFAC} \times 8760$$

$$\text{CHEAT} = \text{QSCP} \times 8760 \times \text{CAPFAC} \times F_{11}$$

$$\text{AFDC} = (1.095^3 - 1)(\text{CTLPIP} + \text{CHX} + \text{CFINV} + \text{CPCS} + \text{CPRJ} + \text{CEI})$$

[a] The variable definitions are given in Table D.9, and the liquid pipeline cost factors are given in Table D.10.

Table D.9. Variable definitions[a]

Name	Units	Definition
t	m (in.)	Pipe wall thickness
p	pa (psi)	Pipe maximum pressure
s	N/m^2 (psi)	Pipe material allowable stress
do	m (ft)	Pipe outside diameter
di	m (ft)	Pipe inside diameter
G_1	1.07 (1.07)	Allowance for turns, bends
F_1	\$/kg (\$/lb)	Unit cost, pipe material
CPIPE	\$/m (\$/ft)	Pipe cost
t_{ins}	m (ft)	Insulation thickness
F_2	$\$/m^3$ ($\$/ft^3$)	Unit cost, insulation
CINS	\$/m (\$/ft)	Insulation cost
CROWD	\$/m (\$/ft)	ROW and damages cost
F_3	\$/m (\$/ft)	Unit cost, ROW and damages
F_4	$\$/m^{1.48}$ ($\$/ft{\cdot}in.^{0.48}$)	Unit cost, pipeline construction
CLAY	\$/m (\$/ft)	Pipeline construction cost
ΔL	m (in.)	Pipe thermal expansion
α	m/m·°C (in./ft·°F)	Thermal expansion coefficient
T_{avg}	°C (°F)	Average pipe temperature
T_{amb}	°C (°F)	Average ambient temperature
L	m (ft)	Pipe length
F_5	$\$/m^{2.797}$ ($\$/in.^{2.797}$)	Unit cost, expansion joints
CJNTS	\$ (\$)	Expansion joints cost
Δp	pa (psi)	Line pressure drop
Q	m^3/s (ft^3/s)	Volumetric flow rate
ηp	0.8 (0.8)	Pump efficiency
BHP	W (hp)	Pump power
F_6	\$/kW (\$/hp)	Unit cost, pumps
F_7	\$/m (\$/in.)	Unit cost, pump station
CPUMP	\$ (\$)	Pumps cost
CCC	$(1.07/1.06)^{10}$[b] $(1.07/1.06)^{10}$[c]	Differential inflation rate factor
CTLPIP	\$	Total pipeline cost
φ_F	kg/m^3 (lb/ft^3)	Fluid density
F_8	\$/kg (\$/lb)	Unit cost, fluid
CFINV	\$ (\$)	Fluid inventory cost
LMTD	°C (°F)	Log mean temperature difference
DTA	°C (°F)	Terminal temperature difference
DTB	°C (°F)	Terminal temperature difference
ASH	m^2 (ft^2)	Heat exchanger surface area
QXFr	W (Btu/h)	Heat exchange rate
U	$W/m^2{\cdot}°C$ ($Btu/ft^2{\cdot}h{\cdot}°F$)	Heat transfer coefficient
F_9	$\$/m^2$ ($\$/ft^2$)	Unit cost, heat exchanger surface
CHX	\$	Heat exchanger cost
CPCS	\$	Process contingency
CPRJ	\$	Project contingency
CEI	\$	Engineering and inspection cost
COPLBR	\$/year	Operating labor cost
CMLBL	\$/year	Maintenance labor cost
CMMAT	\$/year	Maintenance material cost
ADOH	\$/year	Administration and overhead cost
F_{10}	\$/kWh (\$/kWh)	Unit cost, electrical power
CAPFAC	0.7 (0.7)	Capacity factor
QSUP	GJ/h (MBtu/h)	Reactor heat requirement
F_{11}	\$/GJ (\$/MBtu)	Unit cost, reactor heat

Table D.9 (continued)

Name	Units	Definition
AFDC	$	Allowance for funds during construction
CC1	7860 kg/m^3 (490 lb/ft^3)	Density of steel
CC2	1000 W/kW (0.262 $hp \cdot s \cdot in./ft^3 \cdot lb$)	Conversion factor
CC3	1 kW/kW (0.746 kW/hp)	Conversion factor

[a] Cost factors are given in Table D.10.
[b] 1990 start-up.
[c] 2000 start-up.

Table D.10. Liquid pipeline cost factors[a]

F_1	\$5.51/kg	\$2.50/lb, ss pipe
	\$0.71/kg	\$0.32/lb, A106B pipe
F_2	\$810/m^3	\$23/ft^3
F_3	\$6.21/m	\$10,000/5,280 ft
F_4	\$168.45/$m^{1.48}$	\$46,510/5,280/$ft \cdot in.^{0.48}$
F_5	\$57,112/$m^{2.797}$	\$1.973/$in.^{2.797}$
F_6	\$1,073/kW	\$800/hp, salt
	\$536.42/kW	\$400/hp, oil
F_7	\$1,425,196/m	\$36,200/in.
F_8	\$0.66/kg	\$0.30/lb, salt
	\$0.33/kg	\$0.15/lb, oil
F_9	\$355.08/m^2	\$33.00/ft^2
F_{10}	\$0.0243/kWh	\$0.0243/kWh
F_{11}	\$3.96/GJ	\$4.18/10^6 Btu

[a] 1980 costs scaled by the factor CCC: CCC = $(1.07/1.06)^{10}$ for 1990 start; CCC = $(1.07/1.06)^{20}$ for 2000 start.

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Appendix E

GLOSSARY OF ACRONYMS

ANSI	American National Standards Institute
API	American Petroleum Institute
CHP	chemical heat pipe
CRBRP	Clinch River Breeder Reactor Project
cs	carbon steel
DOE	Department of Energy
EPA	Environmental Protection Agency
FCR	fixed-charge rate
FSV	Fort St. Vrain
GA	General Atomic Company
HEU/Th	high enriched uranium/thorium
HTGR	high-temperature gas-cooled reactor
IHX	intermediate heat exchanger
KFA	Institut für Reaktorentwicklung, Kernforschungsanlage, Julich, Federal Republic of Germany
KOH	potassium hydroxide
K-T	Koppers-Totzek
LBG	low-Btu gas
LEU/Th	low enriched uranium/thorium
LWR	light-water reactor
MBG	medium-Btu gas
MFBI	major fuel-burning installations
O&M	operation and maintenance
PBR	pebble-bed reactor
PCB	polychlorinated biphenyls
PWR	pressurized-water reactor
R&D	research and development
ROW	right of way
SASOL	South Africa Synthesis Oil Ltd.
SIC	Standard Industrial Classification
SMSAs	Standard Metropolitan Statistical Areas
SNG	synthetic natural gas
SPE	solid polymer electrolyte
ss	stainless steel
VHTR	very high temperature gas-cooled reactor

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