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STUDY, COST, AND SYSTEM ANALYSIS OF LIQUID HYDROGEN PRODUCTION

FINAL REPORT

by N. C. Hallett

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for

MISSION ANALYSIS DIVISION
AMES RESEARCH CENTER
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ABSTRACT

This report contains information related to contemplated large-scale liquid hydrogen systems. Descriptions of feasible processes and equipment are presented. Information concerning availability and cost of required raw materials and energy are projected. Composite system analyses based on preliminary NASA hypersonic transport (HST) liquid hydrogen requirements indicate estimated average product cost of 7.7 to 8.8 cents per pound.

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STUDY, COST, AND SYSTEM ANALYSIS OF LIQUID HYDROGEN PRODUCTION

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Air Products and Chemicals, Inc.

INTRODUCTION

Air Products and Chemicals, Inc. was retained by the National Aeronautics and Space Administraion (NASA) to perform a study related to systems that produce liquid hydrogen as propulsion fuel in aircraft. The study objective was that of developing information to project hydrogen fuel costs in the post-1980 period. These costs were to reflect production and distribution technology, plus costs of raw materials and process energy. Further, consideration of integrating hydrogen fuel production with other industrial activities was to be considered. NASA proposed a hydrogen fuel requirement schedule based on preliminary projections for hypersonic transport (HST) systems in the post-1980 period.

NOTE

The documents and companies referenced in this report are listed in numerical order in Appendix F. For example, if text contains mention of U.N. Statistical Paper[43], item 43 in Appendix F lists the source of information. References are made also to companies from whom data has been and may be obtained.

ACKNOWLEDGMENT

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Section 1

STUDY DESCRIPTION

1.1 HYDROGEN MARKET DEMAND

Gaseous and liquid hydrogen have been marketable commodities for less than two decades. Demand for hydrogen prior to 1955 was small compared with present standards. These small requirements were satisfied generally by small-scale generation equipment.

Since 1955, however, major developments in two specific areas have resulted in a substantial demand for hydrogen. Development of rocket propulsion by burning liquid hydrogen has resulted in the present demand for liquid hydrogen. Secondly, developments within the process industry in general have established substantial demands for gaseous hydrogen. Agricultural chemical production, for example, utilizes gaseous hydrogen as a basic synthesis component.

Current world production of liquid hydrogen is estimated to be between 200 and 400 tons per day, with most of the liquid being used in rocket propulsion. By comparison, aggregate production requirements for a proposed hypersonic transport system approaches 8000 tons per day. Individual proposed area requirements vary between 200 and 2500 tons per day per facility.

Therefore, this report is directed at satisfying the liquid hydrogen requirement of the proposed HST system. Other liquid hydrogen requirements will exist but probably will be incrementally small when compared to possible HST needs.

1.2 LARGE-SCALE LIQUID HYDROGEN SYSTEMS

Basic questions of raw material supply and process energy source must be considered for the large-scale facilities. Raw material sources for hydrogen can be classified as aqueous solutions or hydrocarbon materials. To date, natural gas and petroleum products have been primary raw materials used for product hydrogen. Raw material for hydrogen production in the quantities studied in this project would be equivalent to slightly more than 1 percent of the world's annual natural gas production. Reserves of natural gas and petroleum products would deplete at accelerated rates with the undertaking of this HST system. To alleviate the raw material problem, aqueous solution processing (water electrolysis for example) may be a hydrogen generation process. Also, carbonaceous materials, namely, coal in coal-water processes, will be developed for use. In any event, the supply of required amounts of raw material will be a major problem for large-scale hydrogen production.

Process energy, likewise, will be required at high rates compared to current process industry standards. Electrical energy will be needed in vast amounts for electrolysis systems.

It is projected that the product hydrogen for the proposed HST system would require as much as 8 percent of all electrical energy now produced in the United States.

In addition to the raw material and process energy problems, there will be radically new applications of technology. The largest currently operating liquid hydrogen facility has a production rate of approximately 50 tons per day. By comparison, proposed production requirements for the HST system will range from 200 to 2500 tons per day. The required scale-up and extension of technology is apparent.

1.3 STUDY PROGRAM DESCRIPTION

In accordance with the stated objectives, the study program was organized into three specific task activities:

Section 3, Parametric Study
Section 4, Logistics
Section 5, Composite System Analyses

The parametric study incorporates extensive detail description of anticipated liquid hydrogen systems. Information was developed with respect to the following four specific operations in a liquid hydrogen system:

(1) Hydrogen gas production and purification

(2) Hydrogen gas liquefaction and conversion

(3) Storage and transfer of liquid hydrogen

(4) Distribution of liquid hydrogen

For each of the four basic operations or subsystems, several techniques, equipment types, or operating modes are possible. The basic objective of the parametric study, then, was to develop "parametric information" for each of the four subsystems. After development of the parametric or "building block" data, liquid hydrogen systems could be constructed in accord with postulated geographic liquid hydrogen requirements.

Data and information developed in the parametric study can be termed "descriptive". That is, information related to process and equipment description was required. Based on the descriptive information, unit requirements, or other characteristics of the respective subsystems, were quantitatively developed. Amounts of raw material, process energy, capital investment, etc., required for unit production from the given subsystem alternates were developed. Cost of the required commodities was not developed. Estimates of the required quantities of respective commodities were generated. Capital investment, for example, was based on Washington, D.C., costs and 1967 United States dollars. In addition, the unit requirements were developed so that the effect of plant size could be reflected.

The descriptive and quantitative information developed is contained in Section 3. From the information developed, it is possible to determine the liquid hydrogen facility requirements as dependent upon chosen "building blocks" and plant production requirements. This information provides the basis for making comparisons between various systems conceived for the same production requirements.

The logistics activity was primarily the development of projected availability and costs of commodities required in liquid hydrogen production. Data was assembled in relation to types and amounts of commodities for various systems at respective geographic locations. Generation of information detailing prospective electrical energy systems was another major part of the logistics activity. Study of current and projected electrical energy supply and demand patterns was completed with detailed analysis concerning liquid hydrogen plant locations. The data developed in the logistics work is described in Section 4.

Composite system analyses, the third major portion of the study program, incorporated extensive numerical analysis of the parametric and logistic information. The analyses were based upon a possible fuel-requirement schedule for HST service. This schedule is presented in table 1-1.

TABLE 1-1. HST FUEL SCHEDULE

Location	Nominal Daily Requirement T/D LH_2
Bangkok, Thailand	200
Honolulu, Hawaii, U.S.A.	200
Johannesburg, Union of South Africa	200
Beirut, Lebanon	300
Sao Paulo, Brazil	400
Sydney, Australia	400
Tokyo, Japan	500
Los Angeles, California, U.S.A.	1300
Paris, France	2000
New York City, New York, U.S.A.	2500

Various liquid hydrogen plants were "built up" from the parametric study information. Based on the types and amounts of commodities required for respective plants, comparative sets of production costs were drawn. From this, the projected unit production cost of liquid hydrogen fuel was developed at the respective locations.

A second set of unit production costs was developed to reflect effects of combining liquid hydrogen production with other industrial activities in the respective areas. This incorporated preliminary projection of the general industrial patterns to be expected in specific areas. Based on this information, reductions in production cost were projected for certain of the areas considered.

Detailed information and calculational results of the composite analyses are presented as Section 5. A sample of calculational detail is appended to the report to document the actual mathematical procedure.

A final activity within the study program was that of defining areas where advancement may reduce the production cost of liquid hydrogen. Information and results of this activity are presented as Section 6.

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Section 2

CONCLUSIONS

2.1 PROJECTED PRODUCTION COSTS

The result of production-cost estimates are shown in table 2-1. Calculation results are presented in Section 5 of this report.

TABLE 2-1. PROJECTED PRODUCT COSTS

Plant Location	Nominal Production Rate (Tons LH_2/Day)	Stand Alone Facility Product Cost[a] (\$/lb LH_2)	Integrated Facility Product Cost[a] (\$/lb LH_2)
Bangkok, Thailand	200	0.113	0.093
Honolulu, Hawaii	200	0.132	0.124
Johannesburg, Union of South Africa	200	0.108	0.100
Beirut, Lebanon	300	0.106	0.087
Sao Paulo, Brazil	400	0.104	0.095
Sydney, Australia	400	0.098	0.083
Tokyo, Japan	500	0.099	0.084
Los Angeles, U.S.A.	1300	0.081	0.071
Paris, France	2000	0.085	0.077
New York City, U.S.A.	2500	0.079	0.073
Total Weighted Average Cost	8000	0.088	0.079

[a] Byproduct credit incorporated where applicable.

2.2 GENERAL OBSERVATIONS AND CONCLUSIONS

Contemplated liquid hydrogen systems will be characterized by their respective gaseous hydrogen generation and purification process. Hydrocarbon processing will be employed in most situations. Projected economics and raw material availability favor the steam reforming or partial oxidation type processes. Electrolysis processes, although not presently projected as being economically attractive will likely become somewhat more competitive in the future. Advances in process technology and materials of construction will be more beneficial to the electrolysis processes than to the hydrocarbon processes. Other process types could be further considered for large-scale systems. For example, hydrogen halide systems, presently undesirable because of poor economics, could prove worthy of consideration as technology changes.

In view of current concern regarding urban environments, hydrocarbon processing may not be desirable in larger, well-developed communities. In those cases, electrolysis processing could be a feasible process, but with approximately 30 percent additional product costs. In view of this, it is suggested that effort be expended in the near future to more fully develop the electrolysis process technology in order to reduce costs for this process in line with competing processes.

The analysis technique developed in this report can be utilized in future evaluations. Incorporation of updated numerical data will yield different results, but the evaluation technique will be valid. For example, an 11 percent annual capital charge rate was used for this report. Government participation in financing could affect a different capital charge rate (6 percent, for example), which would result in a changed set of calculations based on this analysis technique.

Location of the production facilities affects liquid hydrogen production costs. New sets of calculations could be developed for alternate liquid hydrogen use-points. It is expected that significant cost reductions could result from such an analysis.

Integrated facilities, as herein discussed, should result in projected production cost savings of 10 to 20 percent of the stand-alone operations. The cost savings are based on current experience with other integrated facilities. Upon achieving clearer understanding of the projected hydrogen program, a more detailed evaluation of the integrated facility could be made for specific locations. It is projected, however, that detailed integrated-facility studies would show production cost savings ranging from 10 to 30 percent of the stand-alone costs.

SECTION 3

PARAMETRIC STUDY

3.1 OBJECTIVES

The primary objective of the parametric study is to establish unit quantities required for or produced by large-scale liquid hydrogen systems. A secondary objective of this study is to provide a technical description of various components of which, when combined, constitute the total hydrogen supply system.

This report is organized so that technical evaluation of all liquid hydrogen systems is based on four primary subsystem functions:

(1) Hydrogen gas generation and purification

(2) Hydrogen gas liquefaction and conversion

(3) Liquid hydrogen storage

(4) Liquid hydrogen distribution

Several processes and different types of equipment are available for each of the subsystems. The following presentation develops unit consumption and production data for the processes and equipment.

3.2 HYDROGEN GAS GENERATION AND PURIFICATION

An important part of any liquid hydrogen system is the process equipment and related hardware that generate relatively pure hydrogen gas. Gaseous hydrogen is not as readily available as certain other industrial gases; e.g. oxygen or nitrogen. Hydrogen gas must be generated in a combustion or reformer reaction, or it must be produced by the dissociation of water or an aqueous solution.

In addition to producing hydrogen gas, this process must also remove contaminating material. If the generated gas is contaminated, liquefaction and conversion equipment must be designed to remove the contamination. Unchecked contamination generally causes plugging and fouling of cryogenic equipment.

The following paragraphs provide process descriptions of the process that can be employed in large scale hydrogen systems. Unit consumption data is classified as process elements or cost elements. Process elements detail amounts of raw material, process energy, and byproducts related to net unit production of hydrogen. Similarly, cost elements detail typical amounts of capital investment and operating costs associated with the referenced net unit production. This information is presented in a manner so that computer programming can be utilized later for initiating systems and related unit requirements.

3.2.1 STEAM REFORMING

On a volume basis, steam reforming is currently the most widely used process for hydrogen gas generation. In recent years, plants have been built or are being designed with hydrogen gas generation capacity approaching several hundred tons per day. It is probable that steam-reforming will be a major process in the 1980 period.

3.2.1.1 Process Description

Steam reforming includes several reactions involving hydrocarbon compounds and steam. A light hydrocarbon feedstock, which has been desulfurized, is combined with steam in a reformer furnace. The furnace supplies endothermic heat reaction which supports the conversion of steam and hydrocarbons to hydrogen, carbon monoxide, and carbon dioxide. Products from the furnace, combined with additional steam, pass through a catalyst bed which promotes a conversion of carbon monoxide and steam to additional hydrogen and carbon dioxide.

Product gas (hydrogen and carbon dioxide) then passes through an absorber tower with a monoethanolamine (MEA) scrubbing fluid. The product gas is purified by removal of carbon dioxide. At this point, the product gas is 96 percent hydrogen, and the remaining impurities are removed subsequently by cryogenic adsorption as shown in figure 3-1.

Natural gas, refinery gas, propane, butane, and naptha are among the common raw feedstocks that may be utilized in a steam-reforming process. However, all these terms are generic; (i.e., the hydrocarbon content varies greatly within each group). Ultimately, the unit consumption of feedstock material per unit of hydrogen produced is a function of the amount of hydrogen contained in that feedstock.

3.2.1.1.1 Natural Gas Stream.– Natural gas, well gas, or liquefied gas will likely be used for the large-scale systems of the future. Table 3-1 presents the composition of a typical natural gas stream available in the northeast United States.

TABLE 3-1. TYPICAL NATURAL GAS COMPOSITION (U.S. NORTHEAST PIPELINE)

Gas Composition		% Volume
Methane	CH_4	94.60
Carbon dioxide	CO_2	0.65
Nitrogen	N_2	0.46
Ethane	C_2H_6	3.44
Propane	C_3H_8	0.53
Isobutane	C_4H_{10}	0.17
Isopentane	C_5H_{12}	0.10
Hexane	C_6H_{14}	0.05
		100.00

As referenced in table 3-1, methane represents a substantial portion of the natural gas stream. Therefore, the unit data following is based upon a pure methane feed. This provides a common basis for analysis of feed streams of varying hydrocarbon composition throughout the world.

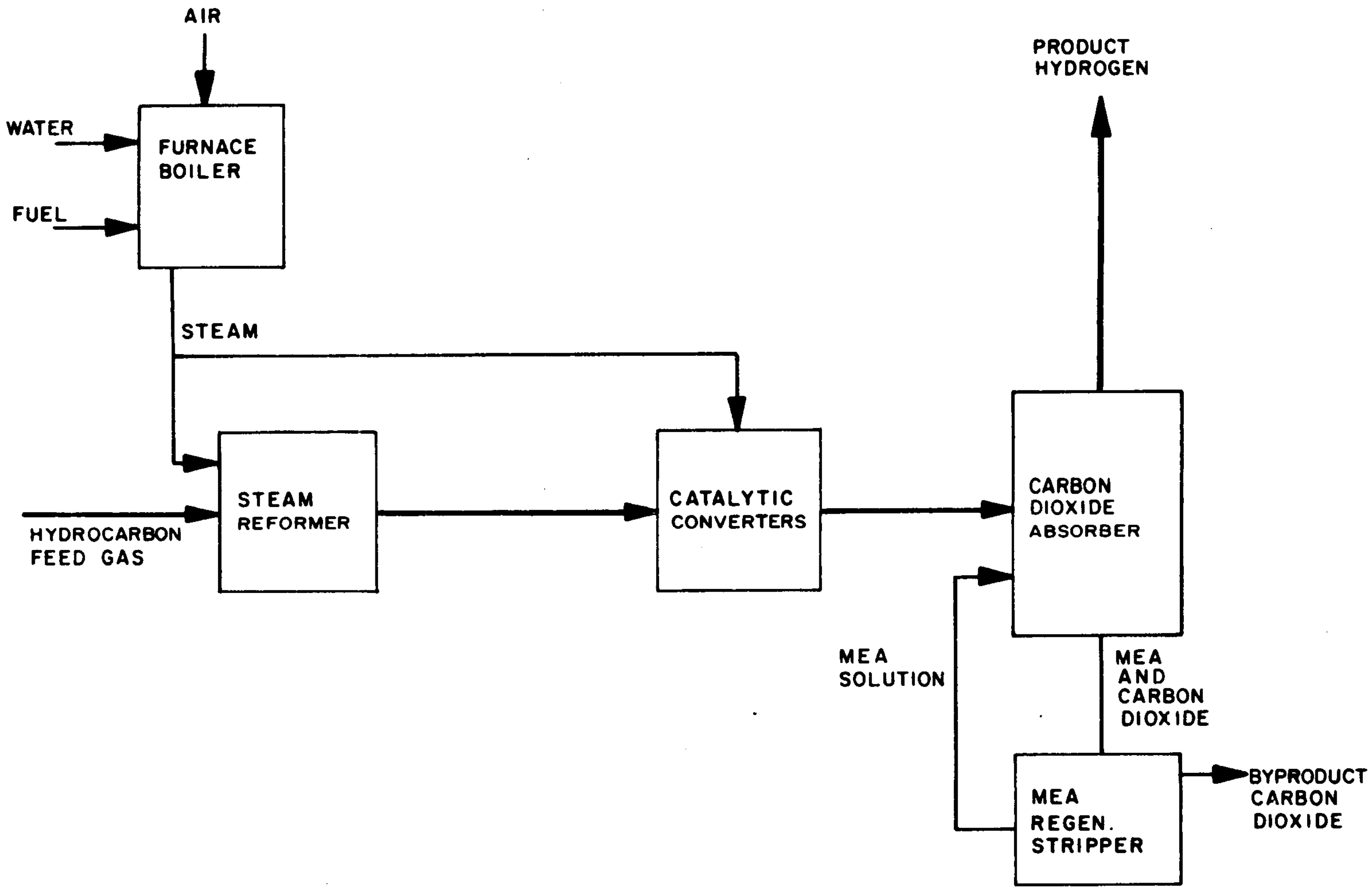


FIGURE 3-1. STEAM REFORMING PROCESS BLOCK DIAGRAM

3.2.1.1.2 Production of Carbon Dioxide. – The results of a carbon dioxide marketability study are presented in Section 5. The potential problem of CO_2 disposal and possible objections to atmospheric venting of the gas have been investigated. It has been found that quantities of CO_2 in excess of the largest quantity contemplated in this study are presently being vented in major metropolitan areas such as New York, Chicago, and Los Angeles from large coal- and gas-fired power plants. Carbon dioxide, which is a combustion product of these plants, is issued at a rate of up to 25,000 tons per day. The hydrogen facility is expected to produce less than 15,000 tons per day at maximum production rates.

The venting of carbon dioxide presently is considered tolerable if concentration at ground level is less than 1 percent. The disposal methods now employed incorporating proper vent-stack design and height are assumed to be suitable for the proposed facilities. Thus, carbon dioxide disposal represents no great difficulty.

3.2.1.1.3 Reactions With Steam – Hydrocarbons react catalytically with steam as follows:

$$C_nH_{2n+2} + nH_2O \longrightarrow nCO + (2n+1)H_2$$

(steam-hydrocarbon reaction)

followed by

$$CO + H_2O \longrightarrow CO_2 + H_2$$

(shift reaction)

resulting in an overall reaction of

$$C_nH_{2n+2} + 2nH_2O \longrightarrow nCO_2 + (3n+1)H_2 \qquad (1)$$

This reaction is endothermic and takes place over a nickel catalyst.

The removal of carbon dioxide is accomplished by absorption of the gas in an amine solution such as aqueous monoethanolamine (MEA) through the reactions

$$CO_2 + 2HOCH_2CH_2NH_2 + H_2O \longrightarrow (HOCH_2CH_2NH_3)_2CO_3$$

and

$$CO_2 + HOCH_2CH_2NH_2 + H_2O \longrightarrow HOCH_2CH_2NH_3HCO_3 \qquad (2)$$

The absorbed carbon dioxide is then separated from the amine solution by heating, which also regenerates the MEA for further carbon dioxide removal.

3.2.1.2 Process Elements

The following process studies show that certain relationships exist between the various process elements and plant capacities.

3.2.1.2.1 Raw Material Unit Consumption. – With proper reformer system design, the decomposition of hydrocarbons is substantially complete at temperatures of 1500°F and above. Thus, the hydrogen produced contains less than 0.1 percent of residual hydrocarbons. Since present reformer designs permit complete reaction, there should be no further reduction in the unit quantity of feed gas required for production of a unit of hydrogen. Therefore, the relationship of feed-gas requirements to plant capacity is that of a constant.

Furnace system operation at pressures above 300 to 350 psig reduces the high costs of compressing light hydrogen gas. The increased pressure required unfavorably affects steam-hydrocarbon reaction and has little effect on the shift reaction.[1]

Reduction in the unit quantity of fuel required per unit of hydrogen is probable in post-1980. Extensive studies are in progress to develop catalysts which are physically more rugged, are less susceptible to feedstock poisoning, and that will completely decompose the feedstock at lower operating temperatures. In addition, as plant size increases, use of multiple-cell reformer furnaces of advanced design will increase thermal efficiency. At least one large engineer-constructor of steam reformer systems is researching this area.[2] Present estimates indicate a 5 to 10 percent decrease in fuel requirements as the complete facility size approaches 2500 tons per day. This decrease will vary linearly in accord with plant size.

Cooling-water makeup requirements are expected to remain constant on a unit basis as plant size increases. This increase will be accomplished by addition of parallel production units. Boiler-water makeup requirements will also remain constant on a unit basis, since the steam consumption is directly tied to the constant feed-stock consumption.

The relation of raw material requirements to plant capacity can be expressed as

$$R_i = R_{B_i} + M_{R_i} (C - C_B)$$

where

R_i = raw material requirement for material i at system capacity (C)

R_{B_i} = raw material requirement for material i at base system capacity (C_B)

M_{R_i} = raw material scale factor for material i

C = system capacity

C_B = base system capacity (250 T/D)

(1)

The above equation when reduced to simplest form becomes

$$R_i = K_{R_i} + M_{R_i} C$$

where

$$K_{R_i} = R_{B_i} - M_{R_i} C_B \qquad (2)$$

The units and estimated values for K_{R_i} and M_{R_i} for the various raw materials consumed in the steam reforming process are tabulated in table 3-2.

TABLE 3-2. UNITS AND ESTIMATED VALUES

Material	K_{R_i}	M_{R_i}
Feed gas (methane)	1.90 lb/lb H_2	0
Fuel gas (methane)	1.44 lb/lb H_2	-4.4×10^{-5} lb/lb T/D
Cooling-water makeup	22.4 lb/lb H_2	0
Boiler-water makeup	16.6 lb/lb H_2	0

3.2.1.2.2 Process Energy Unit Consumption. – Process energy for the steam reforming process is required mainly for amines pumps, feedstock compression, and product compression to liquefier feed pressure. Use of multiple production trains enables pumping requirements to be optimized for each train by utilizing updated pump hardware. Thus, energy requirements remain constant per unit of hydrogen produced. Additionally, feedstock compression energy requirements remain constant per unit of production. Process energy unit consumption is estimated to be 0.47 kwh/lb of hydrogen.

3.2.1.3 Byproduct Unit Production

Removal of CO_2 generated during hydrogen production is accomplished by using an MEA solution. From the reaction;

$$C_nH_{2n+2} \quad + \quad 2nH_2O \longrightarrow nCO_2 \quad + \quad (3n+1)H_2$$

it is apparent that, if the feed were pure methane (CH_4), every four moles of hydrogen produced would be accompanied by one mole of carbon dioxide. Therefore, the byproduct production of carbon dioxide per unit of hydrogen produced will remain constant for any particular feed gas. The carbon dioxide byproduct production is 5.3 lb/lb H_2, if CH_4 feed is utilized.

3.2.1.4 Cost Elements

Capital investment and operating costs, other than raw materials and energy, have been studied over the selected capacity range of 250 T/D to 2500 T/D of gaseous hydrogen production. Capital investment and operating costs, as described in the following paragraphs, were found to vary exponentially with plant size.

3.2.1.4.1 Capital Investment. – Capital investment data has been obtained from several steam-reformer engineer-constructors for plants ranging from 125 T/D to 400 T/D of gaseous hydrogen.[3,4]

It is common in the chemical process industries, to make use of "The Williams Exponential Factor". Correlating the cost of plant and equipment, Williams found that the cost varies as a power of the size parameter;[5,6] that is:

$$I = I_B \left(\frac{C}{C_B} \right)^{M_I}$$

where

I = capital investment required for system capacity (C)

I_B = capital investment required for base system capacity (C_B)

C = required system capacity

C_B = base system capacity

M_I = exponential constant

Williams' law was studied in great detail and its usefulness has been verified for application to complete plant estimates.[7]

The exponential constant (M_I) often has been assigned a value of 0.6. However, as pointed out, discretion should be employed in its use beyond a ten-fold range of capacity. The factor is empirical, and a more exact value for a particular type of plant equipment can be established from available data.[8]

Information available in the industry indicates that a 0.7 factor applies to gaseous hydrogen production equipment. In addition, since the largest plants now operating are smaller than the lower limit of this study (250 T/D), more than a ten-fold range is needed to reach the required 2500 T/D plant size. For these reasons, an exponential factor of 0.7 has been used herein to ensure that the data adequately represents these large plant costs.

The relationship of plant investment with plant size is shown in figure 3-2. The estimated capital investment for the base system (250 T/D) is $\$7.65 \times 10^6$, which compares favorably with available investment figures. A reformer system capable of producing 130 T/D (50 MM/scf/day) of gaseous hydrogen at 96 percent purity and 250 psig has been estimated at 4.5 to 5 million dollars. The feedstock and product hydrogen compression necessary for the plants considered in this study, plus other on-site utility costs, would add an additional 1.2 million dollars to the plant cost.

3.2.1.4.2 Operating Costs. – Operating costs include labor, chemicals and lubricants, catalysts, maintenance costs, general and administrative costs, and home office allocations.[3,4,9,10] It should be noted that taxes and insurance are not included in the referenced cost analyses, but are calculated in the plant capital investment charge.

The requirement of four men for a one man per shift twenty-four hour operation can be demonstrated as follows:

$$1 \text{ operating month} = 30.4 \text{ days} \times \frac{24 \text{ hr.}}{\text{day}} = \frac{730 \text{ hr.}}{\text{mo.}}$$

and for one operator

$$\frac{40 \text{ hours}}{\text{week}} \times \frac{4.3 \text{ weeks}}{\text{man month}} = \frac{172 \text{ hours}}{\text{man month}}$$

$$\frac{730 \text{ hours}}{\text{month}} \times \frac{\text{man month}}{172 \text{ hours}} = 4.25 \text{ men}$$

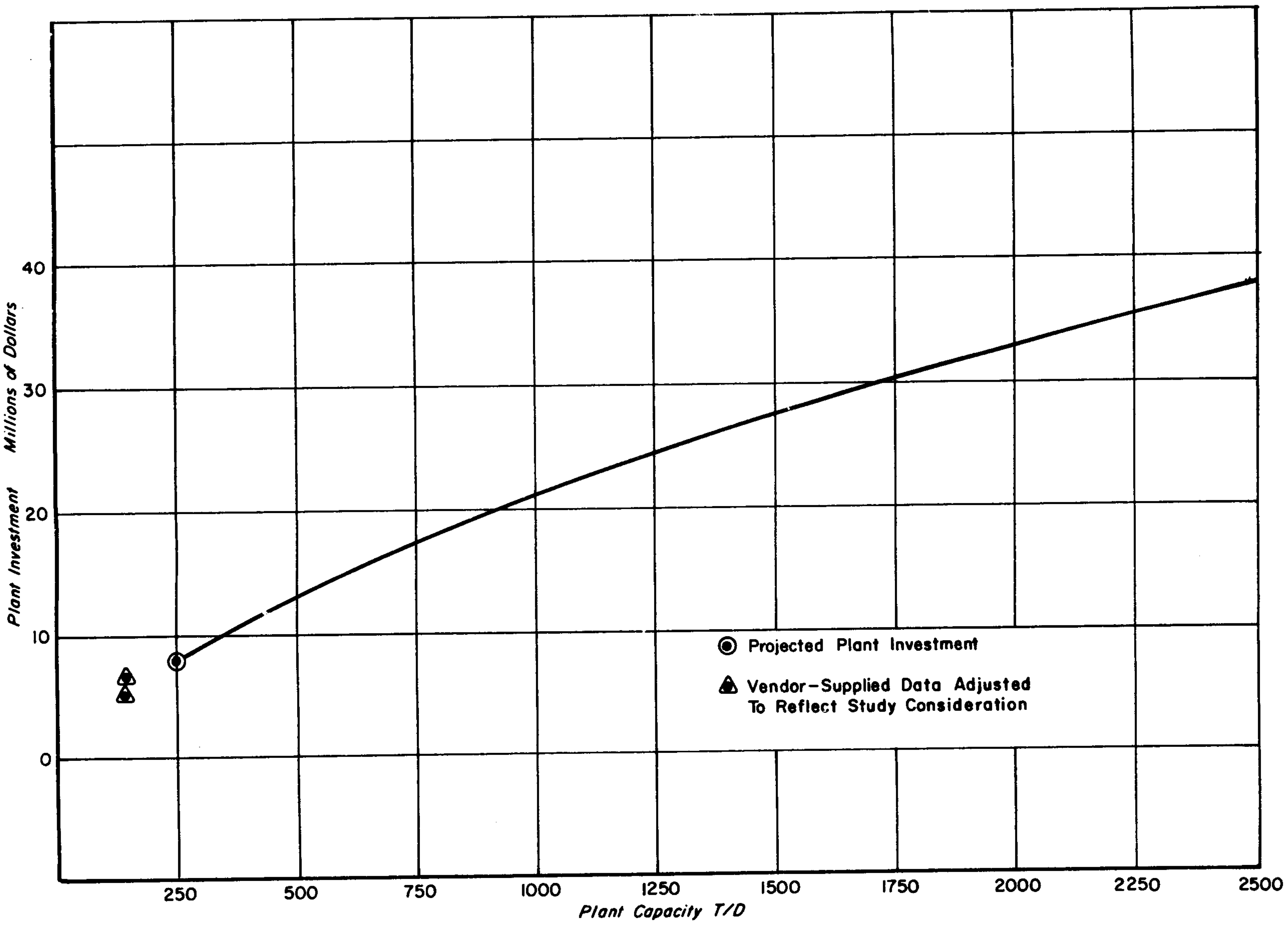


FIGURE 3-2. CAPITAL INVESTMENT vs PLANT SIZE (STEAM REFORMING)

Thus, with nominal overtime, four men are required to staff a one man per shift operation twenty-four hours per day.

At the steam-reforming plant (250 T/D), it was determined that six men would be required. This is equivalent to one man per shift for plant operation and a maintenance pool of one electrical repairman and one maintenance utility man. The total monthly labor cost is based upon each man working an equivalent of 212 straight-time hours at a rate of \$3.90/hr. Fringe benefits are also included at 21 percent of the total wage cost. As the plant capacity increases, the labor force will increase proportionately (at 2500 T/D level, 15 men will be required).

Costs for chemicals, lubricants, and catalysts will increase directly with plant capacity. However, credit has been taken for bulk purchase of these items. Thus, chemicals costs are based on a rate of 0.24 cents per pound of hydrogen produced at the 250 T/D level and 0.19 cents per pound at the 2500 T/D level. Maintenance costs will vary with plant investment rather than capacity, as is common with industrial gas-production units. Maintenance costs for the steam-reforming system are calculated at 4 to 5 percent/year of plant investment. It should be noted that maintenance costs do not include costs associated with major improvement of equipment, which are considered as additional capital investment recoverable through depreciation.

Operations allocations, including home and field office activities, may be either a function of plant labor costs or of plant-level operating costs with 15 year straight-line depreciation of plant equipment. Due to the relatively small cost of labor in the total system economics, the latter method was selected for calculations. Thus, operations allocations are included in the operating costs at 10 percent of straight-line depreciation, labor, chemicals and lubrications, maintenance, and miscellaneous expenses.

General and administration costs are commonly expressed as a percentage of all the costs (calculated as 15 percent in this study) and covers corporate administration, research and development, and other overhead costs. The total for all these costs is \$1.28 x 10^6 per year for a 250 T/D plant. When all the items are compared to plant capacity as a single group, it is found that the relationship is exponential in nature, following the formula

$$O = O_B \left(\frac{C}{C_B} \right)^{M_o}$$

where

O = Operating cost for system capacity (C)
O_B = Operating cost for base system (C_B)
C = Capacity of production and purification system
C_B = Capacity of base production and purification system
M_o = Operating cost exponent

The relationship of plant operating costs with plant size is shown in figure 3-3. It has been determined empirically that the operating cost exponent (M_o) is 0.73.

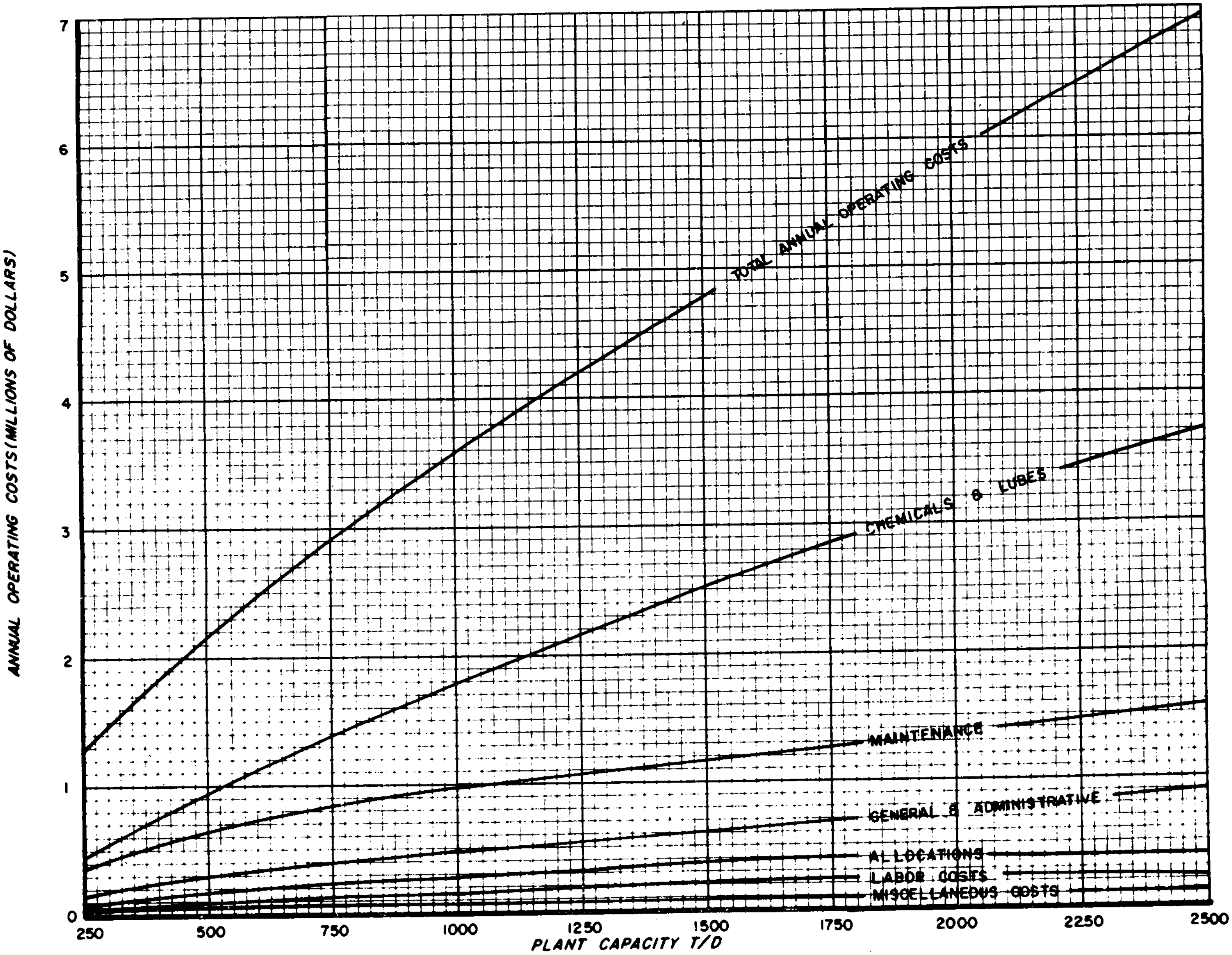


FIGURE 3-3. ESTIMATED OPERATING COST vs PLANT SIZE (STEAM REFORMING)

3.2.2 PARTIAL OXIDATION

The partial oxidation process at present is second only to steam reforming in the quantity of hydrogen produced. Feedstock to partial oxidation plants can range from natural gas to fuel oil. Steam reforming, however, generally proves to be more economical with light hydrocarbons. For this reason, partial oxidation units are usually located where natural gas or other light hydrocarbon feeds are either unavailable or too costly.

There are two major commercial processes utilizing partial oxidation in hydrogen production. These are the Shell Gasification Process, licensed by the Shell Development Company, and the Texaco Process, licensed by the Texaco Development Corporation. It is highly probable that partial oxidation and steam reforming will be the major processes in the post-1980's.

3.2.2.1 Process Description

In a typical partial oxidation process, a crude oil is mixed with steam and fed in carefully controlled quantities to a generator where the fuel reacts with preheated oxygen. This produces a synthesis gas consisting primarily of carbon monoxide and hydrogen.

The reaction takes place in two steps. In the primary exothermic zone of the generator, oxygen reacts with part of the fuel oil to produce carbon dioxide and water vapor. In a secondary endothermic zone, the carbon dioxide and water vapor react with the remainder of the fuel to produce carbon monoxide and hydrogen (synthesis gas). Since the reaction is controlled by limiting the supply of oxygen, up to 3 percent of the fuel may be converted to free-carbon. After combustion, the generated gas is quenched and scrubbed with water to remove the carbon from the system as a slurry. The scrubbed gas then flows to a converter, where the carbon monoxide reacts with additional steam over an iron catalyst to produce additional hydrogen and carbon dioxide. The product gas, hydrogen and carbon dioxide, then passes through an adsorber system in which the carbon dioxide is effectively removed by adsorption on monoethanolamine (MEA). The product gas as shown in figure 3-4 at this point is primarily hydrogen containing only minor impurities, which are subsequently removed by cryogenic adsorption.

The delivery pressure of the hydrogen in this process is 1700 psig. The high pressure operation of the plant eliminates hydrogen compression costs by permitting feed-gas injection at higher pressures. (Feed-gas and oxygen compression can be accomplished more economically than hydrogen compression.)

The carbon removed in the water slurry, after drying, may be recovered as a byproduct or auxiliary fuel. The carbon dioxide adsorbed by the MEA solution is liberated during the regeneration of the amine solution and can be recovered for sale as a byproduct. The quantity of gas emitted from the plant relating to byproduct CO_2 production is described in paragraph 3.2.1.1.

The reaction chemistry of a typical partial oxidation process is as presented:

Primary reaction:

$$2C_nH_m \quad + \quad nO_2 \longrightarrow 2_nCO \quad + \quad mH_2$$

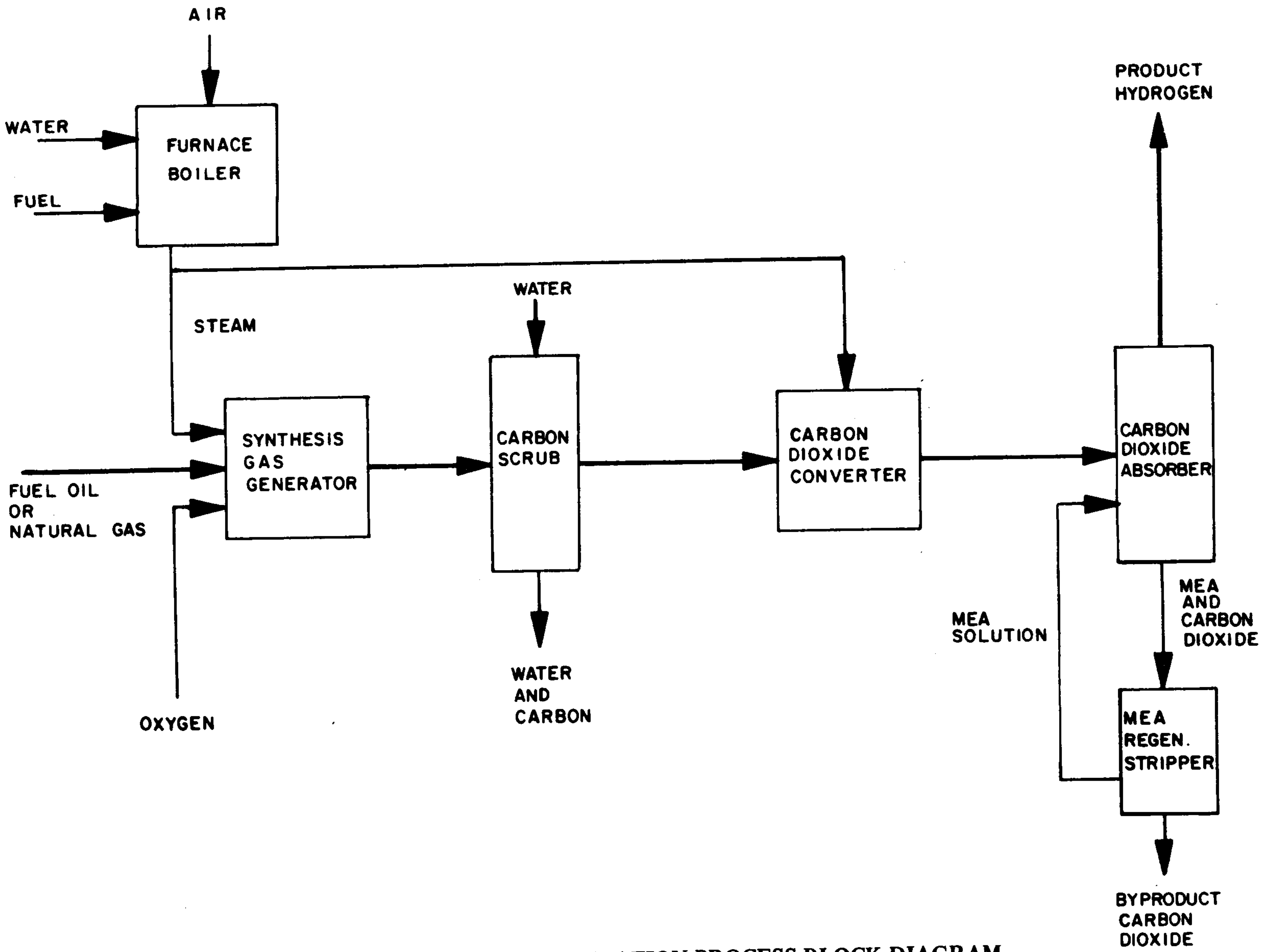


FIGURE 3-4. PARTIAL OXIDATION PROCESS BLOCK DIAGRAM

Shift reaction:

$$CO + H_2O \longrightarrow CO_2 + H_2$$

The removal of carbon dioxide is accomplished by adsorption of the gas in an amine solution such as aqueous MEA through the reactions

$$CO_2 + 2HOCH_2CH_2NH_2 + H_2O \longrightarrow (HOCH_2CH_2NH_3)_2CO_3$$

and

$$CO_2 + HOCH_2CH_2NH_2 + H_2O \longrightarrow HOCH_2CH_2NH_3HCO_3$$

The amine scrub solution is regenerated by heat addition, causing reactions to reverse and releasing the carbon dioxide.

3.2.2.2 Process Elements

Process studies have been completed which show that linear relationships exist between the various process elements and plant capacity. These relationships are discussed in the following paragraphs.

3.2.2.2.1 Raw Material Unit Consumption. – The maximum output of a synthesis gas generator varies directly but not proportionally with the size of the combustion chamber. Since the reactor heat loss is nearly independent of throughput for a given vessel and increases less rapidly with generator size than with the number of vessels, it is advantageous to meet plant needs with only a few generators operating at full capacity. Multiple units are also advantageous where large variations in production rates are expected and where high on-stream factors are of paramount importance. For this reason (i.e., several units operating at maximum capacity in parallel), little reduction in unit consumption of feedstock per unit of hydrogen produced is expected.[11]

The hot synthesis gas leaving the generator must be cooled before entering the shift converter. This cooling is accomplished in small plants by indirect heat transfer in a waste heat-boiler. In large-tonnage plants such as those contemplated in this study, direct quenching with hot water will effect a more economical recovery of the enthalpy of the synthesis gas while simultaneously increasing the steam content of the gas prior to entry into the shift converter.[12] The entire plant efficiency increases with capacity and results in a decrease in fuel requirements for steam production. It is projected that a reduction in fuel requirements on a unit basis with hydrogen production of from 15 to 20 percent can be expected as the amount approaches 2500 T/D.

Cooling water and boiler feed-water requirements will remain constant on a unit basis as plant size increases for reasons discussed.

As discussed in paragraph 3.2.1.2.1, the relation of raw material requirements to plant capacity can be expressed as

$$R_i = K_{R_i} + M_{R_i}C$$

Estimated values for K_{R_i} and M_{R_i} for the various raw materials consumed in the partial oxidation process are listed in table 3-3.

TABLE 3-3. VALUES OF RAW MATERIALS

Material	K_{R_i}	M_{R_i}
Feed (Bunker "C" oil)	3.89 lb/lb H_2	0
Fuel (Bunker "C" oil)	.40 lb/lb H_2	-3.1×10^{-5} lb/lb T/D
Cooling Water makeup	12.5 lb/lb H_2	0
Boiler Water makeup	12.5 lb/lb H_2	0

3.2.2.2.2 Process Energy Consumption. – Process energy for the partial oxidation process is required for feedstock compression and purification-train-pumping requirements (amines, pumps, etc.). By using multiple production trains, pumping requirements can be optimized for each train, ensuring high-efficiency pumping equipment. The energy requirements per unit of hydrogen produced, therefore, will remain constant. Feedstock compression energy requirements will remain nearly constant per unit of production for the reasons presented in paragraph 3.2.2.2.1. The process energy unit consumption is 1.3 kwh/lb of H_2.

3.2.2.2.3 Byproduct Unit Production. – The removal of CO_2 generated in the hydrogen production system is accomplished through the use of an MEA solution as described for the steam reformer process. Similar large quantities of CO_2 will be available either for resale or disposal.

Unconverted carbon from the liquid-fed synthesis gas generator is scrubbed from the gas and discharged as a water slurry. This carbon may be pumped to disposal or may be converted to relatively dry pellets suitable for burning, and the water in the slurry can be returned to the system for reuse.

The quantity of byproduct production for carbon dioxide is 6 lb/lb H_2; for export steam 5 lb/lb H_2; and for carbon 0.11 lb/lb H_2.

3.2.2.3 Cost Elements

Capital investment and operating costs, other than for raw materials and energy as previously discussed, have been studied over for plants producing 250 T/D to 2500 T/D of gaseous hydrogen via the partial oxidation process. Capital investment and operating costs were found to vary exponentially with plant size, as described in the following paragraphs.

3.2.2.3.1 Capital Investment. – Based upon vendor information about partial oxidation units ranging from 125 to 400 T/D capacity,[13] capital investment projections have been developed for plants having a production capacity of up to 2500 T/D. Extensive use was made of the Williams equation.[5] Here again, it was determined that the 0.7 scale best fits the available data. Relationship of capital investment to plant size can be expressed as

$$I = I_B \left(\frac{C}{C_B}\right)^{M_I}$$

A partial oxidation unit capable of producing 130 T/D (50 MM scf/day) of gaseous hydrogen at a purity of 98 percent or better, with a discharge pressure of 1700 psig, has been estimated by an engineering construction firm as costing 7.85 million dollars. A similar unit capable of 390 T/D (150 MM scf/day) was valued at 14.9 million dollars. The relationship of partial oxidation plant investment with plant size is shown in figure 3-5. The estimated capital investment for the base system (250 T/D) is $\$8.46 \times 10^6$, which compares favorably with the data discussed above and included in figure 3-5.

3.2.2.3.2 Operating Costs. – Operating costs described in the following paragraphs exclude those associated with raw materials and process energy. Taxes and insurance are also excluded.

The labor pool for a 250 T/D partial oxidation plant is estimated to be nine men. This is equivalent to one man per shift for plant operation (eight-hour shifts, continuous operation) and a maintenance pool of five men for electrical and mechanical repairs.

Chemical, lubricant, and catalyst requirements will increase directly with plant capacity. Credit has been taken (as for the steam reforming process) for bulk purchase of these items as plant capacity increases. Chemical costs are based on a rate of 0.29 cents/lb of hydrogen produced at the 250 T/D level and 0.22 cents/lb at the 2500 T/D level.

Maintenance, general and administrative, and operations costs are calculated in the same manner as described in paragraph 3.2.1.4.2.

Operating costs for the partial oxidation process can be expressed as a function of the plant capacity as follows:

$$O = O_B \left(\frac{C}{C_B}\right)^{M_o}$$

The value for O_B is $\$1.46 \times 10^6$ per year for the 250 T/D plant and the operating cost exponent (M_o) is 0.73. The relationship of plant operating costs with plant size is shown in figure 3-6.

3.2.3 WATER ELECTROLYSIS.

Present commercial water-electrolytic processes are too expensive to be considered as sources of large quantities of gaseous hydrogen. Recent studies have resulted in modified processes that show promise of water electrolysis becoming economically competitive with steam reforming and partial oxidation within the next decade. One modified water-electrolytic process operates at elevated pressure and temperature (5000 psi and 250°F), resulting in significantly lower unit power consumption. Capital costs are correspondingly greater . A second modified process is currently being developed incorporating porous electrodes. This process requires the same amount of power as the high temperature-high pressure process. The porous electrode feature results in satisfactory operation at lower pressures (300 psia, 250°F) and offers the best potential for economic commercial application. The modified processes are described in the following paragraphs.

3.2.3.1 Process Description

The water-electrolytic process is described in a generally technical matter. The modified processes

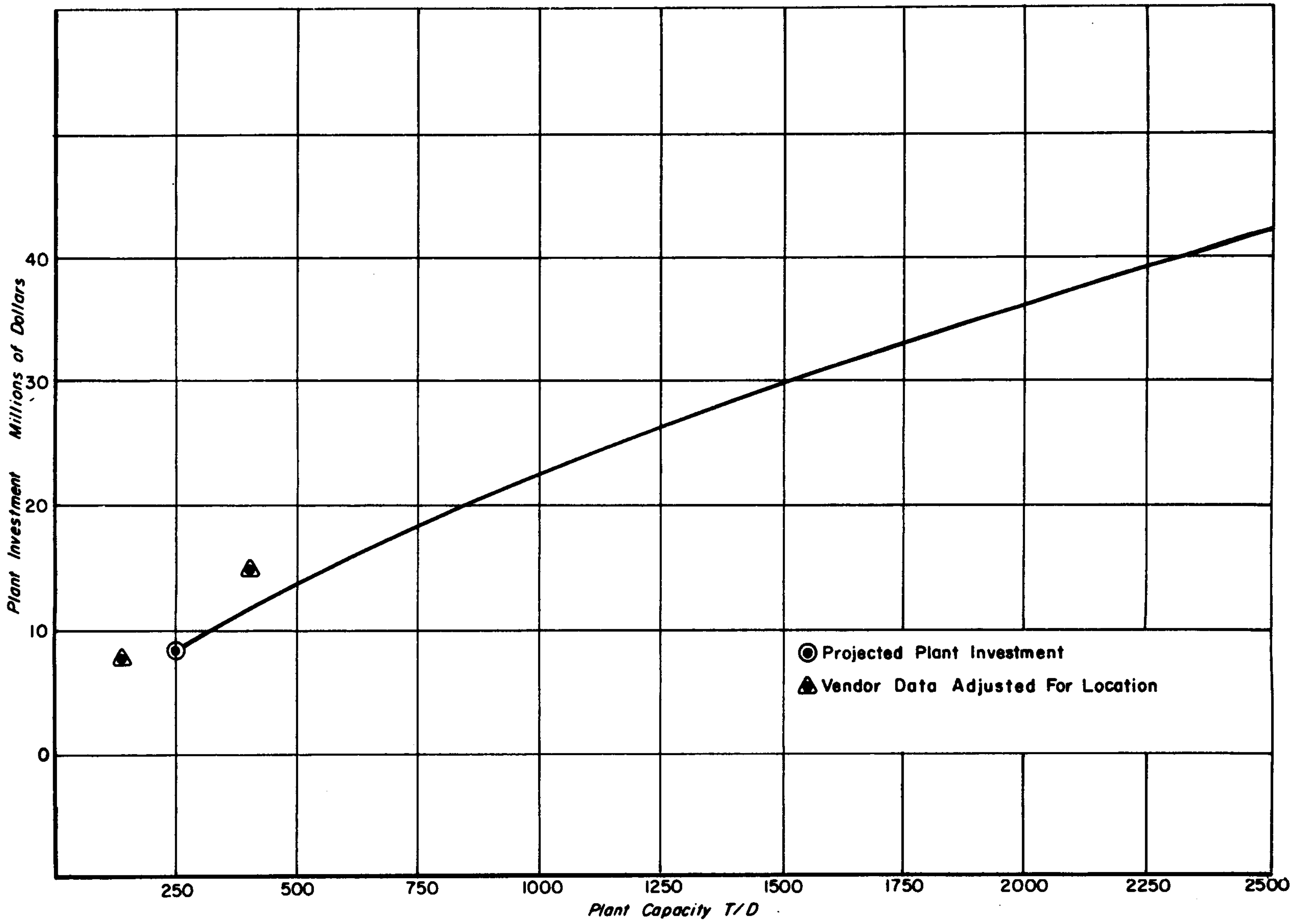


FIGURE 3-5. ESTIMATED INVESTMENT vs PLANT SIZE (PARTIAL OXIDATION)

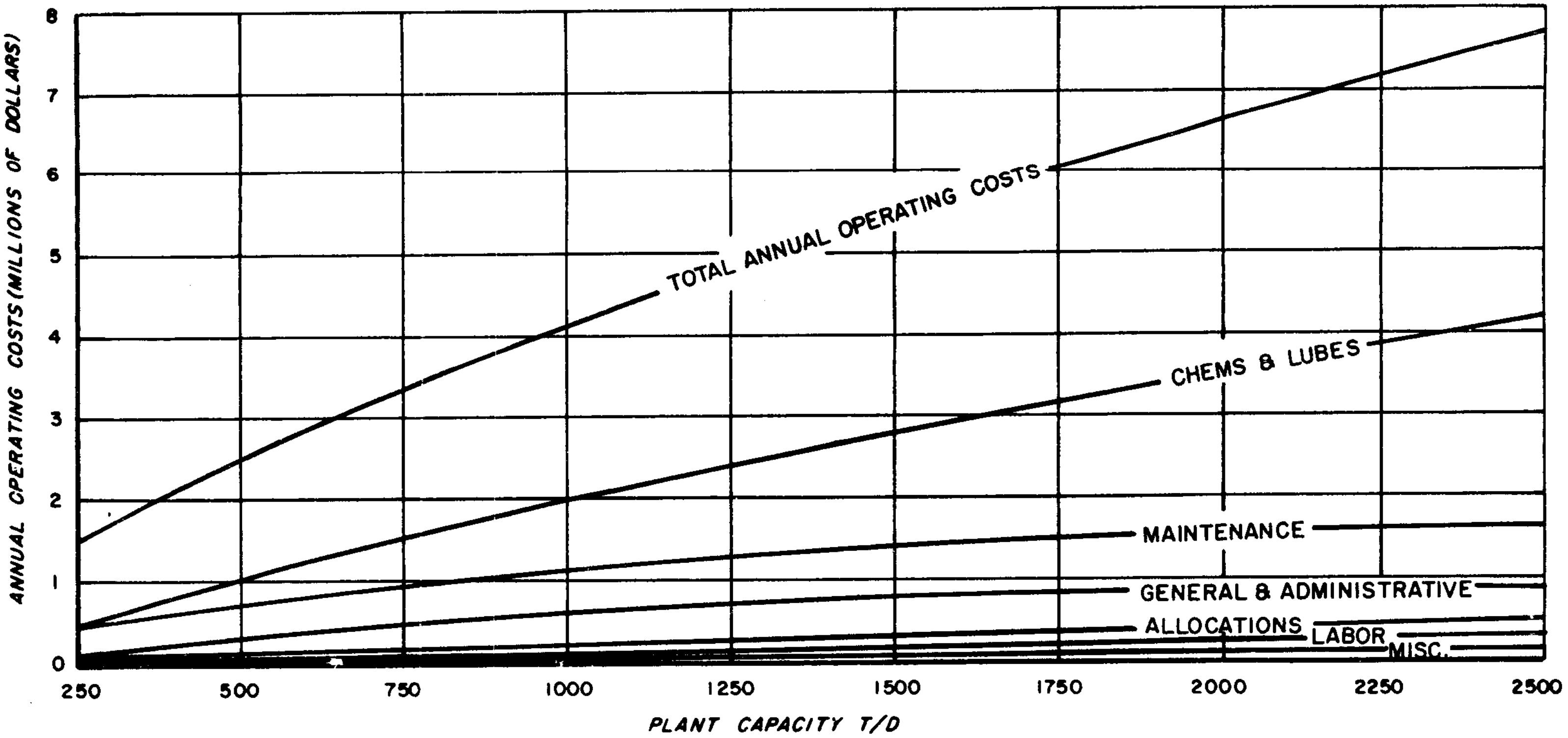


FIGURE 3-6. ESTIMATED COST vs PLANT SIZE (PARTIAL OXIDATION)

(high pressure-high temperature and low pressure-high temperature) are discussed in greater detail in paragraphs 3.2.3.1.2 and 3.2.3.1.3 respectively. Electrolysis of water flow diagram is shown in figure 3-7.

3.2.3.1.1 General Technical Considerations. – Water electrolysis is the term associated with electrolytic reactions in which water dissociates into hydrogen and oxygen. The reaction takes place in a pressure vessel known as an electrolytic cell, which physically contains electrodes. Fresh water feeds into the cell and surrounds the electrodes. By passing current through the electrodes, water is dissociated into hydrogen and oxygen. Direct current flow is such that gaseous hydrogen collects at the cathodic electrodes and gaseous oxygen collects at the anodic electrodes. Protruding into the cell are several sets of electrodes. The fluid in the cell is generally an aqueous solution of a suitable electrolytic salt (potassium hydroxide, for example). The presence of the electrolytic salt greatly enhances the electrical characteristics of the fresh feed-water. Fresh feed-water constantly feeds to the cell and serves as "makeup" as the dissociated water is removed from the cell as gases.

The theoretical energy requirement for electrolytic dissociation of water can be determined by reference to Faraday's Laws. Considering the electrolytic cell, local electrolytic reactions occur in the proximity of the various electrodes. H^+ ions present in the electrolyte migrate toward the cathodes in the cell's electric field. At the cathode, the H^+ ions combine with electrons and form molecular hydrogen. Symbolically, the local electrolytic reaction is

$$2H^+ \quad + \quad 2\bar{e} \longrightarrow (H_2) \text{ gas}$$

Similarly, OH^- ions in the electrolyte tend toward the anodes. At the anodes, the OH^- ions give up electrons and form water and molecular oxygen. This local electrolytic reaction is symbolized as

$$2OH^- \longrightarrow 2\bar{e} \quad + \quad H_2O \quad + \quad (1/2\ O_2) \text{ gas}$$

The energy requirement of 76.2 kwh/1000 cubic feet of hydrogen (60°F and 1 atm) can be obtained from the following equation:

$$Q = \frac{M v \mathcal{F}}{A} \tag{1}$$

where

Q = unit energy requirement for liberation of the element
M = the mass of the element liberated
A = the atomic weight of the element
v = the valence of the element
$\mathcal{F}$ = Faraday's constant $\left(\frac{\text{26.8 amp-hrs.}}{\text{gm-equiv.}}\right)$

For hydrogen liberation, the above factors have the following value:

M = 5.26 lb (1000 cu. ft. H_2 @ 60°F and 1 atm)
A = 1.008 gm/gm equivalent
v = 1
$\mathcal{F}$ = 26.8 amp-hrs/gm equivalent

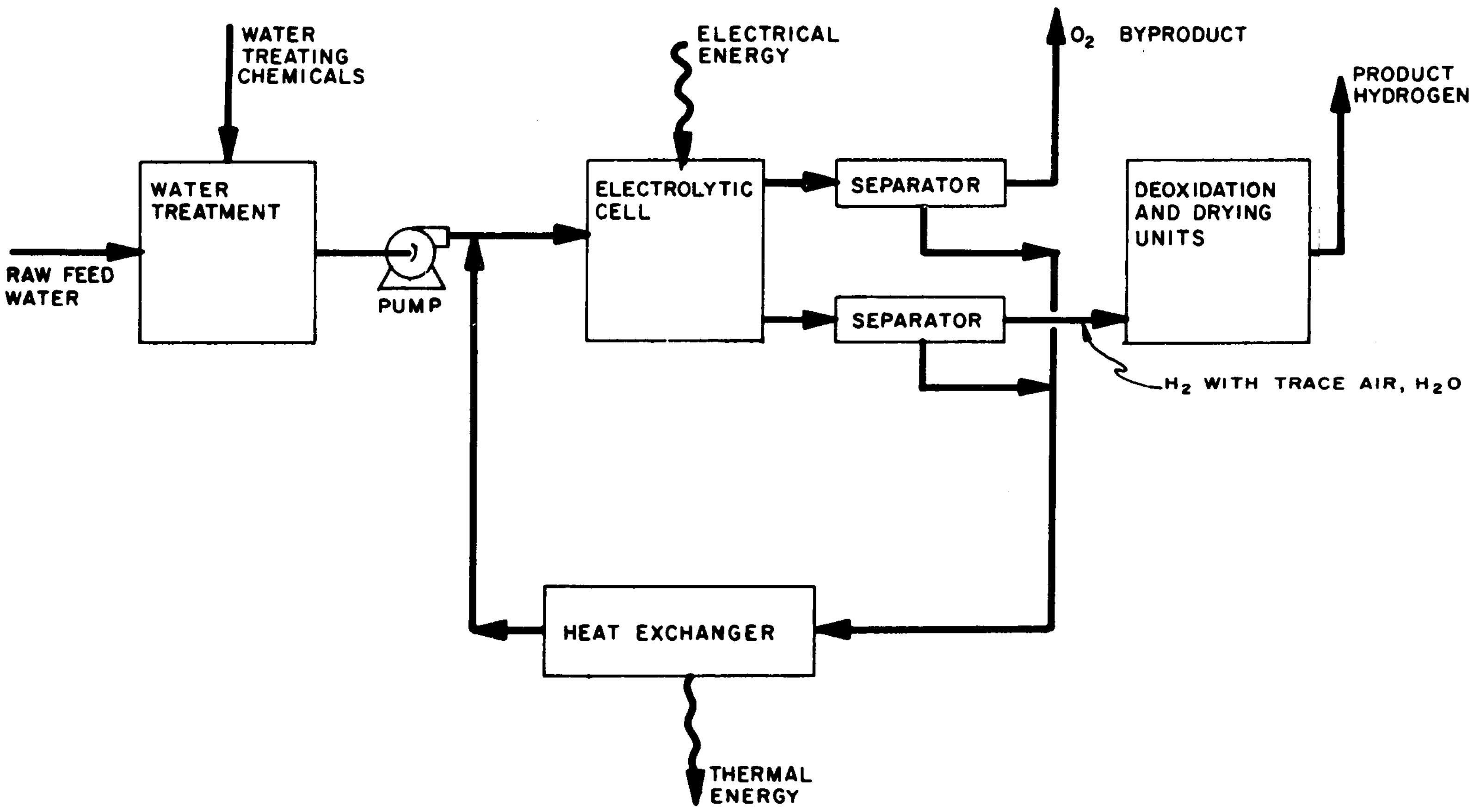


FIGURE 3-7. ELECTROLYSIS OF WATER PROCESS BLOCK DIAGRAM

Therefore, the equation becomes

$$Q = \frac{5.26 \text{ lb} \times 26.8 \frac{\text{amp-hrs}}{\text{gm equiv}} \times 454 \frac{\text{gm}}{\text{lb}}}{\frac{1.008 \text{ gm}}{\text{gm equiv}} \times 103 \text{ cu ft}}$$

$$= \frac{63{,}500 \text{ amp-hrs}}{\text{gm equiv}} \qquad (2)$$

Thus, for a cell potential of 1.20 volts, the power requirement is

$$\frac{63{,}500 \text{ amp-hrs}}{1000 \text{ cu ft}} \times 1.20 \text{ volts} = \frac{76.2 \text{ kwh}}{1000 \text{ cu ft}} \qquad (3)$$

Water-electrolytic cell potential under thermodynamic equilibrium conditions of 25°C and one atmosphere pressure is 1.229 volts. This value of potential is calculated from the standard free energy change required for the formation of hydrogen and oxygen from liquid water. In general practice, cell voltage is taken as 1.2 volts with the usual operating temperature range of 60°C to 80°C. This corresponds to a theoretical energy requirement of 76.2 kwh per thousand cubic feet (60°F in one atmosphere) with the cell operating at one atmosphere pressure.

Operating cells function with electrical currents essentially at the predicted theoretical levels. Actual cell voltage, however, is higher than that predicted for thermodynamic equilibrium conditions. Voltage drops due to electrolyte resistance and polarization effects cause the overvoltages of operating cells. This excess cell potential, termed hydrogen and oxygen overvoltages, results in power efficiencies being significantly below 100 percent based on thermodynamic equilibrium conditions. For example, a commercial production system requires cell voltages on the order of 2.3 volts to generate hydrogen at atmospheric pressure with a 60°C to 80°C operating temperature range. This, in turn, corresponds to an actual energy requirement of 147 kwh/1000 scf (60°F, 1 atm). Comparing this set of conditions with the theoretical requirement, a power efficiency of approximately 50 percent is noted.

3.2.3.1.2 High Pressure-High Temperature Electrolysis. – With reference to the foregoing discussion it is noted that cell overvoltage decreases as the process temperature rises. Temperature rise is associated with increasing vapor concentration of water in the product gas. Further, vapor bubbles rising in the electrolyte solution contribute to the solution resistance and tend to offset the reduced overvoltage. As a result, 80°C is projected as an optimum temperature for electrolytic operation at one atmosphere.

Pressure variation affects the cell voltage characteristics. As the pressure of the water-electrolytic system is increased the theoretical cell voltage also increases, but actual operating cell voltages decrease. The decrease in actual cell voltage requirements results from the decreasing gas bubble size related to increased pressure operation. At a cell pressure of 30 atmospheres, for example, theoretical cell voltage is 1.3 volts. This corresponds to a theoretical energy requirement of 82.5 kwh/1000 cu.ft. Commercial units operating with a 30-atmosphere pressure maintain a cell voltage of 2.0 volts, corresponding to an actual energy requirement of 128 kwh/1000 cu ft. By comparison, the conditions associated with the 340-atmosphere operating pressure are 1.35 volts theoretical voltage, corresponding to 1.78 volts actual operating voltage. Energy requirements correspondingly are 86 kwh/1000 cu ft and 113 kwh/1000 scf.

Besides the lower energy requirements for electrolysis at high pressure, the elimination of compression equipment for product hydrogen provides an additional advantage. For a cell operating at atmospheric pressure, an additional 7.5 kwh/1000 cu ft is required for compression to 5000 psia. Similarly, for the three-atmosphere cell, an additional 3.1 kwh/1000 scf are needed. Data associated with the foregoing discussion is presented in figure 3-8 and table 3-4.

As described, definite advantages exist for operating electrolytic cells at elevated pressures and temperatures. A major advantage is noted in that reduced energy requirements for the dissociation reactions are associated with the increased temperature and pressure level operation. Further, the high-pressure level of the product gases from the cell may prove advantageous in further processing of the product gases. Economic considerations will ultimately determine the most advantageous pressure level for cell operation.

3.2.3.1.3 Low Pressure-High Temperature. – Recent research in fuel-cell technology has resulted in significant advancement in the general state of the art of electrolytic cells. Reduced electrical polarization, improved mechanical cell design, and lower electrical resistive losses are some accomplishments. Further, the use of porous electrodes has been investigated and actual applications developed for fuel-cell construction. Electrode porosity allows reactive gases to pass through the backside of the electrodes to the reaction zone. A similar utilization of porous electrodes in a water-electrolytic cell is possible. It is projected that application of the porous electrode to a water-electrolytic cell will permit operation at substantially lower pressures. Allis-Chalmers, General Electric, and the Allison Division of General Motors are typical of U. S. corporations presently involved in application of fuel-cell technology to water electrolytic systems. Information presented herein on the porous electrode cell has been obtained from reports and articles published by these firms as well as direct communications with Allis-Chalmers personnel.[14,15]

The general technical information related to the porous electrolytic cell is essentially the same as that presented for the high pressure-high temperature cell. The basic difference between the physical characteristics of the cells lies in electrode construction. Electrodes of the high pressure-high temperature cell are solid, that is, non-porous. Product gas is evolved as bubbles adjacent to the solid electrodes. The bubbles are non-conductive and, therefore, increase internal electrical resistance of the cell. High cell operating pressure is required to minimize the effect of this bubble formation and, consequently, reduce the cell power requirements.

An alternate means of reducing the bubble effect without the use of high pressures is the application of porous electrodes. As the gases evolve, they may pass through the porous electrode and out the backside. This eliminates that part of the internal cell resistance attributed to the bubble formation. The porous electrode thus has two main advantages:

(1) Cell power requirement is significantly reduced to the level associated with the high pressure-high temperature cell.

(2) The cell can in fact be operated at relatively low pressures.

The actual operating pressure must, however, be high enough to prevent the electrolyte from boiling . Cell pressure is not required for suppression of bubble size because of the porous electrode construction.

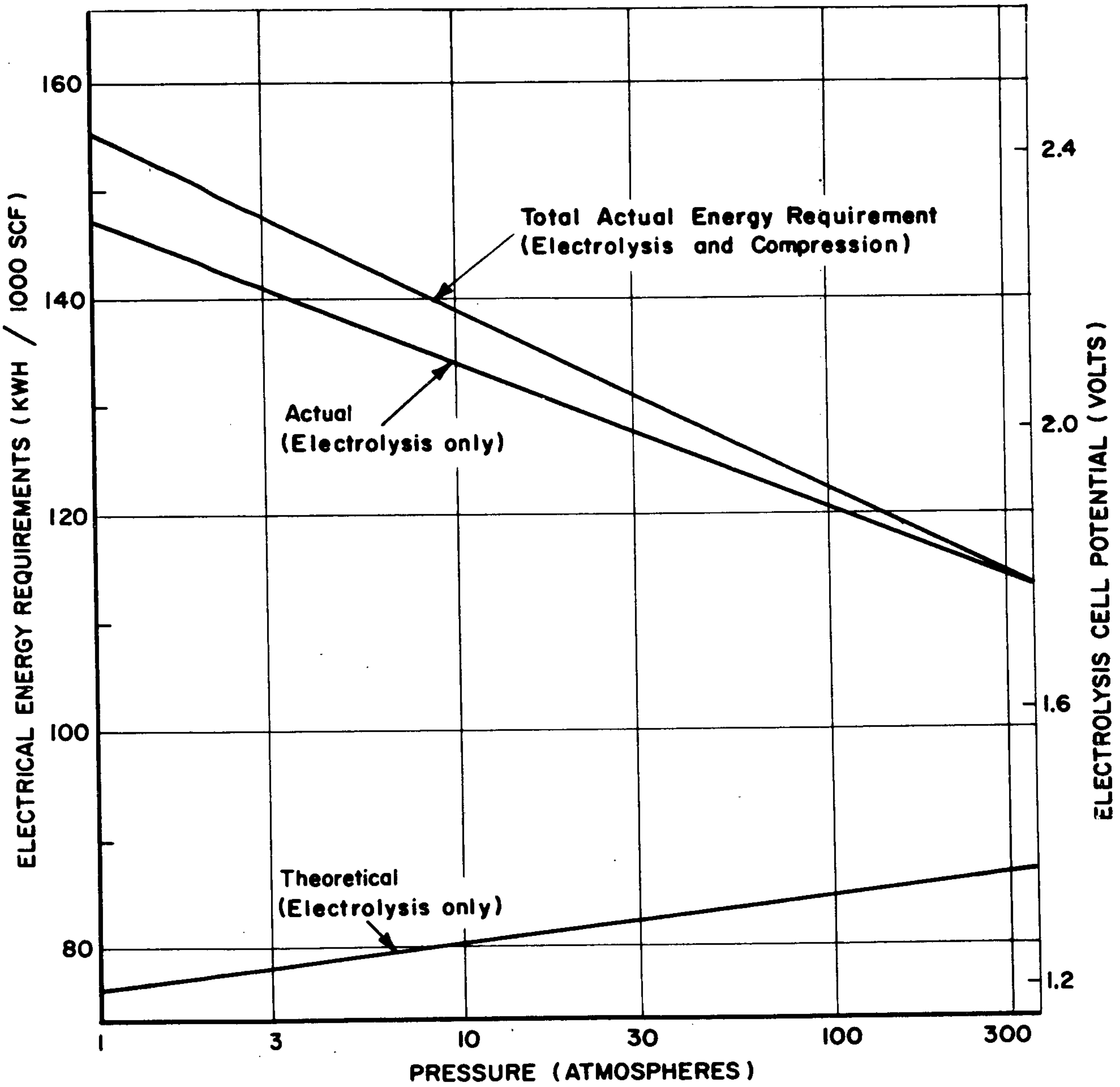


FIGURE 3-8. ELECTRICAL ENERGY AND ELECTROLYTIC CELL POTENTIAL AS A PRESSURE FUNCTION

TABLE 3-4. ELECTRICAL ENERGY REQUIREMENTS FOR HYDROGEN PRODUCTION BY WATER ELECTROLYSIS AS A FUNCTION OF PRESSURE

Pressure	Voltage (Volts) Theoretical	Voltage (Volts) Actual	Electrical Energy (kwh/1000 scfh) Theorectical	Electrical Energy (kwh/1000 scfh) Actual	Compression Energy[a] (kwh/1000 scf) Actual	Total Energy (kwh/1000 scf) Actual	% of 5000 psia case
14.7 (1 atm)	1.2	2.3	76.2	147	7.5	155	137
450 psia (30 atm)	1.3	2.0	82.5	128	3.1	131	116
5000 psia	1.35	1.78	86.0	113	0	113	100

[a] Compression energy from cell working pressure to 5000 psia.

3.2.3.1.4 General Process Description. – Purified water serves as a feedstock to the electrolytic cell. Source of the water will determine the complexity of the purification process. The water purification, however, is taken as an auxiliary process and not considered a main operation within the electrolytic process. Following purification, the feedstock is pumped to a pressure level corresponding to the specific cell-operating pressure. Dissociation of the water into gases occurs in the areas of cathodes and anodes. In the solid-electrode cell, the separated gases are "collected" and removed for further processing. Porous electrodes, however, result in the cell products being two-phase mixtures of separated gases and electrolyte. The two-phase mixtures subsequently are separated into vapor and liquid, the liquid electrolyte being recycled to the cell.

H_2 gas product from the cells contains small amounts of impurities, primarily air and water vapor. Oxygen and water are removed from the H_2 gas as it passes through a catalytic deoxidation reactor and a water adsorption bed. At this point, the product H_2 gas contains only trace amounts of impurities (nitrogen, for example) that are removed in further processing (liquefaction system).

Byproduct O_2 gas is processed in a manner similar to the H_2 gas. The level of impurity removal required is dependent upon the end use of the byproduct. Schematic representation of the impurity removal and other process description is shown in figure 3-7.

Information presented is generally applicable to the two modified water electrolytic processes considered in this report. Particular points related to these processes are as follows.

Increased cell-operating temperature results in lower power requirements. However, the higher temperatures adversely affect certain other cell characteristics. Rates of metal corrosion and other types of general material deterioration are increased as temperature levels rise. In view of this, 250°F is projected as the maximum operating temperature level.

Several acids and alkaline materials could be used in aqueous solution to form the electrolyte. Potassium hydroxide is considered a suitable electrolyte in this study, since it exhibits relatively low unit resistivity. Further, this electrolyte serves as a heat-transfer medium.

As indicated, the high-pressure cell operates at 5000 psia, with the low-pressure unit operating at 300 psia. The 5000-psia level is chosen for the solid electrode cells to provide for low unit-power requirements. The porous electrode unit operates with essentially the same power consumption characteristics over a wide range of cell pressures, namely one to several hundred atmospheres. A practical pressure level of 300 psia has been chosen to prevent the electrolyte from evaporating and to maintain reasonable volume (density) of generated gas. Higher pressure levels for the porous electrode system could be chosen based on economic evaluation of cell design and gas compression requirements for subsequent processing.

For this study, a cell-current density of 800 amperes per square foot (ASF) was selected. This current density has been attained with the porous electrodes but not with conventional electrodes. Present commercial cells employ maximum current densities of 200 ASF. However, Allis-Chalmers personnel have been able to attain current densities of 1600 ASF at 250°F for extended periods of time using the porous electrodes with laboratory scale equipment. Current densities as high as 4000 ASF have been achieved for short periods of time. However, at these high current densities, heat removal from the cell becomes critical (i.e., the electrolyte is not able to remove all the thermal energy generated as a result of the cell inefficiencies). The value of 800 ASF, assumed for both types of cells, represents a compromise choice. The effect of this choice is apparent in mechanical design of the cell.

Future development will have great effect on this mecahnical design, but significant projections of cell development cannot be made at this time.

3.2.3.2 Process Elements

A direct proportionality relationship exists between the process elements and plant capacity for the water-electrolytic process.

The units and quantities required of the various process elements associated with the electrolysis of water are as follows:

(1)	Raw materials (both processes):	
	Feed water	9 lb/lb H_2
	Cooling water	11.5 lb/lb
	Potassium hydroxide	neg.
(2)	Process energy at 800 ASF:	
	High pressure-high temperature	21.5 kwh/lb H_2
	Low pressure-high temperature	21.6 kwh/lb H_2
(3)	Byproduct oxygen:	
	Both processes	7.92 lb/lb H_2

3.2.3.3 Cost Elements

Capital investment and operating costs, other than raw materials and process energy, have been studied over the selected capacity range of 250 T/D to 2500 T/D of gaseous hydrogen production.

3.2.3.3.1 Capital Investment – The capital investment data (250 T/D plant) for the high pressure-high temperature process were developed as part of this study. The investment for the low pressure-high temperature (porous electrode) process was predicated on previous work by Allis-Chalmers.

As plant capacity increases from 250 to 2500 T/D, the capital cost is expected to increase in an exponential manner as described by the Williams equation.[5] Thus, the relationship of capital investment to plant size can be expressed as

$$I = I_B \left(\frac{C}{C_B} \right)^{M_I}$$

The estimate prepared for this study indicates that a 250 T/D system utilizing the high pressure-high temperature process requires a \$36.3 x 10^6 investment. The investment consists of the following major cost areas:

(1)	Electrical equipment	\$10,000,000
(2)	Electrolytic cells	14,460,000
(3)	Process Equipment	4,150,000
(4)	Construction	7,690,000
(5)	Total investment	\$36,300,000

The electrical equipment and electrolytic-cell costs, on a per pound of hydrogen basis, are fairly insensitive to cost reduction as plant capacity increases due to their modular nature. Construction costs are only slightly less insensitive to plant capacity increase. Due to the high-cost factor in these three areas, the value of the exponential constant (M_I) is estimated to be 0.89 for the high pressure-high temperature process.

The capital investment (I_B) for the low pressure-high temperature process is estimated to be \$28.5 x 10^6 and the exponential constant is 0.88. The relationship of plant investment to plant capacity is shown in figure 3-9.

3.2.3.3.2 Operating Costs – Operating costs described in this section exclude those associated with raw materials and process energy. Taxes and insurance are also excluded.

The labor force requirements for the base case facility (250 T/D) indicate a need for a ten-man crew. This is equivalent to one man per shift for plant operation, one man per shift for electrical repairs and maintenance and two maintenance-utility men. The labor requirements will increase with plant capacity; at the 2500 T/D level sixteen men should be required. As in the previous processes studied, the relationship of labor to capacity exists because plant expansion will be accomplished through the parallel operating train concept. These men will also be part of a large labor pool because the gaseous production plant is in combination with a liquefaction plant.

With all processes considered thus far in this study, chemicals, lubricants, and catalysts requirement will increase directly with plant capacity. Credit has been taken for bulk purchase of these items as plant capacity increases. Therefore, chemical costs range from 0.20 cents/lb at the 250 T/D level to 0.14 cents/lb at the 2500 T/D level. Maintenance costs for both processes were calculated at a rate of 3.5 to 5.5 percent/year of plant investment.

Operations allocations and general and administrative costs were calculated in the same manner described in paragraph 3.1.1.4.2.

When all of the items are compared to plant capacity as a single group, it is found that the relationship is exponential in nature, following the form

$$O = O_B \left(\frac{C}{C_B}\right)^{M_O}$$

where the value for O_B is \$3.26 x 10^6 per year for a 250 T/D high pressure-high temperature plant, and the operating cost exponent M_O is 0.73.

The estimated operating costs associated with the low pressure process are \$2.62 x 10^6 per year with a cost exponent of 0.73.

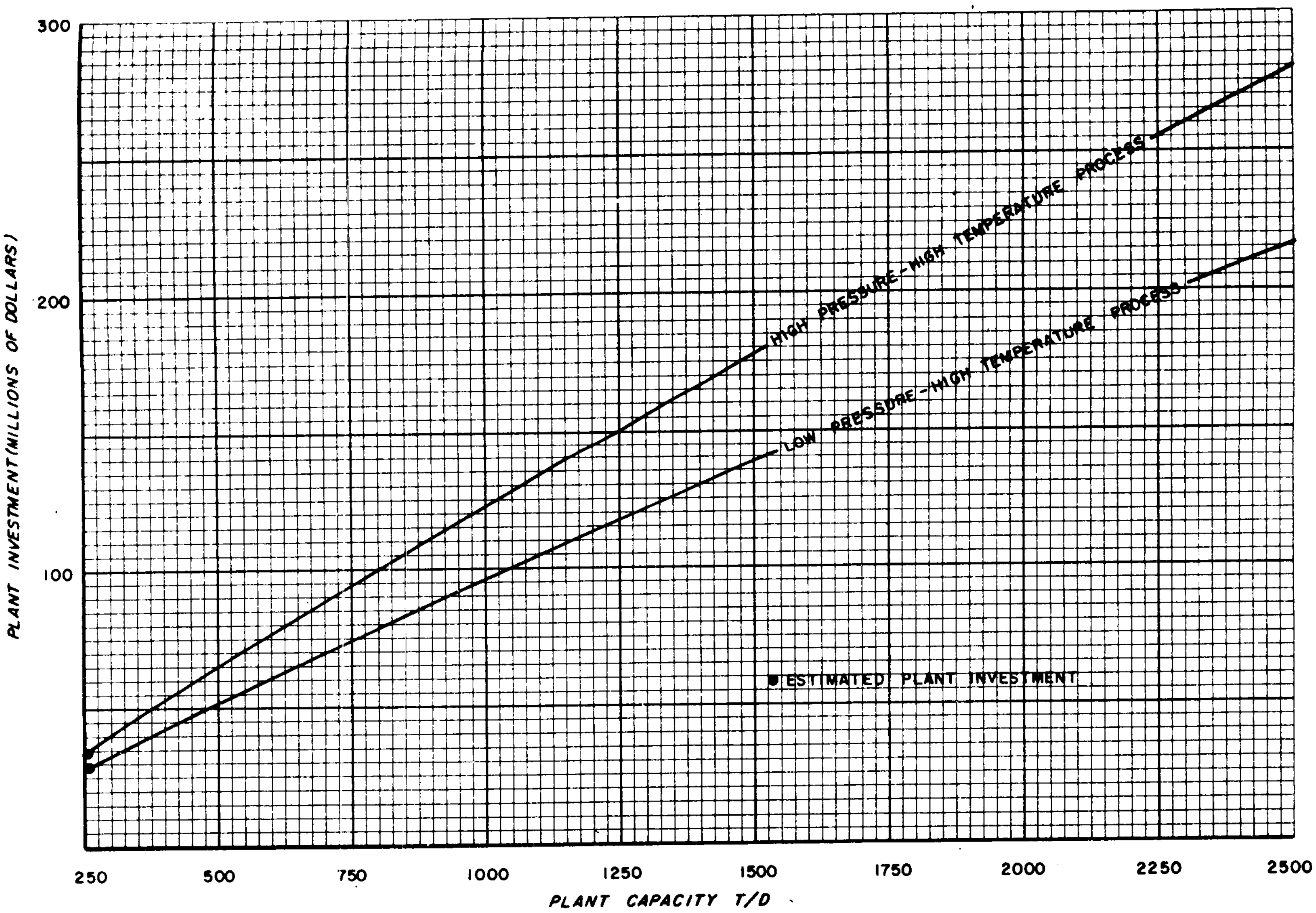


FIGURE 3-9. ESTIMATED INVESTMENT vs PLANT SIZE (WATER ELECTROLYSIS)

Representations of the relationship of operating costs to plant capacity are shown in figures 3-10 and 3-11.

3.2.4 COAL GASIFICATION

Coal must be considered a raw material for processing in future energy systems. Various coal gasification processes in recent years have been studied and definite feasibilities of large scale gaseous hydrogen production have been established. In general, the coal gasification processes have been based on reacting steam and coal at elevated temperatures. These reactions, being endothermic, require large amounts of thermal energy input. This thermal energy is derived from an exothermic combusion process (such as coal and oxygen, for example). The requirement for oxygen results in heavy economic burdens upon these coal gasification processes.

More recently, emphasis has been placed upon development of alternate means of supplying the endothermic heat of reaction. Suitable thermal energy supplies have been sought to economically circumvent the need for a high purity oxygen supply. This has led to a group of systems known as carbon dioxide accepter processes. These processes basically involve an exothermic reaction between CO_2 and an accepter material. Reaction kinetics can be established so that accepter process exothermic heat approximately equals endothermic heat of the coal gasification reaction.

3.2.4.1 Process Description

As indicated, much work has been done relating to the utilization of coal in energy systems.[16 thru 24] In particular, two general types of processes have been considered for this study. Information pertaining to the respective processes is presented in the following paragraphs.

3.2.4.1.1 Steam-Oxygen Process Description – Generation of gaseous hydrogen from coal is based upon an endothermic reformer type reaction. The endothermic heat of reaction is typically supplied by a combustion process of coal and high purity gaseous oxygen. Two basic types of coal are utilized in these reactions, namely, lignite and bituminous. Qualitatively, lignite is a more "reactive" material than bituminous. Further, the bituminous varieties contain volatile hydrocarbons. The volatile hydrocarbons must be removed from the bituminous coal prior to processing for gaseous hydrogen generation. This prevents agglomeration of the coal in the gasification reaction. Both lignite and bituminous coals are being considered in this study. Processing of these materials is described in the following paragraphs.

Crude feedstock coal is pulverized to a prescribed mesh size prior to actual processing. Further drying and removing of volatile hydrocarbons, if necessary, is completed as the feedstock coal is prepared for processing. When the feedstock is properly conditioned, it feeds into a pressurized reactor vessel and forms a fluidized bed. Fluidization results from reactant and product gases flowing upward through the coal. Reaction occurs in the fluidized bed with waste coal solids (ash) falling out at the bed bottom. Reactant gases fed into the bed bottom are steam and oxygen. The steam and oxygen passing upwards through the bed result in two specific reactions.

(1) Carbon in the bed reacts with oxygen to produce carbon monoxide and carbon dioxide in an exothermic reaction:

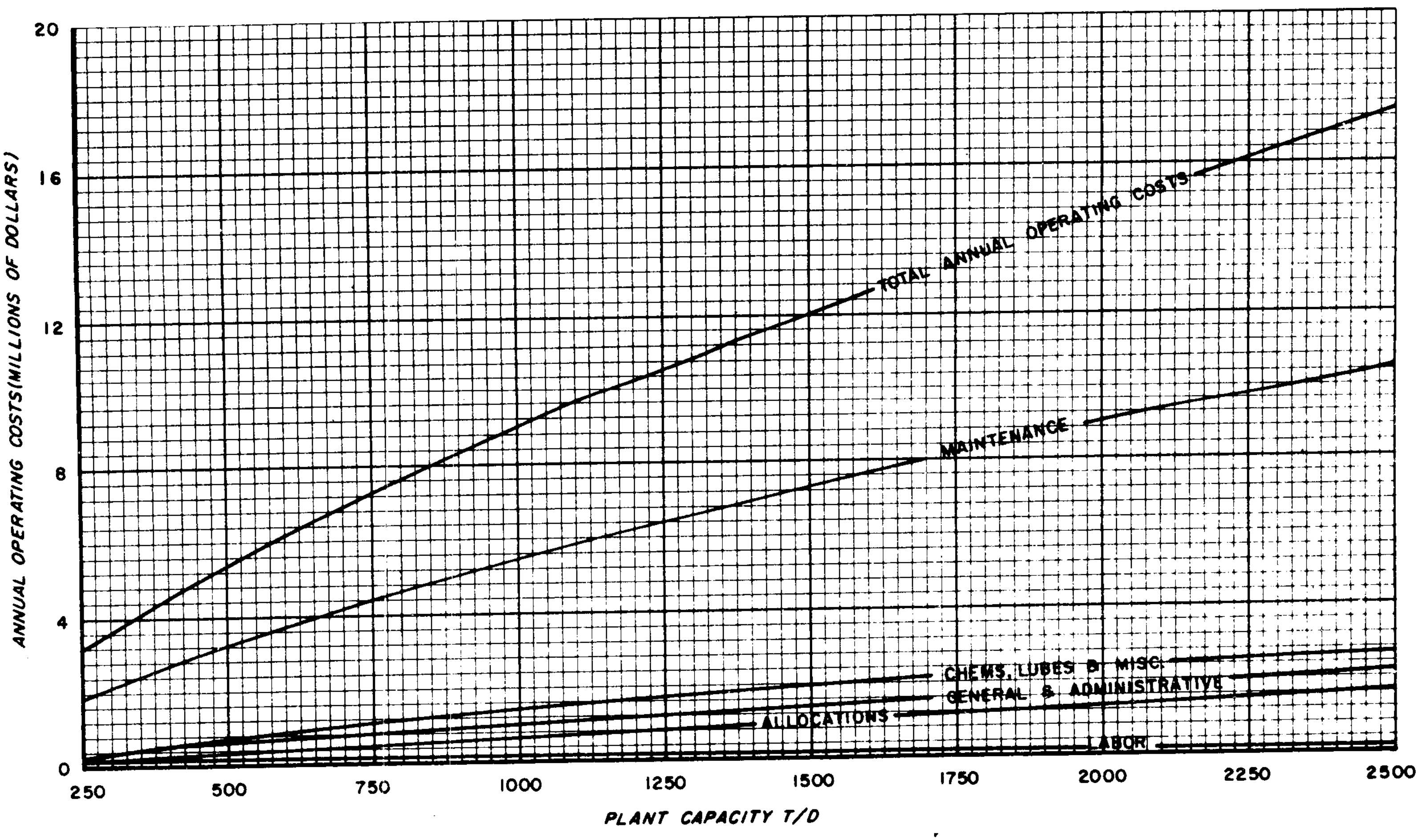


FIGURE 3-10. ESTIMATED OPERATING COST vs PLANT SIZE
(WATER ELECTROLYSIS PROCESS, HIGH PRESS-HIGH TEMP)

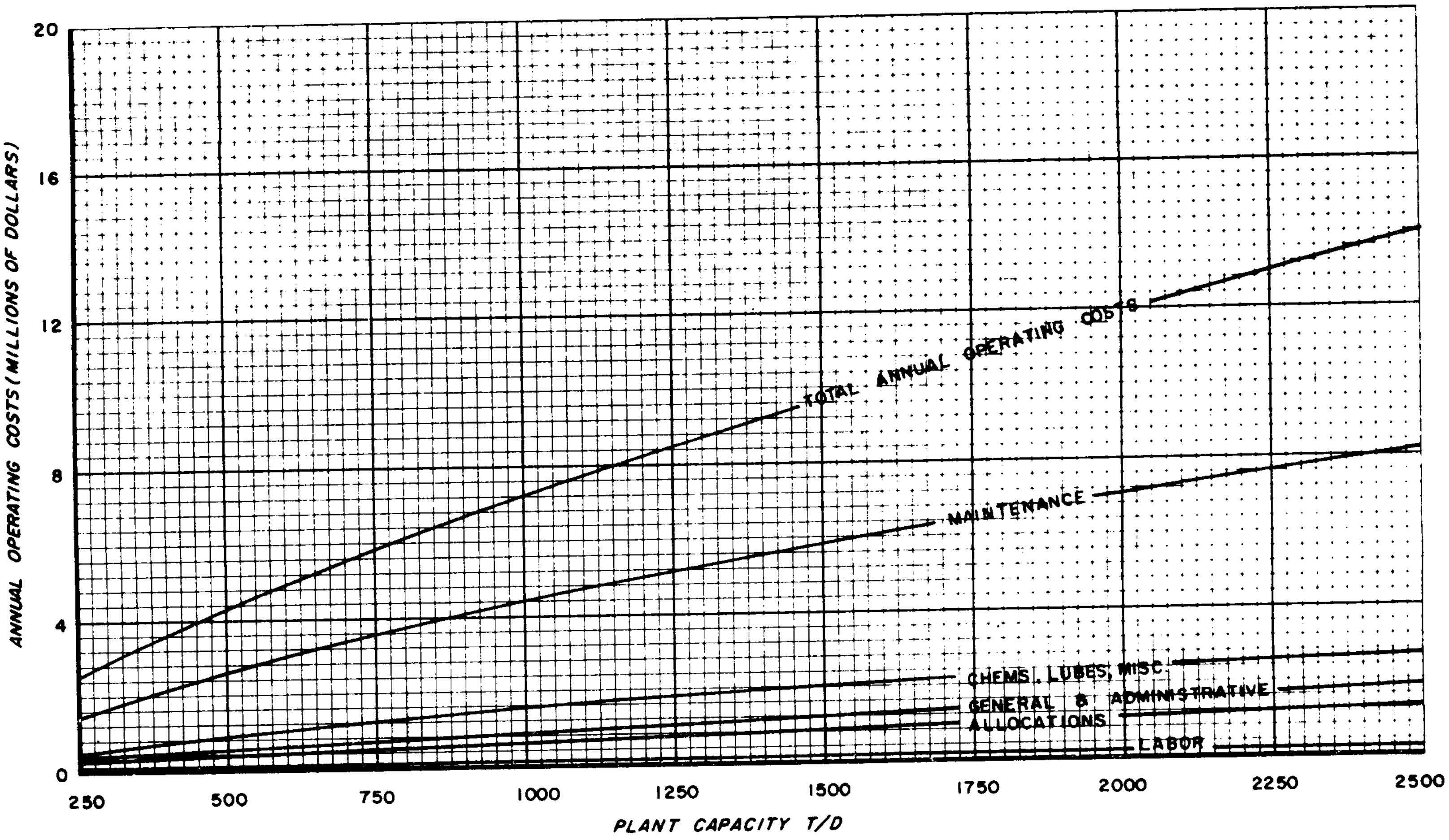


FIGURE 3-11. ESTIMATED OPERATING COST vs PLANT SIZE
(WATER ELECTROLYSIS PROCESS, LOW PRESS-LOW TEMP)

$$3C + 2O_2 \longrightarrow CO_2 + 2CO + \text{heat}$$

(2) The heat of reaction so produced supports the endothermic reaction between carbon and steam:

$$3C + 3H_2O + \text{heat} \longrightarrow CO + CO_2 + H_2 + CH_4$$

The four gaseous products then pass out through the top of the reactor, thus completing the coal gasification phase. Typical fluidized bed reaction conditions are 450 psig pressure and 1600°F or 1800°F for lignite or bituminous feed respectively.

The product gases from the gasification reaction pass through a shift converter, analogous to the catalytic shift converter found in the steam reforming process. Auxiliary steam is fed into the converter and reacts with the carbon monoxide in the product stream;

$$H_2O + CO \longrightarrow CO_2 + H_2$$

producing additional product hydrogen. The product gas from the converter, then, is predominantly H_2 with significant H_2O and CO_2 impurities present as well as lesser amounts of CO and CH_4.

From the converter, the product gases pass into CO_2 removal equipment, again analogous to the MEA scrub system in the steam reforming process. CO_2 removal equipment removes H_2O and CO_2 from the process gases. This yields a final product gas composed of H_2 and impurities of CO and CH_4. Typically, the product gas is 97 percent H_2 with the balance being CO and CH_4. A diagram of this process is shown in figure 3-12.

3.2.4.1.2 CO_2 Accepter Process Description. – As with the steam-oxygen process, several reactions occur in the course of gaseous hydrogen generation via the CO_2 accepter process. The following description relates to the CO_2 accepter process that results from using lignite as the carbonaceous raw material. Bituminous material, because of its lower reactivity, does not maintain as desirable a kinetic reaction balance as the lignite.

Pulverized and dried, lignite feeds to a reacter vessel. Steam is used to fluidize the coal for transfer to this vessel. Simultaneously "CO_2 accepter material", (e.g., CaO) feeds to the same reacter vessel. Accepter material and lignite are mechanically mixed and maintained in a fluidized state. In this condition, several reactions occur simultaneously and on a continuing basis. The reactions presented constitute the gasification of coal.

$$C + H_2O + \text{heat} \longrightarrow CO + H_2, \quad \text{(a)}$$

$$CO + H_2O + \text{heat} \longrightarrow CO_2 + H_2, \text{ and} \quad \text{(b)}$$

$$CaO + CO_2 \longrightarrow CaCO_3 + \text{heat.} \quad \text{(c)}$$

CO_2 formed in shift reaction (b) reacts with the accepter and is removed as quickly as it is generated (reaction c). This drives both reactions (b) and (a) to completion.

It has been observed that the overall reaction,

$$C + CaO + 2H_2O \longrightarrow CaCO_3 + 2H_2$$

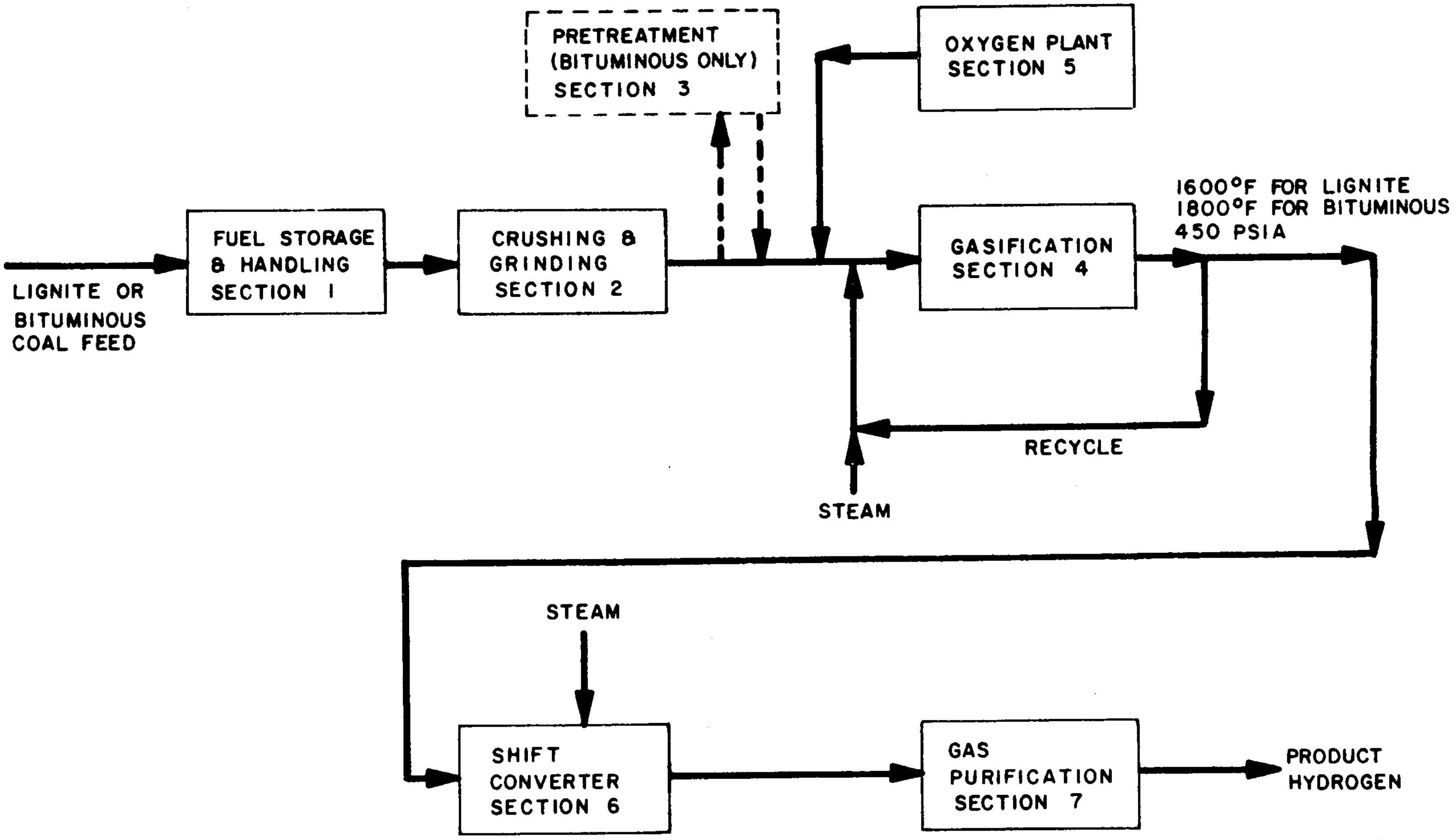


FIGURE 3-12. COAL GASIFICATION STEAM-OXYGEN PROCESS

is essentially thermoneutral. This eliminates the need for thermal energy addition or removal. Further, the number of shift converters required is greatly reduced, or entirely eliminated, as is the CO_2 scrub system. However, the CO_2 cannot be recovered readily as a byproduct as in the steam-oxygen gasification process.

The product gas, containing H_2, CO, CO_2, and CH_4 leaves the gasifier for futher processing. The unreacted coal, ash, and spent accepter are removed from the gasifier and conveyed via air fluidization to the accepter regenerater. The unreacted carbon residue at this point is about 30 percent of the initial feed. In the regeneration vessel, oxygen in the fluidizing air reacts with the residual carbon. The reaction, typically at 450 psia and 1900°F, supplies thermal energy necessary to regenerate the accepter. The CaO is then separated from the ash, unburned coal, and accepter fines, and recycled back to the gasification unit. The gases and the ash from the regenerater are disposed as waste. The utilization of the carbon present in the lignite is approximately 90 percent. Carbon utilization of 80 percent would be expected if bituminous coal were used.

A most critical parameter associated with the gasification reaction is the maintenance of the steam partial pressure in the reaction zone. A partial pressure of 13 atms, or less, must be maintained. Failure to do so will cause agglomeration of the fluidized particles, destroying the fluidized bed.

The product gases leaving the gasifier are processed similarly to those generated in the steam- oxygen process, with the exception that the shift converter and CO_2 scrubber are not required. Typically the product gas is 96 to 97 percent H_2, the impurities being CO, CO_2, and CH_4. A diagram of this process is shown in figure 3-13.

3.2.4.2 Process Elements

Process element requirements have been based on literature data.[18,26] A 250 T/D gaseous hydrogen production facility was used as the base case. The requirements for the two study processes are presented as follows:

(1) Steam-Oxygen Process:

a. Raw Materials	Lignite	Bituminous
Feed Water ($lb/lb\ H_2$)	8.84	20.3
Coal ($lb/lb\ H_2$)	12.10	8.74
Cooling Water Makeup ($lb/lb\ H_2$)	23.5	26.0
Oxygen ($lb/lb\ H_2$)	5.20	5.20

b. Process Energy	Lignite	Bituminous
Power ($kwh/lb\ H_2$)		
Misc.	0.57	0.57

c. Byproduct Production	Lignite	Bituminous
CO_2 ($lb/lb\ H_2$)	24.0	24.0

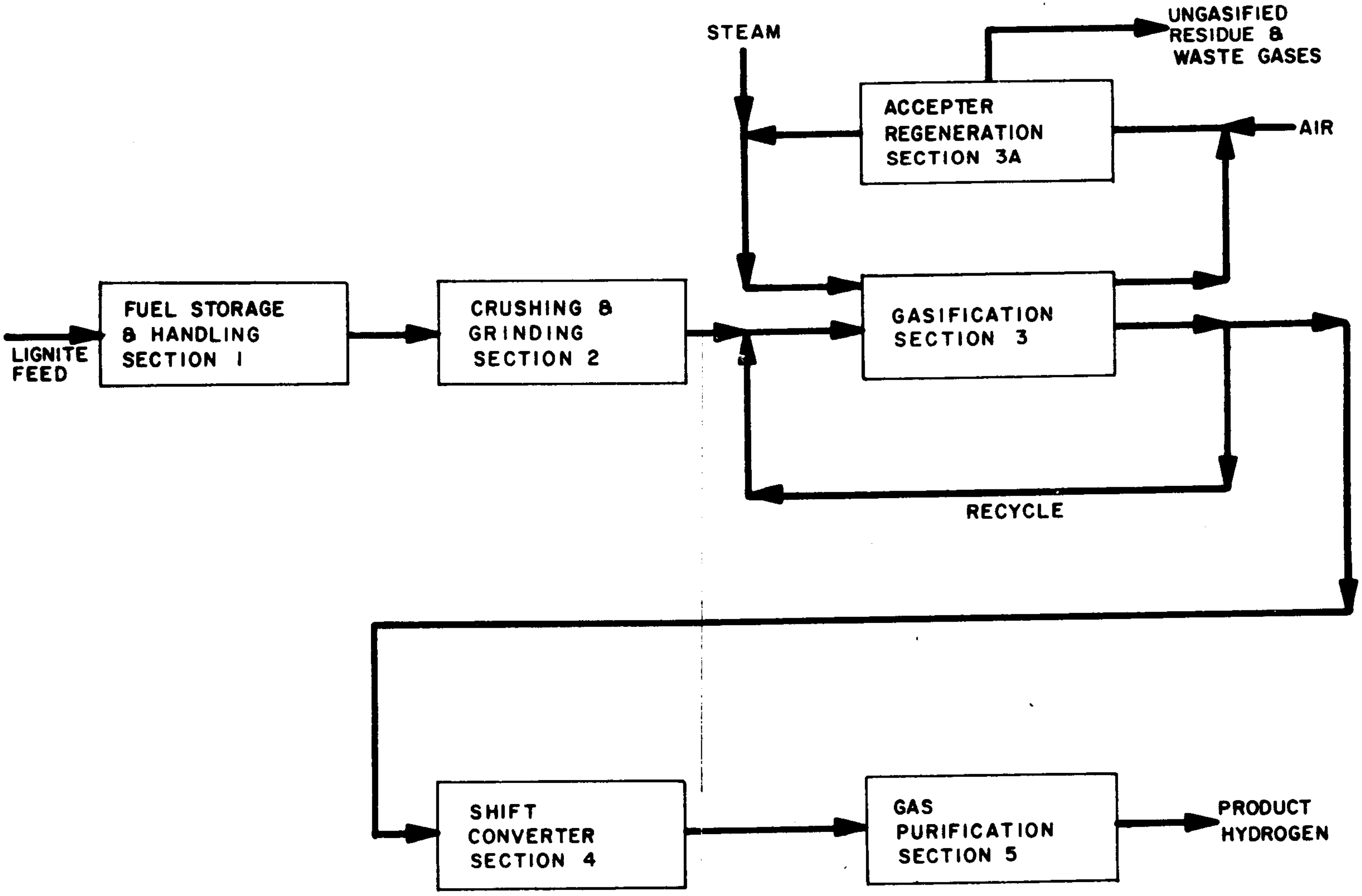


FIGURE 3-13. COAL GASIFICATION CO_2 ACCEPTER PROCESS

(2) CO_2 Accepter Process

a. Raw Materials	Lignite
Feed Water (lb/lb H_2) (Additional H_2O is furnished by the moisture content of the lignite feed)	8.84
Coal (lb/lb H_2)	12.10
Cooling Water Makeup (lb/lb H_2)	24.0
CaO (lb/lb H_2)	neg.
b. Process Energy	
Power (kwh/lb)	1.11
c. Byproduct Production	
CO_2(lb/lb H_2)	0

3.2.4.3 Cost Elements

Cost elements for capital investment and operating expenses have been determined for a plant capacity of 250 T/D of hydrogen.

3.2.4.3.1 Capital Investment – The exponential relationship used for capital investment is:

$$I = I_B \left(\frac{C}{C_B}\right)^{M_I}$$

Due to a lack of detailed data to determine the scale-up factor (M_I), the 0.7 factor used for the steam reforming and partial oxidation processes has been employed. It is felt that the general equipment similarities justify its use.

The estimated cost information presented and shown in figure 3-14 is the result of original engineering estimates of the various coal gasification processes prepared for this study. The estimated capital cost for the steam-oxygen process is \$16.9 x 10^6 for a 250 T/D plant using lignite feed. A similar plant using bituminous feed represents a capital investment of approximately \$20.8 x 10^6. The CO_2-accepter process requires an estimated capital investment of \$21.3 x 10^6.

3.2.4.3.2 Operating Costs – The operating costs were also developed as a result of engineering studies in conjunction with this report. The operating costs discussed exclude raw materials and process energy.

The operating costs can be expressed as an exponential function of plant capacity as with other processes studied:

$$O = OB \left(\frac{C}{C_B}\right)^{M_o}$$

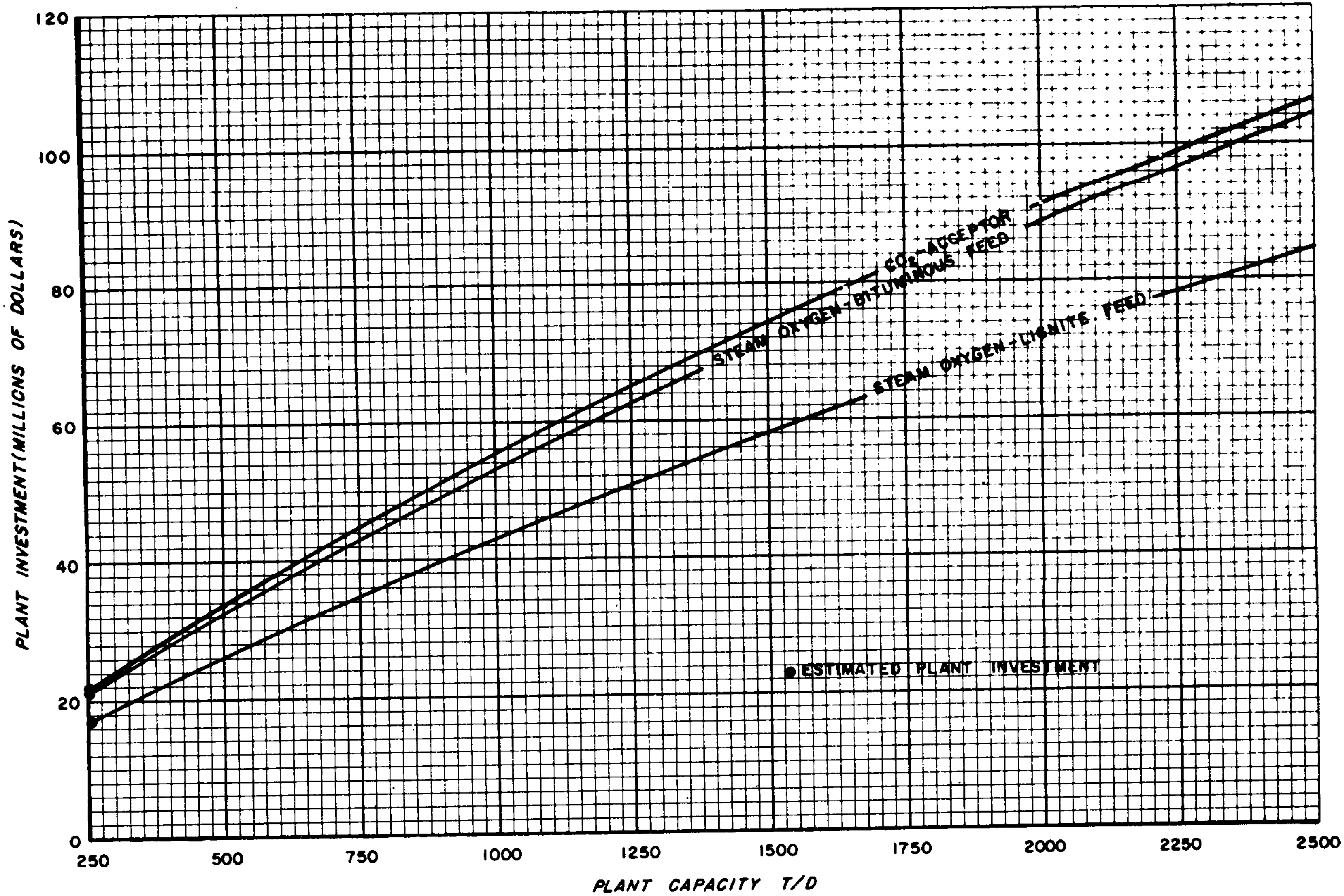


FIGURE 3-14. ESTIMATED INVESTMENT vs PLANT SIZE
(COAL GASIFICATION PROCESS)

Operating costs were calculated for all these processes in a manner similar to that described in paragraphs 3.2.1.4.2 and 3.2.2.3.2. Chemical costs range from 0.30 cents/lb at the 250 T/D level to 0.24 cents/lb at the 2500 T/D level for both steam-oxygen processes and the CO_2-accepter process. Maintenance costs for all three systems were calculated at a rate of 5.0 to 8.0 percent/year of plant investment. Operations allocations and general and administrative costs were calculated in the same manner described in paragraph 3.2.1.4.2. The estimated value for O_B which is the operating cost for the base 250 T/D facility for each of the various processes is as follows:

Steam-Oxygen (lignite feed)	\$2.57 x 10^6 per year
(bituminous feed)	\$2.89 x 10^6 per year
CO_2-Accepter (lignite feed)	\$2.89 x 10^6 per year

The operating cost exponent (M_O) is 0.65 for all three processes. The relationship of the various subcategories to total annual operating costs as plant capacity increases are shown in figures 3-15 and 3-16.

3.2.5 HYDROGEN - HALIDE

A hydrogen-halide process feasibility study for a small scale installation was conducted in 1965 by Air Products and Chemicals.[25] The production of gaseous hydrogen by the hydrogen-halide process consists of two main steps:

(1) The reaction of water and a halogen to form a hydrogen halide.

(2) The electrolytic dissociation of the hydrogen halide to produce gaseous hydrogen and recycle halogen.

The Air Products study included a conceptual process design utilizing 1500°F hydrogen chloride (HCl) as the halogen source. This conceptual process design served as the basis for the hydrogen halide large scale hydrogen production process, as described in the following paragraphs.

3.2.5.1 Process Description

Feedstock water (cleaned, purified and pressurized) enters the process and mixes with HCl and previously liberated Cl_2, as shown in figure 3-17. The mixture, having a Cl_2:H_2O mol ratio of 2.48:1, passes to a reactor vessel. With the addition of thermal energy, a halogenation reaction takes place at 1500°F and 770 psia. The H_2O and Cl_2 react to form HCl and byproduct O_2. It is projected that the catalyzed reaction would result in 55 percent (mol percent) conversion of water to HCl.

Heat is removed from the gaseous reaction products by counter-flow heat exchange with reactants entering the reactor vessel. Upon cooling to 100°F, the products form a two-phase mixture of liquids and gases. The mixture passes to a phase separator where the residual halogen and unreacted water pass-out as liquid for recycle to the reactor. Gaseous overhead from the phase separator, including primarily HCl and byproduct O_2 pass to a condenser.

A liquid phase is formed passing through the condenser, containing HCl and trace amounts of Cl_2. The gaseous phase is essentially pure byproduct O_2, which is separated from the two immiscible

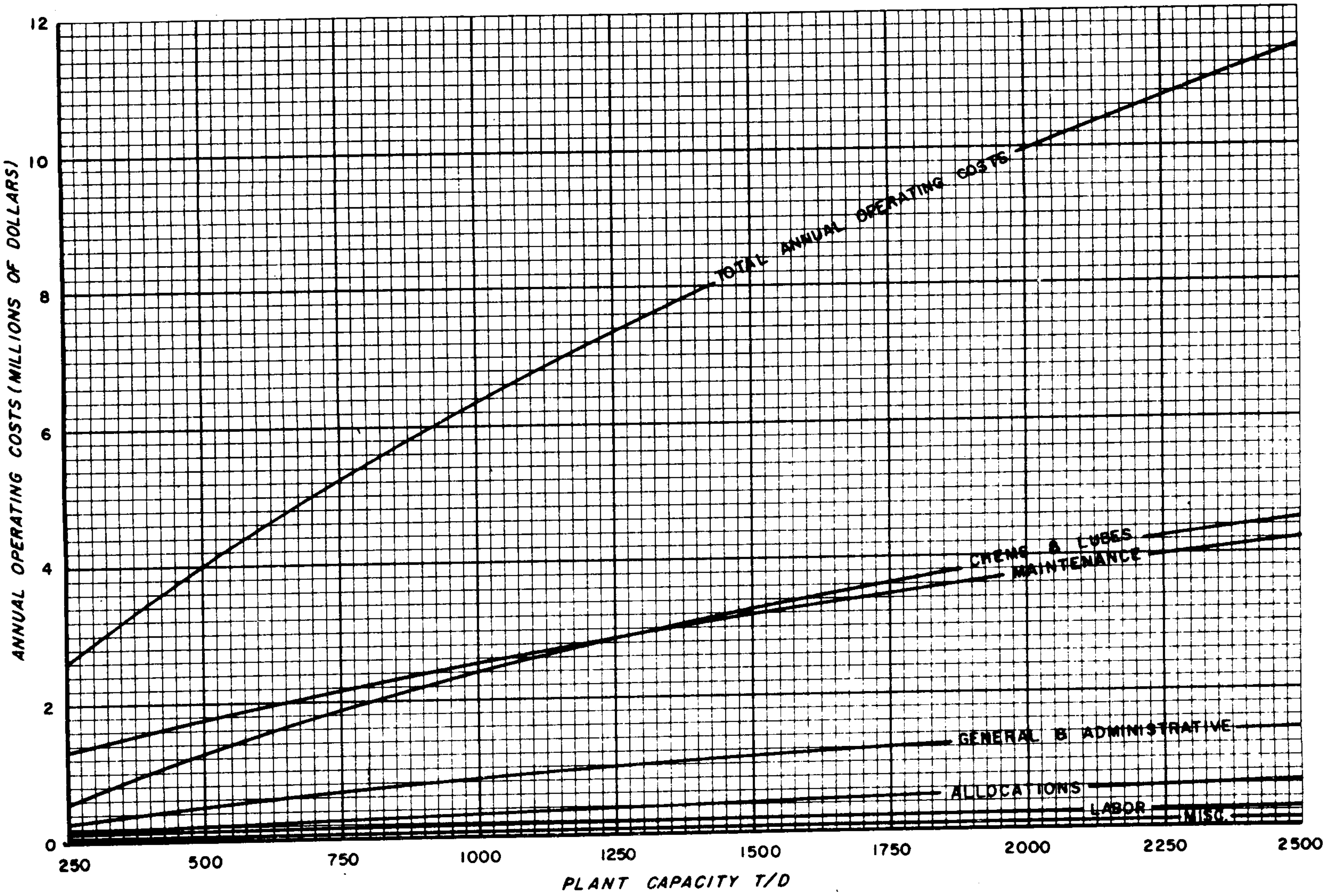


FIGURE 3-15. ESTIMATED OPERATING COST vs PLANT SIZE (COAL GASIFICATION PROCESS STEAM-OXYGEN-LIGNITE)

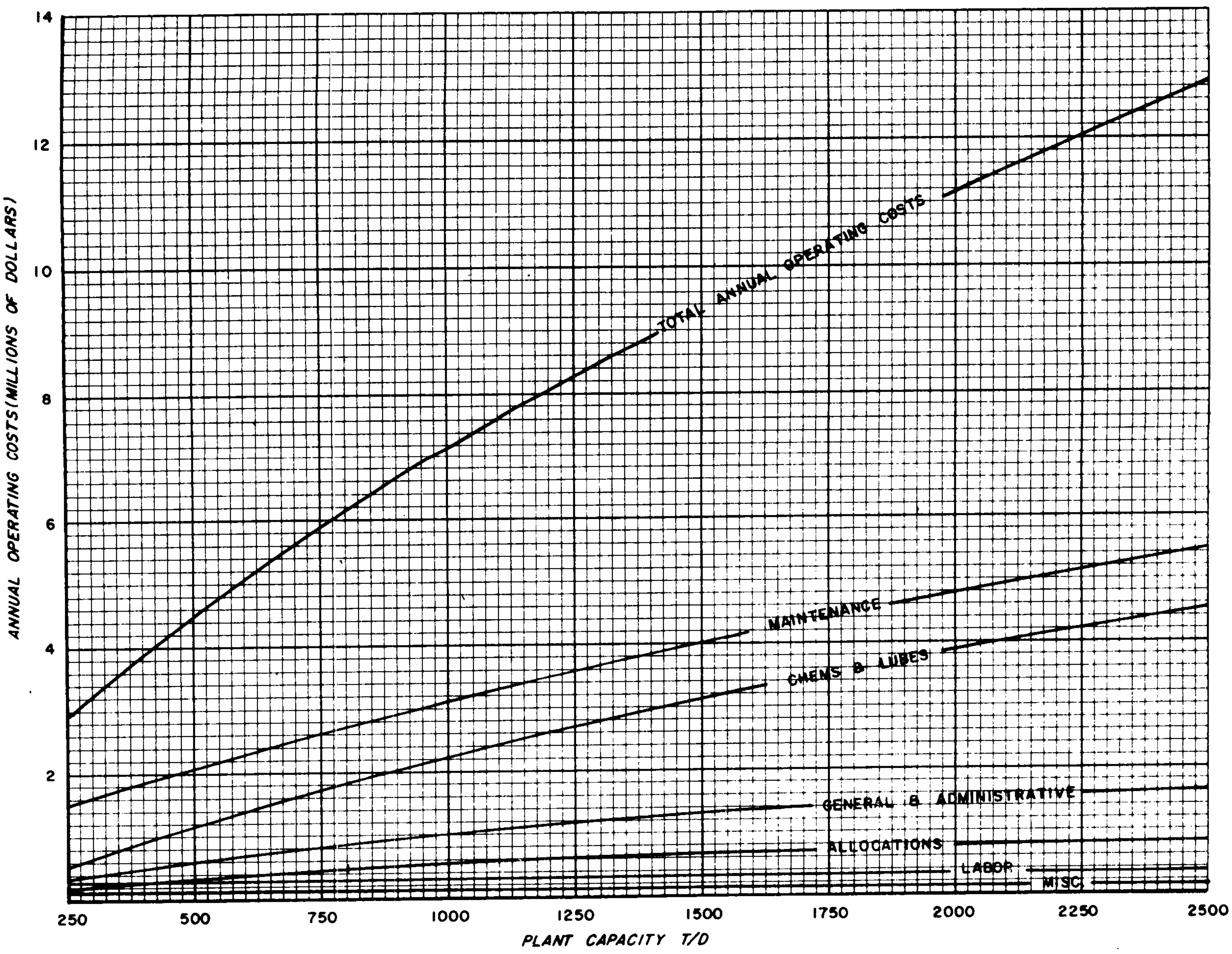


FIGURE 3-16. ESTIMATED OPERATING COST vs PLANT SIZE (COAL GASIFICATION PROCESS (CO$_2$ ACCEPTER)

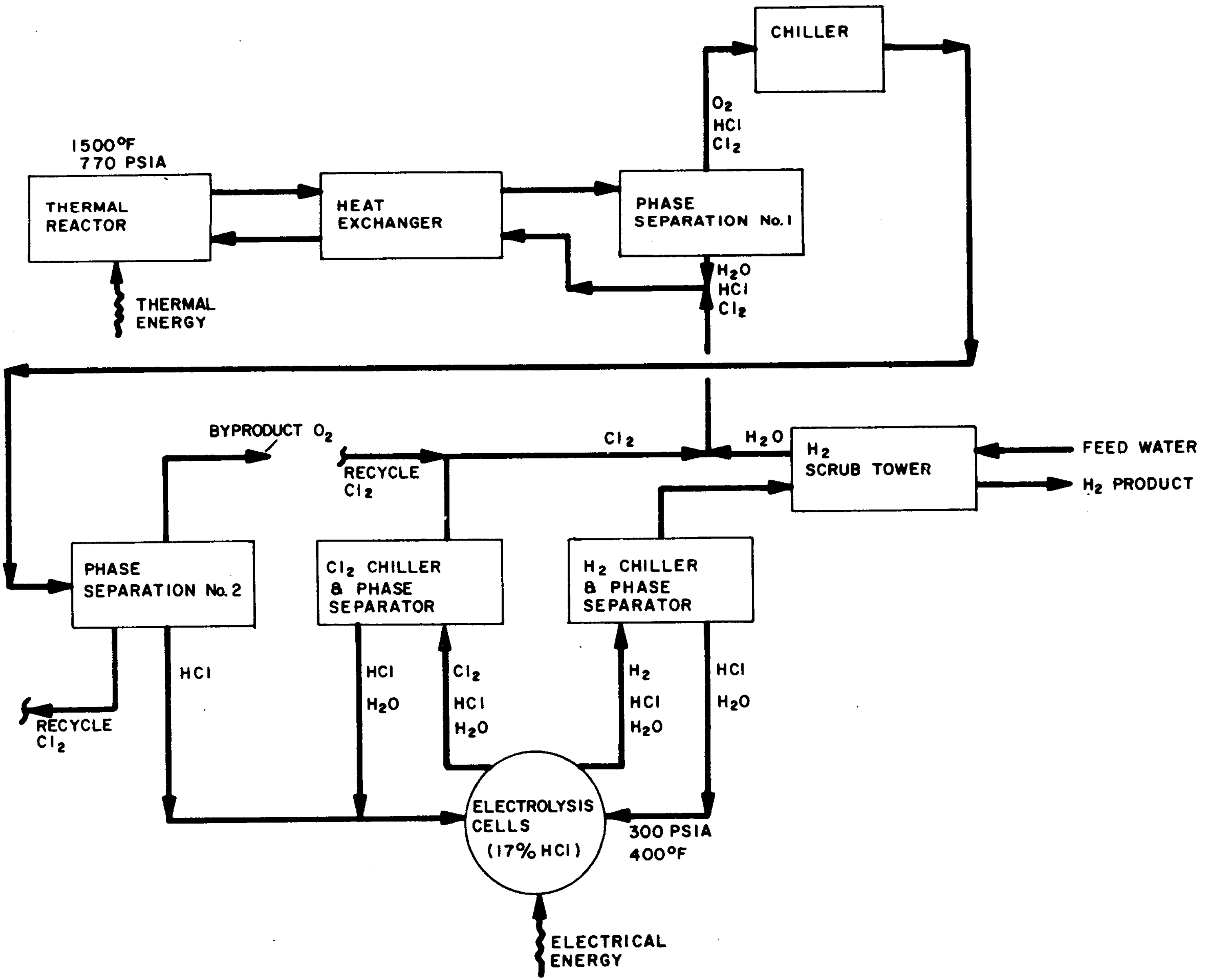


FIGURE 3-17. HYDROGEN-HALIDE PROCESS BLOCK DIAGRAM

liquids (HCl and Cl_2) in a second phase separator. The separated O_2 then is vented or further processed to marketable quality. Mechanical separation of the two immiscible liquids is achieved in the phase separator. Liquid Cl_2 is effectively separated and routed to recycle with incoming fresh water feedstock while the remaining liquid, HCl, is directed toward the electrolytic cells. Enroute, HCl mixes with recycle hydrogen-halide and water, then enters the electrolytic cell system typically operating at 392°F and 300 psia.

The reactant feed to the cell, 17 percent (mol percent) HCl, dissociates in the cell to gaseous product H_2 and gaseous chlorine (CL_2). The product, saturated with electrolyte, passes through heat exchangers which cool the stream and cause residual electrolyte (HCl and H_2O) to condense.

The liquid is separated from the gaseous product in a subsequent phase separator, the product passing on out of the process. Gaseous halogen from the cell is cooled to condense residual electrolyte. After phase separation, the gaseous halogen is routed to recycle with incoming fresh water feedstock and the electrolyte is routed to recycle with feedstock flowing to the electrolytic cell. Final cleanup of gaseous product (H_2) is accomplished by a counterflow scrub with fresh water feedstock to the process. The product hydrogen leaves the process at 50°F, 280 psia.

The halogenation reaction (first reactor vessel) has not been experimentally studied at high temperatures. It is estimated that reaction energy of 2.48×10^4 Btu/lb mol CL_2 reacted is necessary for the reaction. This energy requirement is based on equilibrium conversions using standard electrochemical potential data. A preliminary evaluation was made of the reaction energy requirements as affected by temperature, pressure, and reactants composition. On this basis, reactor conditions of 1500°F and 770 psia were chosen as optimum conditions.

Little significant data is available from the literature regarding the kinetics of the HCl reaction. It has been noted generally that granular solids such as cesium oxide and quartz chips catalyze the bromination reaction several fold. This, being similar to the HCl reaction, would indicate that suitable catalytic materials could be developed for the HCl process. It is expected that the limitation of such a catalyst would be heat transfer characteristics.

Operating conditions for the electrolytic cells were based on reversible and irreversible electrical effects and thermal requirements. Reversible potential for the electrolytic reaction are known with respect to temperature, pressure and reactant composition. Irreversible effects of overvoltage and electrolyte resistance are similarly understood. Initial consideration of this data led to the optimum operating cell conditions of 392°F and 300 psia pressure. The referenced 17 percent (mol percent) HCl reactant mixture is significant in that cell characteristics change for concentrations above 20 percent HCl. That is, for electrolysis of aqueous HCl, reversible potential required decreases with increasing temperature and HCl concentration. However, the irreversible effects of overvoltage and electrolyte resistance increase rapidly as the HCl mole percent concentration exceeds 20 percent. Noting this, the cell operating conditions were chosen, yielding a total electrical power requirement of 15.06 kwh/lb H_2 produced at 300 psia. Compression of the hydrogen to liquefier feed pressure requires an additional 0.24 kwh/lb.

3.2.5.2 Process Elements

The process elements for the hydrogen-halide process are directly proportional to the plant sizes.

Projected unit requirement data is:

(1) Raw Materials

Feed Water (lb /lb H_2)	9.0
Methane (lb /lb H_2)	1.1
Cooling Water Makeup (lb /lb H_2)	13.0
HCl (lb /lb H_2)	neg.
Cl_2 (lb /lb H_2)	neg.

(2) Process Energy

Electrical (kwh/lb H_2)	15.3

(3) Byproduct Oxygen (lb/lb H_2) 7.92

3.2.5.3 Cost Elements

Investment and operating cost characteristics of the hydrogen-halide process have been evaluated for a plant capacity of 250 T/D. All cost data has been developed through engineering studies based upon a 1965 feasibility study conducted for the U. S. Army Engineer Reactors Group.[25]

3.2.5.3.1 Capital Investment. – The investment cost characteristics for the hydrogen-halide process are similar to those for the water electrolysis process. As plant capacity increases, the capital cost increases exponentially. This relationship can be expressed through the use of Williams' equation as:

$$I = I_B \left(\frac{C}{C_B}\right)^{M_I}$$

The value of the exponential constant M_I for the hydrogen-halide process is 0.9. The constant has a high value due to the fact that larger plants cannot employ larger pieces of equipment but rather must employ greater numbers of maximum or optimum sized equipment items. The relationship of plant investment to plant size is shown in figure 3-18. The savings thus realized are limited to those received for quantity purchases only.

The capital investment for the hydrogen-halide process base system 250 T/D has been estimated at $46.5 x 10^6 based upon engineering studies and an economic analysis of the process. While the hydrogen halide process is essentially an electrolytic process the investment is 30 to 60 percent higher than the electrolytic processes discussed earlier in this study, due to exotic material requirements,(i.e., tantalum, Hastalloy "C", and stainless steel), and additional units of equipment.

3.2.5.3.2 Operating Costs. – The hydrogen-halide process operating cost elements are comparible to those of the water electrolysis process. The two systems exhibit basic similarities, i.e., feedwater purification, feedwater pressurization, electrolytic dissociation, and product hydrogen purification. The major dissimilarity between the processes is the halogenation subsystem required in the hydrogen-halide process.

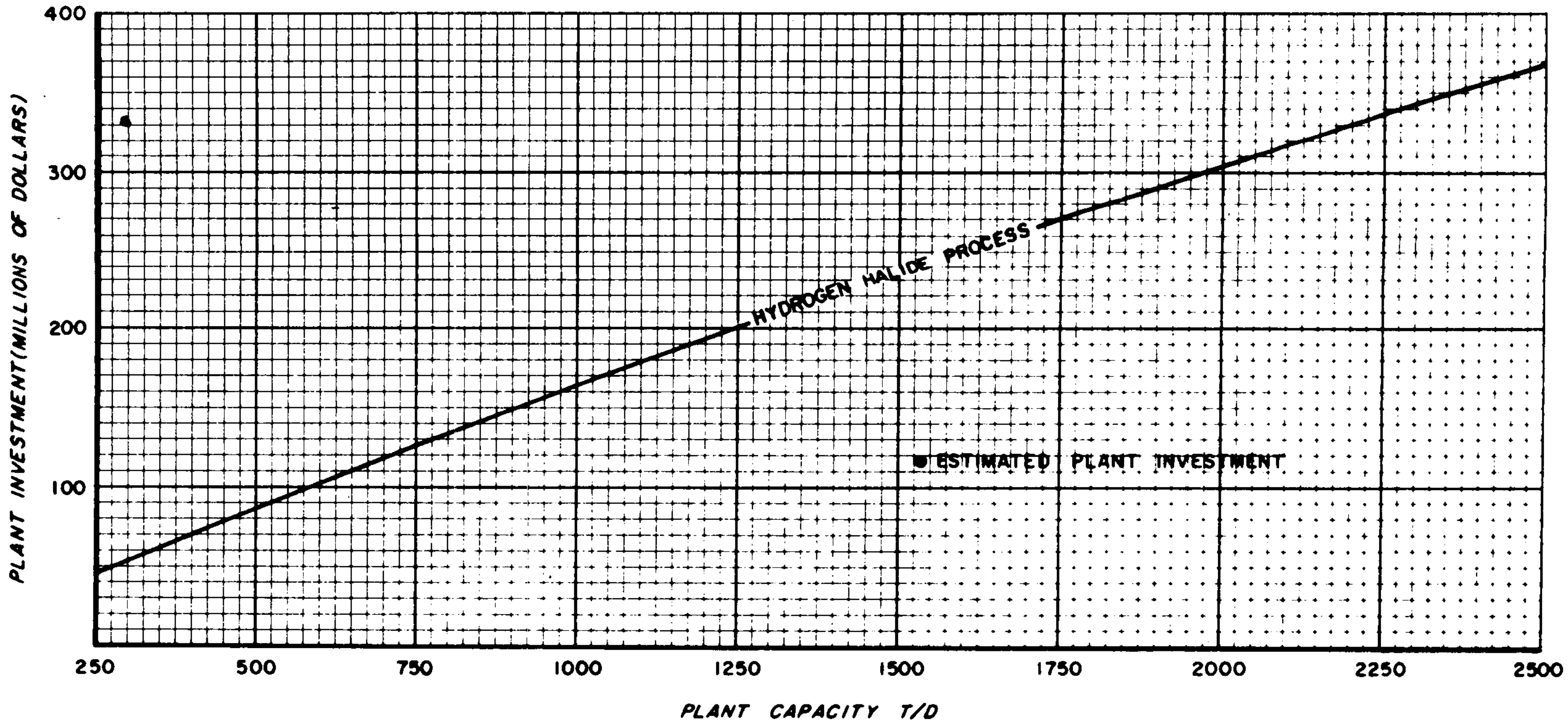


FIGURE 3-18. ESTIMATED INVESTMENT vs PLANT SIZE
(HYDROGEN-HALIDE PROCESS)

Thus, operating costs excluding raw materials and process energy can be expressed as a function of plant capacity as were the costs for water electrolysis. The relationship is exponential and can be stated as:

$$O = O_B \left(\frac{C}{C_B}\right)^{M_o}$$

Cost of chemicals and lubricants will range from 0.14 cent/lb to 0.20 cent/lb. Maintenance costs were calculated at a rate of 4 to 6 percent/year of plant investment. Operations allocations and general and administrative costs were calculated in the same manner described in paragraph 3.2.1.4.2.

The calculated value for O_B is \$4.35 x 10^6 per year for a 250 T/D plant. The scale-up factor for the operating cost is estimated to be 0.76. The relationship of the cost elements to total annual operating costs as plant capacity increases are shown in figure 3-19.

3.3 HYDROGEN GAS LIQUEFACTION AND CONVERSION

The second major subsystem within a large scale liquid hydrogen plant is the process equipment and ancillaries required for the actual liquefaction of the gas. In addition to liquefying the subsystem must be capable of converting orthohydrogen which is a molecular variety of common hydrogen to parahydrogen. This conversion is necessary because the orthopara equilibrium concentration levels vary with the temperature of the liquid. If the conversion were allowed to take place spontaneously the reaction, being exothermic, would release more energy than the heat of vaporization and thus large quantities of liquid product would be lost by vaporization during storage.

The following paragraphs provide process descriptions of the liquefaction and conversion methods that can be employed in large scale hydrogen systems. Unit consumption data is classified as process elements or cost elements. As with the various gas production subsystems, process elements detail amounts of raw material and process energy related to net unit production of the subsystem. Similarly, cost elements detail typical amounts of capital investment and other operating costs associated with the net unit production. The elements are considered for all the processes.

3.3.1 GENERAL DESCRIPTION

The liquefaction of any gas is accomplished by the cooling of the gas until it reaches its condensation temperature. At this point, the latent heat of vaporization is then removed. Thus, the only basic requirement for the liquefaction of a gas is suitable refrigeration.

The difference in the liquefaction of hydrogen to other common industrial gases is the fact that there is a wider range of temperatures at which refrigeration is required and a good deal of this refrigeration is at temperatures below the boiling point of liquid air. More specifically, the liquefaction of hydrogen requires refrigeration for three distinct loads:

(1) Cooling the hydrogen to its boiling point -423°F

(2) Removing the latent heat of vaporization at -423°F

(3) Removing the latent heat of ortho-para conversion

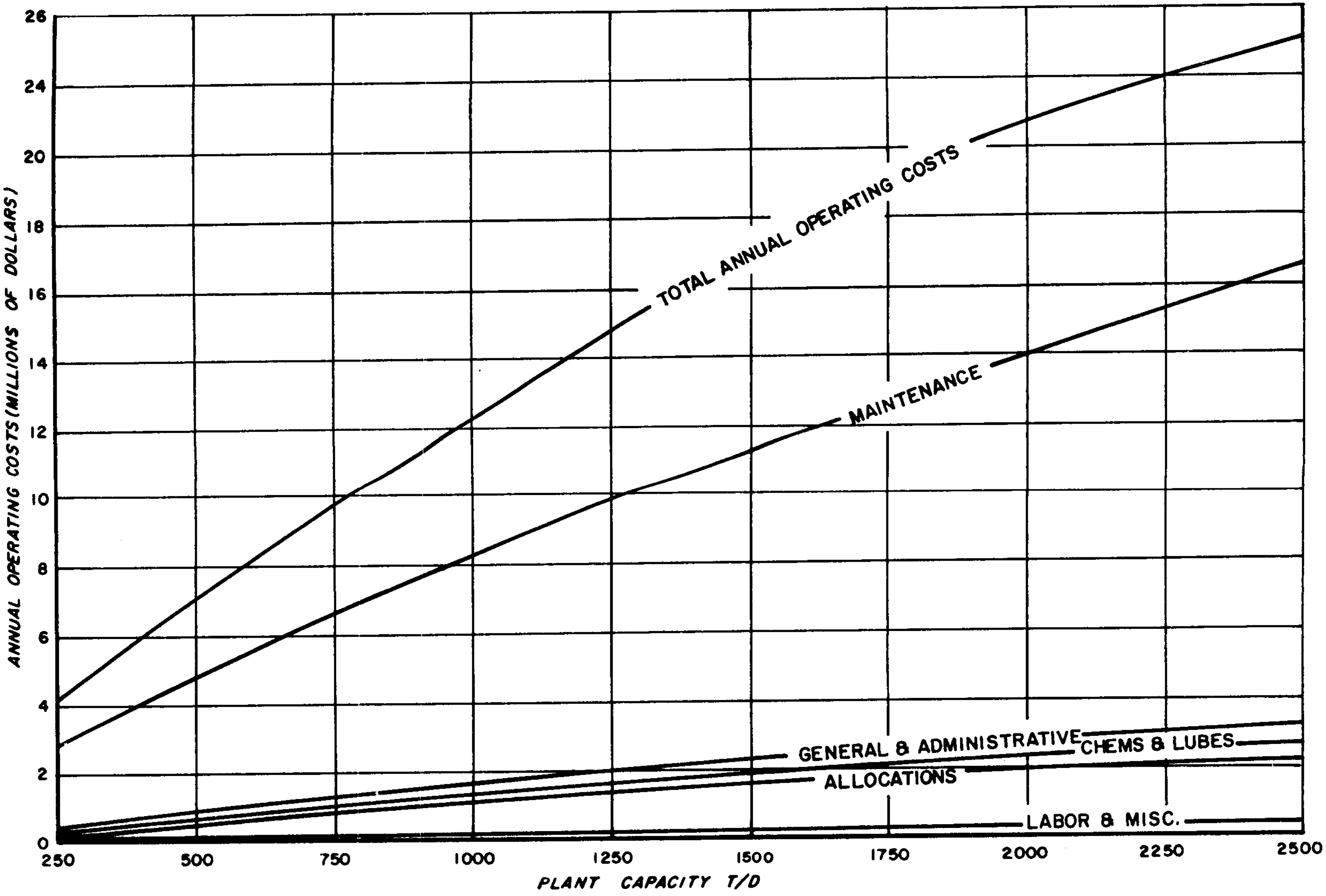


FIGURE 3-19. ESTIMATED OPERATING COST vs PLANT SIZE (HYDROGEN-HALIDE PROCESS)

Refrigeration is supplied at various temperature levels in the liquefaction plant to satisfy the load requirements given. Two levels are commonly delineated for process design purposes. The lower level of refrigeration, from -294°F to -423°F is referred to as "below Lin" and above -294°F as "above Lin" or high level refrigeration. These terms refer to the temperature level only, not the refrigerant; e.g., refrigeration above Lin temperatures may be supplied by argon, methane, or nitrogen.

The high level refrigeration methods selected for intensive study in this report are:

(1) Cascade Refrigeration System

(2) 3000 psig Nitrogen Recycle System

3.3.1.1 Cascade Refrigeration System

The cascade refrigeration system is a system wherein a series of liquids of progressively lower boiling points are condensed under pressure at the temperature produced by the evaporation of the next higher boiling liquid. For the liquid hydrogen system, the refrigerants utilized in the cascade process are propane -43.7°F, ethylene -154.8°F, methane -258.6°F, and nitrogen -320.5°F.

3.3.1.2 3000 PSIG Nitrogen Recycle System

The 3000 psig nitrogen recycle system is a system wherein high pressure nitrogen is recirculated in a closed loop process. High pressure nitrogen from the compressor passes through a heat exchanger where it is cooled by a counterflow of returning low-pressure gas. The high pressure stream is split part-way through the exchanger and the side stream passes through the expander thus providing additional refrigeration to the process. The remaining high pressure stream is further cooled through heat exchange and expanded through a Joule-Thomson valve at which point liquid nitrogen is formed. The heat of vaporization of this liquid is used in conjunction with the expander to provide refrigeration for the cool down of the product hydrogen gas.

The low level refrigeration method consists of a 1500 psia hydrogen recycle system similar to the 3000 psig nitrogen high level refrigeration system.

As indicated, ordinary hydrogen at room temperature consists of two molecular varieties, namely orthohydrogen and parahydrogen. The hydrogen molecule is composed of two hydrogen atoms and each of these atoms consists of a nucleus and one electron. The hydrogen nucleus is a proton. The nuclear spin of the individual atoms are either oriented in the same direction which corresponds to the ortho modification or in opposite directions which corresponds to the para modification.

The equilibrium concentrations of the ortho-para molecules are a function of temperature. If the temperature is changed, the concentrations are also changed. At room temperature the equilibrium mixture is 75 percent ortho and 25 percent para. At liquid nitrogen temperatures, -320°F, the equilibrium is approximately 50 percent for each. At liquid hydrogen temperatures, -423°F, the liquid in equilibrium is 99.8 percent para-hydrogen.

While the ortho-para concentration will naturally come to equilibrium over a period of time, there are two fundamental methods of inducing an ortho-para transition. One method consists of dissociating the hydrogen molecules and allowing the atoms to recombine. When dissociated, the two atomic nuclei no longer are oriented with respect to nuclear spin. Upon reassociation, the molecules are formed according to the equilibrium energy distribution determined by the temperature. The other method involves the interaction between an inhomogeneous magnetic field and the magnetic field associated with the nuclear spin of the hydrogen nuclei. Since a magnetic field is produced along the axis of rotation of the spinning nucleus, an external field which causes a reversal of the direction of this nuclear magnetic field in effect produces a reversal of spin in one of the nuclei. This spin reversal is equivalent to an ortho-para transition.[26]

The first method, dissociation of the hydrogen molecules, is an acceptable approach to ortho-para transition at high temperatures. At low temperatures, however, there is not enough internal energy to promote dissociation and the magnetic interaction method predominates. All tonnage liquid hydrogen plants operate with a catalyst (a typical catalyst consists of molecular nickel silicate), containing a paramagnetic component which induces the required inhomogeneous magnetic field.

The latent heat of conversion is approximately one and one half times as great as the latent heat of vaporization. Therefore, it is economically desirable to complete the conversion as far as possible in the liquefier, rather than in storage after liquefaction has been completed.

The closer the para content of the product approaches the equilibrium para content, the higher the power requirement for liquefaction, but the lower the storage loss. Consequently, the optimum para content of the product is a function of the mean storage time as shown in figure 3-20. Scott[27] gives the uncatalyzed reaction rate as

$$\frac{-dx}{dt} = kx^2$$

where

x = mole fraction ortho

t = time, hours

k = 0.0114/hour

Therefore, assuming initial product para concentrations, and mean storage times, the stored product para concentration and conversion induced boil-off can be calculated.

For the purposes of this study, it has been assumed that storage capacity for two weeks will be required at each location and that product will be drawn out of storage on a continuing basis; i.e., residence time for each pound of product will normally be two weeks. Boil-off rates were calculated from the rate equation results as:

$$\frac{-dx}{dt} = k\,x^2$$

$$\frac{1}{x^2}\,dx = -k\,dt$$

by integration

$$\frac{1}{x_t} - \frac{1}{x_o} = k\,t$$

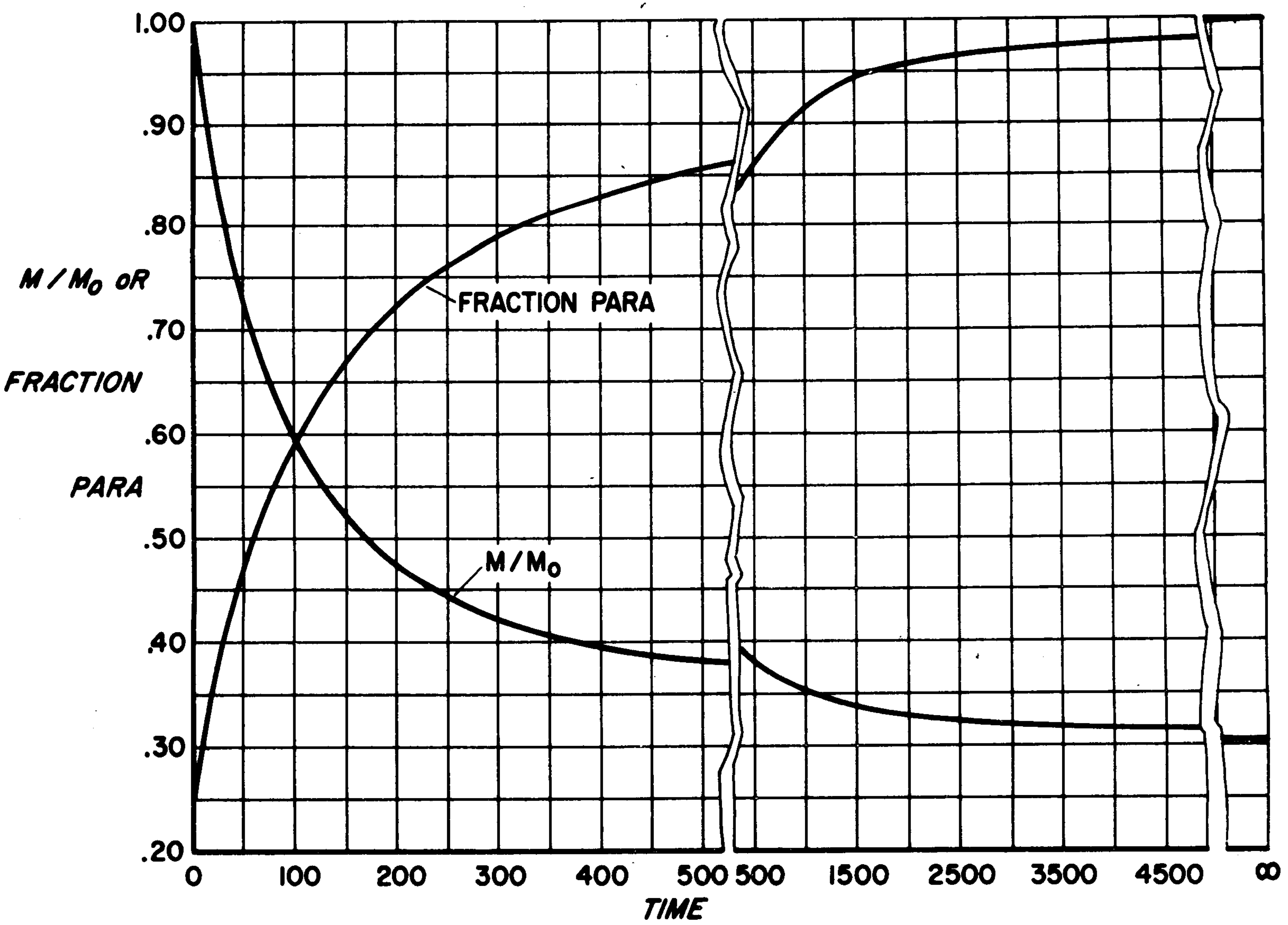


FIGURE 3-20. STORAGE LOSS, UNCATALYZED LIQUID HYDROGEN

where

x_o = ortho concentration at time zero
x_t = ortho concentration at time t
t = time in hours

Assuming a 95 percent para concentration from the liquefier and a mean storage time of two weeks the equation becomes

$$\frac{1}{x_t} - \frac{1}{0.05} = 0.0114\,(336)$$

$$\frac{1}{x_t} = 3.83 + 20.0 = 23.83$$

$$x_t = 0.042$$

therefore, para concentration equals 1.000 - 0.042 = 0.958

$$\Delta x_p = 0.958 - 0.950 = 0.008$$

Boil-off due to conversion is

$$m = \frac{(\Delta x_p)\ h_r}{h_v}$$

where

m = boil-off rate
Δx_p = change in molal per cent para concentration
h_r = heat of reaction for ortho-para conversion = 609 $\frac{\text{Btu}}{\text{mol}}$
h_v = heat of vaporization = 389 $\frac{\text{Btu}}{\text{mol}}$

therefore, $m = (0.008)\ \frac{609}{389} = 0.0124$

These calculations indicate that a 95 percent para concentration is desirable for product hydrogen. This level concentration will result in a conversion boil-off loss of 1.24 percent of plant capacity assuming 2 week residence. If the plant were to produce only 75 percent para product the boil-off losses would be 19.1 percent and this was judged to be excessive.

A plant designed for 95 percent para also permits the production of liquid hydrogen at lower para concentrations by by-passing a portion of the catalyst beds and by turn-down of the low level refrigeration systems to compensate for the lesser amount of heat of conversion being released in the liquefier. For these reasons, the plants discussed in this study are capable of producing 95 percent para product hydrogen.

3.3.2 PROCESS DESCRIPTION

As was stated previously, the actual liquefying process can be discussed in terms of two refrigeration levels.

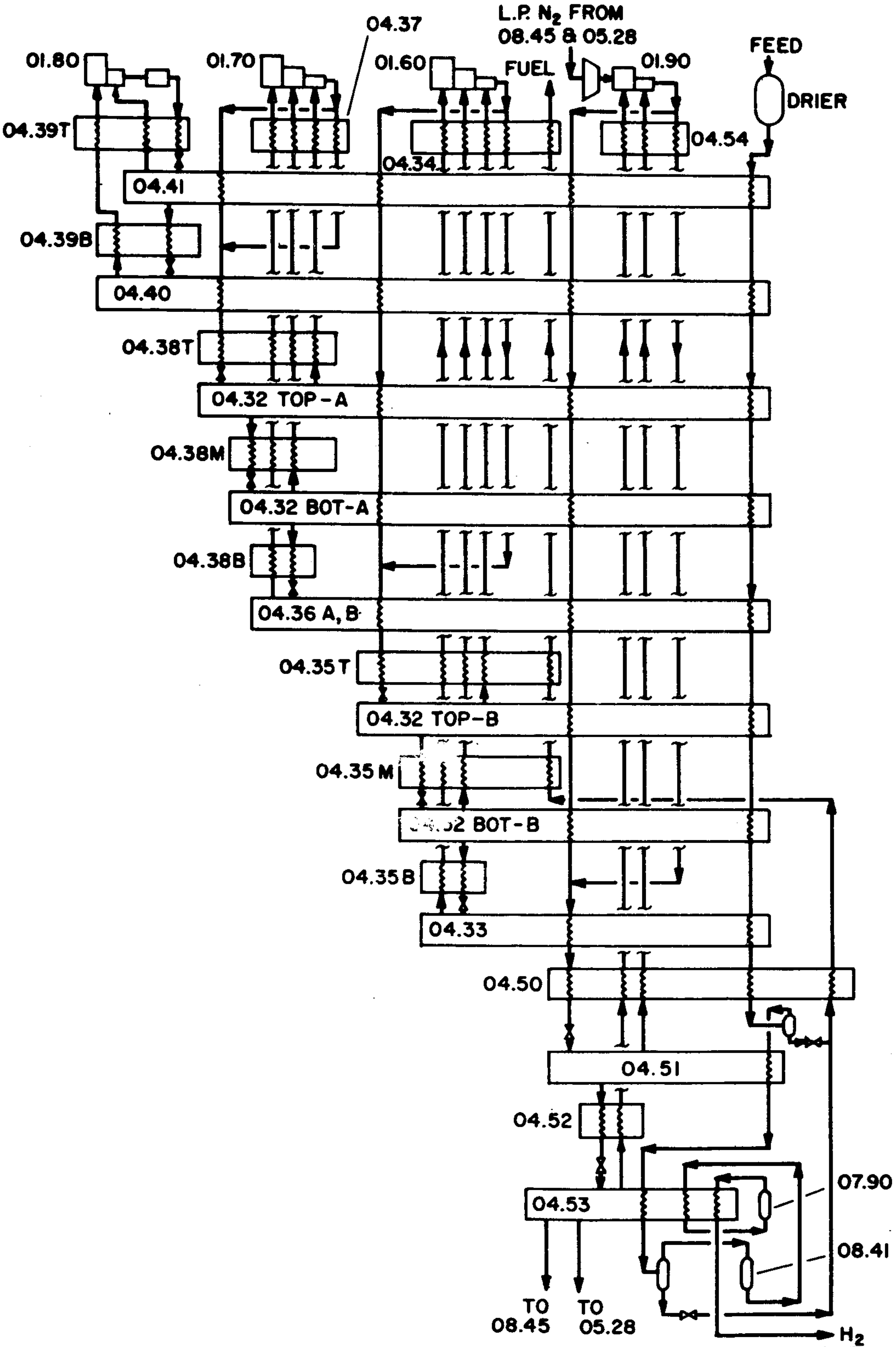


FIGURE 3-21. HIGH-LEVEL REFRIGERATION SYSTEM (CASCADE CYCLE)

LEGEND FOR FIGURE 3-21

01.60	Methane Compressor
01.70	Ethylene Compressor
01.80	Propane Compressor
01.90	Nitrogen Compressor
04.32	TOP-A High-level Ethylene Exchanger
04.32	BOT-A Intermediate-level Ethylene Exchanger
04.32	TOP-B High-level Methane Exchanger
04.32	BOT-B Intermediate-level Methane Exchanger
04.33	Low-level Methane Exchanger
04.34	Main Methane Cooler
04.35 T	High-pressure Methane Liquid Subcooler
04.35 M	Intermediate-pressure Methane Liquid Subcooler
04.35 B	Low-pressure Methane Liquid Subcooler
04.36 A,B	Low-level Ethylene Exchanger
04.37	Main Ethylene Cooler
04.38 T	High-pressure Ethylene Liquid Subcooler
04.38 M	Intermediate-pressure Ethylene Liquid Subcooler
04.38 B	Low-pressure Ethylene Liquid Subcooler
04.39 T	Main Propane Cooler
04.39 B	Low-pressure Propane Liquid Subcooler
04.40	Low-level Propane Exchanger
04.41	High-level Propane Exchanger
04.50	High-pressure Nitrogen Liquid Subcooler
04.51	High-level Nitrogen Evaporator
04.52	Intermediate-pressure Nitrogen Liquid Subcooler
04.53	Intermediate-level Nitrogen Evaporator
04.54	Main Nitrogen Cooler
05.28	Low-level Nitrogen Evaporator
07.90	First Ortho-para Hydrogen Converter
08.41	Feed Gas Impurities Adsorber (Charcoal)
08.45	08.41 Reactivation Cool-down Exchanger

The high-level refrigeration systems studied include the cascade and the high-pressure nitrogen recycle systems. Figures 3-21 and 3-22 are schematic representations of these two systems respectively. The low-level system studied was a 1500 psia hydrogen recycle system. This system is shown schematically in figure 3-23.

3.3.2.1 Cascade Refrigeration System

Prior to entry the gaseous hydrogen is dried and fed into the (+23°F) high-level propane exchanger (04.41, figure 3-21). The feed continues to the low-level propane exchanger (04.40) where it is cooled to -34°F. The refrigerant discharge from the ethylene (01.70), methane (01.60) and nitrogen (01.90) recycle compressors are all in heat exchange with the two propane exchangers. It is from this "cascading" of each constituent against the next higher refrigerant to promote condensing that this cycle gains its name.

The hydrogen feed is cascaded in a similar manner through the high and low level ethylene exchangers (04.32 TOP-A and BOT-A), and methane exchangers (04.32 TOP-B and BOT-B) until the feed is approaching a temperature of -250°F in the last stage methane exchanger (04.33). The cold hydrogen gas is then flashed in a phase separator to remove any hydrocarbon impurities which have liquefied.

The hydrocarbon stream is passed back through the plant to recover refrigeration and is vented as a warm fuel stream.

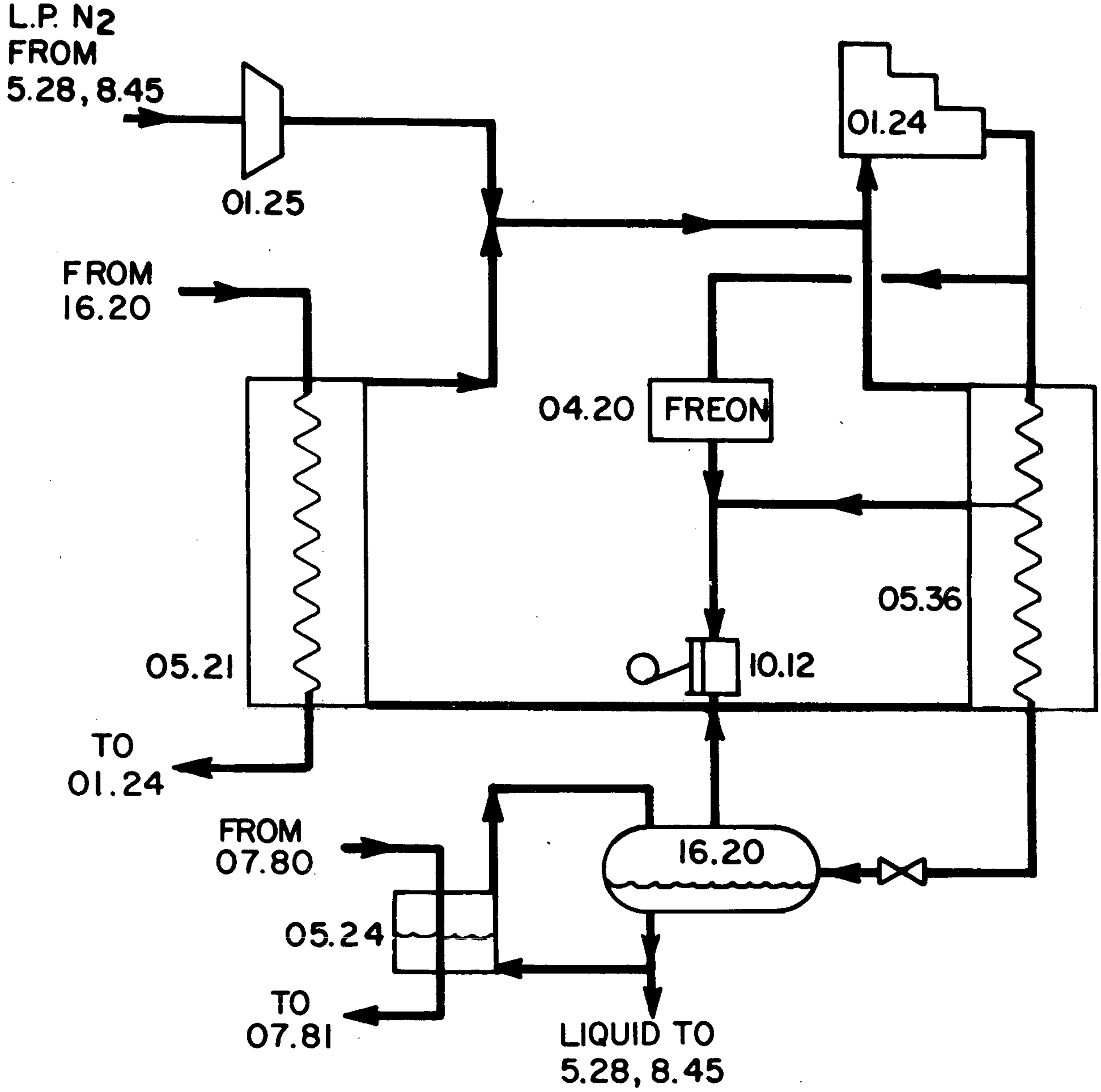
The hydrogen feed then enters the high level nitrogen exchanger (04.51) and makes several passes through the low level nitrogen exchanger (04.53). Any nitrogen impurity in the feed stream is liquefied in this last exchanger and separated from the feed. The liquid nitrogen thus separated is then dumped into the hydrocarbon fuel stream for refrigeration recovery and venting. The feed stream is then cleaned of any remaining impurities in charcoal beds (08.41), and passes to the converter (07.90) for the first stage of ortho-para conversion. The temperature rise due to the conversion reaction requires that the feed gas be recooled once again in the (04.53) LIN bath before entering the low level section of the liquefier.

3.3.2.2 High Pressure Nitrogen Recycle Refrigeration System

The 3000 psi nitrogen recycle system as shown in figure 3-22 is an alternate refrigeration method to the cascade system.

Gaseous nitrogen is compressed to 3000 psia in the nitrogen recycle compressor (01.24). The discharge is separated into two streams. One stream passes through exchanger (05.36) where it is cooled by the returning cold nitrogen vapor. The high pressure stream is then flashed into the liquid nitrogen storage tank (16.20). This tank is a source of refrigeration to exchangers (05.28, figure 3-21 and 08.45, figure 3-23) in the low level section of the liquefier. It also provides refrigeration to the high level refrigeration system via heat exchanger (05.24, figure 3-24).

The second portion of the compressor discharge is cooled against Freon in exchanger (04.20, figure 3-22) and expanded in the turbo-expander (10.12), thus providing a second source of refrigeration. This cold gas is combined with the boil-off from the storage tank (16.20) to provide refrigeration to exchanger (05.21) in the low level section of the liquefier and to exchanger (05.36) which is



01.25	Low-pressure Nitrogen Booster Compressor	05.24	Intermediate Level Nitrogen Evaporator
01.24	High-pressure Nitrogen Recycle Compressor	05.36	Main Nitrogen Recycle Exchange
04.20	Freon Unit With Freon Evaporator	10.12	Nitrogen Expansion Engine
05.21	Main Feed to Nitrogen Gas Exchanger	16.20	Liquid Nitrogen Storage Tank

FIGURE 3-22. HIGH-LEVEL HIGH-PRESSURE NITROGEN REFRIGERATION SYSTEM - RECYCLE, REFRIGERATION SECTION

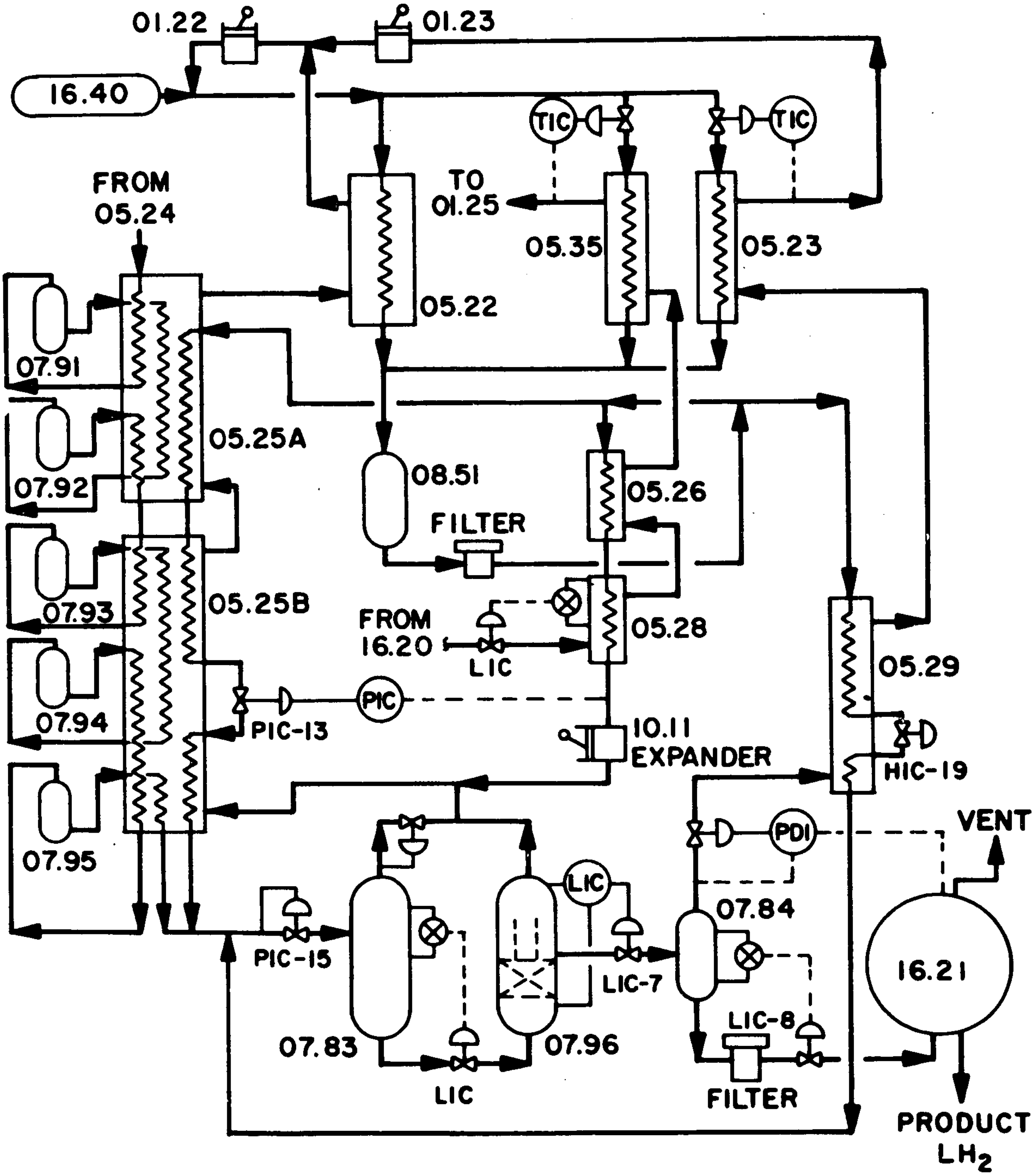


FIGURE 3-23. LOW-LEVEL LOW-PRESSURE NITROGEN REFRIGERATION SYSTEM

LEGEND FOR FIGURE 3-23

01.22	High-pressure Hydrogen Recycle Compressor
01.23	Low-pressure Hydrogen Recycle Compressor
01.25	Low-pressure Nitrogen Compressor
05.22	Main Hydrogen Recycle Exchanger
05.23	Warm Low-pressure to High-pressure Hydrogen Exchanger
05.24	Intermediate Level Nitrogen Evaporator
05.25 A&B	Feed Converter Exchangers
05.26	Hydrogen Expander Precooler Nitrogen Gas Exchanger
05.28	Low-level Nitrogen Evaporator
05.29	Cold Low-pressure to High-pressure Hydrogen Exchanger
05.35	Nitrogen to High-pressure Hydrogen Warm Exchanger
07.83	First Liquid-vapor Hydrogen Separator
07.84	Second Liquid-vapor Hydrogen Separator
07.91	Second Ortho-para Hydrogen Converter
07.92	Third Ortho-para Hydrogen Converter
07.93	Fourth Ortho-para Hydrogen Converter
07.94	Fifth Ortho-para Hydrogen Converter
07.95	Sixth Ortho-para Hydrogen Converter
07.96	Final Ortho-para Hydrogen Converter
08.51	Guard Adsorber
10.11	Hydrogen Expansion Engine
16.20	Liquid Nitrogen Storage Tank
16.21	Liquid Hydrogen Storage Sphere
16.40	High-pressure Hydrogen Gas Storage Bottle (for 08.41 Purge)

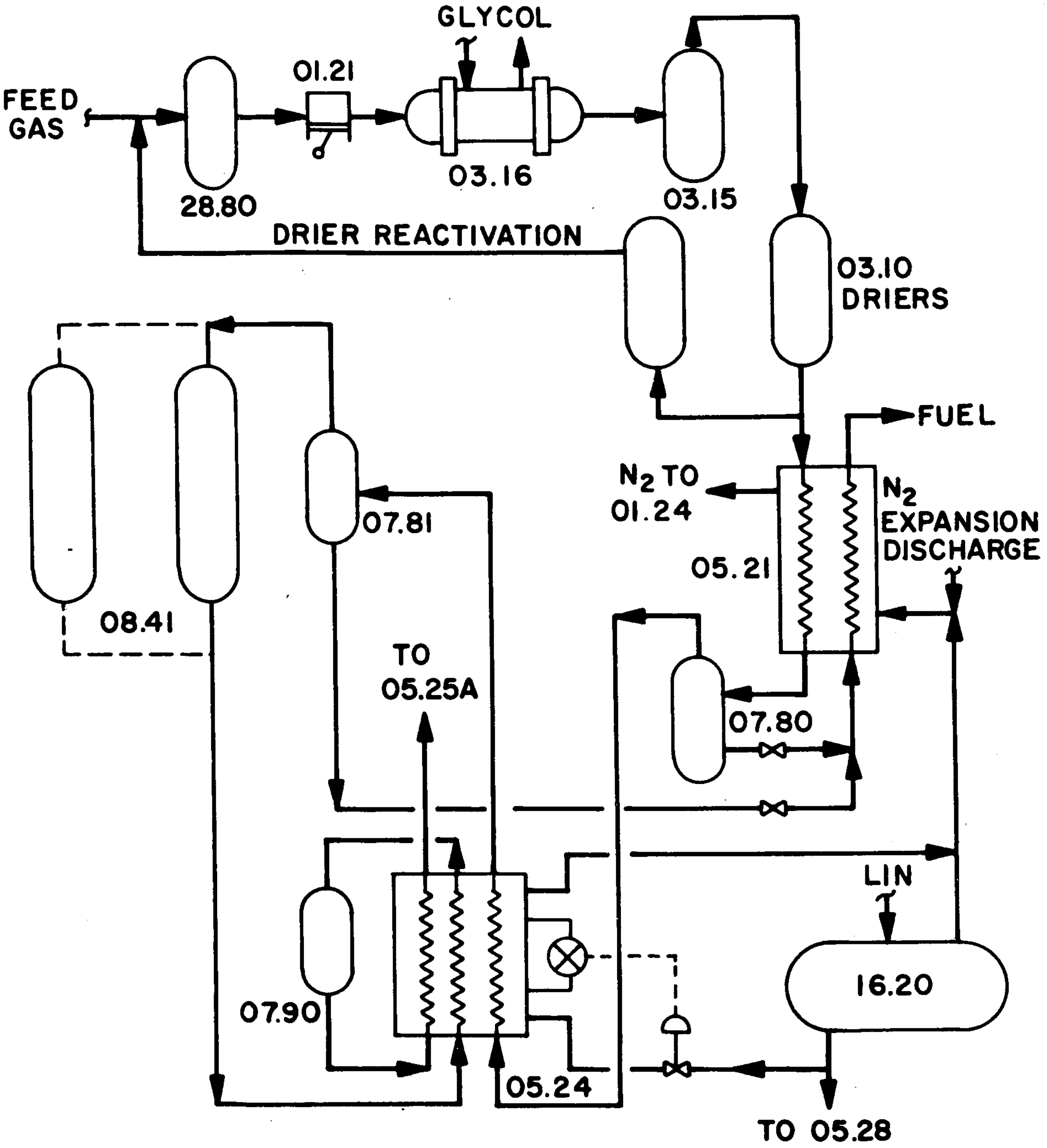


FIGURE 3-24. HIGH-LEVEL HIGH-PRESSURE NITROGEN REFRIGERATION SYSTEM - RECYCLE FEED PURIFICATION SECTION

LEGEND FOR FIGURE 3-24

01.21	Feed Gas Compressor
01.24	High-Pressure Nitrogen Recycle Compressor
03.10	Feed Pressure Driers
03.15	Feed Pressure Condensate Separator
03.16	Feed Drier Precooler
05.21	Main Feed to Nitrogen Gas Exchanger
05.24	Intermediate Level Nitrogen Evaporator
05.25 A	Feed Converter Exchanger
05.28	Low-level Nitrogen Evaporator
07.80	First Feed-Liquid Methane Phase Separator
07.81	Second Feed-Liquid Methane Phase Separator
07.90	First Ortho-para Hydrogen Converter
08.41	Feed Gas Impurities Adsorber (Charcoal)
16.20	Liquid Nitrogen Storage Tank
28.80	Feed-Condensate Separator

the primary recycle system exchanger.

Returning low pressure nitrogen from exchangers (05.28, figure 3-21 and 08.45, figure 3-23) is boosted to (01.24, figure 3-22) suction pressure in booster compressor (01.25) prior to recirculation in the refrigerator.

3.3.2.3 Low Level Refrigeration System

Hydrogen feed gas at liquid nitrogen temperatures from exchanger (05.24, figure 3-23) in the high level section of the liquefier is cooled by expander (10.11) exhaust and converted in a series of five adiabatic reactors (07.91 through 07.95). The heat of reaction is continuously removed during the ortho-para conversion in exchangers (05.25 A&B). The 95 percent para hydrogen is then flashed into reactor (07.83) where liquid hydrogen is formed. The liquid formed in (07.83) passes into (07.96) which is a final conversion catalyst bed. The product stream leaves (07.96) and is flashed into (07.84) thus providing the refrigeration needed to remove any heat of reaction generated in (07.96) and to sub-cool the product prior to entry in storage.

Flash gases from (07.83 and 07.96) combine with the expander exhaust, providing refrigeration to the feed stream in exchangers (05.25 A&B). In this manner the gas is warmed to ambient temperatures and is compressed for recycling in compressor (01.22). From the compressor the recycle stream is cooled to approximately -292°F in exchangers (05.22, 05.23 and 05.35). It then passes through an adsorber (08.51) where trace impurities are removed. The stream is then divided into three portions. The major portion is precooled against liquid nitrogen in exchanger (05.28) and then expanded in the hydrogen expander (10.11). The second stream is cooled in the exchanger (05.25) and combined with the feed. The third stream is cooled down in the exchanger (05.29) and also combined with the feed prior to entry into phase separator (07.83).

3.3.2.4 Process Elements

Process studies have been completed to show certain relationships exist between the various process elements and plant capacities. The nature of those relationships are described in the following paragraphs.

3.3.2.4.1 Raw Material Unit Consumption. – The raw material requirements for the liquefier differ somewhat from the requirements of the gaseous production systems of the total facility. The primary raw material input to the liquefier is the gaseous hydrogen stream itself. This hydrogen is a "cost" to the liquefier only to the extent that the gas is consumed or lost in the process. The total hydrogen loss in the liquefaction process is estimated to be 4 percent.

A certain quantity of hydrogen is lost due to leakage around the hydrogen recycle and booster compressors. Hydrogen is required for continuous cold box purge, and finally hydrogen gas is required as a clean and purge medium for the low-temperature adsorbers. The compressor losses are a direct function of hydrogen flow rates, the cold box purge is a function of the cold box volume, and the adsorber purge is also a function of product flow through the liquefier. Therefore, with the possible exception of the small cold box purge requirement, the hydrogen losses are a direct function of plant size and are constant.

Another process element which can be considered as a raw material requirement is the makeup refrigerant. Here again a slight loss must be anticipated around the compression equipment. These losses are expected to amount to 0.1 percent of the compressor flow rates and also will be constant with relation to plant size. Certain quantities of nitrogen are also required for purge and general utility usage.

The quantities of the various raw materials required for the liquefaction process are shown in table 3-5. The process selected consists of the cascade system and the 1500 psia hydrogen recycle system. This combination offers the economic optimum with respect to power consumption, capital cost, raw material requirements and operating costs.

TABLE 3-5. RAW MATERIALS FOR LIQUEFACTION

Raw Material	Units
Feed Gas (Hydrogen)	1.04 lb/lb
Refrigerant Makeup	
Propane	0.02 lb/lb
Ethylene	0.02 lb/lb
Methane	0.01 lb/lb
Nitrogen	0.04 lb/lb

3.3.2.4.2 Process Energy Unit Consumption. – Process energy for the liquefaction system represents one of the largest single cost items in the entire purification, and liquefaction process train. For this reason, considerable emphasis has been placed upon the study of the refrigeration system to be employed and the optimization of compression equipment to minimize the power consumption rates per unit of product.

In making this study, an attempt was made to predict the maximum probable advance in compressor technology which may occur in the next fifteen years.

The reciprocating compression equipment necessary for various services will be available in 1980. The efficiency values used to calculate the power requirements are today's industry standards. It is anticipated that the same values will be used over the next 10 to 20 years. Advancements to be made in reciprocating compressor technology are expected to be primarily metallurgical and will permit greater rod loads and higher discharge pressures. It is also probable that the state-of-the-art for non-lubricated services will be increased. Technology has increased to the point where all of the services considered in this study will employ non-lubricated cylinders.

Assuming that high strength materials technology continues to advance at its present pace, that reasonable lead time is provided and that the compressor industry is properly stimulated, it appears that centrifugal compressors for this service can be available by the early 1980's. Current industrial multistage gas compressors use closed type backward swept bladed impellers operating at approximately 1050 ft/sec tip speed. High head rise compressors use open type radial bladed impellers and run at tip speeds of 1250 to 1350 ft/sec. Experimental impellers of this type have been successfully tested to 1900 ft/sec. For this system, a tip speed of 1700 ft/sec was selected as realistic for the sizes required. Open radial vaned wheels most likely would be used. This means that in multistage casings, considerable clearance allowance must be provided between the impeller and the side wall to avoid rubbing during thermal transients.

The centrifugal compressor horsepowers have been calculated using efficiency values ranging from 5 to 12 percent higher than industrial practices in use today. We expect that over the next 10 to 15 years the efficiency will be raised with technological advances to the levels we have assumed.

The compression equipment study indicates that the cascade system offers the most economically attractive approach to the high-level refirgeration requirement. The horsepower requirement associated with this method is 44,200 hp for a 250 T/D plant. Centrifugal compressors can be utilized for all four recycle services (propane, ethylene, methane, and nitrogen) in the cascade process. By way of contrast, the high pressure nitrogen recycle system requires approximately 64,700 Btu for a 250 T/D plant.

The low level refrigeration system, for a 250 T/D plant, requires an additional 69,400 hp for the hydrogen recycle and hydrogen booster compressors. These machines would both be of the reciprocating type. Thus, the gross energy requirement for liquefaction is 113,600 hp. During plant operation, the hydrogen expander will return 4200 hp to the system. The return of horsepower is achieved by loading the expander with an electrical generator. Therefore, the net energy requirement for the system is 109,400 hp for a 250 T/D plant.

An analysis was made to determine the value of the facility power requirements for recovering the energy of high pressure feed from the electrolysis process. It was determined that with a feed gas stream at 100°F and 5000 psia being expanded to approximately 640 psia, 34.8 bhp/TD (0.034 kwh/lb) of liquid hydrogen leaving the liquefaction plant can be recovered from the feed. Also, the additional plant investment and associated maintenance costs, attendant to the equipment necessary to the recovery of this power, far outweighed the cost saving in recovered power. Therefore, no credit has been taken for high pressure feed gas streams.

Liquefaction process energy will remain constant per unit of production as the plant size increases from 250 T/D to 2500 T/D because plant expansion will be effected through the use of multiple production trains. Therefore the process energy unit consumption is 4.46 kwh/lb.

3.3.2.5 Cost Elements

Capital investment and operating costs, other than raw materials and energy, have been studied over the selected capacity range of 250 T/D to 2500 T/D of liquid hydrogen production. Both capital investment and operating costs were deemed to vary with plant size exponentially. The following paragraphs discuss these relationships.

3.3.2.5.1 Capital Investment. – Capital investment data for the complete liquefaction system were developed, based on historical costs for plants utilizing similar equipment methods. The historical data has been adjusted and refined to reflect those changes in process and equipment technology as discussed in other paragraphs of this study.

As plant capacity increases, the capital costs are expected to increase in an exponential manner. This relationship is best expressed by Williams' formula;

$$I = I_B \left(\frac{C}{C_B}\right)^{M_I}$$

The estimated capital investment (I_B) for the base system utilizing the cascade/1500 psia hydrogen recycle system is \$31.5 x 10^6. The value of the exponential constant (M_I) for the liquefier is 0.80. These cost figures are the result of a detailed engineering design and estimate for a 250 T/D facility.

3.3.2.5.2 Operating Costs. – Operating costs described in this paragraph include labor, chemicals and lubricants, catalysts, maintenance costs, general and administrative costs, and home office allocations. It is significant to note that taxes and insurance are not included in this area but are contained in the plant capital charge.

At the base case (250 T/D) it has been determined that the liquefier would require seventeen men. This is equivalent to three men per shift for plant operation and maintenance, four shift supervisors and one plant superintendent. As the plant capacity increases, the labor force will increase proportionately so that at the 2500 T/D level twenty five men will be required. This relationship exists because plant expansion will be accomplished through the parallel operating train concept. These men will also be part of a larger labor pool because the liquefaction plant is in combination with the gaseous production plant.

Chemicals, lubricants and catalysts requirements will increase directly with plant capacity as is common with all industrial gas and liquefying facilities. Chemicals and lubricants have been included at the rate of 0.08 cent/lb at the 250 T/D level and 0.07 cent/lb at the 2500 T/D level. Maintenance costs will vary directly with plant investment at a rate of 2 to 3.5 percent/year rather than with capacity. It must be noted that maintenance costs are those costs associated with keeping the plant in good repair and do not include costs associated with major improvement of equipment. Major equipment improvement costs are considered to be additional capital investment and are recovered through depreciation.

Home office allocation has been included as a function of plant level operating costs including straight line depreciation of the plant equipment at a rate of 5 percent.

General and administrative costs are commonly expressed as a percentage of all the above costs and is a method of recovering plant level overhead costs at a rate of 15 percent.

When all of the above items are compared to plant capacity as a single group it is found that the relationship is exponential in nature, following the form:

$$O = O_B \left(\frac{C}{C_B}\right)^{M_O}$$

The estimated value for O_B is \$1.91 x 10^6 per year for a 250 T/D plant. The operating cost exponent M_O is expected to be 0.65. The relationship of the various elements to total annual operating cost as plant capacity increases is shown in figure 3-25.

3.4 STORAGE OF LIQUID HYDROGEN

The third major subsystem of a liquid hydrogen production facility is the product storage and transfer equipment. The following paragraphs provide a technical description of the various system

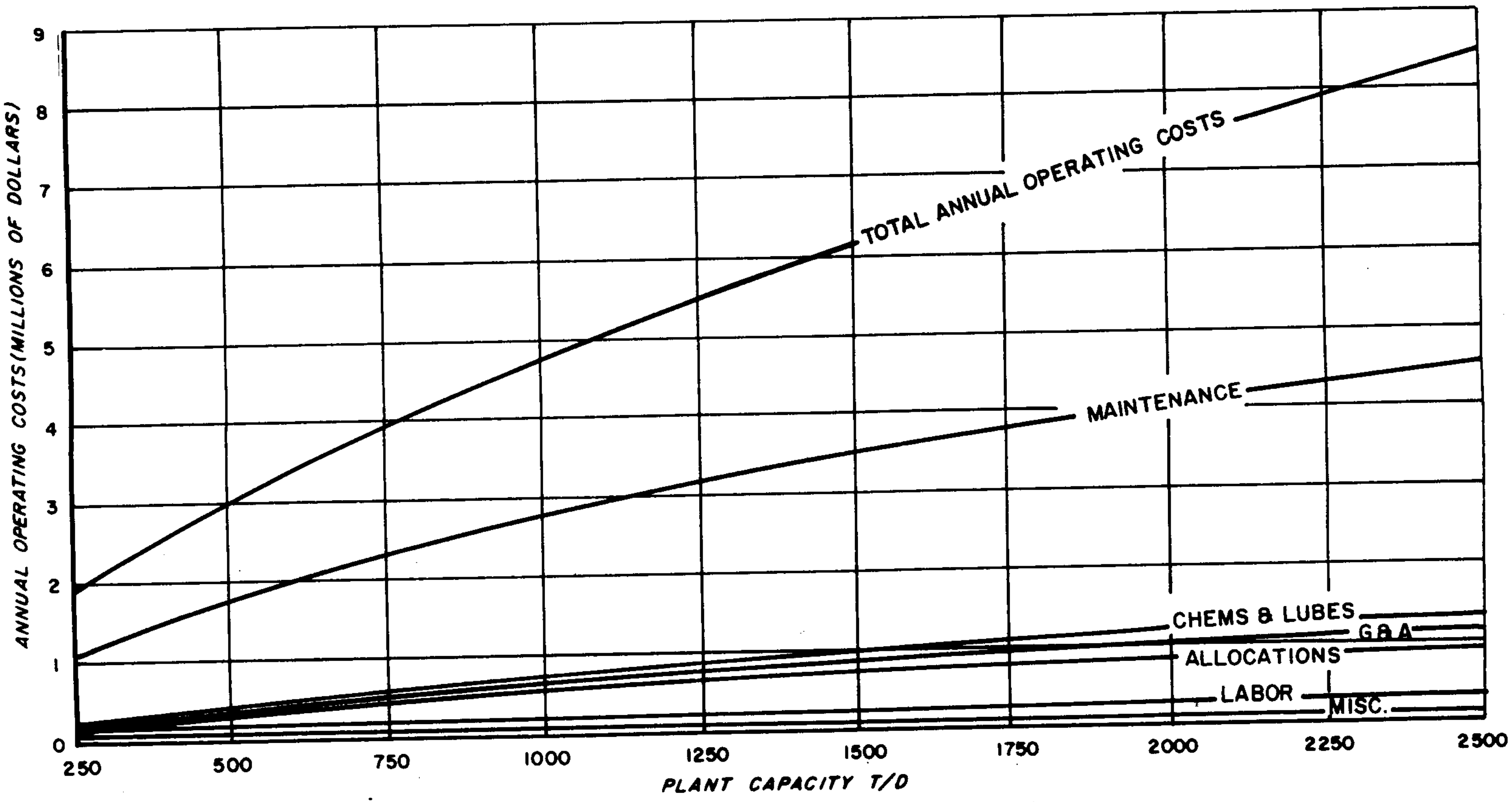


FIGURE 3-25. ESTIMATED OPERATING COST vs PLANT SIZE

configurations available for storage of large quantities of liquid hydrogen. Each configuration has been evaluated to determine salient features such as system heat-leak and corresponding product boil-off rates, insulation characteristics and requirements, and capital costs associated with each system configuration.

The study of storage equipment has been segregated into two distinct categories:

(1) Underground storage

(2) Above-ground storage

3.4.1 ABOVE-GROUND STORAGE OF LIQUID HYDROGEN

Considered for above-ground storage are four basic configurations. These are:

(1) Single Wall Tanks –
  a. Externally insulated
  b. Internally insulated

(2) Double Wall Tanks –
  a. Purged powder insulated
  b. Vacuum insulated

3.4.1.1 Single Wall Externally Insulated Storage Tank

The single wall externally insulated tank as shown in figure 3-26 incorporates the use of standard bulk type insulating materials. These materials can be applied in board form, strapped in place, or can be applied as foamed material which is sprayed or poured in place. The common insulating materials include foamglass, fiberglass, urethane, styrene, epoxy, and silicone rubber.

The present method of applying board type insulation utilizes the use of bands to hold the insulation against the tank wall. In order to hold the boards in place as tank contraction occurs, the bands are prestressed. Use of this type insulation system would require a high degree of prestressing for hydrogen service, because of the greater tank movement. For this reason, board type insulation is not desirable for hydrogen service.

Sprayed-on foam, poured-in-place foam, or adhered boards will eliminate the banding problem. Through proper bonding techniques the insulating material could be made to move with the tank as it contracts. However, since the insulating materials have coefficients of thermal expansion differing from that of the tank, the insulation must be capable of absorbing the high differential movement. Brittle materials, such as foamglass, would be prone to break-up or delaminate under these conditions and would therefore be unacceptable for this service.

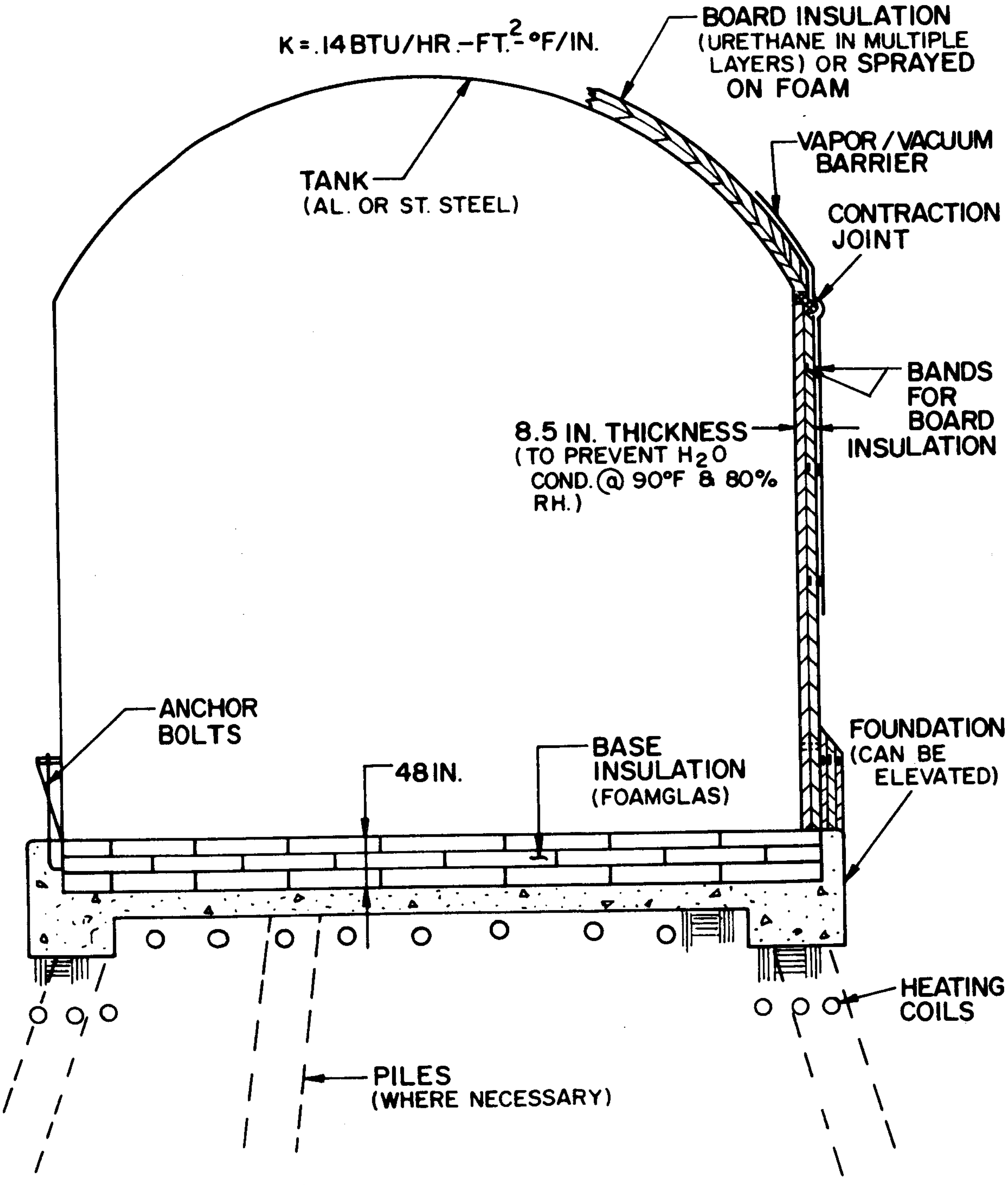


FIGURE 3-26. SINGLE WALL EXTERNALLY INSULATED TANK

The proper choice of insulation material for single wall externally insulated tanks (if considered) would appear to be urethane foam that is either sprayed or poured-in-place. This material exhibits low thermal conductivity (0.14 Btu/hr ft^2 -^{o}F/in at 70^{o}F), has a low density (2.0 lb/ft^2), and has a good resistance to water vapor transmission. It also has the compressive strength (20 to 40 psi) to withstand the vacuum loading which will occur. Urethane exhibits superior properties when compared to foamglass for this service in all areas except flammability. However, new flame retardant additives for urethanes are constantly being developed.

The significant factors associated with the externally insulated (urethane foam) storage tank of various capacities is shown in table 3-6. This information is based upon an insulation thickness sufficient to prevent water condensation from forming on the outside surface of the vapor barrier under conditions of 90^{o}F and 80 percent relative humidity.

TABLE 3-6. URETHANE FOAM INSULATION (EXTERNAL)

Capacity	Insulation Thickness	Heat Leak Btu/hr	% Boil-Off[a] Per Day	Boil-Off[b] Cost $/yr	Storage System Cost $
6×10^5 lb H_2	8.5 in.	1.16×10^5	2.49	5.3×10^5	5.4×10^5
18×10^5 lb H_2	8.5 in.	2.14×10^5	1.53	9.8×10^5	9.6×10^5
36×10^5 lb H_2	8.5 in.	3.43×10^5	1.22	16.5×10^5	15.6×10^5
72×10^5 lb H_2	8.5 in.	7.17×10^5	0.95	24.2×10^5	27.5×10^5

[a] Assuming 10 cents/lb liquid hydrogen value.

[b] Boil-off due to heat leak only. Losses due to ortho-para conversion are not included.

The cost data presented is based upon cost information supplied by an erector-contractor[28], or original estimates prepared for this study. In all cases, it has been assumed that operating and/or maintenance costs associated with these storage systems are negligible. The costs reflect the projected 1980 technology and are expressed in present dollars.

3.4.1.2 Single Wall Internally Insulated Storage Tank

The internally insulated single wall storage tank as shown in figure 3-27 utilizes bulk type insulation placed on the inner surface of the tank shell. The insulation is either adhered to the wall, sprayed-on, or poured-in-place. Block insulation can be applied by merely stacking and is held in place with a liner which is required as a liquid-vapor seal.

Insulating a large liquid hydrogen storage tank internally represents a significant advance in the state-of-the-art, and would require acute detail to obtain a satisfactory installation. This particular concept was considered for the Saturn S-IV and S-IVB.[29] The insulation used is a rigid polyurethane reinforced with fiberglas threads. The liner is No. 116 Fiberglas cloth bonded to the urethane with a polyurethane resin, and sealed with six spray coats of polyurethane resin. The reinforced foam density was 5.2 lb/ft^2 and exhibited thermal conductivity (K values) in the range of 0.24 Btu/hr ft^2 ^{o}F/in.

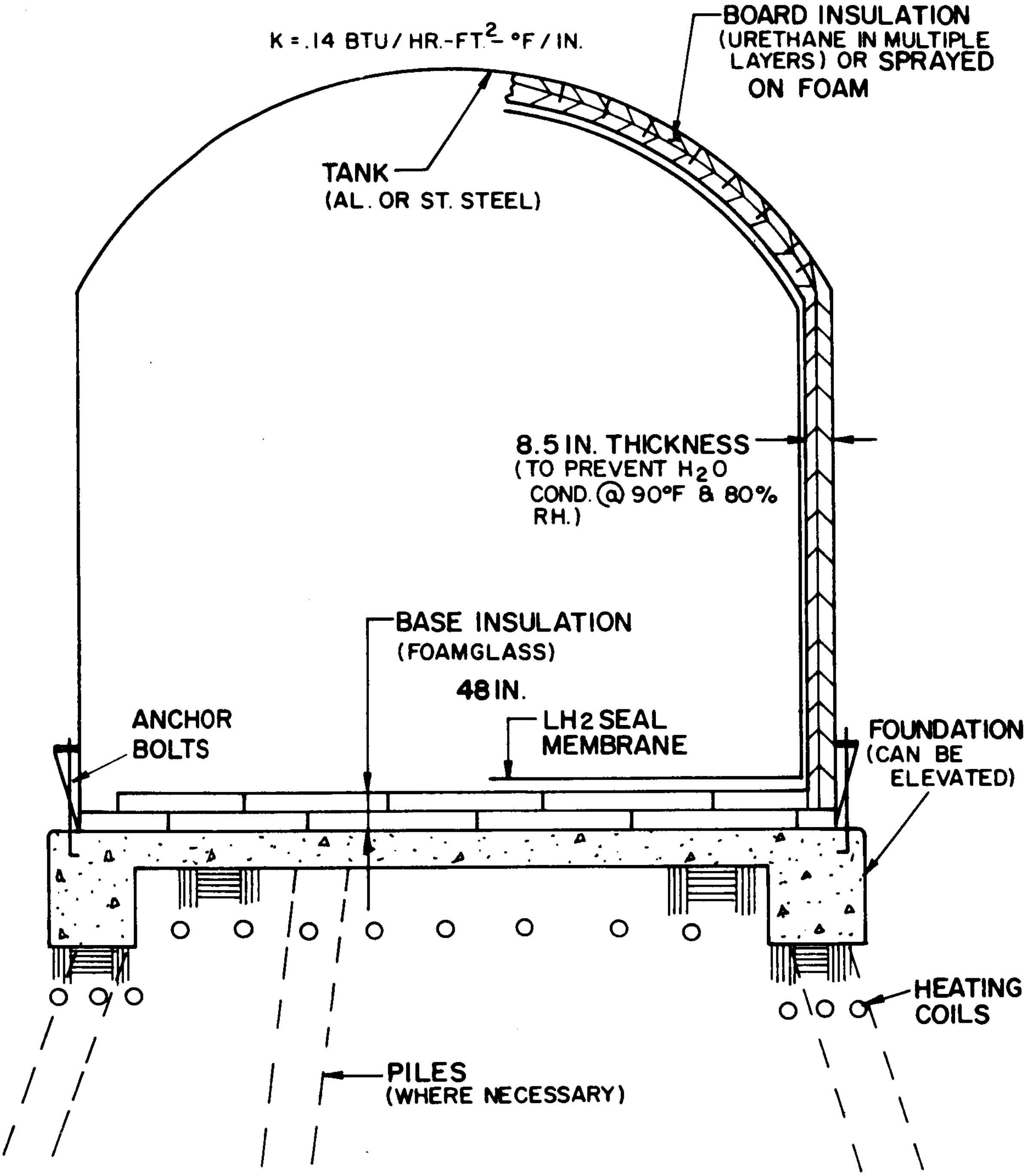


FIGURE 3-27. SINGLE-WALL INTERNALLY INSULATED TANK

to 0.48 Btu/hr ft^2 $^\circ$F/in. This method is not suitable and would require approximately three feet of insulation to prevent condensation on the outer surface at 90°F and 80 percent relative humidity, which is the basis for one set of heat leak calculations.

The material which shows the most promise for this insulation method is rigid urethane foam.

The data for an internally insulated storage tank utilizing the urethane foam material for various capacities is shown in table 3-7.

TABLE 3-7. URETHANE FOAM INSULATION (INTERNAL)

Capacity	Insulation Thickness	Heat Leak Btu/hr	% Boil-Off Per Day	Boil-Off[a] Cost \$/yr	Storage System Cost \$
6×10^5 lb H_2	8.5 in	1.16×10^5	2.49	5.3×10^5	5.7×10^5
18×10^5 lb H_2	8.5 in	2.14×10^5	1.53	9.8×10^5	10.4×10^5
36×10^5 lb H_2	8.5 in	3.43×10^5	1.22	16.5×10^5	17.0×10^5
72×10^5 lb H_2	8.5 in	7.17×10^5	0.95	24.2×10^5	30.0×10^5

[a] Based on product hydrogen cost of 10 cents/lb.

3.4.1.3 Double Wall Tank - Purged Perlite Insulation

The double wall, purged perlite storage system employs an insulation concept as shown in figure 3-28, that has been proven in actual cryogenic service above liquid hydrogen temperatures. The system does not require any significant advance in the state-of-the-art. The major drawback to a purged system which must use hydrogen or helium for purge gas is the fact that these gases exhibit large K values (1 Btu/hr ft^2 $^\circ$F/in and 0.93 Btu/hr ft^2 $^\circ$F/in respectively).

While the system is technically feasible and lies within the state-of-the-art, it is presently undesirable from an economic viewpoint. The system exhibits heat leak characteristics similar to systems previously discussed at a cost in excess of 120 percent of those other systems. The development of a purge system in combination with a thin layer of economical bulk insulation on the outside surface of the inner tank could make the system economical. By increasing the allowable temperature in the purge space, gases with a lower conductivity such as nitrogen ($K = 0.22$ Btu/hr ft^2 $^\circ$F/in) could be used. Nitrogen would also lessen purge gas costs and would be readily available in facilities using the partial oxidation process.

The data for the double wall purged perlite storage system is shown in table 3-8. The purge gas assumed in this data is helium. Cost of purge gas is not included in the cost information presented.

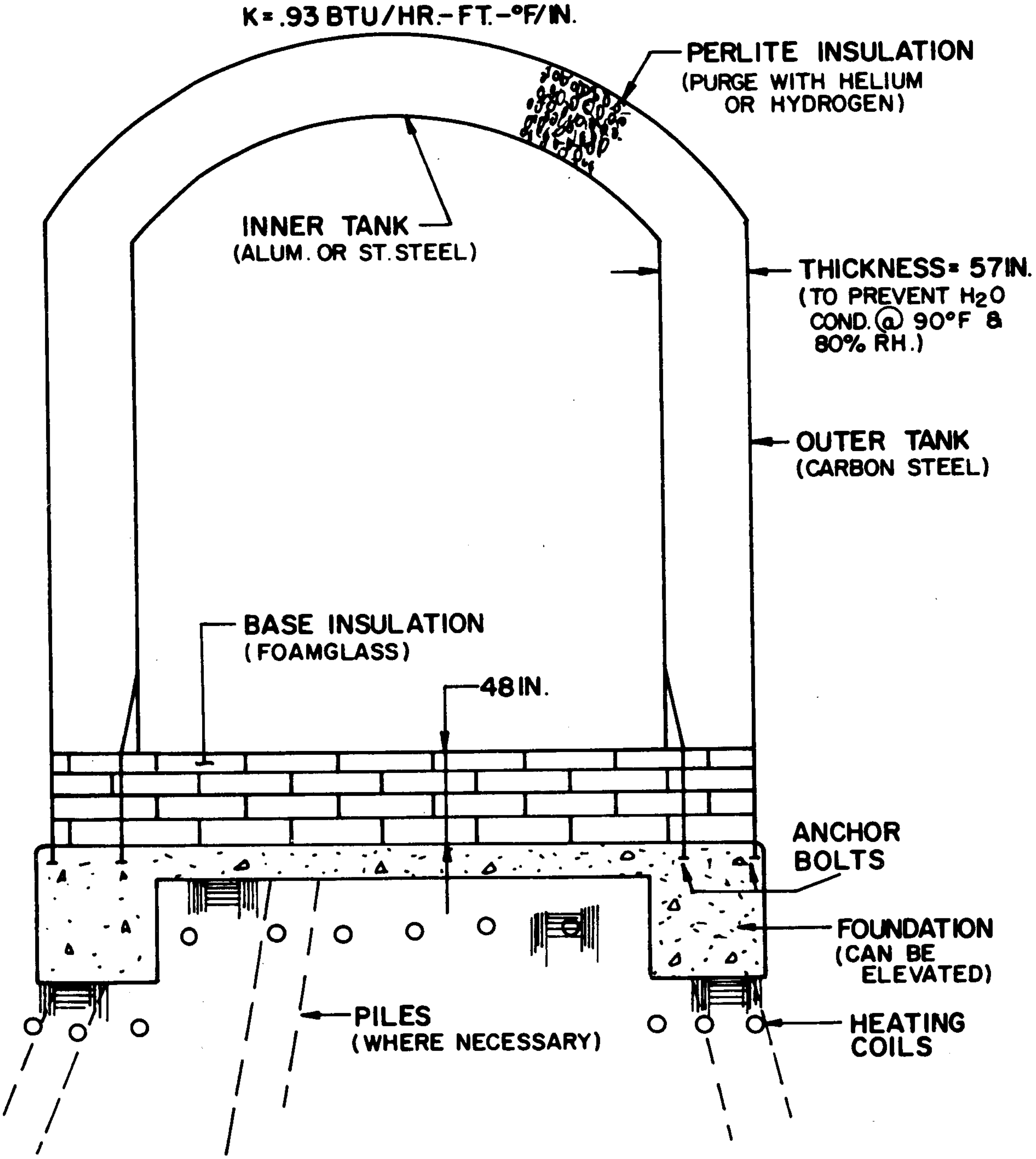


FIGURE 3-28. DOUBLE WALL PURGED PERLITE INSULATION

TABLE 3-8. DOUBLE WALL PURGED PERLITE STORAGE SYSTEM

Capacity	Insulation Thickness	Heat Leak Btu/hr	% Boil-Off Per Day	Boil-Off[a] Cost \$/yr	Storage System Cost \$
6×10^5 lb H_2	57 in	1.2×10^5	2.64	5.7×10^5	6.7×10^5
18×10^5 lb H_2	57 in	2.3×10^5	1.58	10.2×10^5	12.0×10^5
36×10^5 lb H_2	57 in	3.5×10^5	1.25	15.8×10^5	19.6×10^5
72×10^5 lb H_2	57 in	7.2×10^5	0.97	24.7×10^5	35.4×10^5

[a] Assuming a value for liquid hydrogen of 10 cents/lb.

3.4.1.4 Double Wall Tank - Evacuated Perlite Insulation

The double wall, evacuated perlite storage system as shown in figure 3-29 utilizes an insulation concept that has been used almost exclusively for the storage of liquid hydrogen. The insulation system has an extremely low K value (0.0084 Btu/hr-ft^2-^{o}F/in). Technical feasibility of this type storage system has been field proved with many tanks in use at present.

The insulation system exhibits an inherent safety factor in that if the vacuum on the annular area is lost, the perlite continues to exhibit sufficient insulating characteristics to allow in-service tank repair without major loss of product.

A projected variation of this system as shown in figure 3-30 would be the evacuated super insulation system utilizing multiple radiation shields in conjunction with an evacuated space. The thermal conductivity of the evacuated super insulation system is one third that of the evacuated perlite method. This increase in insulation efficiency represents a correspondingly higher initial cost.

The data for the double wall evacuated perlite storage system is shown in table 3-9.

TABLE 3-9. DOUBLE WALL EVACUATED PERLITE STORAGE SYSTEM

Capacity	Insulation Thickness	Heat Leak Btu/hr	% Boil-Off Per Day	Boil-Off[a] Cost \$/yr	Storage System Cost \$
6×10^5 lb H_2	36 in	1.6×10^3	.034	7.3×10^2	7.4×10^5
18×10^5 lb H_2	36 in.	3.2×10^3	.023	14.7×10^2	19.8×10^5
36×10^5 lb H_2	36 in	5.1×10^3	.018	22.7×10^2	37.5×10^5
72×10^5 lb H_2	36 in	8.2×10^3	.015	37.5×10^2	72.0×10^5

[a] Based on liquid hydrogen value of 10 cents/lb.

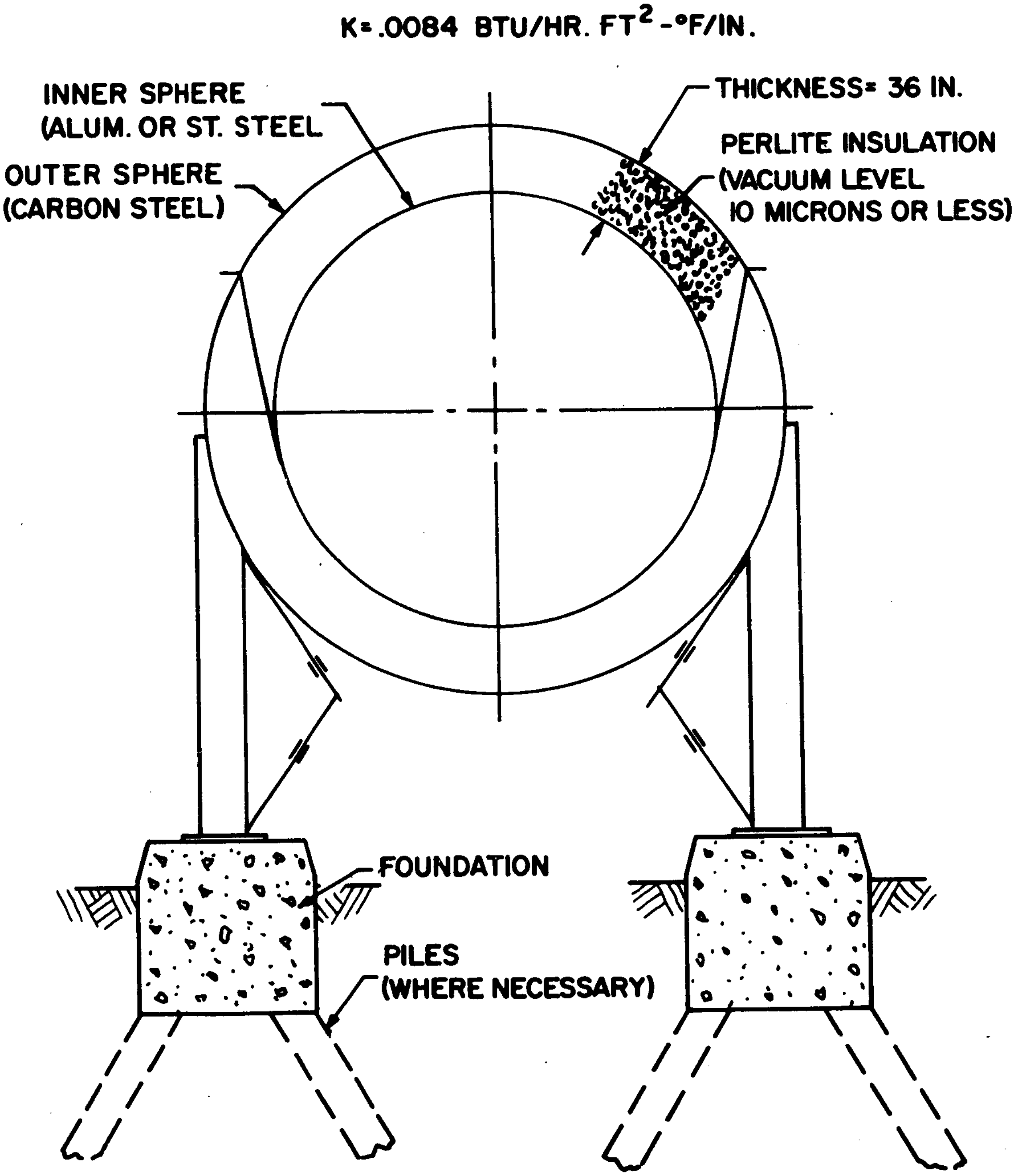


FIGURE 3-29. DOUBLE WALL EVACUATED PERLITE STORAGE SYSTEM

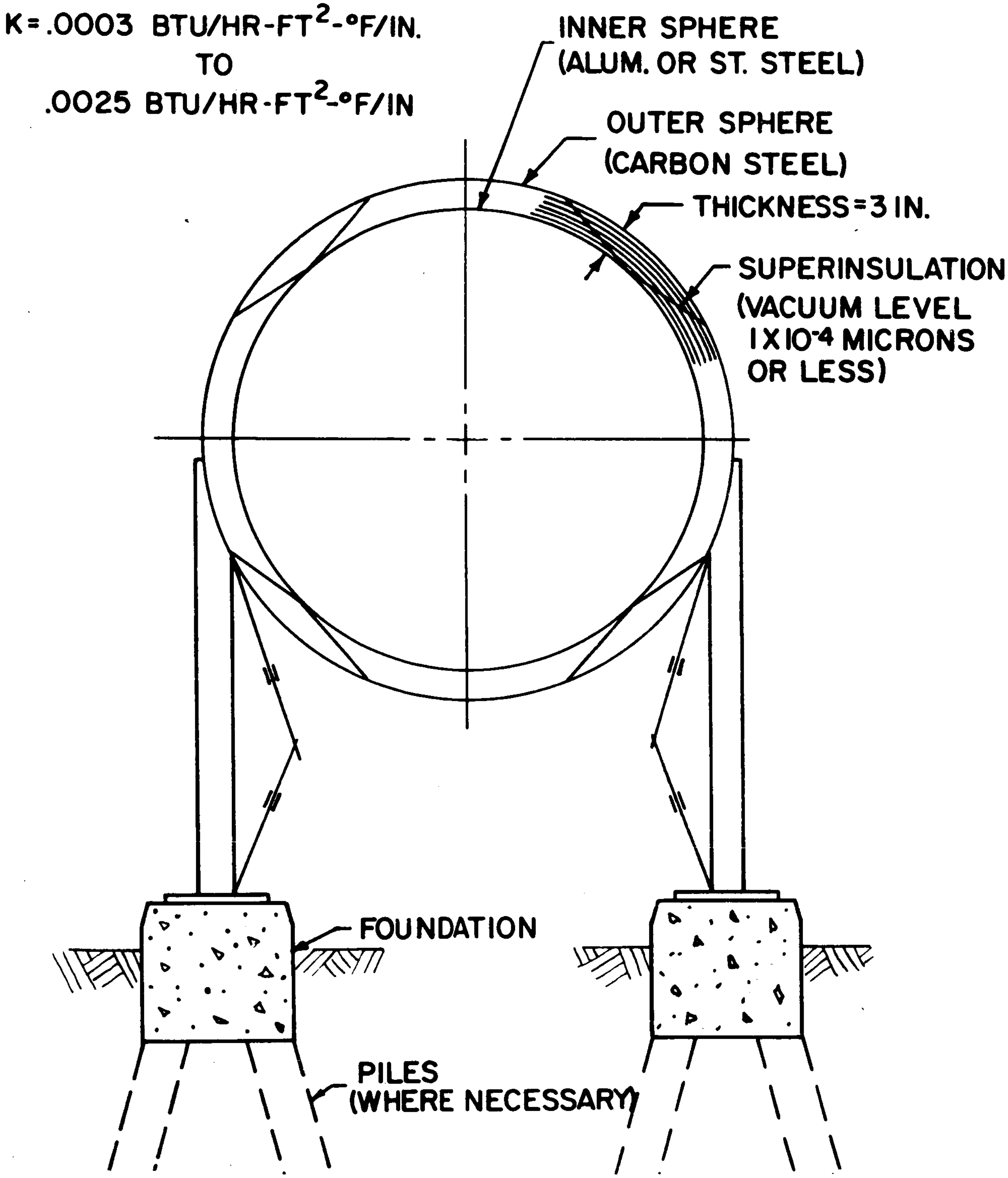


FIGURE 3-30. DOUBLE WALL EVACUATED SUPER-INSULATION STORAGE SYSTEM

3.4.2 UNDERGROUND STORAGE OF LIQUID HYDROGEN

The following paragraphs are presented for technological and background information purposes only. No cost information nor definitive heat leak data is presented. Investigation of the possible use of underground storage requires detailed subterranean information as to type of earthen materials in the area, rock formation, etc.

The investigation of various methods of underground storage in recent years was brought about primarily by the need to store liquefied natural gas (LNG) in large quantities. This need developed when it was discovered that LNG could economically be used for peak shaving purposes by utilities in the United States and for supplying base load requirements in foreign countries. Numerous underground storage facilities are now in use for the storage of LNG and more are in the planning and construction stages.

Below ground storage systems, in general, exhibit some inherent differences from above ground storage systems. They are:

(1) Gradual freezing of the soil adjacent to underground storage areas causes lower temperature gradients than with above ground systems.

(2) The frozen area also contributes some additional thermal insulating value beyond any insulation which is applied.

(3) Liquid spill problems are reduced due to the frozen earth wall surrounding the area.

(4) Underground systems require more soil investigation and are more site dependent than are above ground systems.

Most of the present underground facilities are in the capacity range of 285,000 barrels (approximately 12,000,000 gallons) or greater. This appears to be the range of greatest economic advantage. Underground storage for LNG has apparently been proven satisfactory in large volumes and is seeing increased use.

The change from storage of LNG to that of liquid hydrogen is great since material requirements will differ due to the lower operating temperature. Also, the economics of liquid hydrogen storage will require insulation where LNG is stored without insulation. In addition a liner which is impermeable to hydrogen gas will be required to prevent gas from permeating into the insulation. There should be no appreciable difference in the affect of the liquid hydrogen on the soil from that of the LNG, except for the fact that greater areas around the storage site will be affected due to the greater temperature gradient.

The three variations of underground storage which have been investigated for LNG storage are:

(1) Frozen In-Ground Storage (Excavated Pit).

(2) Prestressed Concrete Tank (Underground).

(3) Underground Cavern Storage.

3.4.2.1 Frozen In-Ground Storage

The frozen in-ground storage system as shown in figure 3-31 is now in use for the storage of LNG. This system has shown an economic superiority to the other underground storage systems to date. It consists primarily of a large open pit with a ring of frozen soil which acts as the insulation and side walls. This wall is previously frozen through pipes which are installed around the perimeter. Brine or propane refrigeration is used to freeze the earth initially and the central unfrozen portion of ground is excavated. The roof is covered with an externally insulated steel cap and backfilled to either partially or completely cover the roof. Generally, only the roof and floor are insulated due to the problems and added expense involved in insulating the walls.

Advantages of this system include:

(1) Safety hazards minimized.

(2) Ground heating not required.

(3) The heat leak characteristics improve as the earth freezes.

Disadvantages of this system include:

(1) Frozen soil is not an effective insulator, therefore, boil-off will initially be quite high, unless insulation is added. Addition of insulation will complicate system and increase the cost of this method.

(2) The system requires special soil conditions of water saturated silty clay or water saturated rock, plus a layer of rock or impenetrable soil directly below the projected base.

(3) With liquid hydrogen temperatures, the soil would eventually be frozen for a considerable distance from the tank. This could cause ground heaving in these areas so that foundations, pipelines, etc., would be affected.

3.4.2.2 Prestressed Concrete Tank Storage Systems

There are two types of concrete tank storage systems. One type of system employs external insulation while the second type of system utilizes internal insulation. The following paragraphs give in detail these two types of systems.

3.4.2.2.1 Externally Insulated Prestressed Concrete Tanks. – The externally insulated prestressed concrete storage system as shown in figure 3-32 has been investigated for the storage of LNG. It incorporates a concrete ring wall prestressed and poststressed in the circumferential and vertical directions, lined with a liquid barrier, and covered with an insulated dome roof. It can be placed totally or partially in-ground. It is constructed in a manner similar to above ground single wall tanks in that the tank walls are supported on a concrete ring wall and the floor is placed over load bearing insulation which has been placed on a prepared subbase. Heating coils are normally required as in above ground tanks which are placed at ground level. Tanks of this type could be built for the sizes being considered in this study.

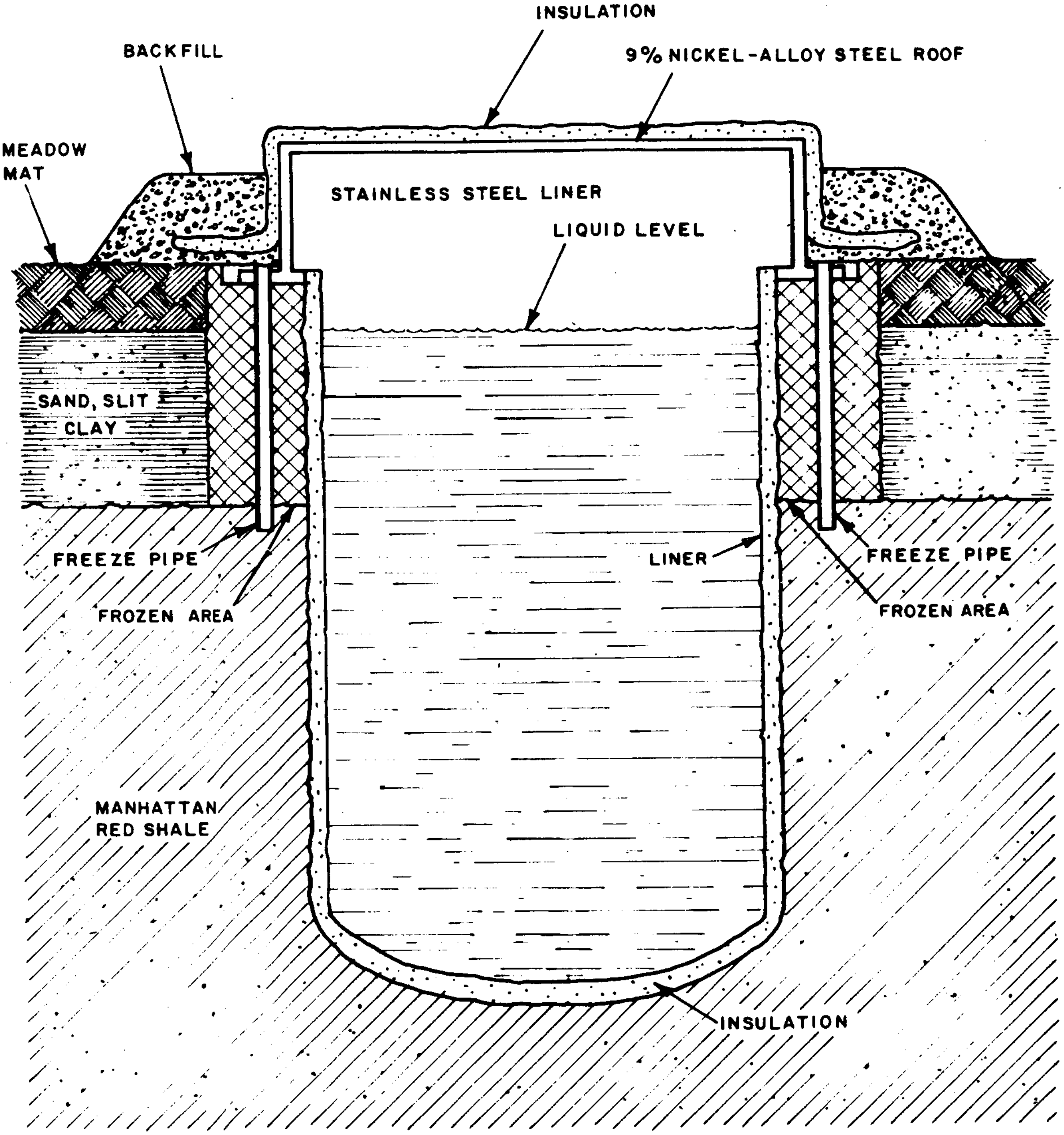


FIGURE 3-31. FROZEN IN-GROUND STORAGE

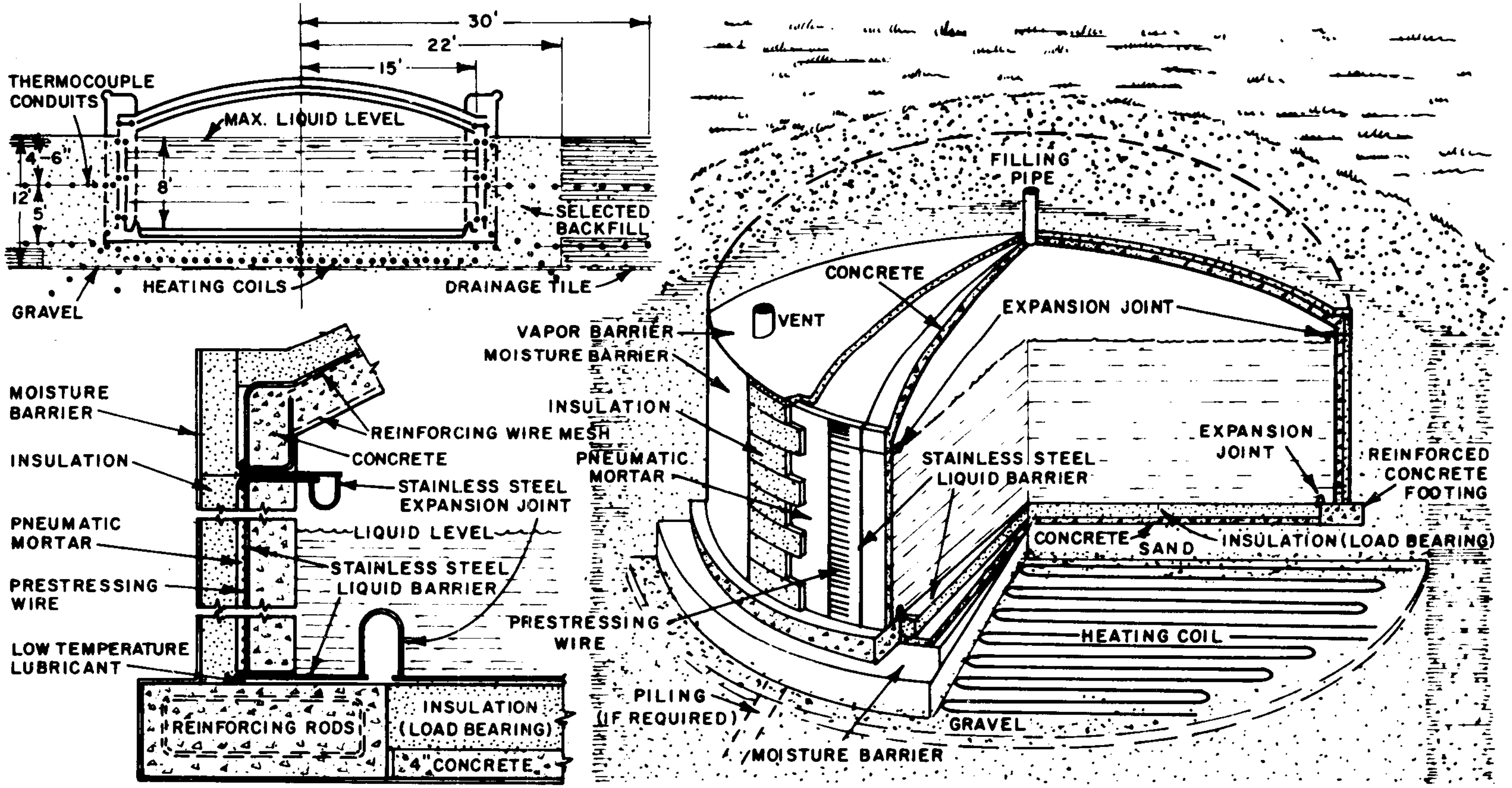


FIGURE 3-32. EXTERNALLY INSULATED PRESTRESSED CONCRETE TANKS

The advantages of this type of construction are:

(1) The heat leak characteristics improve as the surrounding earth becomes frozen.

(2) Underground storage could improve the safety factor.

The disadvantages of this type include:

(1) The concrete ring wall is in contact with the low temperature and the temperature drop across the wall is relatively high. This requires greater prestress loads since larger stresses are induced in the wall due to this temperature differential. This also requires that a controlled cooldown rate be maintained since cold shocking with liquid hydrogen could fracture the wall. The cooldown loss would be high since the large concrete mass is cooled to a low mean temperature.

(2) Tank shrinkage is greater since the tank wall is cooled to low temperatures with the insulation on the outside. If the insulation is made to follow the tank, the soil adjacent to the tank and not yet frozen will flow in and fill the void left by the tank shrinkage. After the tank has been in operation for a while this soil will become frozen. If the tank is then allowed to warm up the frozen soil ring will restrict this movement and the resulting stresses could crush the tank walls. The use of soil stabilizers or a layer of crushable insulation could help to solve this problem.

(3) The insulation requires a vapor barrier to prevent soil moisture from migrating into the insulation and causing an increase in the thermal conductivity.

(4) The insulation and vapor barrier must be rugged enough to take soil back-filling and soil pressures. They are also subject to attack by soil moisture and salts in the soil.

(5) Heating coils are required to prevent floor upheaval of the tank and selected fill is required along the sides to prevent ice lenses from forming.

(6) A frozen soil zone will extend out from the tank so that the area must be kept clear of foundations or anything which would be adversely affected by the frozen condition. Heating coils could be used to limit this migration.

(7) A liquid seal liner is required inside the concrete tank. This has been a stainless steel liner in the past. Invar liners which exhibit a low coefficient of thermal expansion (0.9×10^{-6} in/in $^{\circ}$F vs. 7×10^{-6} for 304 St Stl in the range 70°F to -420°F) are also available and would probably be used for liquid hydrogen tanks. The cost is high, however. The cost of the liner could be up to 40 percent of the tank cost from past experience. Laminate materials could probably be developed, however, which could lower this cost.

3.4.2.2.2 Internally Insulated Prestressed Concrete Tanks. – Internally insulated concrete tanks as shown in figure 3-33 have become the most thoroughly investigated type for underground storage system of LNG. At present a 25 million gallon storage tank is being designed by Texas Eastern Transmission Company for the storage of LNG. This tank is insulated internally with rigid polyurethane foam located inside the concrete walls, floor, and roof. Individual boards 4 by 8 feet are used in three 2-inch layers for a total of 6 inches.

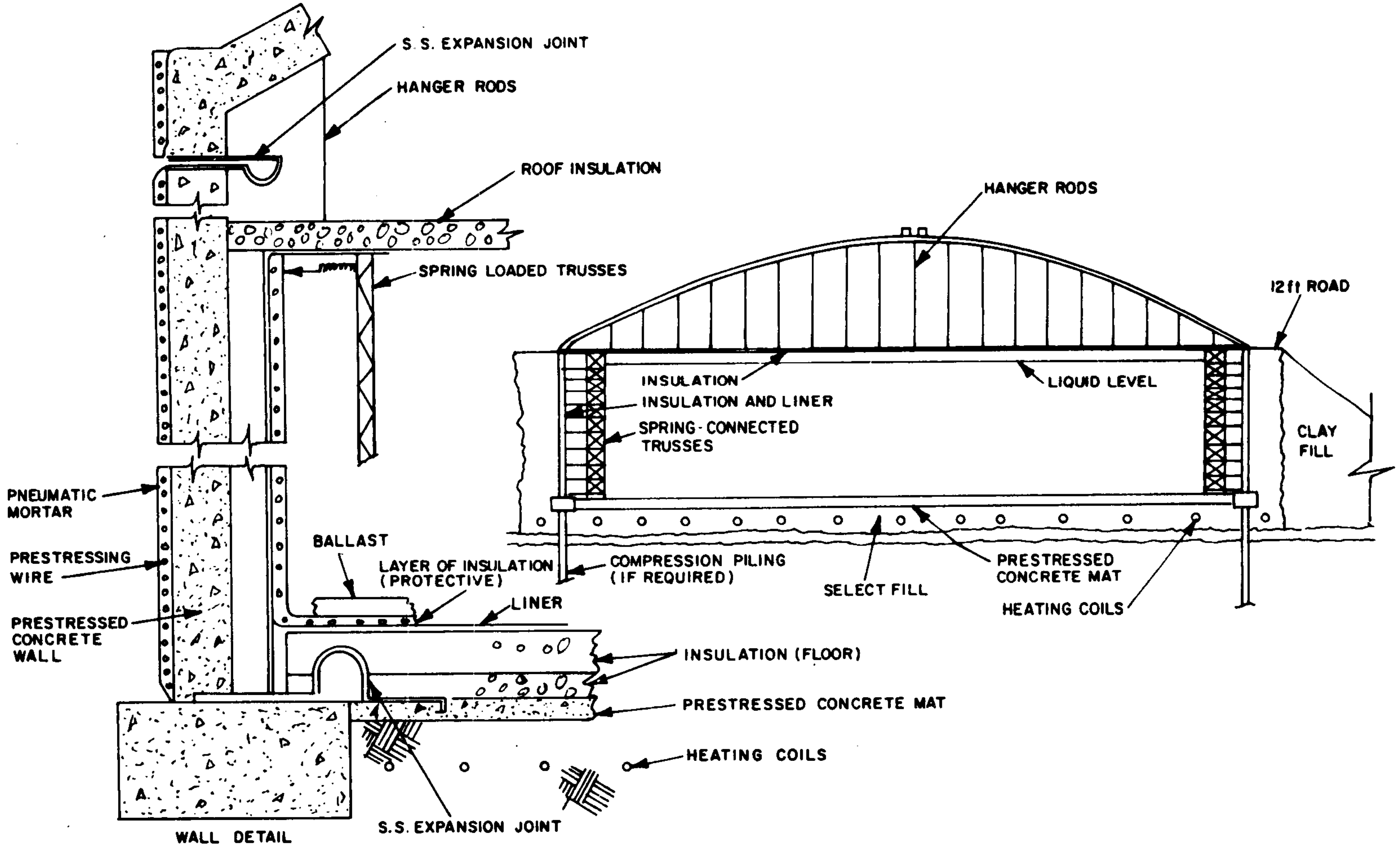


FIGURE 3-33. INTERNALLY INSULATED PRESTRESSED CONCRETE TANKS

Advantages of an internally insulated system would be:

(1) The internal insulation prevents large temperature differentials across the concrete wall. The amount of prestress and poststress required would not be as great and the wall could be slightly thinner since the temperature differential is a major factor in the amount of stress in the wall.

(2) Tank cooldown can be faster since the insulation prevents extreme thermal shock in the concrete wall.

(3) Concrete wall temperatures are higher.

(4) The insulation system does not need added protection from external sources such as backfilling, soil pressure, or soil minerals.

(5) Cooldown losses are lowered somewhat since the concrete mass is not cooled to as low a temperature as externally insulated tanks.

(6) The effective thermal conductivity of the underground system will increase with time.

(7) The tank wall helps to act as a vapor barrier although an additional vapor barrier is still required to gain an adequate perm rating.

Disadvantages to this system are:

(1) The tank has to be free to move in order to eliminate excessive bending stresses in the walls.

(2) Requires heating coils to prevent soil upheaval plus selected soil backfill along the sides.

(3) Soil investigation must be more detailed and successful operation of the tank is somewhat site dependent.

(4) A liquid and gaseous seal is required to prevent hydrogen from permeating the insulation system. This is not as critical as the above ground tanks, however, since the frozen soil will become an insulator and the detrimental effect of the hydrogen gas on the total insulation value would not be as great as with an above ground tank.

(5) The insulation and liner are subject to the hydrostatic pressure of the liquid hydrogen and the liner must be strong enough to withstand this pressure at the seams.

3.4.2.3 Mined Cavern Storage

This method of storage consists of a cavern mined in rock through an inclined passageway capable of admitting trucks and heavy equipment, thereby cutting excavation costs to a minimum. The cavern as shown in figure 3-34 is supported by pillars which remain during the excavation. The passageway is sealed with a liquid barrier and also a vapor barrier further up the passageway. This system has seen a slower development than other methods due to the fact that there is increased surface area

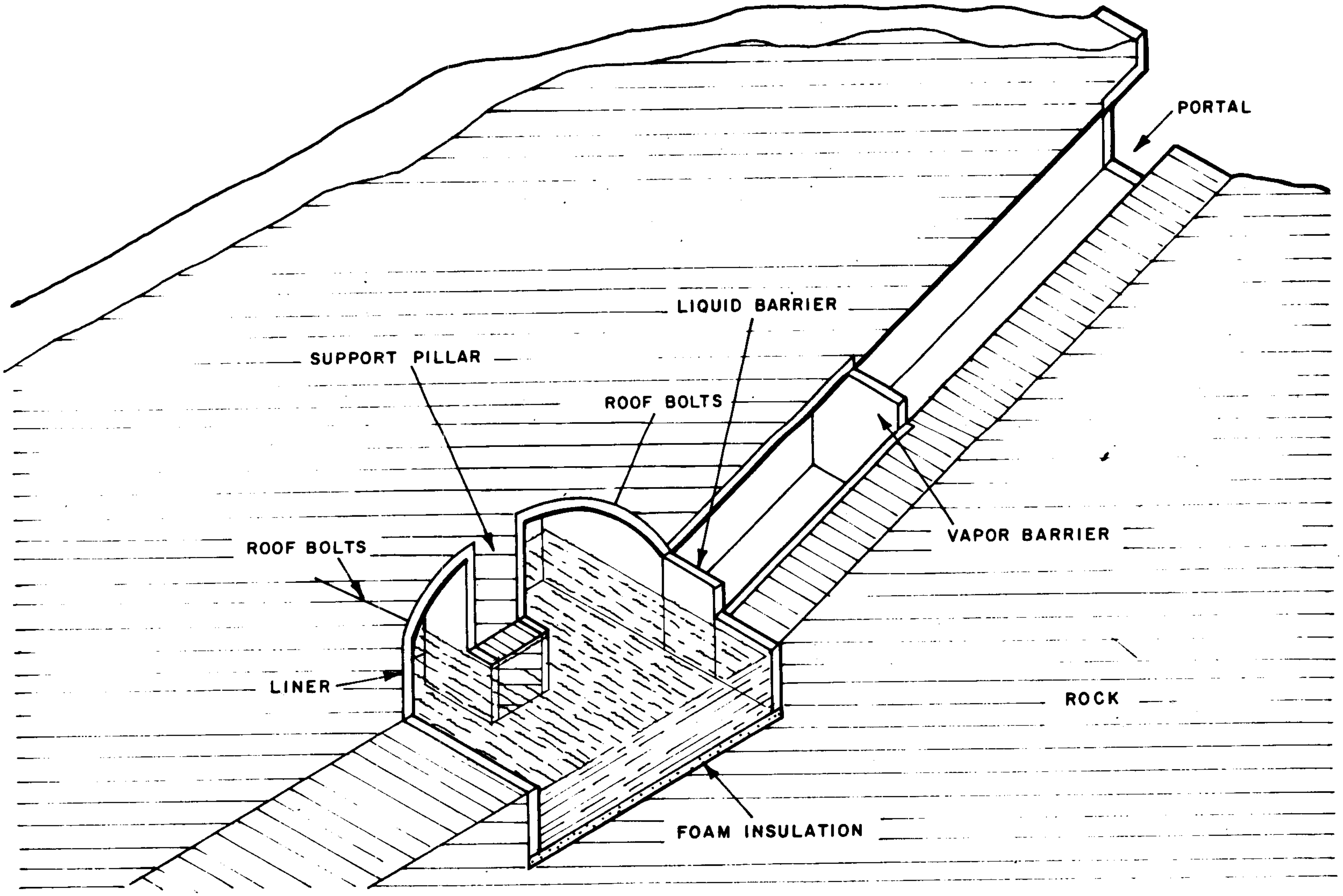


FIGURE 3-34. MINED CAVERN STORAGE

due to the pillars which increases the heat leak into this storage system. The cost of excavating a cavern of this nature has historically been higher than the other methods considered due to the fact that the excavation method involved mining through a small vertical shaft and all equipment had to be disassembled to enter and leave the excavation area through this small access shaft. New methods of excavation have been developed which have lowered the cost. Efforts are being made to develop an insulation system to keep the heat leak within bounds. A test cavern for storage of LNG using the latest known methods is now being completed by the Institute of Gas Technology for the Lowell Gas Company of Lowell, Massachusetts.[30]

Advantages of an excavated cavern would include:

(1) Underground cavern storage would appear to be good from a safety standpoint.

(2) Failure of the insulation system would not necessarily mean that liquid air could form since the vapor seal can be placed in the passageway at a point which is removed from the liquid seal or a double vapor seal could be employed.

(3) Heating coils are not required since the rock is not subject to frost heaving as is the soil in which other storage systems are generally located.

(4) Cavern storage could be designed to store liquid at higher pressures if there were some advantage to it.

Disadvantages of cavern storage would include:

(1) Heat leak through an uninsulated cavern wall and the large surface area presented by the pillars would exclude an uninsulated cavern. This means that an insulation system with a hydrogen seal must be employed as in other systems previously discussed. The insulation and sealing of a cavern would appear to be much more difficult than for a smooth cylindrical surface such as would be the case with a cylindrical tank. Sprayed foam and a laminated liner would probably constitute the most simple system. A system using rigid urethane board insulation fastened to a framing system which, in turn, has been fastened to the rock wall has also been considered.

(2) This system is definitely site oriented in that it requires a rock formation of suitable proportions.

(3) The rock formation must be solid to prevent moisture and water from seeping into the area. Freezing of this seepage water could cause high pressures in the rock and promote cracks in the insulation system and eventually cause the insulation system to fail, thus increasing the heat leak considerably.

(4) Natural phenomena such as earthquakes could cause complete failure of this system.

3.5 DISTRIBUTION OF LIQUID HYDROGEN

Liquid hydrogen has four distinct methods for distribution on a large scale. Each of the methods has distinct characteristics that are reflected in the economic analysis of any composite liquid hydrogen system.

The four distribution methods are as follows:

(1) Truck Transport

(2) Railroad Transport

(3) Marine Transport

(4) Pipeline Distribution

These distribution systems are described in the following paragraphs.

3.5.1 TRUCK TRANSPORT

Current liquid hydrogen systems characteristically have a small number of production facilities. and several remote usage points. Because of this fact, overland transportation employing truck transport has been the primary method of distribution. Use of truck transport has provided vast flexibility in these existing liquid hydrogen systems.

Future large scale hydrogen systems will incorporate a set number of large volume use points and several comparatively small volume use points. To supply liquid hydrogen to the large volume use points, associated large scale production units would be constructed, or possibly a hyper-large production unit centrally located. Considering the large volume use points, and associated large scale production units, it can be seen that these systems will require less overall flexibility as compared with present day production point-use point "networks". However, reduction of system flexibility is offset by the fact that volume distribution requirements will be orders of magnitude greater than present day. It is felt that the present level of technology in truck transport is "sophisticated" with no major performance "breakthroughs" forecast.

3.5.1.1 System Configuration

Presently used semi-trailers transport liquid hydrogen in batch quantities over public roadways between production points and use points. Liquid hydrogen trailers so employed are best described as mobile storage tanks. Typically, such a storage tank consists of concentric inner and outer tanks separated by an evacuated space. Liquid hydrogen is carried in the inner tank and effectively insulated from the ambient by the annular evacuated space.

3.5.1.2 System Capabilities

System capabilities are determined by the numbers and capacities of semi-trailer units, and the distance that liquid hydrogen product must be transported. Gross capacities of liquid hydrogen trailers are primarily governed by design restrictions imposed by vehicular law. In the U.S.A., for example, current "standard" semi-trailer units have gross transport capacities of 7100 to 7700 pounds of hydrogen. Practical "oversize" semi-trailer units range to 9600 pounds of hydrogen gross capacity.

An advance design currently operating 8300 pound capacity unit is taken as representative of the future 10,000 pound capacity units from a design standpoint. Rapid fill-discharge and remote operation capabilities make these units somewhat different from the standard over-the-road units. The additional instrumentation and flow capabilities provide that off-loading be accomplished in a few minutes. Additional complexities introduced by these special units are easily offset by the time savings in respect to fill and discharge operations.

Transportation distance is a characteristic basic to system capability. Truck transport, railroad, and marine transport will likely be responsible for distribution in those cases where pipeline becomes impractical. Design and safety technology have been developed so that long distance truck transport would be a safe, operational distribution system. No new developments are judged necessary or anticipated in this respect.

The basic disadvantage with a truck transport system is that the number of semi-trailer vehicles in a given system could be quite large. For example, a single aircraft fueling could be equivalent to 20 to 25 trailer loads at 10,000 lb each. A system solely dependent upon truck transport could involve complex and perhaps costly operation.

3.5.1.3 Product Distribution

Truck transport distribution system losses can be categorized as follows:

1. Cool down losses
2. Steady-state heat leak equivalent losses
3. Transfer losses

3.5.1.3.1 Cool Down Loss.– Cool down loss is the amount of product cryogen required to refrigerate a trailer to equilibrium operating conditions. This amount of product is that vaporized in cooling the tank inner shell, the inner tank support system, the interconnecting piping, and the insulation. When all components reach thermal equilibrium, cool down is complete.

Evaluation of trailer cooldown losses frequently becomes an academic exercise. In most cooldowns, the vaporized hydrogen is recycled to the hydrogen liquefier. The tank, a horizontally oriented cylindrical configuration, is analogous to a transport trailer. It was determined that 600 to 720 pounds of liquid hydrogen were consumed in the cooldown. That is, approximately 8 percent of the nominal tank capacity was required to complete the cooldown of a tank similar to a trailer. It is reasonable to assume that trailer cooldown would be completed in a fashion less rigorous than this experiment. Thus, 15 percent of nominal trailer capacity is being taken as the cooldown loss, i.e., 1500 pounds for a 10,000 pound capacity unit.

The number of cooldowns that a trailer experiences in several years operating is extremely small. Considering operating experience with a whole fleet of 20 trailers, the average number of cooldowns is less than two per year.[33] For the purposes of this report, it is assumed that trailer cooldown averages will be two per year, and that average cooldown losses will be 1500 pounds per cooldown per a 10,000 pound unit. This yields a unit cooldown loss of 3000 pounds of liquid hydrogen per year.

3.5.1.3.2 Steady State Heat Leak Loss – Steady state heat leak equivalent losses result from the regular energy influx from the ambient to the contained liquid hydrogen in a trailer. The liquid acts as a heat sink for the energy influx by converting liquid to vapor in proportion to the energy flow. The equivalent liquid loss boil-off is primarily dependent upon the quality of the insulating medium and ambient conditions. High level vacuum insulated trailers have high quality insulation characteristics compared to other insulation methods. Recent experimental measurements indicate that boil-off is equivalent to 0.5 percent of nominal capacity per twenty-four hour day. The boil-off rate varies slightly with the actual level of liquid within the tank. However, 0.5 percent of capacity per day is a reasonable average to characterize the high level vacuum insulated liquid hydrogen trailer.

It should be noted that the boil-off is not truly a "loss" when a trailer is in operation. The boil-off vapor is retained within the trailer tank, and permits "building" tank pressure over a period of time. Frequently, no gas is actually "vented" from a trailer between charging or discharging. At the discharge point, depending upon the receiving facility, gas may be vented, or truly lost. For the purposes of discussion, however, it will be assumed that the 0.5 percent of capacity is lost.

In addition, it is considered that the boil-off rate is applicable to each day of the year, excepting a nominal two days per year for major maintenance. The two day exception corresponds to the referenced two cooldowns per year per unit. The boil-off rate applies to trailers when returning empty from deliveries. Returning trailers (empty) are not truly emply, but carry a "heel". The heel is an amount of liquid left in the trailer tank to maintain a cold condition. Thus, boil-off occurs with both full and empty tanks, basically all the time. Noting this, the yearly heat leak equivalent losses can be calculated as 0.5 percent of nominal capacity per trailer per 363 days per year, or 18,150 pounds per year per trailer for a 10,000 pound capacity unit.

3.5.1.3.3 Transfer Losses. – Transfer losses are associated with the operations of loading and offloading a liquid hydrogen trailer. The losses may be considered as arising from two specific sources:

(1) Flash loss, including heat leak equivalent loss

(2) Pressurization loss

Flash Loss. – Flash loss relates to the fact that the liquid hydrogen as stored or produced exists at a saturation pressure somewhat higher than the saturation pressure of the product in the loaded condition on board the trailer. Typically, liquid stored or produced has a saturation pressure of 3 to 10 psig. Liquid on board a trailer, however, generally has initial saturation pressures of less than 3 psig. The liquid arriving at the trailer, then, is superheated and not in thermal equilibrium with the trailer pressure. In order to establish equilibrium, spontaneous "flashing" (vaporization) of the liquid takes place. The liquid continues to vaporize until the saturation temperature of the liquid-vapor mixture corresponds to the saturation pressure in the receiving trailer. The amount of liquid that vaporizes and bacomes "flash vapor" is that amount of product considered as flash loss.

Further, flash losses are increased due to the energy influx to the liquid as it moves to the trailer from storage or plant. The energy input results directly from ambient heat leak through the insulation of the transfer piping. This energy input results in the vaporization of a portion of the transferring liquid. To evaluate the magnitude of the heat leak equivalent loss, it is necessary to understand (or assume) characteristics of the transfer piping between the plant or storage and the trailer.

Two cases are considered in order to demonstrate the significance of flash losses. These cases are shown in table 3-10.

TABLE 3-10. FLASH LOSSES

Item	Case I	Case II
Flow Rate - (Out of Storage of Plant)	600 Gal/Min (Approx. 250 T/D)	6000 Gal/Min (Approx. 2500 T/D)
Apparent Operating Pressure Drop	1.0 psid	6.0 psid
Trailer Pressure	1.0 psig	1.0 psig
Transfer System Heat Leak	1410 Btu/hr	1410 Btu/hr

Considering one pound-mol of hydrogen as it transfers from the plant or storage to the trailer, the First Law of Thermodynamics can be written in simplest form as:

$$\bar{h}_{\text{Trailer}} = \bar{h}_{\text{Storage}} + \bar{Q}_{\text{Heat Leak}}$$

Evaluation of the First Law is:

Case I

$$\bar{h}_{\text{Storage}} = 2240 \text{ Btu/lb mol} \qquad \text{(Saturated liquid at 2.0 psig)}$$

$$\bar{Q}_{\text{Heat Leak}} = 1410 \text{ Btu/hr/600 gal/min} = 0.013 \text{ Btu/lb mol}$$

$$\bar{h}_{\text{Trailer}} = 2240 \text{ Btu/lb mol} + 0.013 \text{ Btu/lb mol} \approx 2240 \text{ Btu/lb mol}$$

$$\bar{h}_{\text{Trailer Vapor}} = 2614 \text{ Btu/lb mol (Saturated vapor at 1.0 psig)}$$

$$\bar{h}_{\text{Trailer Liquid}} = 2238 \text{ Btu/lb mol (Saturated liquid at 1.0 psig)}$$

Now

$$\bar{h}_{\text{Trailer liquid}} + X\ (\bar{h}_{\text{Trailer vapor}} - \bar{h}_{\text{Trailer liquid}}) = \bar{h}_{\text{Trailer}},$$

Where X = pound-mol vapor/pound-mol vapor-liquid mixture,

Or

$$X = \frac{(2240 - 2238)\ \text{pound-mol vapor}}{(2619 - 2238)\ \text{pound-mol mixture}} = 0.00525\ \frac{\text{pound-mol vapor}}{\text{pound-mol mixture}}$$

and 1-X = 0.99475 pound-mol liquid/pound-mol mixture.

This yields a flash loss of 0.525 percent for Case I, and a flash loss of 2.85 percent for Case II.

For purposes of evaluation, a standard flash loss of 4 percent of nominal semi-trailer capacity is suggested here. This loss is projected as typical of that which might be expected from a 5 psig pressure drop system. In terms of the 10,000 pound semi-trailer units, a flash loss of 400 pounds would be expected for each loading. Further, it should be noted that the loss is identified only with the "on loading" of the semi trailer units. An "off loading" loss would be expected at the using agency. However, the using agency loss normally is accounted for in that agency's liquid hydrogen requirements.

3.5.1.3.4 Pressurization Loss–Pressurization losses are those losses associated with the generation of gas required to provide pressure for transfer of liquid from storage to the trailer. Typical current operating techniques at one plant include filling trailers from one of two storage tanks while "making" liquid hydrogen into the other. Facilities, in general, do have the capability of making directly into trailers, but this is not usually the desired way of operating. For purposes herein, it is assumed that all loading of trailers is done from storage. As with the ullage vapor build up in trailer traveling between points, the pressurization gas is not always truly lost. In certain cases, the pressurization gas is withdrawn as the storage tank is replenished with liquid, the gas so withdrawn being returned to the production facility or routed elsewhere. However, many installations simply vent the tank gas and do not reclaim it. The worst case approach is taken here, i.e., the pressurization gas is considered as truly lost.

Depending on facility operating characteristics, pressurization gas will be at pressures just slightly higher than the tank saturation pressure, or significantly higher. The better operating technique is that of having just slightly higher pressure operating gas. This results in less heat and mass transfer at the liquid-vapor interface. Several evaluations have been made of pressurization techniques.[29,31,32] In those cases, relatively high pressure hydrogen gas was used. Typical pressurization gas requirements were experimentally determined, and are shown in table 3-11.

TABLE 3-11. PRESSURIZATION GAS REQUIREMENTS

Gas Pressure	Gas Inlet Temperature	Gas Required/Liquid Transferred
58.1 psia	517°K	1.24 pounds/99.8 pounds
56.8 psig	5.2°K	1.76 pounds/98.1 pounds
45.0 psia	200°K	2.81 pounds/254.7 pounds

In the actual operation of a pressurization system, it is likely that a 300°K inlet temperature would result. In view of this, the value of 2.81 pounds pressurization gas for 254.7 pounds discharge is analogous to an actual operation at a facility. This data yields a pressurization requirement of 1.08 percent. Since all of the data is tenuous, it is assumed here that 1.5 percent is actually representative of the pressurization loss.

3.5.1.3.5 Summary of Losses. – In summary, then, Product Distribution Losses may be as listed for 10,000 pound capacity units:

Item	Product Loss
Cool Down Losses	3000 pounds per year per trailer
Steady State Heat Leak Equivalent Losses	18,150 pounds per year per trailer
Transfer Losses	
Flash Losses	400 pounds per trailer loading
Pressurization	150 pounds per trailer loading

Absolute amounts of loss encountered in a truck transport distribution system would be dependent upon the number of vehicles and scheduled loading frequency. The unit quantities shown can be employed to determine these absolute amounts.

3.5.1.4 Cost Elements

Two basic modes of operation of truck transport systems are projected to be incorporated in future liquid hydrogen systems. In one system the semi-trailers and tractors are owned, operated, and maintained as an integral part of the liquid hydrogen facility. In the second system the semi-trailers are owned, operated, and maintained by the liquid hydrogen facility with the hauling of the units being contracted to a separate agency. This second mode of operation separates the trucking service from the cryogenics facility.

3.5.1.4.1 Capital Investment – In the tractor service method of operation, capital investment is that associated only with the semi-trailer units. Capital associated with piping, connection bases, and other facility supporting equipment is accounted as a part of the facility capital investment.

It is projected that unit capital investment will be independent of the number of units procured at a respective location. This results from the fact that a large number of units will be produced initially. Developmental costs can be correspondingly distributed over the large number of units. Future developments, however, may in fact result in alternate types of semi-trailers being developed. Although it is not forecast herein, special equipment for a given location may be required. If this were the case, the analysis based on uniform capital investment for 10,000 lbm capacity units considered for this study would be invalid. A special evaluation would be required.

In particular, capital investment data presented herein relates to 10,000 lbm capacity units with semi-remote operating, fast unloading capabilities. Experience with the previously mentioned 8300 lbm capacity semi-remote operating unit has been reviewed. Extrapolating the experience to quantity production of the 10,000 lbm capacity semi-remote operating unit, it was felt the larger units would reflect a capital investment of 130 to 150 percent that of standard units of similar capacity. For purpose of this study it is projected that the 10,000 lb capacity units would have a capital investment 150 percent of that for standard units.

Data regarding costs of currently operating liquid hydrogen semi-trailer units hold to no particular pattern. Further, much information regarding costs of such equipment is held confidential by respective manufacturers. This has not led to a thorough evaluation of capital costs.

As an example, accumulated data of recent months is presented in table 3-12.

TABLE 3-12. COST DATA

Semi-Trailer Description (all MLRS Standard Design)		
Nominal Capacity	**Operating Pressure**	**Aver. Capital Investment Adjusted U.S.A. 1966 $**
5900 pounds	100 psig	110,000/unit
4720 pounds	50 psig	87,000/unit
7700 pounds	50 psig	85,000/unit

It is projected that a reasonable estimate for capital investment for the 10,000 pound units (standard design) might be $100,000 each. Considering this number as a base cost, the capital investment for the projected 10,000 pound capacity, semi-remote operating units would then be $150,000 each. This value is thus proposed as the unit capital investment to be considered for truck transport distribution of liquid hydrogen.

3.5.1.4.2 Operating Costs. – Two specific explanations of operating costs are related to the "tractor service" method of operation. A freight rate is applicable to the labor and equipment used in the hauling of the semi-trailer units. The second set of costs are those related to maintenance costs associated with the semi-trailer unit itself.

Considering the first set of costs, current U.S.A. practice is for regulatory agencies to establish unit fees of rates for common carrier firms providing transportation service to the LH_2 facility. Current rates based upon U.S.A. LH_2 facility operations range from $0.26 to $0.44 per hauling mile. Although the rates are set by regulatory agencies, they reflect all of the tractor service costs including profit for the hauling agency. These approved rates are negotiated between the shipper and carrier and filed with the regulatory agency for approval. In this sense, then, it is reasonable to project an analogous fixed rate that is associated with the 10,000 lbs capacity units.

For purposes of this study, a rate of $0.44 per hauling mile is being taken. The rate is applicable to the U.S.A. locations. Adjustments of the rate would be made elsewhere to reflect various geographic locations. It is noteworthy that a large portion of the stipulated rate results directly from taxes or other governmental fees. With relief from these burdens, significantly lower rates could result. However, the rate of $0.44 per hauling mile is taken as a representative working base for this study.

The second set of costs is that related to maintenance of the semi-trailer units. The maintenance costs include insurance, dispatching, repair, inspection and other costs generally associated with operation readiness of the semi-trailers. Industry practice is such that the aggregate maintenance costs are generally prorated on equipment use mileage. Considering recent U.S. experience with 7100 and 7700 lb capacity units, maintenance costs range to $0.10 per mile. The projected semi-remote operating 10,000 lb capacity units will have an inherently higher degree of mechanical complexity. It would be thought that this degree of complexity should increase maintenance costs. However, it must be noted that respective fleets in future systems will incorporate greater numbers of units. Here, an economy from numbers should be realized.

It is projected that the aggregate maintenance cost components will be $0.06 per operating mile. Thus, a total operating cost of $0.50 per vehicular mile is projected for a future truck transport system. This means that a 200 mile distance between production and use point would result in $0.020 operating cost per nominal pound capacity, based on a round trip for a 10,000 lb unit.

3.5.2 RAILROAD TRANSPORT

Liquid hydrogen systems currently employ railroad transport as one type of product distribution. The degree of application of this distribution method, however, is relatively small when compared with other distribution systems. It is projected that railroad transport would find more application in future systems. It is conceivable, for example, that a special railroad system could be built between a specific production point and its associated use point. Special tank trains then would ply the route between the production and use points. Since present day employment of railroad transport is limited, much of the material presented must be taken as conjecture.

3.5.2.1 System Configuration

It is not anticipated that future rail cars will differ basically from those currently in operation. Respective sizes (capacities) may increase, but no change in basic design and construction is foreseen. Improvement in manufacturing technique may result as production of more units proceeds. This type of improvement will be reflected, primarily, in unit cost of the rail cars.

3.5.2.2 System Capabilities

Two major, inter-related factors define the system capability of railroad transport. One of the factors is simply the unit capacity of the rail cars. This capacity, however, is dependent upon the second factor, that is the railroad routings to be used in railroad transport. The restrictiveness of these two factors in current operations is in part responsible for the relatively small application of railroad transport for liquid hydrogen.

Railroad operations currently in the U.S.A. are such that no railroad car can exceed a width of 10 ft 8 inches. This restriction is imposed so that railroad rolling stock can negotiate curves in tracks as built or proposed in accordance with current design practice. Further, for long railroad cars (cars having truck centers separated by more than 51 ft 3 in), the car width must be reduced accordingly. The width reduction must be that no part of a long car displaces more than 5 ft 9¾ inches from the truck center line in negotiating a 13 degree turn. A car built within these restrictions will be able to negotiate most U.S.A. railroads without special routings. Here, then, the effect of railroad routing is apparent; tank design (capacity) must be that the rail car can be routed over the U.S.A. railroad network.

Railroad routings also have a second effect upon railroad transport. The effect is that resultant from lengthy, often devious, routes of travel necessary to move a railcar from one location to another. Frequently, several days of travel time are required for a specific liquid hydrogen railcar in transit between two points. It is likely that a large fleet of railcars would have to be "oversized" to cope with the unpredictably long periods of travel. Liquid hydrogen, as a product, becomes costly as time between production and use is increased. This results from losses that must be provided for by over production.

Capacity of liquid hydrogen railcars could vary so long as the basic restrictions as imposed were not violated. Current operating stock has been developed to the point that capacity beyond that now used would likely result in violation of the present restrictions. In other words, present vacuum jacketed super insulated units are as large as possible under the current circumstances. These units, nominally of 17,000 pounds capacity, are in use for transcontinental movement of liquid hydrogen in the U.S.A.

As with truck transport, safety aspects of railroad transport are well understood. No specific problem relating to safe operation of this equipment is foreseen at this time.

It is projected that future railroad transport distribution systems could not economically function in accord with the current framework. Rather, it is expected that specific new railroad routes would be developed to directly serve the needs of the liquid hydrogen facilities as required. In the process of developing the new railroad routes, railroad engineering concepts would be devised such that larger capacity rolling stock could be used. It is difficult to project, in this instance, how large a capacity unit would be desirable. Conceivably, the units would be double present capacities. As an example, capacity of 25,000 pounds is assumed for the future railroad cars. Data based on this capacity will be developed in the following paragraphs.

3.5.2.3 Distribution Losses

All of this can be evaluated in a fashion analogous to that presented for truck transport. Detail information presented in paragraph 3.5.1.3 is not repeated herein, but reference to that data is made as it applies to railroad transport.

Railroad transport distribution losses can be categorized as:

(1) Cool down losses

(2) Steady-state heat leak equivalent losses

(3) Transfer losses

3.5.2.3.1 Cool Down Losses. – It has been noted that the railcar unit is analogous to the truck unit with respect to the storage tank. In view of this, cool down losses developed for the truck transport are being projected for the railcars. The cool down loss, then, is taken as 15 percent of nominal rail car capacity, or 3750 pounds for a 25,000 pound unit.

Similarly, the annual number of cool downs per railcar is being taken as 2 per year. No real operating basis exists to support this number, but indications are such that railcars should perform as well or better than the semi-trailer units. Thus, two cooldowns per year can be viewed as a conservative number. In turn, this yields an annual cooldown loss of 7500 pounds per year per unit.

3.5.2.3.2 Steady State Heat Leak Loss. – Again the analogy between the truck transport and railroad transport can be drawn with regard to steady state heat leak loss. It has been suggested the railroad transport loss rate would be 0.3 percent capacity per day. This compares with a corresponding loss rate of 0.5 percent capacity per day for truck transport. Support for the rail transport's lower loss rate is non-existent, but the speculation stems from anticipated improved heat leak characteristics of the larger capacity railcar tanks. It should be noted, however, that "beefier" mechanical design of the rail units could in fact result in poorer heat leak characteristics of the larger units. Thus, without any firm data, the value of 0.5 percent capacity per day can be taken as qualitatively representative.

Considering the 0.5 percent capacity per day loss, then, an annual loss of 45,375 pounds is projected for a 25,000 pound capacity railroad transport unit (363 days).

3.5.2.3.3 Transfer Losses. – Transfer losses could be analyzed for railroad transport systems in a fashion corresponding to that for truck transport. However, it is expected that equipment and attendent losses considered for the truck transport system would be the same for the railroad transport system. For a detailed development of loss data refer to paragraph 3.5.1.3.3.

3.5.2.3.4 Summary of Losses. – In summary, the losses projected for 25,000 pound capacity units are:

Item	Product Losses
Cool down losses	7,500 pounds per year
Steady state heat leak equivalent loss	45,375 pounds per year
Transfer losses	1,000 pounds per railcar loading
Flash	1,000 pounds per railcar loading
Pressurization	375 pounds per railcar loading

3.5.2.4 Cost Elements

To date, six liquid hydrogen railcars have been put in service in the U.S.A. Cost elements of capital investments and operating costs for future systems is not easily projected based on the limited present operation. Following estimates of cost elements must be considered as conjecture.

3.5.2.4.1 Capital Investment – Cost data relating to the six liquid hydrogen railcars presently operating has not been obtained from the manufacturer. Such information is currently considered proprietary. Economic analysis have estimated the average cost to be $140,000 per unit. The units are standard design, 17,000 pound capacity.

The projected units to be used in future systems would be of the semi-remote operating design, 25,000 pound capacity. Noting the above capital investment, an approximate value of $200,000 capital investment would be projected for a 25,000 pound capacity, standard design unit. This value results from approximate direct scaling based on capacity. For truck transport, a semi-remote operating unit was valued at 150 percent that of the standard units of the same capacity. In considering the railroad units, the corresponding factor is postulated to be 125 percent. Thus, the capital investment associated with the semi-remote operating, 25,000 pound capacity unit is taken as $250,000 each.

3.5.2.4.2 Operating Costs – Operating costs for the railroad transport system can be divided into two categories:

(1) Freight

(2) Maintenance

Railroad freight tariffs in the U.S.A. are established by governmental agencies (ICC). Established rates reflect costs and reasonable profits for respective carriers handling a given commodity between two points. Rates established for current liquid hydrogen operations are not complete enough to cite as future costs. Estimate per mile rates range linearly $1.25 to $0.50 corresponding to travel ranging 50 to 1000 vehicular miles. These rates are based on vehicular mileage. The rate for the round trip of a given unit would be based on the total mileage covered by the unit in making the round trip.

Maintenance costs are projected as being approximately the same as that for truck transport.

As an example, consider distribution of liquid hydrogen over a distance of 200 miles. For this, the linearized freight rate would be $0.97 per mile (400 miles), or a total operating cost of ($0.97 + $0.06) per mile = $1.03 per mile. Considering the 25,000 pound capacity units, the unit operating cost would be $0.0155 per nominal pound capacity.

3.5.3 MARINE TRANSPORT

The regional or local distribution of liquid hydrogen by marine transport has been employed in servicing NASA's Mississippi Test Facility. Long distance marine distribution, on the other hand, has not been undertaken for liquid hydrogen. However, other cryogenic products, including liquefied natural gas, presently are distributed by long distance marine transport.

Experience to date has not been satisfactory for the local distribution of liquid hydrogen by marine transport. The existing liquid hydrogen marine units, which consist of 160,000 pound capacity tanks mounted on modified barge hulls, are towed between respective production and use points. In this operation the hydrogen barges serve as storage and distribution equipment. Overland transport and additional storage equipment would have been a better distribution-storage system for the facility because of costly long delivery time due to the awkward movement of the barges in the channel. The economics of regional or local movement of hydrogen product at the different sites will require very detailed evaluations of marine distribution systems.

Long distance transportation of liquid hydrogen may, however, prove to be extremely beneficial to the hydrogen program. For example, it may be economically feasible to construct a large liquid hydrogen facility near an abundant raw material source such as a natural gas field. There the production cost of the liquid hydrogen would be substantially lower than costs at some of the proposed production sites (i.e., New York City). One possibility is to produce liquid hydrogen in the Venezuelan gas fields and transport product hydrogen from Curaco to New York and Sao Paulo. The merits of this method take on added significance in the event numerous use points for liquid hydrogen and other cryogenic products would develop along the East Coast of the U.S.A. Shuttle service from such a central producing point may then become the most practical and economic system.

3.5.3.1 Ship Configuration

To date no significant work has been done in design of liquid hydrogen marine tankers. However, design of liquefied natural gas tankers, the forerunners of LH_2 tankers, has been progressing rapidly. Some design problems which must be overcome for cryogenic vessels include special tank insulation requirements, structural design and ship stability. The density of LH_2 is 4.42 lb/CF, LNG is 26.5 lb/CF and oil about 56 lb/CF which greatly affects design of the respective vessels. In addition, the proposed employment of these mammouth LH_2 or cryogenic vessels in the post-1980 tanker fleet will generate investigation of other design concepts such as hydrofoil or ground effects principles. Use of boiloff liquid for propulsion fuel versus reliquefaction and return to storage must also be evaluated in ship design.

Two future marine cryogenic transport configurations can be visualized. One would be a vessel designed exclusively for liquid hydrogen cargo. This ship would be a large barge-type vessel with numerous modular storage tanks. The second configuration would be a huge conventional type ship with multi-purpose modular storage tanks. This ship would have the capability for transporting a variety of cryogenic fluids simultaneously, providing flexible use of the vessel.

3.5.3.2 Ship Capacity

Projections for tanker ship capacity in the near future have been increasing rapidly. One projection indicates oil tanker capacity of 500,000 tons in the 1970 period. This tanker would be a "super barge" capable of sustaining a sea cruise speed of 17 knots. Based upon the dimensions proposed for this tanker, a liquid hydrogen tanker with a capacity of approximately 25,000 tons or larger could be projected.

3.5.3.3 Distribution Losses

Cooldown losses will be essentially negligible for the marine units. A residual amount of liquid will be kept in the modular storage tanks. The heel will maintain the cold condition of the storage units. These tanks, except for rare isolated instances, are cooled down only once, that being when put in service.

Steady state heat leak loss can be evaluated with respect to facility storage equipment. It is expected that steady state heat leak losses for marine transports will be somewhat higher than facility loss rates. The marine tanks would have loss rates of up to 0.04 percent per day. Based on a nominal 25,000 ton capacity unit, the daily heat leak loss would be 10 tons per day.

Transfer losses are dependent upon an understanding of the overall marine transport system. Transfer loss calculation can be assumed comparable to pipeline distribution loss.

3.5.3.4 Cost Elements

Capital investment for a marine distribution system can be estimated based on current cost data. For example, the famed Torrey Canyon had an estimated capacity of 36,000,000 gallons of crude oil or about 850,000 barrels with a cruise speed of 17 knots. She was listed as the 13th largest merchant ship in 1967 with an investment of 16.5 million dollars and a crew of 36 men. The largest tanker afloat is reportedly the Idemitsu Maru at 120,000 tons or twice that of the Torrey Canyon. LNG tankers including the Methane Progress and Methane Princess with capacities of about 180,000 barrels are presently in service. The proposed Phillips-Marathon venture in the Alaskan gas fields will utilize two 450,000 barrel LNG vessels at an estimated total cost of $43 million. Based on scaling the Torrey Canyon investment by the ratio of ship size to the 0.7 power, yields an equivalent investment of about $10.6 million for a 450,000 barrel oil tanker. LNG tanker costs are presently, therefore, about twice the cost of oil tankers in this size range. Assuming LH_2 tankers will require correspondingly larger investments than LNG tankers, perhaps by a factor of 1.2, would yield a cost of $26 million for the 450,000 barrel or 11,000 ton capacity LH_2 tanker. Assuming the 0.7 power is an appropriate scaling factor, then a 25,000 ton LH_2 vessel would cost about $46 million. Other capital requirements include harbor facilities, tanker loading and unloading facilities, and storage facilities which will vary for each proposed system.

Operating cost data will include crew's wages, overheads, maintenance and repair, stores and supplies, subsistance, port charges, ship insurance, taxes, fuel and miscellaneous costs. These costs will vary depending on ship and system size.

Detailed cost data for marine distribution systems is not considered to be within the scope of this project. However, it is apparent that a detailed analysis of this type of LH_2 production and distribution method may provide very low cost LH_2 especially if multi-air terminal sites are to be serviced.

3.5.4 PIPELINE DISTRIBUTION

Liquid hydrogen systems generally incorporate pipelines in one or more product transfer functions. Future systems will certainly employ pipelines for inplant transfer and, in certain instances, will employ pipelines as the primary volume distribution device. It is necessary to comprehensively understand pipeline distribution since it will be incorporated in future liquid hydrogen systems. Characteristics of the pipeline systems will have major effects upon other portions of the overall liquid hydrogen system. The product losses associated with pipeline distribution must be accounted for by production "oversizing", this requiring significant additional capital and operating costs.

Historically, much concern has developed about the type of pipeline for a given system. Effort generally is expended reviewing all available methods for the liquid hydrogen application. The data presented details the qualifications of various methods for liquid hydrogen service. Further, cost data is developed so it will be possible to evaluate economics for distribution of liquid hydrogen by pipeline or other means.

3.5.4.1 System Configuration

Several types of pipeline can be considered for application in liquid hydrogen systems. As with storage equipment, basic differences between the various pipeline types relate to respective insulation techniques employed. Also, provision for expansion and/or contraction in the respective pipeline systems is a source of other basic differences.

The types of pipelines considered for liquid hydrogen distribution can be classified according to insulation techniques. These classifications are:

(1) Uninsulated (no insulation, direct burial of pipeline underground)

(2) Bulk-type insulated (fibrous glass, styrofoam, polyurethane foam)

(3) Vacuum insulated (evacuated powder, high vacuum, superinsulated)

All of these pipeline types have been incorporated in distribution systems for various cryogens. The respective pipeline systems as related to liquid hydrogen distribution are covered in the following paragraphs.

3.5.4.1.1 Uninsulated Pipeline. – Uninsulated pipeline is frequently used in limited distance distribution of warmer cryogens. (Warmer cryogens have normal boiling points equal to or above that of liquid nitrogen.) Normally, transfer of warmer cryogen through an uninsulated pipeline results in atmospheric condensation on the cold pipeline surface. If the cryogen temperature is lower than the normal boiling point of oxygen ($\approx$ -297°F), oxygen enriched air condenses on the pipeline surface. The presence of the liquid air creates an extremely hazardous situation by providing a highly concentrated oxidant for a combustion reaction.

It has been suggested that direct burial pipeline could be utilized for liquid hydrogen distribution. The major problem that negates further consideration of this system is expansion and contraction allowances for the pipeline. For example, consider a one mile length of Invar pipe direct buried for liquid hydrogen distribution. Invar pipe, an extremely low specific thermal expansion material, would exhibit an average contraction of 1×10^{-6} ft/ft°F over the temperature range of ambient to liquid hydrogen level. The length of pipe would contract approximately 2.6 feet in the 500°F cooldown. Provision for movement of this type, plus heaving and other movement caused by the frozen earth layer, would make the direct burial method moderately complex. No further consideration of the direct burial system is given in this report.

3.5.4.1.2 Bulk-Type Insulated Lines. – Bulk-type insulations have been used extensively in both high and low temperature applications. Cryogenic systems operating at, or above, liquid nitrogen level have employed bulk-type insulated pipelines. Application of this insulation technique has not been pursued on a large scale for lower temperature cryogenic systems.

Description of Insulating Materials. – The vapor barriers for liquid hydrogen application have not proved successful. Excepting metallic barriers (double wall pipelines), no satisfactory vapor barrier has been found for long term liquid hydrogen service. A major failing of many proposed vapor barriers is the effect of thermal cycling, that is, repeated pipeline cooldowns and warmups. Movements of the insulating material in the annular space result in movements or stressing of the vapor barrier. No further consideration of bulk-type insulating materials is given in this report.

Insulating characteristics of the bulk materials are satisfactory for many cryogenic applications. However, certain other properties pose particular problems in liquid hydrogen distribution applications. For example, compatibility of the insulating material with liquid oxygen is prerequisite. Fibrous glass and some foam glass materials are now deemed liquid oxygen compatible.

3.5.4.1.3 Vacuum Insulated Pipeline. – The basic characteristics of these vacuum insulated pipelines are:

(1) A metallic vapor barrier (outer pipe)

(2) Reduced pressure (vacuum) in the annular space of the pipeline configuration

This type of system, in varied forms, is the primary device employed currently for pipeline distribution of colder cryogens, such as liquified neon, hydrogen, and helium.

Vacuum Insulation Techniques. – Distinction among the various vacuum-jacketed pipeline systems lies with variations in the annular space insulating characteristics.

One type of vacuum insulated pipeline system employs no insulating material. The residual gas in the annular space is evacuated until an extremely low pressure is achieved. By establishing the low pressure (hard vacuum), heat transfer losses by gaseous conduction and convection are minimized. The major component of heat transfer results from radiation between the two walls of the annular space. An amount of heat transfer can occur by conduction through mechanical paths connecting the inner pipe to the outer pipe. To further reduce radiation heat transfer, it is common to wrap several layers of metallic foil around the inner pipeline prior to assembly.

The major problem with this insulation system is that of maintaining a suitable vacuum level. It is necessary to maintain a residual gas pressure of 10^{-5} millimeters of mercury, or lower, for an effective vacuum insulation.

A second type of vacuum insulated pipeline is that now generally termed as "Multi Layered Radiation Shielded" (MLRS) pipe. The basic configuration of this system is the same as described for a hard vacuum system. However, MLRS pipe generally employs several layers of reflective foil wrapped around the inner pipe. The layers, typically metallic deposit on one side of an insulating foil are wrapped until a desired number of reflectors are accumulated. The radiation heat transfer is approximately proportional to the reciprocal of the number of reflective layers. Conduction heat transfer, on the other hand, tends to increase as the layers are built up.

Various manufacturers have proprietary methods of building the MLRS pipe. All of the variations have exceptionally good heat leak characteristics, that is, low apparent thermal conductivity. At the present time, MLRS pipeline is the most often employed system used for direct distribution of liquid hydrogen.

3.5.4.2 System Capabilities

System capability is basically determined by the allowable volume rate of flow and associated pressure drops. In addition, energy influx heat leak to the flowing fluid is a major concern. Consideration will determine system economics and ultimately determine if pipeline distribution can be considered practical.

Calculations of flow characteristics of 10, 18 and 24 inches IPS have been made for comparison purposes. Calculations are based on liquid hydrogen properties for conditions of 30 psia and -423°F (subcooled liquid). It is expected that the 20 psid maximum allowable pressure drop in distribution would be consistent with other facility equipment and operation. Considering the 20 psid pressure drop in transfer, and a nominal facility production of 500 tons/day, distribution requirements are such that a one-day production must be transferred in a 12-hour period, over a distance of 75 miles. Consider the 75-mile distribution pipeline as being 400,000 equivalent feet in length. For this length and with an allowable 20 psid pressure drop, a unit pressure drop of 0.025 psid per 1000 ft equivalent is required. The nominal liquid hydrogen mass flow rate is 500 tons per 12 hrs or 1390 lbs per min. It would appear that a 20 inch IPS inner line would be satisfactory, as shown in figure 3-35.

3.5.4.3 Product Distribution Losses.

Pipeline distribution system losses are analogous to the transfer losses referenced in paragraph 3.5.1.3.3, Flash Loss, except as noted in the following paragraphs.

3.5.4.3.1 Flash Losses. – During the course of transfer, the liquid remaining in the storage tank will tend to warm up. The liquid will warm to a temperature level approaching the saturation temperature corresponding to the overhead saturation pressure (ullage vapor pressure). The liquid hydrogen, then, exists at a saturation pressure equal to or greater than the saturation pressure at the delivery point of the pipeline system.

It is being considered that the saturation pressure change of the transferred liquid is in fact 20 psid. It should be noted that the storage liquid, although pressurized 20 to 30 psia above initial saturation pressure, does not reflect immediately a saturation pressure characteristic equivalent to the ullage pressure. The liquid is in a subcooled condition and will slowly warm its saturation pressure toward the ullage pressure. Taking this approach, then, the transferred liquid enters the pipeline as a saturated liquid at delivery pressure plus the 20 psid differential. Now, further flash losses will occur based upon the energy influx through the pipeline system walls.

Heat leak is dependent upon the insulation characteristics of the pipeline system. Such characteristics, however, generally are proprietary information developed by various manufacturers. Certain

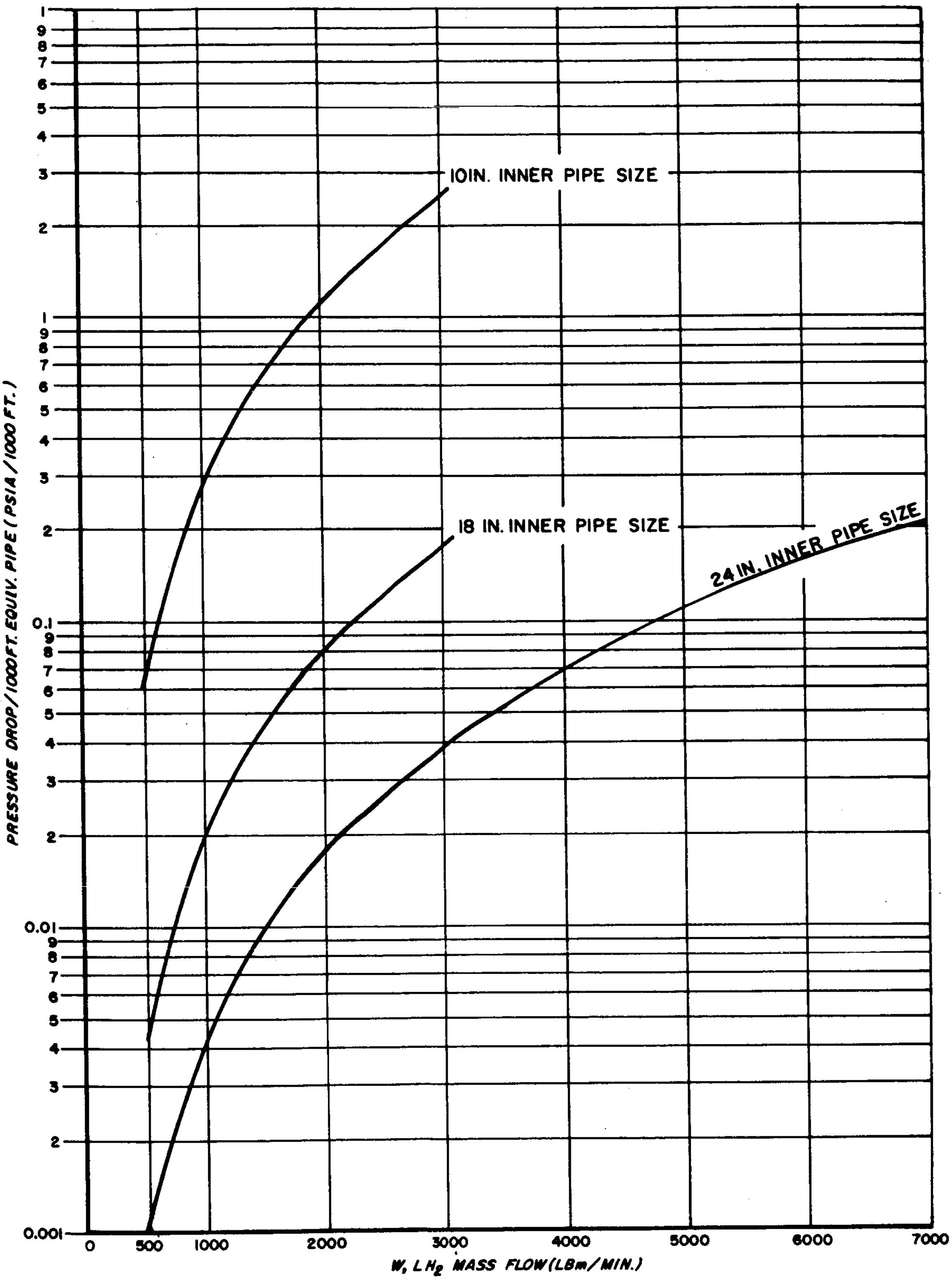


FIGURE 3-35. PRESSURE DROP PER 1000 FEET

data that is available[33] does not necessarily represent current designs. Calculations have been made to develop an amount of data for use in this report. Projections have been made based on specifications and design information for MLRS pipeline systems in liquid hydrogen service.

Evaluation of heat leak for any hydrogen service MLRS pipeline requires consideration of individual components. The system components have been evaluated relatively with respect to heat leak characteristics of the pipeline joints. In simplest form, this consideration results in a series of heat transfer factors identifiable with the line components. The component heat leak characteristics can be ascertained from the knowledge of system joint heat leak values.

Component Heat Transfer Factors

Item	Factor
Joint	1.00
100 ft Line	0.30
Fitting (all)	0.38
Valves (all)	5.00

Future efforts will be directed toward reduction of heat leak characteristics of components. Data is graphically presented for respective line size joint characteristics in figure 3-36. This data is representative of average characteristics of currently manufactured equipment.

Difficulty in understanding system heat leak characteristics is experienced since the values necessarily reflect a specific design. An attempt to generalize is here made to provide a simplified analysis. First, it is assumed that a cross-country pipeline can be considered in terms of two-mile increments. Further, each two-mile increment will be made up of components as follows:

Standard Two-Mile Line Segment

Item	Heat Transfer Factor	
106-100 ft pipe	106 x 0.30 =	31.8
1 - valve	1 x 5.00 =	5.00
106 - Joints	106 x 1.0 =	106.00
30 - Fittings	30 x 0.38 =	11.4
Total		154.2
		≈ 154.0

The two-mile typical line segment has a characteristic heat transfer factor approximate value of 154. Actual heat leak characteristics of a given line size can be determined from reference to figure 3-36.

The combination of the heat leak data and pressure drop related losses will permit tabulation of the system flash losses.

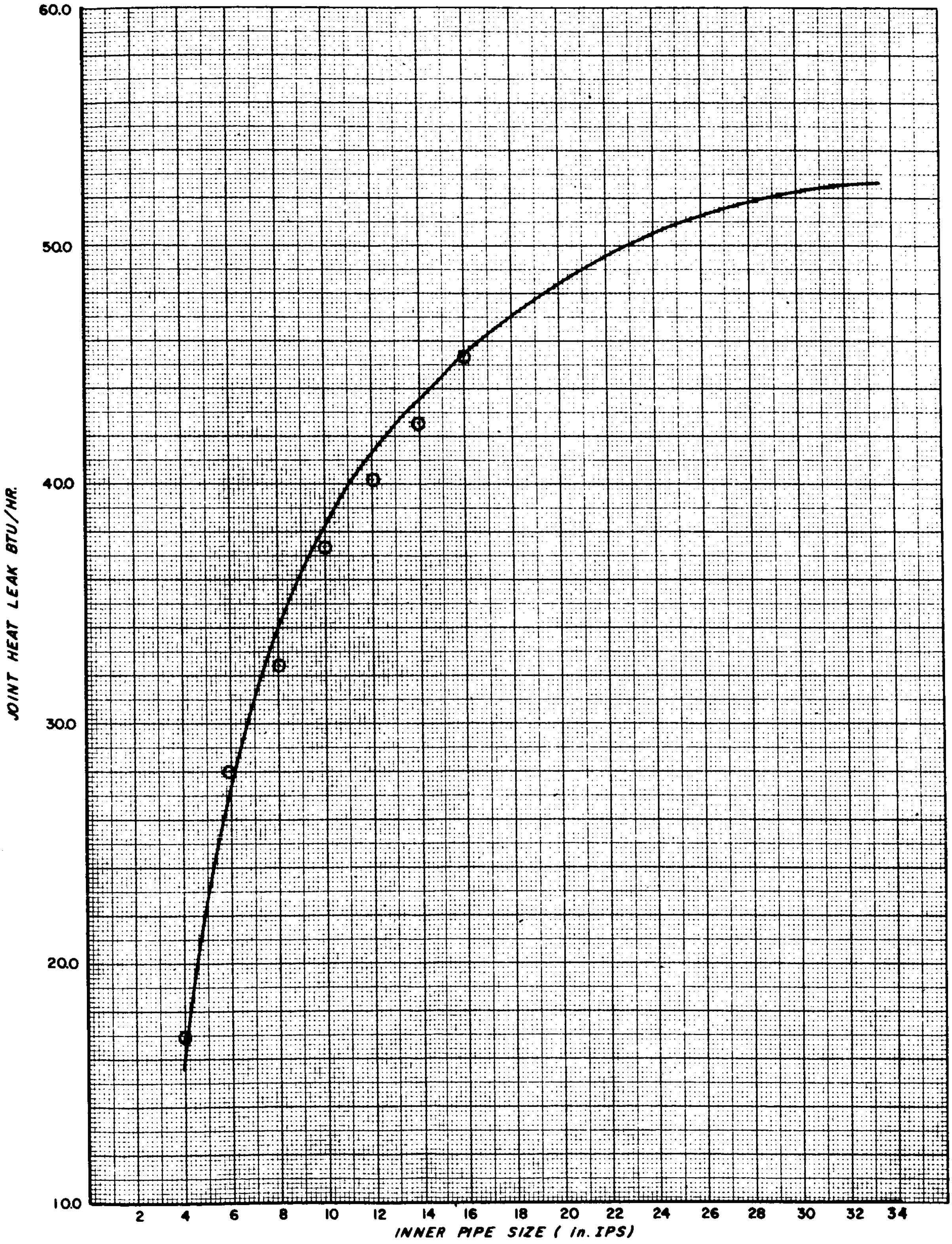


FIGURE 3-36. JOINT HEAT LEAK BTU PER HOUR

3.5.4.3.2 Pressurization Losses. – The pressurization losses in a pipeline distribution system are analogous to the truck transport system as referenced in paragraph 3.5.1.3.3, Pressurization Loss.

3.5.4.3.3 Summary of Losses. – A general survey of losses cannot be made. An example of loss calculation is presented as a guide to establish the general calculational technique.

(1) Distribution Requirements:

Mass Flow - 3500 lb/min

Distance - 16 miles (including expansion loop, etc.)

Max. Allowable ΔP - 20 psid

Exit Pressure (Saturated) - 3 psig

(2) Pipeline Selection:

Equivalent pipeline length of 97,000 feet, approximately 115 percent of actual line length.

Thus $\Delta P/L = 20$ psid/97,000 feet = 0.206 psid/1000 feet

See figure 3-35

$\Delta P/L = 0.206$ psid/1000 feet $\quad w = 3500$ lbm/min

Inner pipe size of 18-inches IPS is indicated.

(3) Heat Leak Calculation:

Unit Heat Transfer Factor is 154 for two-mile segments, or a total HTF of 2464 for the 16-mile pipeline.

See figure 3-36, the joint heat leak for an 18-inch IPS line is 47 Btu/hr

Thus, heat leak, Q, is

$$Q = 2464 \times 47 \text{ Btu/hr} = 115{,}808 \text{ Btu/hr} = 1{,}930.1 \text{ Btu/min}$$

Transfer Losses (First Law of Thermo Dynamics)

$$H_{exit} = H_{entrance} + Q,$$

or $$h_{exit} = h_{entrance} + Q/w$$

$$h_{entrance} = 2270 \text{ Btu/lb mol}$$

$$(P_{sat} = 37.7 \text{ psia}$$

$$h_{ent} = h_{liq} \begin{array}{l} 25\% \text{ para} \\ 75\% \text{ ortho}) \end{array}$$

$$w = \frac{3500 \text{ lbm}}{\text{min}} \times \frac{\text{lb mol}}{2.016 \text{ lbm}}$$

Now

$$h_{exit} = 2270 \frac{\text{Btu}}{\text{lb mol}} + \frac{1930 \text{ Btu}}{\text{min}} \Big/ \left(\frac{3500 \text{ lb}}{\text{min}} \times \frac{\text{lb mol}}{2.016 \text{ lb}} \right)$$

$$\text{and} = 2270 \frac{\text{Btu}}{\text{lb mol}} + \frac{1.11 \text{ Btu}}{\text{lb mol}} \approx 2271 \frac{\text{Btu}}{\text{lb mol}}$$

$$h_{exit} = h_{liq\ exit} + x\,(h_{vap\ exit} - h_{liq\ exit}),$$

$$\therefore x = \frac{h_{exit} - h_{liq\ exit}}{h_{vap\ exit} - h_{liq\ exit}}$$

$$h_{vap\ exit} = 2621 \text{ Btu/lb mol}$$

$$h_{liq\ exit} = 2242 \text{ Btu/lb mol}$$

(Sat. liq.—vap at 3.0 psig)

$$\text{or} \quad x = \frac{2271 - 2242}{2621 - 2242} = \frac{29}{379} = 0.0767$$

That is a transfer loss of approximately 7.7 percent could be anticipated in this system.

If straight vacuum-jacketed pipeline were used the heat leak would be thus,

$$Q = 10 \times 1.11 \text{ Btu/mol and}$$

$$x = \frac{39}{379} = 0.104$$

It can be seen that use of MLRS pipe results in substantially lower losses. Generally, the straight vacuum-jacketed line proves uneconomical due to increased losses. For this reason, only MLRS pipe can be considered for liquid hydrogen.

Pressurization Losses

As indicated, a value of 1.5 percent of nominal flow is taken.

Total Distribution System Losses

Losses will calculate to 7.7% + 1.5 percent or a total of 9.2 percent of inlet.

The presented example of the pipeline distribution information is for purposes of this report. A second set of conditions will be established for on-site pipeline distribution. That is, for those situations wherein the hydrogen production is located adjacent to the aircraft fueling area, a pipeline distribution system would be utilized. In this instance, the pipeline system conditions are revised to:

(1) Mass Flow - Multiple values of 120 T/hr. (consistent with one aircraft fueling per hour or multiples thereof).

(2) Distance - 2 miles (distance is storage to aircraft servicing manifold, but not manifold).

(3) Max. allowable ΔP - 5 psid

(4) Exit Pressure (Saturated - 3.0 psig)

In a fashion analogous to the previous example, line sizes and associated losses have been calculated. The values are shown in table 3-13.

TABLE 3-13. LINE SIZES

Mass Flow (T/hr)	Line Size (in)	Pressure Drop (psid)	% Total Calculated Loss
120	18	4.4	4.0
240	24	5.0	4.1

It should be noted that straight vacuum-jacketed line used in this example would yield total losses approaching 5.0 percent.

For study purposes, on-site distribution total loss values of 5 to 7 percent could be anticipated. A value of 7 percent is being suggested for use in subsequent analyses. Further, the use of straight vacuum jacketed line, in this application, could account for approximately 1 percent additional loss, or a total loss value of 8 percent.

3.5.4.4 Cost Elements

Cost elements of capital investment and operating costs could be determined for an installed and operating pipeline distribution system. However, experience with currently installed pipeline distribution systems is that identifiable operating costs are vanishingly small in relation to operating costs for other parts of a given liquid hydrogen facility.

In view of this, operating costs for pipeline distribution systems can be considered as incorporated within the operating costs of other portions of the liquid hydrogen system.

3.5.4.4.1 Capital Investment. – Significant historical data related to pipeline distribution of liquid hydrogen is not generally available. Data that is available relates to several specially designed and constructed systems. The systems currently in service do not have distribution distances comparable to projected future systems.

Capital investment analysis was made to provide a basis for pipeline distribution cost elements. A major premise was that of choosing a minimum system distribution distance of one mile. It was felt that systems of lengths less than one mile would result in premium unit costs. This has been demonstrated to some degree in the costs of various short distance systems compared with larger systems of the same line size. For lengths of one mile or more, the capital investment per foot of pipeline should remain essentially constant.

A proportional relationship exists between inner line size and per foot capital investment for a specific manufacturer, i.e., for a given design. It would be expected that costs of future systems would be directly proportional to a diametric size, say, inner line size. All elements of construction, e.g., material, inches of weld, etc., are related in terms of diametric size.

As a base, then, a 6-inch IPS inner line size pipeline system has been chosen. Historically, 6-inch IPS MLRS pipeline had installed costs ranging from $100 to $300 per foot of line. The higher costs are those associated with older, relatively short distribution systems. The lower costs, however, reflect more recently installed systems. Absolute projection of future costs for 6-inch IPS systems is not possible. A value of $90 per feet, however, is considered representative of future installed systems. Considering the diametric proportionality, it is suggested that the value of $15 per foot per diametric inch be taken. This value of $15 per foot per diameter inch is applicable to an Invar-stainless steel MLRS pipeline system.

It should be noted that the investment related to a pipeline distribution system (local distribution) will be small related to overall system investment. In view of this, the absolute value of dollar per foot per diametric inch could be doubled with only a minor effect upon the end cost of liquid hydrogen.

Section 4

LOGISTICS

4.1 GENERAL REQUIREMENTS

The plant requirements described in Section 3 identify the need for raw material and power. Potentially suitable commodities (if available at a competitively favorable cost) include coal, petroleum, natural gas, electricity, water power, water, uranium, and heavy water. It is the purpose of this section to summarize the raw material and power requirements for the various size and type liquid hydrogen plants. Also cited are their availability and assignment of, as accurately as possible, their contribution to the post-1980 hydrogen cost.

The rapid growth of world population, coupled with comparable increases in per capita consumption of energy, will place stringent demands on the energy market in the 1980 to 2000 period. The demand for fossil fuels will be most competitive, as will be the demands for fresh water. The increased demand will tend to drive the price of these commodities upward if the supply remains constant. This demand will also increase the incentive to find substitutes, more efficient use, and new sources, thus reducing unit costs of energy. The past levels and trends of production and consumption of potential material resources are intended to show the background for projection of price trends of raw materials.

4.1.1 FOSSIL FUELS AND ENERGY

Today, almost all of the energy consumed in the world is derived from fossil fuels. A small percentage comes from water power and nuclear plants. Energy is consumed in four general and rapidly growing sectors of the economy; industry, transportation, domestic uses, and the electric utilities. Thus, aircraft fueling, as part of transportation, competes with these other sectors and is confronted also with rapidly growing automotive demands. It is generally estimated that world demand for energy by 1980 will be more than twice today's level.[34 thru 38] With regard to fossil fuels, the need to exploit marginal or less accessible geological and geographical deposits, plus more efficient use of this energy potential, will become even greater economic factors.

At the present time, oil is the most widely used fossil fuel for energy. The use of natural gas is increasing most rapidly, and coal appears to be the most abundant from the standpoint of total resources. The knowledge that fossil fuel existence as natural resources is not infinite has supplied the incentive for the development and application of nuclear power sources, just as in the past (and continuing today) geographical shortages of fossil fuels have stimulated the development of hydroelectric power. Ultimately, energy converted from nuclear and other known but undeveloped sources must displace fossil fuels if world population and attendant socia-industrial trends are to continue.[39] To what extent these sources will be applied in the early post-1980 period poses a difficult question in evaluating raw material and, more particularly, power costs for hydrogen fueling. Present indications are that non-fossil fuel energy will make strong inroads on the electric utility sector which will, to some extent, alleviate fossil fuel demands in the domestic-use sector. However, if this results in expected lower-cost power, there will be greater demands for fossil fuels in the industrial and transportation sectors. Complete evaluation of these cost effects depends to a great degree on proper assessment of the rate of change and the rate of technological development, which is difficult to predict. It is anticipated that the fossil fuels, based on traditional and current trends, will continue as the major energy source in the near post-1980 period. However, any particular demand situation, geographical or otherwise, may swing the balance to a nuclear (or other) energy source; for example, the use of water as the feed stock to obtain hydrogen.

4.1.2 WATER

Although water is abundant throughout the world, the demand for fresh water in urban and other areas is becoming critical and creating severe supply problems. This increases the cost of water, since demand tends to outstrip supply. Therefore, a liquid hydrogen production facility may have to provide its own water purification system, with attendant investments and operating costs, or pay an operating premium for makeup water. This demand for fresh water may provide an opportunity to combine a liquid hydrogen plant with a water desalinization system, and despite potential local scarcities, availability of raw water is not considered a logistics problem.

4.1.3 NUCLEAR POWER RAW MATERIALS

Materials such as uranium, thorium, and heavy water, which may be required for nuclear reactors, do not appear to be in great demand at the present time. The location of uranium deposits and their estimated magnitude are given in Appendix A. In the event that reactors requiring heavy water are to be considered for wide use, additional heavy water production plants will be needed. At the present time, however, existing heavy water plants have been idled for lack of demand.

4.2 SPECIFIC REQUIREMENTS FOR LIQUID HYDROGEN PRODUCTION

Raw materials required for hydrogen production can be assigned to four categories:

(1) Fossil fuels - which can be used as feed stock, fuel, or for the production of power

(2) Nuclear power raw materials - such as uranium, thorium, and heavy water used for the generation of electricity

(3) Water - to be used as process water for cooling or as a source of steam. Demineralization of this water for use in electrolytic cells is included in investment and operating costs associated with that process.

(4) Other raw materials required in lesser quantities; notably nitrogen, oxygen, and refrigerants

Of these, only the fossil fuels present a general problem in logistics; i.e., the movement of large quantities of material to the plant site. The movement of electricity generated at the mine mouth or at a hydroelectric site to the usepoint by high voltage transmission lines will pose logistic problems at several locations.

4.2.1 FOSSIL FUELS

The annual fossil fuel requirements for each proposed liquid hydrogen plant location are shown in table 4-1. Using current information, it appears that oil would be used as fuel in the steam reforming process because of the relatively high cost of gas at all tentative plant locations. The table was designed primarily to show the maximum effect of this program on the fossil fuel market at each locality. Oil is assumed to be the fuel for power production, except at Johannesburg and Sydney where coal is favored. Actually, non-fossil fuel power generation is desirable in most instances.

4.2.2 NUCLEAR POWER RAW MATERIALS

In this report, uranium, thorium, or heavy water requirements for the nuclear power plants are considered part of initial investment or as replacement materials in operation costs rather than as raw materials. This method of consideration is most convenient since there is presently no substantial trade in these commodities. Quantities required are not categorized as bulk transport. Their delivery to the use-point would be in a purified and modified form such as fuel elements, which more closely resemble manufactured products than raw materials.

4.2.3 WATER

Although water requirements for process, cooling, and generation of steam are high, it is unlikely that water will pose a logistics problem, since it will undoubtedly be acquired locally. Therefore, water will be a principal factor in plant location. Water costs are, however, considered in subsequent paragraphs of this section.

4.2.4 OXYGEN

In addition to major raw material requirements for hydrogen gas generation, the partial oxidation and coal gasification processes analyzed for this study require oxygen. For partial oxidation, it has been more convenient to include the cost of the oxygen (air separation) plant as part of the overall hydrogen facility. Hence, the cost of oxygen is included as part of capital investment and operating costs. In the case of coal gasification, process cost analysis was simplified by treating oxygen as a raw material.

4.2.5 REFRIGERANTS

The liquefaction process for all locations embodies a cascade refrigeration cycle requiring nitrogen, methane, propane, and ethylene. If the partial oxidation process is employed, nitrogen is available for the liquefier from the air separation plant associated with that process.

4.3 RAW MATERIAL AVAILABILITY

The first factor to be investigated is the geographical distribution and magnitude of resources of coal, petroleum, natural gas, uranium, and water power. Will the increasing world energy demands

TABLE 4-1. ANNUAL RAW MATERIAL REQUIREMENTS FOR LIQUID HYDROGEN PRODUCTION (U.S. Tons x 10^3)

Site and Type	Steam Reforming	Partial Oxidation	Electrolysis	Coal Gasification	Liquefaction	Total (Maximum)	Total (Probable)
Bangkok - 200 T/D							
Natural gas	157	-	-	-	-	157	157
Oil	152	353	-	-	-	995	568
Coal	-	-	-	-	-	-	-
Electricity (gwh)	78	215	3,564	-	709	-	-
Oil equivalent	18	50	830	-	165	-	-
Johannesburg - 200 T/D							
Natural gas	157	-	-	-	-	157	157
Oil	152	353	-	-	-	353	353
Coal	-	-	-	721	-	1,965	1,090
Electricity (gwh)	78	215	3,564	94	709	-	-
Coal equivalent	36	99	1,639	43	326	-	-
Honolulu - 200 T/D							
Natural gas	157	-	-	-	-	157	157
Oil	152	353	-	-	-	995	568
Coal	-	-	-	-	-	-	-
Electricity (gwh)	78	215	3,564	-	709	-	-
Oil equivalent	18	50	830	-	165	-	-
Beirut - 300 T/D							
Natural gas	232	-	-	-	-	232	232
Oil	223	529	-	-	-	1,490	851
Coal	-	-	-	-	-	-	-
Electricity (gwh)	116	321	5,335	-	1,062	-	-
Oil equivalent	27	75	1,243	-	247	-	-

TABLE 4-1. ANNUAL RAW MATERIAL REQUIREMENTS FOR LIQUID HYDROGEN PRODUCTION (Continued)

(U.S. Tons x 10^3)

Site and Type	Steam Reforming	Partial Oxidation	Electrolysis	Coal Gasification	Liquefaction	Total (Maximum)	Total (Probable)
Sao Paulo - 400 T/D							
Natural gas	314	-	-	-	-	314	314
Oil	298	706	-	-	-	1,990	1,135
Coal	-	-	-	-	-	-	-
Electricity (gwh)	155	429	7,128	-	1,414	-	-
Oil equivalent	36	100	1,661	-	329	-	-
Sydney - 400 T/D							
Natural gas	314	-	-	-	-	314	314
Oil	298	706	-	-	-	706	706
Coal	-	-	-	1,997	-	3,365	2,603
Electricity (gwh)	155	429	7,128	125	1,414	-	-
Coal equivalent	61	169	2,808	49	557	-	-
Tokyo - 500 T/D							
Natural gas	391	-	-	-	-	391	391
Oil	368	880	-	-	-	2,486	1,417
Coal	-	-	-	-	-	-	-
Electricity (gwh)	194	536	8,899	-	1,766	-	-
Oil equivalent	45	125	2,074	-	412	-	-
Los Angeles - 1300 T/D							
Natural gas	1,017	-	-	-	-	1,017	1,017
Oil	922	2,274	-	-	-	6,453	3,666
Coal	-	-	-	-	-	-	-
Electricity (gwh)	502	1,391	23,112	-	4,585	-	-
Oil equivalent	117	324	5,385	-	1,068	-	-

TABLE 4-1. ANNUAL RAW MATERIAL REQUIREMENTS FOR LIQUID HYDROGEN PRODUCTION (Concluded)
(U.S. Tons x 10^3)

Site and Type	Steam Reforming	Partial Oxidation	Electrolysis	Coal Gasification	Liquefaction	Total (maximum)	Total (Probable)
Paris - 2000 T/D							
Natural gas	1,564	-	-	-	-	1,564	1,564
Oil	1,390	3,465	-	-	-	9,929	5,609
Coal	-	-	-	-	-	-	-
Electricity (gwh)	774	2,140	35,554	-	7,060	-	-
Oil equivalent	180	499	8,284	-	1,645	-	-
New York - 2500 T/D							
Natural gas	1,954	-	-	-	-	1,954	1,954
Oil	1,700	4,308	-	-	-	12,407	6,986
Coal	-	-	-	-	-	-	-
Electricity (gwh)	967	2,674	44,431	-	8,822	-	-
Oil equivalent	225	623	10,352	-	2,055	-	-

NOTES:

Fossil fuel requirements for electricity have been computed using 8,750 Btu/kwh.

Fuel requirements are based on the following conversion factors:

Natural gas	35.3 x 10^3 Btu/cubic meter
Crude Oil (Bunker C)	41.7 x 10^6 Btu/metric ton
Coal	26.5 x 10^6 Btu/metric ton

In the event that the Btu content of any of the fuels delivered to the plant sites differ from these factors, a correction must be applied.

Natural gas feed stock requirements are based on pure methane. Correction to actual natural gas composition is also necessary.

exceed available resources or the technology of recovering them at acceptable costs? Are the resources located sufficiently close to the proposed liquid hydrogen plant sites so as to eliminate excessive transportation costs? Secondly, are the rates of production, consumption, exports, and imports (of fossil fuels and electricity in the country or vicinity of each hydrogen plant location) such that this program will not significantly alter the supply and demand balance or require special production and handling facilities?

4.3.1 WORLD DISTRIBUTION OF RESOURCES

Fossil fuels are of prime importance as a major source of energy for transportation. These fuels could be used also as a source of hydrogen and generation of electric power.

Since fossil fuels are transportable at relatively reasonable cost, their availability can be considered on a world-wide as well as on a local basis. The same can be stated for uranium and thorium, but since they are not required in bulk, their transportation costs become comparatively insignificant. Water power, even when converted to electricity, needs to be relatively close to proposed locations in order to be economically feasible. The availability of fossil fuel resources has been cited in several ways, as given in Appendix A.

4.3.2 PRODUCTION, CONSUMPTION, AND TRADE

The level of production, consumption, and trade of raw materials and electricity for the world and the countries in which liquid hydrogen plants are suggested is shown in table 4-2. It also indicates whether an abundance or deficit of the commodity exists in a particular location.

4.3.2.1 World and Local Production of Coal, Lignite, Petroleum, and Natural Gas

The historical production of fossil fuels from 1948 to 1964 are presented in tables 4-3 through 4-6. These tables show not only the growth trends of each fuel, but also the relative levels of production in each of the countries for which liquid hydrogen plants have been suggested. In effect, table 4-2 relates production to consumption in the countries of interest, thus indicating dependency on outside sources for the various fuels to be considered in LH_2 production. The factors used in converting to coal equivalent using United Nations data is as follows:

	Coal Equiv. (Metric Tons)
Coal, anthracite, or bituminous (metric tons)	1.0
Brown coal or lignite (metric tons) (France)	0.6
Brown coal or lignite (metric tons) (other countries)	0.33
Crude petroleum (metric tons)	1.3
Natural gas (1000 cubic meters)	1.332
Hydro or nuclear electricity (1000 Kwh)	0.125

TABLE 4-2. PRODUCTION[a], CONSUMPTION[b], AND TRADE OF ENERGY, 1964
(In Millions of Metric Tons of Coal Equivalent)

Country or Area	Total Energy				Solid Fuels		Liquid Fuels		Gas		Hydro & Nuclear Elec	
	Produced	Consumed	Net Imports	Bunkers	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed
WORLD	5093	4965	-36	150	2241	2227	1867	1757	880	875	105	105
United States	1580	1685	119	19	456	403	515	662	586	597	23	23
Brazil	10.9	28.8	16.6	0.4	1.7	3.2	5.7	22.1	0.7	0.7	2.8	2.8
France	69.7	142.1	78.0	3.6	54.4	73.1	4.1	57.4	6.8	6.9	4.4	4.7
Lebanon	0.05	1.65	1.76	0.33	-	0.01	-	1.59	-	-	0.05	0.05
Japan	63.6	161.1	111.6	14.7	51.2	64.7	0.9	85.0	2.7	2.7	8.8	8.8
Thailand	0.05	3.17	3.36	0.21	0.03	0.04	0.01	3.13	-	-	-	-
Australia	35.5	49.6	14.6	2.4	34.3	29.0	0.2	19.7	0.004	0.004	1.0	1.0
South Africa	44.9	50.5	6.3	1.4	44.9	43.6	0.01	6.9	-	-	-	-

[a] Production - refers to the first stage of production; i.e., coal at minehead, oil or gas at wellhead. Electricity, in this instance, is limited to non-fossil (or non-thermal) production, since any coal, gas, or oil used to produce electricity has already been counted, as has been fuel used to produce coke, gasoline, and manufactured gas.

[b] Consumption - derived by subtracting exports and additions to stocks or bunkers from production plus imports.

TABLE 4-3. PRODUCTION OF COAL
(Metric Tons x 10^3)

Location	1948	1955	1956	1957	1958	1959	1960	1961	1962	1963	1964	1965	1966
WORLD	1,410,700	1,598,500	1,687,800	1,735,300	1,818,600	1,894,300	1,985,300	1,811,300	1,857,100	1,929,000	-	-	-
Australia	15,020	19,584	19,583	20,239	20,770	20,624	22,931	24,391	24,863	25,256	27,840	32,016	33,852
Brazil	1,424	1,676	1,550	1,466	1,312	1,284	1,277	1,256	1,599	1,536	1,680	1,764	
France	43,291	55,335	55,129	56,795	57,721	57,606	55,960	52,358	52,369	47,754	53,028	51,348	50,340
Japan	33,726	42,423	46,555	51,732	49,674	47,258	51,967	54,484	54,399	52,052	50,928	49,536	51,254
South Africa	24,017	32,147	33,602	34,769	37,085	36,453	38,173	39,565	41,281	42,455	44,916	48,540	48,205
United States	592,911	442,410	477,992	467,595	389,355	390,134	390,134	391,526	395,522	430,450	454,704	475,284	491,364
Lebanon	-	-	-	-	-	-	-	-	-	-	-	-	-
Thailand	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE 4-4. PRODUCTION OF LIGNITE
(Metric Tons x 10^3)

Location	1948	1955	1956	1957	1958	1959	1960	1961	1962	1963	1964	1965	1966
WORLD	316,200	534,600	564,200	592,800	613,400	617,200	639,900	662,500	685,100	719,800	-	-	-
Australia	6,800	10,274	10,729	10,913	11,831	13,244	15,207	16,540	17,412	18,754	19,344	21,048	22,152
Brazil	-	-	-	-	-	-	-	-	-	-	-	-	-
France	1,837	2,053	2,252	2,294	2,318	2,175	2,276	2,906	2,882	2,475	2,244	2,688	2,625
Japan	2,575	1,368	1,552	1,662	1,582	1,469	1,409	1,309	1,111	914	696	576	446
South Africa	-	-	-	-	-	-	-	-	-	-	-	-	-
United States	2,799	2,872	2,611	2,365	2,202	2,384	2,491	2,738	2,771	2,454	2,676	2,760	
Lebanon	-	-	-	-	-	-	-	-	-	-	-	-	-
Thailand	-	41	88	-	103	110	108	108	135	137	104	125	-

TABLE 4-5. PRODUCTION OF NATURAL GAS
(Cubic Meters x 10^6)

Location	1948	1955	1956	1957	1958	1959	1960	1961	1962	1963	1964	1965	1966
WORLD	-	-	-	-	-	-	-	-	-	-	-	-	-
Australia	-	-	-	-	-	-	-	-	3	3	4	6	-
Brazil	3	62	84	158	300	429	535	527	511	503	532	683	-
France	160	256	306	439	682	1,645	2,846	4,010	4,740	4,862	5,088	5,052	5,148
Japan:													
(1) Petroleum and gas fields	56	145	168	230	356	468	676	894	1,167	1,678	1,836	1,776	1,810
(2) Coal mines	-	41	74	99	123	136	148	158	200	208	240	240	270
South Africa	-	-	-	-	-	-	-	-	-	-	-	-	-
United States	145,776	266,331	285,491	302,433	312,344	341,110	361,637	375,314	392.944	417,581	437,844	454,200	478,428
Lebanon	-	-	-	-	-	-	-	-	-	-	-	-	-
Thailand	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE 4-6. PRODUCTION OF CRUDE PETROLEUM
(Metric Tons x 10^3)

Location	1948	1955	1956	1957	1958	1959	1960	1961	1962	1963	1964	1965	1966
WORLD	467,100	771,700	839,400	884,600	906,500	978,600	1,053,900	1,122,000	1,216,200	1,303,500	-	-	-
Australia	-	-	-	-	-	-	-	-	-	-	190	334	447
Brazil	19	264	531	1,321	2,473	3,083	3,870	4,549	4,366	4,667	4,356	4,488	5,273
France	70	885	1,264	1,410	1,387	1,621	1,983	2,163	2,370	2,522	2,844	2,988	2,928
Japan	159	314	310	321	367	406	526	657	760	785	658	676	777
South Africa	40	37	37	25	24	26	25	6	6	6	-	-	-
United States	273,007	335,744	353,698	353,646	330,955	347,929	347,975	354,303	361,658	372,001	376,608	384,984	410,000
Lebanon	-	-	-	-	-	-	-	-	-	-	-	-	-
Thailand	-	-	-	-	-	4.6	6.1	7.6	7.6	12.2	6.9	-	-

4.3.2.2 Major Sources

It is evident from analysis of table 4-2 that most countries in which liquid hydrogen plant locations are being considered use more energy than they produce. This being the case, fuels are being imported from areas having an excess of one or more fuels. The major producers of oil, coal, and natural gas in 1964 are shown in table 4-7.

4.3.2.3 Solid Fuels

Primarily all Western European exports of solid fuels consist of interchanges between the Western European countries and themselves. Polish and Russian exports remain largely within the Eastern European block. The United States is the world's major exporter of coal, exporting 45.4 x 10^9 metric tons in 1964. Australia and South Africa exported 5 x 10^9 and 3 x 10^9 metric tons, respectively, which makes them economically feasible sources insofar as geography permits.[40,41]

4.3.2.4 Crude Oil

Oil comes from four main areas of the world: The United States, the Caribbean, the Near East (extending into North Africa), and Russia. Indonesian fields may become significant. Except for the United States, and to a lesser degree Russia, the oil-rich countries are relatively small consumers; hence, oil is far more available than coal in the world fuel market. The oil demand for large liquid hydrogen producing plants could be readily met from a number of sources, with far less dependency on geography than in the case of coal.[40,42 thru 49]

4.3.2.5 Natural Gas

The trade in natural gas is presently almost non-existent, except for a modest amount flowing from Canada to the United States. Future developments already being planned will include pipelines from Eastern to Western Europe, utilization of North Sea finds in Western Europe and the United Kingdom, and transport of substantial quantities of liquefied natural gas by tanker from fields to markets in Japan, Western Europe, and the United States.[37,40 thru 50]

4.3.3 COMPARISON OF REQUIREMENTS WITH RESOURCES, PRODUCTION, CONSUMPTION, AND TRADE

The comparison impact that hydrogen plant requirements will have on each country's ability to supply the desired raw materials in light of their own resources or their 1964 trade is shown in tables 4-8 through 4-10. Production, consumption, and trade figures of 1964 were used because they were the latest available on a comparable basis. For lack of separate data, Hawaii has been included in the United States evaluations, although, obviously, supply from the mainland requires separate consideration. Requirements and resource data are taken from Appendix A and table 4-1. Production, consumption, and trade figures come from U.N. Statistical Paper.[40] Data differs slightly from production figures previously cited from other U.N. publications. In general, resource data does not reflect recent discoveries not yet fully measured. Hence, Australia and South Africa show a small production of oil but no resources. Also, the gas discoveries in Australia have been too recent to reach statistical

TABLE 4-7. MAJOR FUEL PRODUCERS
(Other Than Hydrogen Plant Location Countries)

Country or Area	Coal Producers		Country or Area	Crude Oil Producers	
	Quantity (Metric Tons x 10^3)	Exports[a]		Quantity(Metric Tons x 10^3)	Exports
Western Europe	440	43	Caribbean Area	213	130
Common Market	229	35	Venezuela	179	122
West Germany	143	26	Middle East	428	369
United Kingdom	197	7	Iran	84	63
			Iraq	62	59
Poland	117	23	Kuwait	107	95
			Saudi Arabia	86	72
U.S.S.R.	214	27	Libya	41	41
China, Mongolia[b]	305	1	Indonesia	23	15
			Algeria	26	25
			U.S.S.R.	224	37

Country or Area	Natural Gas Producers		Country or Area	Lignite Producers	
	Quantity (Metric Tons x 10^3)	Exports		Quantity(Metric Tons x 10^3)	Exports[a]
Canada	39	11	Western Europe	59	43[b]
Mexico	11	1.5	Common Market	36	35[b]
Western Europe	18	0	West Germany	35	26[b]
Common Market	16	0	Yugoslavia	14	0.1
Rumania	14	0.2	Czechoslovakia	45	5
U.S.S.R.	109	0.3	East Germany	77	5
			U.S.S.R.	212	27[b]

[a] Includes all solid fuels measured in coal equivalent.

[b] Also reported as coal exports.

TABLE 4-8
COMPARISON OF ANNUAL REQUIREMENTS OF RESOURCES, PRODUCTION, CONSUMPTION, AND IMPORTS OF OIL
(All Units in Metric Tons x 10^3)

Country or Area	Max. Probable Req'ment	Resources	1964 Production	1964 Supply	1964 Imports
Thailand	516	-	10	680	670
South Africa	321	-	10	4,860	4,850
Lebanon	773	-	-	1,310	1,310
Brazil	1,032	48,000	4,350	15,160	10,800
Australia	642	-	190	15,300	15,110
Japan	1,288	3,900	680	63,000	62,320
France	5,099	30,000	3,140	52,420	49,270
United States:					
Honolulu	516	-	-	-	-
Los Angeles	3,333	49,500,000	396,330	456,860	60,830
New York	6,351	-	-	-	-
WORLD	19,871	81,600,000	1,436,390	1,434,880	-

TABLE 4-9
COMPARISON OF ANNUAL REQUIREMENTS WITH RESOURCES, PRODUCTION, CONSUMPTION, AND IMPORTS OF NATURAL GAS
(All Units in Metric Tons x 10^3)

Country or Area	Max. Probable Req'ment	Resources	1964 Production	1964 Supply	1964 Imports
Thailand	143	-	-	-	-
South Africa	143	-	-	-	-
Lebanon	211	-	-	-	-
Brazil	285	7,170	530	530	-
Australia	285	[a]	-	-	-
Japan	355	95,100	1,400	1,400	-
France	1,422	124,300	3,400	3,400	-
United States:					
Honolulu	143	-	-	-	-
Los Angeles	925	5,080,000	294,000	299,000	8,400
New York	1,776	-	-	-	-
WORLD	5,688	34,100,000	441,000	438,000	-

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[a] Most discoveries have been recent and are still in the process of being measured.

TABLE 4-10
COMPARISON OF ANNUAL REQUIREMENTS WITH RESOURCES, PRODUCTION, CONSUMPTION, AND IMPORTS OF COAL

(All Units in Metric Tons x 10^3)

Country or Area	Max. Probable Req'ment	Resources	1964 Production	1964 Supply
South Africa	991	21,443,000	44,916 b	43,630
Australia	2,366	1,791,000	27,840 19,344[a]	28,960
WORLD	3,357	771,000,000	1,801,110 439,960[b]	2,247,470

[a] All solid fuels, coal plus lignite plus small amounts of peat and wood in terms of coal equivalent.

[b] Lignite

TABLE 4-11. COST OF OIL AT SOURCE
(As of 3-15-67, Per Market Prices, Petroleum Times, 3-17-67)

Source	Type	Sales Conditions	Posted Price U.S. $/bbl	Last Change In Price
United States:				
New Orleans	Bunker C	Bulk export, cargo lots	2.00 - 2.20	-
New Orleans	Bunker C	Bulk export, bunker lots	2.19	-
New York	Bunker C	Bulk export	2.25	-
Mid-continent	Crude API below 29	Wellhead	2.68 - 2.69	2-24-67
Mid-continent	Crude API 40+	Wellhead	3.02 - 3.08	2-24-67
Signal Hill, Calif.	Crude API 18 - 18.9	Wellhead	2.13	2-24-67
Signal Hill, Calif.	Crude API 30 - 30.9	Wellhead	2.93	2-24-67
Conroe County	Crude Flat	Wellhead	3.25 - 3.35	2-24-67
East Texas	Crude Flat 40 - 44.9	Wellhead	3.15	2-24-67
Panhandle, Texas	Crude Flat 34 - 34.9	Wellhead	2.82	2-24-67
Gulf Coast, Texas	Crude Flat below 20	Wellhead	3.26	2-24-67
Gulf Coast, Texas	Crude Flat 40+	Wellhead	3.68	2-24-67
Middle East:				
Abadan, Iran	Bunker C	Bulk shipment, F.O.B.	1.55	8-17-66
Abadan, Iran	Light fuel	Bulk shipment, F.O.B.	1.55	6-15-66
Ras Tanura, Iran	Bunker C	Bulk shipment, F.O.B.	1.65	6-27-66
Ras Tanura, Iran	Light fuel	Bulk shipment, F.O.B.	1.65	3-1-66
Ras Tanuar, Iran	Heavy fuel	Bulk shipment, F.O.B.	1.60	3-1-66
Kharg Island, Iran	Crude API 31 - 31.9	(At port, exclusive of taxes and post charges. $0.02 barrel escalation.)	1.63	4-1-66
Abadan, Iran	Crude API 31 - 31.9		1.58	4-1-66
Abadan, Iran	Crude API 34 - 34.9		1.73	4-1-66
Kharg Island, Iran	Crude API 34 - 34.9		1.79	4-1-66
Bandar Mashur, Iran	Crude API 34 - 34.9	As above for all Middle East listings except Bunker C, light fuel, and heavy fuel.	1.78	12-15-66
Tripoli	Crude API 34 - 34.9		2.07[a]	1-27-65
Fao	Crude API 35 - 35.9		1.72	1-27-65
Khor-al-Amaya	Crude API 35 - 35.9		1.72	1-27-65

TABLE 4-11. COST OF OIL AT SOURCE (Continued)
(As of 3-15-67, Per Market Prices, Petroleum Times, 3-17-67)

Source	Type	Sales Conditions	Posted Price U.S. $/bbl	Last Change In Price
Middle East: (Cont'd)				
Tripoli/Banias	Crude API 35 - 35.9	(At port, exclusive of taxes and post charges. $0.02 barrel escalation.)	2.19[a]	2-5-64
Tripoli/Banias	Crude API 36 - 36.9		2.21[a]	2-5-64
Halul Island	Crude API 36 - 36.9		1.83	6-15-66
Das Island	Crude API 37 - 37.9		1.86	1-27-65
Jebel Dhanna	Crude API 38 - 38.9	As above for all Middle East listings except Bunker C, light fuel, and heavy fuel.	1.86	2-5-64
Jebel Dhanna	Crude API 39 - 39.9		1.88	2-1-65
U. Said	Crude API 41 - 41.9		1.95	2-5-64
Ras Tanura, Iran	Crude API 27 - 27.9		1.47	10-1-60
Ras Tanura, Iran	Crude API 31 - 31.9		1.59	10-1-60
Ahmadi	Crude API 31 - 31.9		1.59	8-16-60
Ras Tanura, Iran	Crude API 34 -34.9		1.80	10-1-60
Sidon	Crude API 34 - 34.9		2.17	10-1-60
Mina Saudi	Eocene Crude API 16.5 - 17.4		1.28	10-1-60
Mina Saudi	Ratawi Crude API 23.5 - 24.4		1.41	10-1-60
Mina Saudi	Burghan Crude API 23.5 - 24.4		1.48	10-1-60
Ras al Khafji	Khafji Crude API 26 - 26.9		1.42	10-20-60
Far East:				
Singapore	Medium fuel	Bulk shipment, F.O.B.	1.95	1-18-67
Sungei Gerong, Sumatra	Light fuel	Bulk shipment, F.O.B.	1.95	5-2-66
Sungei Gerong, Sumatra	Medium fuel	Bulk shipment, F.O.B.	1.90	2-25-66
Sungei Gerong, Sumatra	Heavy fuel	Bulk shipment, F.O.B.	1.90	5-2-66
Lutong, Sarawak	Light Seria curde API 37	Bulk shipment, F.O.B.	2.15	11-1-65

TABLE 4-11. COST OF OIL AT SOURCE (Concluded)
(As of 3-15-67, Per Market Prices, Petroleum Times, 3-17-67)

Source	Type	Sales Conditions	Posted Price U.S. $/bbl	Last Change In Price
North Africa:				
Bougie, Algeria	Hassai Messaoud - Crude API 40+	F.O.B. - port	2.35	1-1-64
La Skhirra, Algeria	Zarzaitine - Crude API 40+	F.O.B. - port	2.30	1-1-64
Arzew, Algeria	Arzew - Crude API 40+	F.O.B. - port	2.365	1-1-64
Sider, Libya	Libyan High Pour API 36 - 36.9	F.O.B. - port	2.10	3-29-66
Sider, Libya	Light Libyan API 36 - 36.9	F.O.B. - port	2.15	3-29-66
Sider, Libya	Light Libyan	F.O.B. - port	2.21	3-29-66
Sider, Libya	Light Libyan	F.O.B. - port	2.23	3-29-66
Caribbean:				
Aruba	Bunker C	Bulk shipment, F.O.B.	2.00	3-29-66
Curacao	Bunker C	Bulk shipment, F.O.B.	2.00	4-6-66
Punta Cardon, Ven.	Bunker C	Bulk shipment, F.O.B.	2.00	3-29-66
San Lorenzo, Ven.	Bunker C	Bulk shipment, F.O.B.	1.97	4-6-66
Porto Miranda, Ven.	Crude (Boscan) API 10	Bulk shipment, F.O.B.	1.57[b]	1-4-62
Porto Miranda, Ven.	Crude (Laguna) API 11	Bulk shipment, F.O.B.	1.62[b]	1-4-62
Porto Miranda, Ven.	Crude (Pesado) API 12 - 13	Bulk shipment, F.O.B.	1.67[b]	1-4-62
Porto Miranda, Ven.	Crude (Bachaquero) API 13	Bulk shipment, F.O.B.	1.70[b]	1-4-62
Puerto La Cruz	Crude (Merey) API 17	Bulk shipment, F.O.B.	1.88[b]	1-4-62
Porto Miranda, Ven.	Crude (Tamare) API 18	Bulk shipment, F.O.B.	1.87[b]	1-4-62
Caripito	Crude (Pilon) API 14 -15	Bulk shipment, F.O.B.	1.70[b]	2-15-63
La Salina	Crude (Bachaquero) API 17 - 18	Bulk shipment, F.O.B.	1.85[b]	2-15-63
La Salina	Crude (Tia Juana-heavy) API 18	Bulk shipment, F.O.B.	1.90[b]	2-15-63
Capure	Crude (Pedernales) API 21-21.4	Bulk shipment, F.O.B.	1.90[b]	2-15-63

[a] Mediterranean port - end of pipeline

[b] Irrespective of gravity - increase or decrease by $0.025/bbl for each 0.5° API change from listing

summaries which are usually a year or two in preparation. Since the coal gasification process does not appear attractive other than in Australia and South Africa, table 4-9 has been restricted to these locations. An extrapolation of production and/or consumption data was anticipated to predict the 1980 and 2000 usage of these materials in each country for purposes of comparison with plant requirements. However, further review of factors influencing their growth became too complex to be worthwhile for this purpose, and, therefore, was limited to electricity.

4.4 RAW MATERIAL COSTS

Except for a few instances where public utilities have reported their fuel costs, little price data is available on negotiated large-scale purchases. Hence, average or posted prices have been cited in most cases, although it is known that negotiated bulk purchases are made at 10 to 25 percent discounts. Another area of cost quite difficult to evaluate, particularly with future projection in mind, is that of government-imposed royalties and duties at points of export and import. They tend to vary and to be subject to special conditions. Insofar as possible, without making a special study of duties and royalties, their effect has been considered in the raw material costs.

4.4.1 RAW MATERIALS AT SOURCE

Recent price postings of oil at prominent points of origin are given in table 4-11. These prices, with addition of appropriate freight costs, have been used to determine the cost of oil at the various use points.

Very few sources of natural gas for export have been developed; hence, there are no generally published postings at present.

In the case of coal, the export price at the country of origin, plus the best freight data available (without extensive study), does not seem to add up to the price paid at the point of use. The difference may occur as a result of duties and local handling costs, which tend to be high because of the bulk and non-fluid qualities of coal.

4.4.2 RAW MATERIALS AT USE-POINT

4.4.2.1 Fossil Fuels

Since raw material costs used are based on published data and current practices, no credit has been taken for special arrangements beyond that of assuming that large quantity discounts may offset duties and port charges. Table 4-12 serves as an adjunct to table 4-13, since tonnage costs of oil are derived from the per barrel cost by applying a density factor related to the degrees API or the heating value shown in table 4-13.

The cost of coal, oil, and gas at each proposed plant location is listed in Table 4-13. Origin of the costs is shown in the "Justification" column of table 4-12. Slightly lower raw material costs may be attainable through competitive negotiations for the very-large-scale requirements and through special concessions on the part of the host government insofar as duties and taxes are concerned.

TABLE 4-12. HEAT VALUES OF PETROLEUM OILS

Deg API at 60°F	Density, Lb per Gal [a]	High Heat Value at Constant Volume Qv, Btu	
		Per Lb	Per Gal
10	8337	18,540	154,600
20	7787	19,020	148,100
30	7305	19,420	141,800
40	6879	19,750	135,800
50	6500	20,020	130,100
60	6160	20,260	124,800
70	5855	20,460	119,800
80	5578	20,630	115,100

[a] The values in this column are true values obtained by correcting all weights for the buoyancy of air.

4.4.2.2 Nuclear Raw Materials

All costs, including uranium, thorium, heavy water, or other materials necessary for the production of electricity by nuclear reaction, are covered in the section which develops power costs.

4.4.2.3 Water

Costs for the supply and purification of water have been placed at three probable levels. Where fresh water is considered plentiful, a cost of 0.72 cent per ton has been applied for nominal treatment and pumping. Where fresh water must be piped a modest distance, a nominal figure of 1.2 cents/ton has been applied. Where fresh water is known to be scarce, the approximate Los Angeles cost of 7.2 cents/ton has been applied. More analysis of water costs to obtain greater price accuracy is not warranted since the effect on the liquid hydrogen cost is negligible.

4.4.2.4 Water Power

The availability of water power for hydroelectricity has been explored at most locations.[51] Since application is limited to the generation of electricity, its cost, as covered later in this section, is applicable. A brief discussion on its availability follows.

TABLE 4-13. COSTS OF RAW MATERIALS AT USE-POINT

Plant Location	Cost $ Per MCF	Type	Justification
		Natural Gas	
Honolulu (access to sea)	0.45	Alaskan LNG, 1000 Btu/scf	No commercial basis. Assumes price in Hawaii to be comparable to Japanese price without duty
Bangkok (access to sea)	0.45	LNG from Indonesia 1000 Btu/scf	No commercial basis. Assumes liquefaction in Indonesia and tanker transport
Johannesburg	0.45	LNG from Middle East or North Africa, 1000 Btu/scf	No commercial basis. Assumes liquefaction at source plus tanker transport
Beirut	0.35	Piped from Arabia or Iraq, 1000 Btu/scf	No commercial basis. Assumes pipeline from source, about 400 mi.
Sao Paulo	0.40	Piped from Bolivian fields, 1000 Btu/scf	No commercial basis. Assumes pipeline from source, about 700 mi.
Sydney	0.35[a]	Gippsland or other Australian fields, 1000 Btu/scf	Reflects anticipated well price of 29 cents plus piping to Sydney
Tokyo (access to sea)	0.368	LNG from Sakhalin Island, 1000 Btu/scf	Reflects current Russo-Japanese negotiations
Los Angeles (access to pipelines)	0.338	Pipeline gas, 1077 Btu/scf	Best actual price per MM Btu on annual basis by power plant in Los Angeles County
Paris	0.48	LNG from Algeria, 1000 Btu/scf	40.9 cents, cif Medit. port, plus pipeline cost to Paris; some mixture with 43.6-cent domestic gas
New York	0.363	Pipeline gas, 1045 Btu/scf	Best actual price per MM Btu on annual basis at New York power plant

TABLE 4-13. COSTS OF RAW MATERIALS AT USE-POINT (Continued)

Plant Location	Cost: \$ Per Short Ton	Type	Justification
		Coal	
Honolulu (access to sea)	15.00	Australian bit., 11,100 Btu/lb	Referenced to actual cost in Japan
Bangkok (access to sea)	12.50	Australian bit., 11,100 Btu/lb	Referenced to actual cost in Japan
Johannesburg	1.80	So. African bit., mine-mouth	Government-controlled mine price of 1.76/T plus handling and freight for 100-mi. average
	2.40	Johannesburg, 9500 Btu/lb	
Beirut	15.00	Polish or Russian	Little basis for price since no trade exists; assume higher than Western Europe
Sao Paulo	14.00	U.S. bit., 13,300 Btu/lb	\$8.50/T from Norfolk plus \$5.50 freight; no duty included
Sydney	2.30	Australian black, minemouth	Anticipated minemouth price at Lidell power plant plus handling and freight to Sydney
	3.00	Sydney, 11,100 Btu/lb	
Tokyo (access to sea)	13.42	Australian black, 11,100 Btu/lb	Average price, quantity purchase to offset duty and handling
Los Angeles	9.85[b]	Colorado lignite, 10,000 Btu/lb	\$4.50/T loaded in Colorado plus \$5.35/T freight
Paris	13.36	U.S. bit., 13,300 Btu/lb	\$11.36/T for steaming coal, allow \$2.00 for freight and customs
New York (access to sea)	7.92	U.S. bit., 13,300 Btu/lb	Best actual annual average price at New York power plant

TABLE 4-13. COSTS OF RAW MATERIALS AT USE-POINT (Concluded)

Plant Location	Cost $ Per Bbl	Type	Justification
		Oil	
Honolulu (access to sea)	2.25	155,200 Btu/gal[c]	Average actual cost at plant, no additional quantity discounts assumed.
Bangkok (access to sea)	2.15	Middle East crude or heavy fuel from Indonesia	Posted price f.o.b. port of origin and freight; quantity discount to offset duty and port fees
Johannesburg	2.25[d]	Middle East or Nigerian crude	Same as above
Beirut	1.74	No. 6 from Sidon, 19° API	From pipeline at Sidon at 65 cents plus 9 cents freight; assumes no additional duties
Sao Paulo	2.11[e]	Crude from Curacao or Venezuela, 15° API	Posted price plus freight to Santos plus pipe to Sao Paulo; quantity discount to offset nominal duty
Sydney (access to sea)	2.12	Crude from Middle East or Indonesia, 35° API	F.o.b. port side average price; quantity discount to offset handling charges; no duty in Australia
Tokyo (access to sea)	2.27	Heavy or medium fuel from Indonesia, 18° API	Posted price plus freight; quantity discount to offset duty and port fees
Los Angeles (access to pipelines)	1.88	152,500 Btu/gal	Best actual price on annual basis per MM Btu to power plant in Los Angeles County
Paris	2.39	Libyan crude, 37° API	F.o.b. Dunkerque or LeHavre at $2.39; quantity discount to offset duty and piping cost to Paris
New York (access to sea)	1.85	148,900 Btu/gal	Best actual price on annual basis at New York power plant

[a] No positive evidence of adequate supply

[b] Burning of coal probably not legal in Los Angeles County

[c] Bbl = 42 U. S. Gal.

[d] High duty to protect coal interests may cause increase

[e] High duty to protect government-owned oil industry may cause increase

(1) For the Honolulu, Beirut, and Johannesburg area, water power is virtually nonexistent.

(2) For New York, Los Angeles, Tokyo, Paris, and Sydney, the 1980 and 2000 grid systems will undoubtedly include some input from hydroelectric plants, but increased capacity for the proposed hydrogen plant would emanate from a local nuclear plant (or coal-fired plant for Sydney) rather than the more distant hydroelectric plant.

(3) For Sao Paulo, where hydroelectric power is now prevalent and ample means for expansion are available, projects such as the Furnas Dam on the Rio Grande River or the Tres Marias Dam on the Sao Francisco could readily be expanded to meet the increased requirements. The power cost summary for Brazil envisions such action, specifically from Rio Grande River plants having a potential capacity of 7000 mW, of which the 1100mW Furnas Plant is the first step.

(4) For Bangkok, hydroelectric power is also a predominant source of electricity. The Yanhee project about 430 km north of Bangkok and the Nam Pung project at several locations on the lower Mekong are already partially completed and supplying power to Thailand. The Yanhee project, ultimately a 560-mW plant, is already connected to Bangkok, and the Nam Pung and other Mekong River[52,53] stations are scheduled for incorporation in the Yanhee grid. The proposed 1500 mW Pa Mong Dam on the Thailand-Laotian border will add considerably to the availability of hydroelectric power by 1980.

Based on these developments and the lack of substantial quantities of domestic fuel, the Bangkok hydrogen plant power requirements and costs are based on hydropower from a grid fed by these projects.

4.4.2.5 Oxygen

The oxygen for the partial oxidation process is supplied by an integral air separation plant, but the oxygen for coal gasification is considered as a raw material. The cost of oxygen at Johannesburg and Sydney is considered to be about $5.00 and $4.50 per short ton, respectively. These costs are in line with current oxygen prices for guaranteed output of plants of comparable size using low cost electric power.

4.4.2.6 Refrigerants

Nitrogen gas for make up to the cascade refrigeration system can be supplied from an inert gas generator followed by a relatively crude purifier to remove high boiling impurities, mainly water. The cost of methane as a refrigerant is considered to be the same as the cost of natural gas at each location. If large quantities of natural gas are not employed in the process, the methane refrigerant cost may be higher, but the addition to the per pound hydrogen cost will be unnoticeable.

The propane and ethylene refrigerant cost used at all locations is presumed to be the same as bulk price at New York. Once again, substantial variations in this price at any given location will have no significant effect on the per pound hydrogen cost.[54]

4.4.3 AREA SUMMARIES

4.4.3.1 New York

A 2500 T/D liquid hydrogen plant will require raw materials, fuel, and power having an energy requirement equivalent to several major power plants in the area, and would represent a significant addition to energy consumption. The demand for gas and oil in the New York area will continue to be more highly competitive than the demand for coal. The use of coal and oil, particularly with a high sulfur content, will be subject to more and more limitations by air pollution controls. The highly competitive nature of the market, the need for total importation of all fossil fuels, and the potential restrictions on their burning close to the city provide a strong stimulus to the utmost use of nuclear power.[41,55 thru 58] Since fuel is not produced in the immediate environs of New York, substantial increases in importation are indicated. This has been the standard pattern in the past as the New York City industrial complex expanded. The following contain details of the present and projected status of trade in each commodity useful to liquid hydrogen production for which a logistic problem may exist.

4.4.3.1.1 Natural Gas – Natural gas is supplied currently to New York by pipelines from Texas. Present pipeline capacity is heavily taxed, requiring construction of liquefaction and storage facilities to allow for system peak shaving during high demand periods. Interest is also mounting in liquefied natural gas (LNG) transport by tanker from fields having excess gas which is wasted by venting and flaring, such as Venezuela. It would appear that natural gas supplies can meet New York requirements. However, more costly transport and storage facilities should be necessary. The 1964 price for natural gas by large-scale users (power stations using up to 18 billion cubic feet per year) ranged from 36.3 to 38.8 cents per 1000 cubic feet for gas averaging slightly below 1050 Btu/cu ft. The 1966 price reportedly averaged 38 cents for similar gas. It is interesting to note that in 1964, two power plants on the New Jersey side of New York Harbor, at Sewaren and Newark, used 4.3 billion and 1.9 billion cubic feet of gas averaging 1050 Btu/cu ft at 29.1 and 31.7 cents per cubic foot, respectively. Thus, plant location may have a significant effect on fuel costs.

4.4.3.1.2 Oil – Most oil is imported to New York from the Caribbean area as heavy fuel oil or crude. The 1964 price paid by New York power stations using up to 4 million barrels annually varied from $1.85/bbl to $2.28/bbl for oil with a heating value very close to 149,000 Btu/gal. The 1966 average is considered to be approximately $1.98/bbl for comparable oil. The world-wide price trend for oil may not change significantly in the near future. Expansion of relatively new fields should heighten competition and have a modest tendency to reduce price, but increases in labor and equipment costs will at least offset other savings.

4.4.3.1.3 Coal – In 1964 New York's power stations used more than 5,000,000 tons of coal, with the best price being $7.92/T at a plant consuming about 500,000 tons per year. Most coal used is bituminous, from West Virginia and western Pennsylvania, with a heating value of about 13,300 Btu/lb. The average 1964 price paid by power companies was about $8.25/T, whereas it has been reported as about $8.50/T in 1966. Again, it is interesting to note that in 1964 the Sewaren, New Jersey, plant on the opposite side of New York Harbor obtained almost 500,000 tons of comparable coal at $7.75/T.

4.4.3.2 Paris

A 2000 T/D liquid hydrogen plant in the vicinity of Paris would have a similar impact on the fossil fuel market as a 2500 T/D plant in New York. Present supply mechanisms would be inadequate for the most part, but adjustment to the new demand would fit the continuing economic growth pattern of fossil fuel consumption in that area. In view of the highly competitive market and present need for importation of fossil fuels, nuclear energy should be used as much as possible in the Paris basin.[59 thru 64]

4.4.3.2.1 Natural Gas – The recent natural gas supply to the Paris area (since 1959) has been almost exclusively from the Lacq fields in southwestern France. Since 1962 the supply has leveled at about 5 billion cubic meters per year, which is distributed over an area extending from Brittany to the Rhone and as far north as St. Quentin. This natural gas, even when added to the traditional supply of about 6 billion therms (1 French therm = 3968 Btu's) from coke ovens and more than 2 billion from refineries (1964 levels), has been inadequate to meet French industrial requirements. These demands have increased 340 percent from 1951 to 1964. Domestic requirements, constituting about one-third of the total supply, have risen also. Future requirements will be supplemented from at least three sources:

(1) Expansion of domestic fields (recent discoveries of higher quality gas have been made at Meillon) will supplement the Lacq supply.

(2) Imports by pipeline from the Netherlands (beginning in late 1967 or early 1968 under a 20-year contract) are expected to reach 5 billion cubic meters by 1975, with 2 billion marked for industrial use.

(3) Imports of LNG by tanker from North Africa, primarily from Algeria, where 1968 shipments should reach 3.5 billion cubic meters, and possibly from Libya.

The price of Lacq gas, which will decline in availability if present estimates of reserves are valid, runs close to 44 cents/1000 cu ft. It is predicted that the cost of Dutch gas in Paris will be about 55.5 cents/1000 cu ft. LNG prices (in effect or under discussion) range from about 41 cents for Libyan gas (specifically quoted at Barcelona) to almost 50 cents/1000 cu ft f.o.b. Bougie, or Azrew, Algeria. The latter gas can be supplied through a 204-km pipeline from LeHavre to Paris at a price competitive (and quite probably advantageous) with local and Dutch supply. The costs at Paris are cited generally in terms of average price to consumers. The price invery large quantities becomes somewhat speculative. A cost of 48 cents/1000 cu ft seems reasonable in view of future supplies of LNG and piped gas from promising North Sea discoveries.

4.4.3.2.2 Oil – Practically all oil used in France is imported. Earlier colonial ties have caused the majority of oil dealings to be associated with Algeria and other North African sources. The price of crude oil (Libyan - 37° API) at LeHavre is reported at \$2.39/bbl; the heavy (No. 2) fuel oil price at Paris averages \$3.14/bbl in barge lots and \$3.54/bbl in rail car lots, presumably based on \$2.75/bbl from refineries at LeHavre or Dunkerque. Lower transfer costs, as a result of using pipeline, may reduce the price to \$2.60. The contemplated pipeline would be similar to the 20-inch crude oil line which now runs from LeHavre to supply the 80,000 barrels per day (B/D) Grandpuits refinery, a few miles east of Paris.

4.4.3.2.3 Coal – Although France and its neighbors of the European Economic Connumity (Common Market) have substantial coal deposits, mining costs have made imports attractive. In 1965, coking coal from the U.S. (bituminous at 13,300 Btu/lb) was imported at an average cost of $13.20 per U.S. ton, with large quantity purchases on the order of $11.80 per U.S. ton, both prices c.i.f., Atlantic port before customs. Steam coal from the U.S. averaged $11.36 per U.S. ton, with quantity purchases reported at approximately $10.00 per U.S. ton c.i.f., Atlantic port. Prices of other imports were $14.70 per ton from the United Kingdom and $16.10 from Belgium. This compared with $18.80 for coking coal at mine mouth in the Saar Basin and $13.70 to $15.60 per U.S. ton for domestic steam coal at mine mouth. Based on $2.00 per U.S. ton for freight and customs, U.S. coal could be delivered to a Paris plant site for about $13.36 per ton. Given a modest improvement in mining costs, domestic coal should be attainable at about the same prices.

4.4.3.3 Los Angeles

Installation of a 1300 T/D liquid hydrogen plant near Los Angeles would influence the economics of fossil fuel supply similarly to the effects of the large plants in Paris and New York. It would manifest itself most strongly in the area of natural gas supply. Increasing restrictions on the use of oil and coal, coupled with the high demand for gas, suggest once again strong consideration of nuclear power for the Los Angeles LH_2 plant, if improved costs are to be realized.[41,55 thru 58]

4.4.3.3.1 Natural Gas – Air pollution regulations have favored the use of gas in Southern California in preference to liquid and solid fuels. Hence, demand for gas has caused the substantial supply from California fields to be supplemented by a pipeline from Texas. The requirements for hydrogen liquefaction would be imposed on a demand which in 1966 was about 800 billion cu ft. In 1964, the electric utilities in the Los Angeles basin consumed almost 80 billion cu ft for power generation, a quantity almost identical to the 1300 T/D hydrogen plant requirement if the stock and fuel were both natural gas. The best price in 1964 was 33.8 cents/1000 cu ft for gas averaging 1080 Btu/cu ft, with the cost ranging from 31.2 to 31.4 cents/million (MM) Btu. The average price reported in 1966 was 31.4 cents/MM Btu.

4.4.3.3.2 Oil – Although subject to restrictions, substantial quantities of oil are consumed in burning processes in Southern California. The utilities probably represent large-scale commercial buyers as well as anyone. In 1964, the best price paid by a power station in Los Angeles county for a Bunker C type crude oil of about 153,000 Btu/gal was $1.79/bbl, with most other locations ranging 10 cents/bbl higher. Stated as an energy cost, this ran 29.6 cents/MM Btu at the less favorable locations. Across the county line at Fontana, a station obtained 151,600 Btu/gal oil at $1.77/bbl, or 27.6 cents/MM Btu. The average price in Los Angeles county in 1966 has been reported as 30.7 cents/MM Btu.

4.4.3.3.3 Coal – Remoteness of supply and attendant transportation costs, even more than air pollution restrictions, have limited the use of coal in Los Angeles and vicinity. Coal of a 10,000 Btu/lb quality is used by power companies in Colorado at prices as low as $4.54/T. Using unit train concepts for delivery to California, the price at plant site would be on the order of $9.75 to $10.00/T.

4.4.3.4 Tokyo

As with New York, Paris, and Los Angeles, Tokyo is located in an area where fossil fuel energy sources are in short supply, but the level of economic activity has been to create continuing incentives to meet these demands. The requirements of a 500 T/D hydrogen plant, while draining some of the short supply of fuels, would be in line with the normal annual supply increases. In the case of oil, for example, Japan's supply, almost entirely imported, is increasing at a rate of about 10 million metric tons per year. The 500 T/D hydrogen plant yearly requirement on an all-oil basis would be less than 10 percent of this amount of annual increase. The short supply and high demand for fuels strongly suggest fullest application of nuclear power for LH_2 plant requirements.[53,60,65 thru 71]

4.4.3.4.1 Natural Gas – The inadequate local supply of natural gas in Japan, obtained partially from gas wells and partially from coal mines, has stimulated importation of LNG from outside sources. Plans to import LNG from Alaska at 52 cents/1000 cu ft (MCF) are nearing completion. Also, negotiations with Russia are in progress for the supply of LNG from Sakhalin Island. The price under consideration has been reported to be 38.6 cents/MCF. It is questionable at this time that sufficient quantities will be available at the lower price. However, it indicates a direction to be followed in LNG competition in the Far East. Siberian, Indonesian, and possibly Chinese sources could be tapped in response to Japan's requirements.

4.4.3.4.2 Oil – As stated, almost all of Japan's oil is imported, more than 80 percent of it from the middle East. The average price of oil imported from July 1965 to June 1966 was $2.51/bbl. Thus, assuming that quantity discounts would offset duties and port charges, the price of $2.27/bbl, at a waterside hydrogen plant seems likely. Heavy fuel (No. 6) would also be available from Indonesia at a comparable price; for example, at $1.90/bbl posted plus 38 cents/bbl freight.

4.4.3.4.3 Coal – Japan imported almost 14 million metric tons of coal in 1964, with the quantity continuing to increase. More than 40 percent has come from the U.S.; about one-third from Australia. On an all-coal basis, the Tokyo hydrogen plant would need almost 15 percent of Japan's present annual imports, or about 5 percent of its total production (about 50 million metric tons per year), which is declining slowly. The average price of all coking coal imported from June 1965 to July 1966 was $18.74 per U.S. ton on a wholesale basis. The average price, c.i.f., for all coal in 1965 was $15.82 per U.S. ton, with all coking coal at $15.73 per U.S. ton c.i.f. United States coking coal averaged $18.51/ton c.i.f. Although small quantities (475,000 metric tons or about 3 percent), were obtained from mainland China at $13.11/U.S. ton c.i.f. average, the pattern of coal importation is shifting toward the use of Australian coal, which in 1965 averaged $13.42 per U.S. ton c.i.f. To what extent local coal of lower quality, described as Kyuska coal dust, could be used is not known without further investigation, but its average wholesale price was $12.08/U.S. ton for the July 1965 to June 1966 period. Unless otherwise determined, the use of coal for a hydrogen plant should be premised on the Australian price of $13.42 per U.S. ton, on the assumption that quantity discounts will offset duties and port handling.

4.4.3.5 Sydney

The annual energy consumption of the countries previously considered (U.S., France, and Japan) is in excess of 150 million metric tons of equivalent coal. For these locations, the proposed hydrogen plant requirements, even if erected today, would add less than 5 percent to the annual consumption. A larger impact would be felt if natural gas were selected for both power and feed stock requirements. For the same situation to exist in Australia, one must assume that its rate of economic growth will allow the country to reach or exceed the current levels in Japan or France. This is a reasonable assumption overall, even though today the requirements of the proposed 400 T/D hydrogen plant, if expressed in terms of all gas, would exceed the total Australian gas consumption by almost 250 times. The excess of coal in Australia, both black and brown, suggests a coal gasification process. The use of coal for fuel and power also should be promising for LH_2 production near Sydney.[59.60,71 thur 79]

4.4.3.5.1 Natural Gas – Exploration and discovery of natural gas in Australia is in its infancy, so at this time there must be some doubt as to its ultimate avaialbility. Nevertheless, promising deposits have been tapped on both sides of Sydney, with Melbourne and Adelaide already receiving some supply. It has been predicted that the Gippsland Shelf field south of Sydney will yield gas at 29 cents/MCF in modest quantities. The opening of additional sources and larger pipelines should lessen this cost to some extent, bringing gas into a competitive posture with oil at about 35 cents/MCF.

4.4.3.5.2 Oil – Most of Australia's oil is imported from the Near East, although imports from Indonesia may be expected to rise. Local production is at present insignificant. The average price of all imported crude oil in 1965-1966 (year ending June 30) was $2.12/bbl f.o.b., port. Crude oil enters Australia duty free; hence, if quantity discounts offset port handling costs, the average f.o.b. price should be attainable at the hydrogen plant site.

4.4.3.5.3 Coal – Australia is one of the world's major coal producers and exporters; thus, the use of coal is favored to the fullest extent at the Sydney location. The price of coal in New South Wales is in the range of $2.24 to $5.75 per ton. The new mine-mouth power plant at Lidell, for example, expects its coal costs to be between $2.24 and $2.80 per U.S. ton from an open cut mine. Other prices at the mine have been cited in the $2.50 to $3.80/U.S. ton range, whereas the average price for New South Wales black coal at the colliery was $5.75/ton in 1964. Allowing for nominal handling and freight charges, New South Wales black coal should be available at a hydrogen plant near Sydney at $3.00/T. Brown coal is also mined in Australia, primarily in Victoria, and has already been used in gasification processes. This coal, somewhat cheaper than black coal at the mine, would also cost close to $3.00 per ton near Sydney because of longer transport distances, about 400 miles compared to 50 for black coal.

4.4.3.6 Sao Paulo

The annual consumption of energy in Brazil is substantially less than in Australia (about two-thirds) in spite of a much larger population. Moreover, Australia produces about 70 percent of the basic energy it consumes, whereas Brazil imports almost 60 percent. On the other hand, about 80 percent of Brazil's industrial activity is centered in the three states of Minas Gerais, Rio de Janeiro, and Sao Paulo, making the economic activity of the general area of Sao Paulo more comparable to Sydney than the overall country statistics would suggest. By today's standards, the installation of a 400 T/D liquid hydrogen plant in Sao Paulo would require a significant increase in the consumption of the basic energy source selected. Yet, the country's economic growth, particularly in the Sao Paulo area, suggests that the 1980 to 2000 economy could adjust to the raw material demands of a 400 T/D hydrogen plant without too much difficulty. Price data for fossil fuel raw materials in Brazil seems less reliable than that obtained for other locations, partially because of the government's changing policies, with respect to resources and the instability of its currency. Cost projections in the 1980 to 2000 period must be considered highly speculative.[59,60,80,81]

4.4.3.6.1 Natural Gas – Brazil's consumption of natural gas in 1964 was 500 million cubic meters as compared to hydrogen plant requirements on an all-gas basis of close to 700 million cubic meters. Hence, new sources and new transfer or transport systems must be obtained, if natural gas is to be a liquid hydrogen raw material in Sao Paulo. At this time, the most promising source appears to be in Santa Cruz Department, Bolivia, from which a 700-mile gas line to Sao Paulo has been discussed. No history exists on which to base a natural gas price at Sao Paulo. Comparison with cost of transmission at similar distances in the U.S., with consideration of the competitive oil situation, suggests a natural gas price at a Sao Paulo hydrogen plant to be on the order of 40 cents/MCF.

4.4.3.6.2 Oil – Brazilian oil fields are presently in the immediate vicinity of Bahia, more than 500 miles north of Sao Paulo. Two-thirds of the oil consumed in Brazil is imported, about 40 percent from Venezuela and about 40 percent from the Middle East. The posted price for Bascan crude oil is $1.57/bbl to which freight costs close to 50 cents/bbl must be added. Crude oil from Curacao (15°API) costs $2.11/bbl f.o.b. port at Santos without duty and handling charges. Quantity discounts cannot be expected to offset duties because the government-owned oil entity, Petrobras, controls all oil distribution in Brazil and tends to be highly protective of local production.

4.4.3.6.3 Coal – Brazil's coal has a high ash and sulfur content, reducing the volume of cleaned coal to about 60 percent of mined raw coal and requiring mixture with imported coal to produce suitable coke. Imports, almost entirely from the U.S., constituted nearly half of the coal consumed in Brazil in 1964. However, steam coal exists in excess, and the government is planning projects for its increased utilization. U.S. coal should cost about $14.00/T, based on an export price of $8.50/T and $5.50/T for freight from Middle Atlantic ports. The average value of all coal and coke imported in 1963 was very close to $15.00/U.S. ton. No price information has been observed for domestic Brazilian coal. But in spite of the fact that only 60 percent of the coal mined is usable, the price at Sao Paulo should be less than imported coal. This is because most production mines lie less than 300 miles to the south in the state of Santa Catarina. Extensive lignite deposits have been reported in the reaches of the upper Amazon valley, but no exploitation has occurred. Price data is non-existent, but would, in any event, be subject to high freight costs.

4.4.3.7 Beirut

Of all the locations considered, with respect to the impact of raw material requirements on existing or projected level of trade in the basic fuel commodities, the 300 T/D liquid hydrogen plant at Beirut will demand the most severe adjustment. The total energy consumption of Lebanon in 1964 was equal to 1.6 million metric tons of equivalent coal, whereas the energy requirement of a 300 T/D liquid hydrogen plant approaches 70 percent of that amount. Despite the rapid growth rate of Lebanon's energy market, the impact of a 300 T/D liquid hydrogen plant, even in the 1980-2000 period, would be significant. Therefore, facilities for the importation and local handling of fuels require consideration. Cost consideration needs to be given to specific on-site storage and handling facilities for bulk shipments specially received in Lebanon for the 300 T/D liquid hydrogen plant. The normal Lebanese market will probably not be geared to handle the quantities in question.[50,60,82,83]

4.4.3.7.1 Natural Gas – At present, natural gas is not used or produced in Lebanon. However, if a demand exists, its supply does not appear unrealistic. Gas is abundantly available in the Near East oil fields of Iraq, Arabia, Kuwait, and Iran, about 400 miles away. The well-head price could be extremely low, contingent on rayalties imposed by producing countries. Gas pipelines alongside existing oil lines to Tripoli or Sidon could be justified, particularly if Lebanon were used as a trans-shipping point to Europe as in the case of oil. Although attaching a price to a currently non-existent product is highly speculative, it is believed that natural gas from the Near East fields could be available at 35 cents/MCF, or even less if pipeline costs could be shared with a broad market. Another possibility is LNG from North Africa, particularly Libya, although every indication, based on propositions to European users, would suggest a price in excess of 40 cents/MCF for such gas to Beirut.

4.4.3.7.2 Oil – Several pipelines from Iraq and Saudi Arabia now terminate in Lebanon's Mediterranean port cities of Tripoli and Sidon, where transhipment to Europe occurs. The price of heavy fuel oil (No. 6 - 14° API) at Sidon has been reported at $1.65/bbl from the refinery there. Freight to Beirut would be 9 cents/bbl yielding a total price of $1.74/bbl at plant site, providing a Lebanese duty is not levied to an extent greater than can be offset by quantity discounts.

4.4.3.7.3 Coal – The production and use of coal in Lebanon is virtually non-existent. The most probable source of coal would be Poland or Russia, but, since no trade exists, the price for such coal is highly speculative. Fifteen dollars per ton is comparable to the price of Polish or Russian coal in Western Europe.

4.4.3.8 Johannesburg

The level and pattern of energy consumption in South Africa is very similar to that of Australia, except that coal is even more abundant and oil is used somewhat more sparingly. Since the proposed LH_2 plant size of 200 T/D is only half that of Sydney, the Johannesburg area should have no difficulty in supporting raw material supply for the operation. Government policy, availability, and price strongly favor the use of coal in South Africa.[59,60,84 thru 86]

4.4.3.8.1 Natural Gas – There is no trade in natural gas in South Africa, and in spite of the extensive coal and steel activity, the production of manufactured gas is relatively small. If natural gas were to be required for hydrogen production in South Africa, importation as LNG would need to be considered. North Africa and the Middle East would be the most probable sources, although gas from Argentina, Chile, the Caribbean, and Indonesia would also be a possibility. Comparing this requirement with Mediterranean and Japanese indices, it seems unlikely that LNG could be supplied to South Africa for less than 45 cents/MCF.

4.4.3.8.2 Oil – South Africa produces very little oil and imports its requirements. For the Johannesburg area, oil would normally be delivered through the port of Durban. From that point a 440-mile long pipeline is capable of carrying initially 31,500 bbl/D of Durban refinery products to Germiston, a suburb of Johannesburg within ten miles of the Jan Smuts international airport. The cost of this line has been reported as $28 million. If oil is to be used in the vicinity of Johannesburg, a similar line, with its cost shared by other users, needs to be considered. The price of Middle Eastern oil at Durban is $2.12/bbl, based on the posted price of $1.80/bbl plus 32 cents/bbl for freight. If quantity discounts are to offset pipeline charges and duties, special arrangements must be made, for it is customary for the coal and coal gasification projects of South Africa to be well protected. For instance, $3.90/bbl is a typical price for fuel oil (No. 6) delivered in Johannesburg. Current indications are that pipeline costs for transfers from Durban to Johannesburg will include a temporary fee to support the loss of rail revenues. Upon price stabilization, it would seem that oil could be piped to a large volume user at about $2.25/bbl, barring a high coal-mine products protective duty.

4.4.3.8.3 Coal – About 85 percent of the energy consumed in South Africa originates from its coal mines. The majority of the mines are within a 150-mile radius of Johannesburg, which to a large extent has fostered its pre-eminence as the industrial center of South Africa. Major coal fields and their approximate distance from Johannesburg are the Vereeniging, 40 miles; Springs-Heidelberg, 25 miles; Ermelo-Breyton, 120 miles; and Witbank-Middleburg, 75 miles. Thus, at the closer fields, a mine-mouth plant is feasible. In any event, coal transportation cost will be minimal. The price of coal in 1966 was reported fixed at $1.76/T at the colliery by the government. In 1965, the average price of bituminous at the mine was $1.78/T. Assuming use in or near the coal fields, as must be the case in the vicinity of Johannesburg, coal will be available for use at the hydrogen plant for about $1.90/T.

4.4.3.9 Bangkok

The total energy consumption of Thailand is about twice that of Lebanon, but, having a much larger population, the per capital consumption is only 15 percent that of Lebanon. Thus, of all countries considered for a liquid hydrogen plant location, Thailand has the least developed economy. Its accessibility to oil is not as good as Lebanon's, but the availability of hydroelectric power is an advantage. Since the Bangkok hydrogen plant, 200 T/D, will be smaller than the 300 T/D Beirut plant, the economic impact of the raw material requirements should be somewhat less than in Lebanon. Furthermore, the economic growth rate of Thailand appears to be more rapid. In addition to the cited cost of raw materials, the probability of costs for special port and handling facilities for the selected raw material needs to be considered. Even at the rapid economic growth rate of Thailand, there is some doubt that normal handling facilities could absorb the fuel and feed stock requirements for a 200 T/D hydrogen plant.[52,53,59,60,65,87 thru 90]

4.4.3.9.1 Natural Gas – No natural gas is produced or used in Thailand. If required, transport, most likely as LNG, would have to be originated from Indonesia. The economics of this supply, it is surmised, would be comparable to supply across the Mediterranean. A price of 45 cents/MCF is the best approximation that can be made at this time.

4.4.3.9.2 Oil – Practically all the energy consumed in Thailand is derived from imported oil. Despite the proximity to Indonesia, Middle Eastern oil is competitive in Thailand. With quantity discounts offsetting duties, oil should be available at a harbor-side hydrogen plant for $2.15/bbl, based on the $1.80/bbl posted price of Middle East crude plus 35 cents/bbl for freight. Heavy or medium fuel oil is posted at $1.90/bbl, f.o.b. ship in Sumatra, and since transport distance is shorter, this oil would be competitive.

4.4.3.9.3 Coal – A modest amount of lignite is mined in Thailand. Aside from domestic uses, it provides fuel for several small remote (southern peninsula) power plants. The requirements for substantial quantities would require importation, with Australia being the most likely source. Absence of historical data makes price estimation speculative. Since shipping distance is a little less than from Australia to Japan, a slightly better price than that paid in the Japanese market may be expected, probably no less than $12.50/U.S. ton.

4.4.3.10 Honolulu

The location of a 200 T/D liquid hydrogen plant in Honolulu presents a special problem of supply. The island of Oahu is isolated, relatively small, and therefore engages in a rather limited economic activity. Virtually all energy is imported at a level necessary to sustain about 500,000 people. Although the island's technology can be considered modern, a massive industrial complex would not readily be supported by existing or normally planned facilities. Because of long shipping distances for any raw material used, the production of liquid hydrogen on Oahu will be more expensive than at most other locations. There will, therefore, be greater incentive here for avoiding the use of fossil fuels.[56,58]

4.4.3.10.1 Natural Gas – No natural gas is used in Hawaii, but the index set for the supply of Alaskan gas to Japan may provide the pattern for a gas supply to Hawaii. It is possible, however, that broader use of Alaskan gas coupled with additional experience and competition may reduce the presently considered 52 cents/MCF price of the Alaskan-Japanese agreement, perhaps on the order of 45 cents/MCF.

4.4.3.10.2 Oil – Among the energy fuels, oil is the major Hawaiian import, and the price actually paid by a major Oahu power plant for 800,000 bbls in 1964 was $2.25/bbl at the plant site. The origin of this oil was not cited, but its heating content was listed as 155,168 Btu/gal.

4.4.3.10.3 Coal – No significant quantity of coal is consumed in Oahu, and it is hard to conceive of its use as being economically feasible. If large quantities were brought to the island, the U.S., Canada, and Australia are potential sources. The price would be somewhat less than the delivered cost of U.S. and Canadian coal to Japan and somewhat more than Australian coal to Japan, about $15.00/T.

4.5 PROJECTED POWER AVAILABILITY

4.5.1 AVAILABILITY FACTORS

To provide maximum reliability of electric power supply at minimum cost, the plant providing power to each liquid hydrogen facility should be part of an existing power network. Part of the problem of determining the cost of power for each location, therefore, hinges on fitting the hydrogen plant requirements into the regional power network requirements in the 1980 to 2000 period. To provide background for power plant analysis, the following paragraphs define the environs of each plant location which may be served by a common power network and projects the 1980-2000 period demand for that network from existing data.[91 thru 94]

4.5.2 UNITED STATES

The Federal Power Commission (FPC)[92] divides the United States into 48 power supply areas. Area 4 consists of New York City and some of its suburbs. Los Angeles falls into area 47, which includes most of Southern California and central Nevada. The FPC[91] predicts 1980 energy requirements for Area 4 as 85.1 billion kwh and 129.5 billion kwh for area 47. For areas 1 through 4, which are already interconnected, 271.2 billion kwh is the projected requirement. (This does not include southern Ontario, which is presently part of the grid in which New York City is located.) The Pacific Southwest Interconnected System Group encompasses FPC supply Areas 46, 47, and 48, for which the projected 1980 requirements are 296.9 billion kwh. Using 1964 rates, production in the United States runs a little above 50 percent of the total attainable from the installed capacity. If this trend were to continue, the 271.2 billion kwh to be supplied in the interconnected New York area in 1980 would be generated from an installed capacity on the order of 62,000 mW, of which about one-third would be in the immediate New York City power supply area. It is possible that the maximum liquid hydrogen plant requirement could be incorporated into the local system by the addition of large-sized plants without unduly stressing back-up capabilities in the event of an outage. As we progress toward the year 2000, and local power system requirements grow, the hydrogen plant portion of the load becomes relatively smaller. Since installed capacity in the United States has been doubling approximately every ten years, the year 2000 should find an installed capacity well in excess of 120,000 mW in the New York City grid.

In the case of Los Angeles, where the anticipated interconnect system size for 1980 is slightly larger than for the New York area, and the hydrogen plant requirement somewhat smaller, a similar conclusion can be drawn. For example, the addition of large plants to the existing system will not violate accepted practices for the supply of back-up power in the event of a plant outage.

Hawaii must be considered in another manner. The FPC projection [92] does not include Hawaii, and statewide statistics from other sources, which are somewhat sparse, do not pinpoint the island of Oahu. Since interconnects with the other islands do not appear feasible, the power area for the Honolulu plant of this study must be considered to be Oahu. The installed capacity of Oahu is 602 mW, of which 208 mW was installed in 1920, 231 mW in 1938, and 163 mW in 1963-1964. Projection of this growth rate (i.e., tripling in approximately 40 years) would result in a predicted capacity of 1800 mW by the year 2000, whereas projecting the curve (through the three data points) yields a value on the order of 750 mW for 1980. A separate prediction [95] based on contacts with Hawaiian power officials, forecasts an installed capacity of 1000 mW for 1975. It can be assumed with reasonable certainty, therefore, that the installed capacity of Oahu will be on the order of 1000 mW in 1980 and in excess of 1800 mW in the year 2000.

4.5.3 OTHER COUNTRIES

Appendix B of this report shows two projections in figures B-1 through B-8 of installed capacity for the various countries under study. Actual capacities for the period 1948 to 1964 have been plotted from data in the 1964 U.S. Statistical Yearbook.[92] Assuming that installed electrical power capacity for each country will continue to grow at a constant annual rate equal to the 1955-1964 average rate, the upper or straight line projection will apply. If it is assumed that installed capacity will be increased only by a constant annual increment, the lower (drooping) curve will apply. However, generalized analysis of historical data shows that as a country approaches economic maturity, from a supply of power standpoint, the annual rate of growth decreases. The upper curve in Appendix B figures may tend to present an exaggerated growth rate. Conversely, the lower curves present an unrealistically low picture, because as a country's power base expands the annual incremental growth tends to increase. The actual growth curve (which is dependent on population growth, discovery of resources, industrialization, improvement of standards of living, and other socio-economic factors) lies somewhere between the two. The median value on the logarithmic ordinate has been selected as the most reasonable projection in determining the probable area grid size into which hydrogen liquefaction power requirements would fall in the 1980-2000 period.

4.5.4 JAPAN

Power statistics available in United Nations data do not treat the island of Honshu or the Tokyo area separately. Since a very large percentage of the industrial and domestic market for power is concentrated on Honshu in the Tokyo-Yokohama and Kobe-Osaka areas, it can be predicted that about 75 percent of Japan's power in the 1980-2000 period will be available to these two areas, at least half of it connected to Tokyo supply systems. Projecting Japan's rate of growth of installed capacity from 1955 to 1964 (11.4 percent per year) yields a 1980 availability of 210,000 mW and a 2000 capacity of 1,800,000 mW. Using the far more conservative straight-line extrapolation (i.e., continuation of the average growth for the past five years at equal annual increments), Japan's power availability would be 93,000 mW in 1980 and 162,000 mW in 2000. If only 30 percent of this power is available in the Tokyo area, the Tokyo capacity would be from 31,000 mW to 70,000 mW in 1980 and 54,000 mW to 600,000 mW in 2000. The median values for Tokyo in 1980 and 2000 are 39,000 mW and 168,000 mW respectively.

4.5.5 PARIS

If France's power-producing capacity continues to grow at the average rate of 5.7 percent per year as it has from 1955 to 1964, its installed capacity will be about 65,000 mW in 1980 (about 195,000 mW in the year 2000). By straight-line extrapolation of the 1959-1964 period in equal annual increments, capacity in 1980 will be 47,500 mW (75,000 mW in 2000). At least 40 percent of this power should be available to Paris and its environs. Thus, the level should fall in the range of 11,000 mW to 26,000 mW in 1980 (30,000 to 78,000 mW in 2000). The median value for Paris in 1980 should be 20,800 mW and 48,000 mW in 2000.

4.5.6 SYDNEY

The installed capacity of the Australian power industry grew from 1955 to 1964 to 8.2 percent per year. At this rate, 20,000 mW should be available by 1980 (136,000 mW by the year 2000). By straight line extrapolation, there should be 15,700 mW in 1980 (25,500 mW in 2000). In 1966, the Electricity Commission of New South Wales, which supplies Sydney, had a capacity of 2000 mW, plus a share of the Snowy Mountain production and an interconnect with Victoria and the Melbourne area. Moreover, the growth rate from 1965 to 1966 exceeded 10 percent, and commitments have been made to install another 3400 mW of producing capacity before 1975. The largest incremental plant size currently committed is 500 mW, of which four are scheduled for completion prior to 1975.[96] It seems safe to assume, therefore, that about 40 percent of Australia's power can be delivered to the Sydney area in the 1980 to 2000 period. Capacity between 6400 mW and 12,000 mW will be attained by 1980 (between 10,200 mW and 54,400 mW by 2000). The median value for Sydney in 1980 is 8400 mW and 24,000 mW in 2000.

4.5.7 SAO PAULO

At Brazil's current rate of growth (averaging 9.6 percent per year from 1955 to 1964), the country's installed capacity will be 31,000 mW in 1980 and 194,000 mW in 2000. Using the straight line extrapolation, it is forecasted at 16,200 mW in 1980 and 27,500 mW in 2000. According to the Worldmark Cyclopedia of Nations, 80 percent of Brazil's power is utilized in the neighboring states of Sao Paulo, Rio de Janeiro, and Minas Gerais. Continuation of current massive federal power programs suggests strongly that the capability of these three states will be largely interconnected in the 1980 to 2000 period. Power available at Sao Paulo, therefore, should be on the order of 13,000 mW to 24,000 mW in 1980, and 22,000 mW to 155,000 mW in 2000. Using the median principle, installed power in the Sao Paulo grid in 1980 should be 17,600 mW and in 2000 should reach 56,000 mW.

4.5.8 BEIRUT

For all practical purposes, the Beirut power supply area can be considered contiguous with the boundaries of the country of Lebanon. United Nations data from 1955 to 1964 shows an average annual increase in electric power capacity of 17.5 percent, but none from 1962 to 1964. Nevertheless, the annual production increased 13.2 percent, 10.9 percent, and 10.4 percent for the calendar years 1963, 1964, and 1965, respectively. The average annual increase in production from 1955 to 1964 was 14.8 percent. Hence, despite the irregular nature of the increase of installed capacity in Lebanon, a projection based on 17.5 percent per year seems reasonably valid. At this rate, the 1980 capacity would be on the order of 3600 mW and the year 2000 capacity would be 95,000 mW. Using the five-year average straight-line extrapolation, 1980's capacity wold be 700 mW and 1220 mW would be available in 2000. The median method indicates 1600 mW in 1980 and 10,400 mW in 2000.

4.5.9 JOHANNESBURG

The average rate of growth of installed electric power production capacity in South Africa from 1955 to 1964 has been 5.8 percent, but in the latter half of that period increased to 8.2 percent. Using the more conservative number, a projection at constant annual rate of growth shows a 1980 installed capacity of 16,000 mW and a year 2000 capacity of 49,000 mW. The straight-line projection yields 9200 mW in 1980 and 12,600 mW in 2000. Only about one-third of South Africa's industry is concentrated in the Johannesburg area, but 60 percent of the present electric power capacity is available there. Installed capacity to accomodate Johannesburg hydrogen plant power needs would, therefore, be on the order of 5500 mW to 9600 mW in 1980 and 7600 mW to 28,400 mW in 2000. These figures could be even higher if the more recent 8.2 percent growth rate persists. The median value is 7200 mW in 1980 and 15,000 mW in 2000.

4.5.10 BANGKOK

Thailand's capacity to produce electricity increased at the very rapid rate of 19.5 percent during the 1955-1964 period. Continuation of this rate of growth to 1980 would yield an installed capacity of 8800 mW, and in 2000 about 357,000 mW would be available. Straight-line extrapolation, using the average incremental growth for the 1959-64 period, projects a level of 1700 mW in 1980 and 2740 mW in 2000. Since the Bangkok-Thonburi metropolitan area consumes 75 percent of Thailand's available power, the 1980 availability in Bangkok will range from 1280 mW to 6600 mW, and in 2000 will be between 2100 mW and 270,000 mW. The median value for 1980 is 2800 mW (24,000 mW in 2000).

4.5.11 SUMMARY

Projected electric power availability in 1980 and 2000 for each of the suggested plant locations is shown in table 4-14. Group A in the table is obtained by applying the ratio of power available or planned to be available at the city plant location to the logarithmic median of the projected country levels of Appendix B figures. Group B utilizes the FPC projections.

4.6 METHODS OF POWER GENERATION

Various methods of electric power generation will be available by the year 1980 to supply the projected requirements shown in table 4-1. A brief description of each method is contained in the following paragraphs.

4.6.1 HYDROELECTRIC

This method is used widely today in locations conducive to construction of dams for both water control and energy production purposes. Hydro plants utilize the potential energy of stored water to move turbines which, in turn, convert mechanical energy into electricity by driving generators. No fuel cost is incurred with this method of generation. However, the large capital investment normally required partially offsets the substantial fuel savings. Moreover, the effects of drought, which may occur for extended periods, could result in expensive back-up power such as pumped storage systems or use of relatively inefficient gas turbine and/or steam turbine systems.

4.6.2 PUMPED STORAGE

Similar in principle to hydroelectric generating plants, this method stores available off-peak power (as water power) to meet peak loads in excess of normal demands. This is accomplished by installation of reversible pump turbines which transfer water to a reservoir during low power demand. During peak demands, the stored water is released to generate power. Pumped storage is not necessarily a part of a hydro project and can be constructed as a separate system. The principal use of this method is for peaking service or reserve backup. Therefore, it must be weighed carefully against the cost of reserve power from other sources. A rule of thumb for this method is that 1.4 kwh of off-peak power yields 1 kwh of energy during peak periods.

TABLE 4-14. PROJECTED ELECTRIC POWER AVAILABILITY, 1980 AND 2000

Country	Medians 1980 (mW)	Medians 2000 (mW)	Central City	Applicable %	Projected Grid Capacity 1980 (mW)	Projected Grid Capacity 2000 (mW)
		Group A (From U.N. and FPC Data)				
Thailand	3,800	32,000	Bangkok	75	2,800	24,000
South Africa	12,000	25,000	Johannesburg	60	7,200	15,000
Lebanon	1,600	10,400	Beirut	100	1,600	10,400
Brazil	22,000	70,000	Sao Paulo	80	17,600	56,000
Australia	21,000	60,000	Sydney	40	8,400	24,000
Japan	130,000	560,000	Tokyo	30	39,000	168,000
France	52,000	120,000	Paris	40	20,800	48,000

Group B (Primarily from FPC Data)

City	Grid Size	Projected Capacity 1980	Projected Capacity 2000
Honolulu	Oahu	1,000	1,800
Los Angeles	FPC Areas 46, 47, 48	60,000	120,000
New York	FPC Areas 1, 2, 3, 4	60,000	120,000

4.6.3 CONVENTIONAL GAS, OIL, AND COAL

These methods presently form the backbone of world power generation. Each system consists of a central station fossil-fuel-fired steam boiler. The boiler generates steam for use in steam turbines. These turbines, in turn, drive the generators which produce the required electricity. Each of these systems has been consistently improved over the years, resulting in lower costs for power output. Improvements have included automation features, higher operating pressures and temperatures, and better materials for use in supercritical boilers. Temperatures above 1050°F are not presently economically attainable and are one of the principal obstacles to higher efficiencies. Thermal efficiencies of 40 percent have been attained with well-designed supercritical boilers.[58]

4.6.4 CONVENTIONAL NUCLEAR POWER

Nuclear plants operate in a manner similar to the above, although at lower steam temperatures and pressures. The uranium-fueled reactor replaces the central station boiler in the neclear plants. An example of steam conditions for an 1100 mWe reactor is 965 psig and 540°F.[97] Overall nuclear thermal efficiency is about 32 percent. Nuclear plants have the advantage of incorporating reactor design improvements at any time. Also, new advances in fuel enrichment and fabrication techniques will enhance their competitive position. Three types of conventional nuclear reactors are in use today:

(1) Light water reactors (LWR): Boiling water reactors (BWR) and pressurized water reactors (PWR)

(2) Advanced gas cooled reactors (AGR)

(3) Heavy water reactors: Natural uranium-fueled reactors moderated and cooled by heavy water (CANDU)

Of the three types mentioned above, the LWR have the lowest initial investment and are the most feasible where fixed charges on plant investment are high. Reactors in the United States are normally in this group. The AGR and CANDU reactors have higher initial investments. However, CANDU does not require costly fuel enrichment due to the use of natural uranium. Also, heavy water is one of the best neutron moderators and results in more energy from a given quantity of fuel for this type of reactor.

4.6.5 ADVANCED NUCLEAR POWER

Three broad categories of future reactors in various stages of development include those described in paragraphs 4.6.5.1 through 4.6.5.3.

4.6.5.1 Near-Breeder Reactors

High-temperature gas cooled reactors (HTGCR), advanced cooled dragon reactors, heavy water moderated organic cooled reactors (HWBLW), and steam-generating heavy water reactors (SGHWR) are examples of the convertor reactors.

Conversion ratios for these reactors are 0.8 to less than 1.0. The fuel cycle for the convertor reactor can be based on either uranium or thorium, but both fuel cycles will require U-235 for about 8 to 10 years to initiate the reaction. Thereafter these reactors will produce more fuel (which must be recovered by process) than they consume. The demand for uranium ore reserve will be less for the convertor reactor than for the LWR. The need for this type, as well as for the LWR, will be reduced when breeder reactors are fully developed and established as reliable economic power producers.

4.6.5.2 Low-Gain Breeder Reactors

These plants are a part of the fast-breeder family. Conversion ratio (or breeding ratio) is 1.1 to 1.25, with specific power of about 0.25 mW/kg of fissionable fuel. Doubling time is 10 years or more. These reactors could be in commercial operation by 1975.

4.6.5.3 High-Gain Breeder Reactors

These plants constitute that part of the fast-breeder reactor family (FBR) with breeding ratios of 1.4 and a specific power of 0.33 mW/kg of fissionable fuel. Doubling time is about 7 to 10 years. If placed in commercial operation by the year 1980, these reactors can be complementary to the water reactors by the year 2000. In other words, these fast breeders will be able to produce almost all of the enrichment fuel required by the water reactors by that time.[98]

4.6.6 COAL OR OIL GASIFICATION

A description of the process and developments necessary to utilize this method of generating power with oil or coal fuels has been proposed. The proposed power cycle consists of a gasifier plant which takes residual oil or coal and converts these raw fuels to clean fuel gas. Removal of hydrogen sulfide (H_2S) occurs in the gasifier, thereby eliminating harmful air pollution side effects normally associated with fossil-fueled plants.

The clean fuel gas is then used with steam and oxygen (or air) as feed to gas turbine combustors. Perfection of system design results in an overall efficiency approaching 46 percent, as compared to a central station boiler plant efficiency of 40 percent. Turbines at 1500°F are operational today, and 2000°F units are believed attainable prior to 1980. The development of gasifier apparatus to provide clean, dust-free combustible fuel will be the major obstacle. Preliminary economic studies show this method to have substantial merit.

Additional systems and concepts obviously will emerge for evaluation prior to the year 1980. However, 4 to 6 years are normally required for planning, procurement, and construction of large power plants. Therefore, the systems noted will most likely be the facilities in commercial operation by 1980. The following systems will continue to undergo evaluation and testing: Magnetohydrodynamics (MHD), electrogasdynamics (EGD), thermoelectric, thermionics, fuel cells, solar energy, nuclear fusion, nuclear molten salt reactors, and seed and blanket reactors. These latter systems are not expected to be in commercial operation by 1980. However, the emergence of some of these will be seen by 1985 or shortly thereafter.

4.7 SELECTION OF POWER PLANT SIZE

To determine the optimum plant size, a thorough analysis of grid loads, usage patterns, transmission capabilities, and system flexibility must be completed for each site. In the United States, plant size is normally limited by the premise that a system must have enough reserve capacity to supply loads when the largest generating unit is out of service. This is about 15 to 20 percent of the total installed capacity. For example, a system with a peak load of 1200 mWe would, with a 15 percent reserve, have an installed capacity of 1380 mWe. The largest single unit allowed would then be 180 mWe. It should be noted that this premise is based on firm power requirements from the plant.

To improve reliability and obtain low-cost reserve power, several utilities have constructed transmission line inter-ties or grids with neighboring systems. This resulted in the formation of reserve power pools. In the event a member system suffers greater than 15 percent outages, grid reserves are available to ensure continuity of service.

Based on this grid concept and future progress in grid and transmission design, it is most probable that by 1980 construction of plants will be based on grid instead of local system analysis. When justified economically, all the electric systems in an entire nation or area may be joined in a single interconnected network. Construction of generating, transmission, and distribution systems as part of a total network is envisioned. This will eliminate costly and wasteful duplications and take advantage of full utilization of the power capability of all plant sites. Also, the capabilities of the integrated network to take advantage of diversity in load patterns over large areas by effective dispatching will result in higher total network efficiencies. This will result in lower required reserve generating capacity.[92] Therefore, larger size plants approaching 1100 to 3000 mW should be attainable in 1980 systems. As noted in paragraphs 4.5 through 4.5.11, these large size plants will be a part of large fully integrated networks. Determination of the optimum plant size which can be constructed for each site has therefore been based on using the grid concept for each general area. The largest single 1980 fossil and nuclear plants are estimated to be 1100 mW and 3000 mW respectively. Also, the largest size unit that can be installed is assumed to be no larger than 15 to 20 percent of total installed 1980 grid capacity. A summary of plant sizes is shown in table 4-15, using projected grid capacities from table 4-14.

4.8 DETERMINATION OF PLANT INVESTMENT

Plant investment is dependent on plant size, design, and location. Design is influenced by plant location, utilization, efficiency, and fuel cost. A plant designed for high efficiency will require operating conditions which increase the cost of machinery and associated equipment. The resultant savings in fuel cost, therefore, must be evaluated against higher capital costs to determine optimum design characteristics. Plant construction, labor, and material costs will be dependent upon location of the plant. The expected plant utilization factor will be dependent on the function of the plant for either base load or peaking service and will also have a direct effect on plant design.

Plant investment in dollars per kilowatt of installed capacity versus size is shown in figure 4-1. The design criteria for the fossil fuel plants was based on supercritical boilers with reheat cycles using comprehensive investment data developed in a recent study.[99] Nuclear plant investments, shown in figure 4-2, are from various sources, with the base curve derived from the study data. The nuclear plant curves are considered representative of 1980 investment costs.[100] Hydro plant investments shown in figure 4-3 are plotted as actual installed costs.[101] All curves have been based on ideal construction conditions, using the Washington, D.C., area as the construction base. Area adjustment

TABLE 4-15. PROJECTED GRID CAPACITIES AND PROPOSED PLANT DESCRIPTIONS

Location City	mW Grid Capacity		LH_2 Load mW	Process Type	Power Plant Size, mW	Description Type	Assumed Plant Location
	1980	2000					
Bangkok	2,800	24,000	97	Steam reforming	600	Hydro	350 miles north at Pa Mong Dam
			113	Partial oxidation	600	Hydro	
			530	Electrolysis	1500	Hydro	
Honolulu	1,000	1,800	97	Steam reforming	150	Oil	Power plant at the air terminal
			113	Partial oxidation	150	Oil	
			530	Electrolysis	600	FBR	
Johannesburg	7,200	15,000	87	Coal gasification	1100	Coal	50 to 100 miles from air terminal
			97	Steam reforming	1100	Coal	
			113	Partial oxidation	1100	Coal	
			530	Electrolysis	1100	Coal	
Beirut	1,600	10,400	145	Steam reforming	300	Oil	Power plant at the air terminal
			171	Partial oxidation	300	Oil	
			990	Electrolysis	600	Oil	
Sao Paulo	17,600	56,000	193	Steam reforming	1100	Hydro	175 miles to the Furnase Dam at the site
			226	Partial oxidation	1100	Hydro	
			1050	Electrolysis	2000	FBR	
Sydney	8,400	24,000	193	Steam reforming	1100	Coal	50 to 100 miles from terminal at the site
			226	Partial oxidation	1100	Coal	
			1050	Electrolysis	2000	FBR	
Tokyo	39,000	168,000	232	Steam reforming	1100	FBR	Plants at the air terminal
			272	Partial oxidation	1100	FBR	
			1300	Electrolysis	2000	FBR	
Los Angeles	60,000	120,000	625	Steam reforming	2000	FBR	Plants at the air terminal
			735	Partial oxidation	2000	FBR	
			3400	Electrolysis	3000	FBR	
Paris	20,800	48,000	965	Steam reforming	2000	FBR	Plants at the air terminal
			1135	Partial oxidation	2000	FBR	
			5220	Electrolysis	3000	FBR	
New York	60,000	120,000	1220	Steam reforming	2000	FBR	Plants at the air terminal
			1430	Partial oxidation	2000	FBR	
			6600	Electrolysis	3000	FBR	

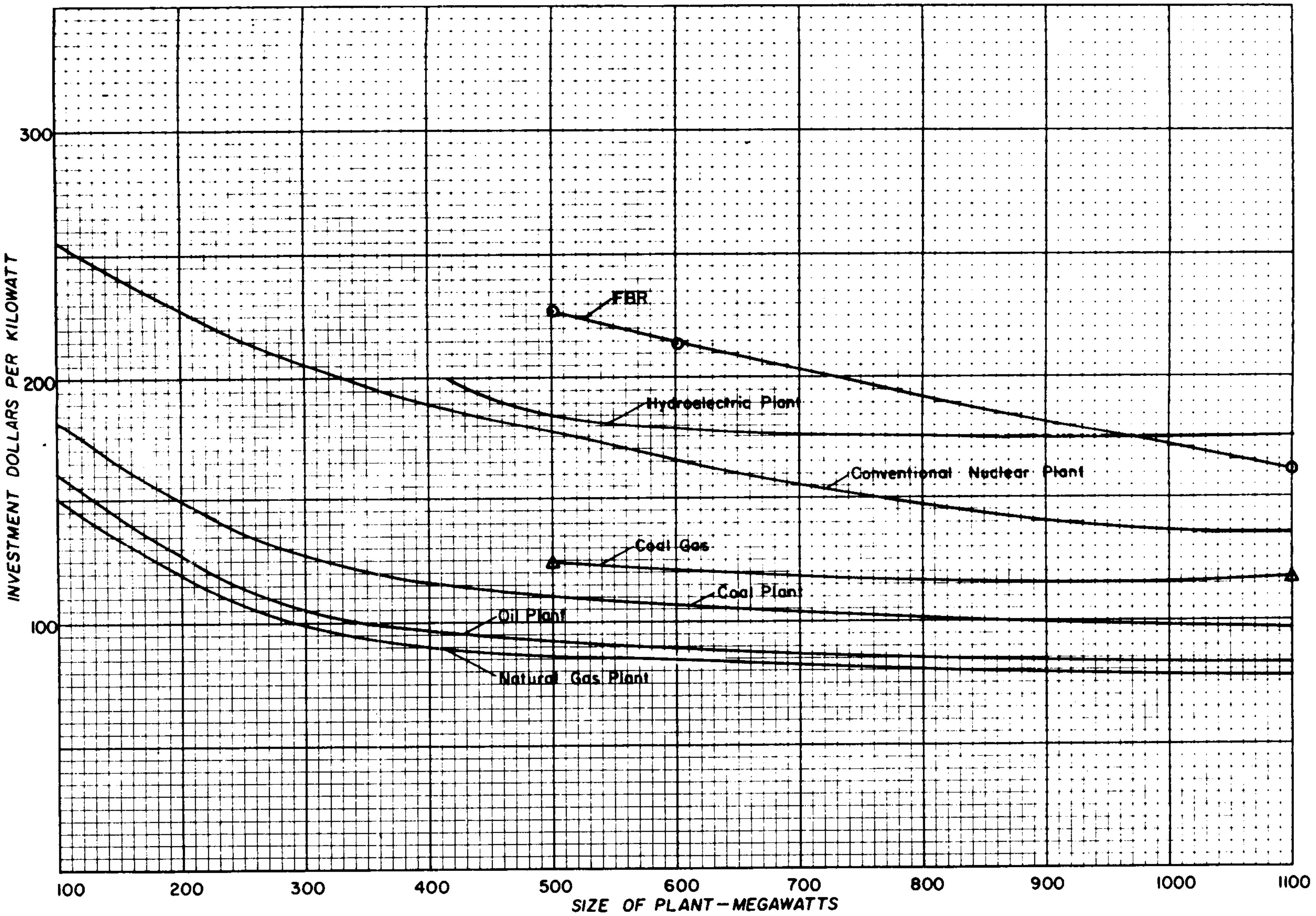

FIGURE 4-1. PLANT INVESTMENT vs PLANT SIZE

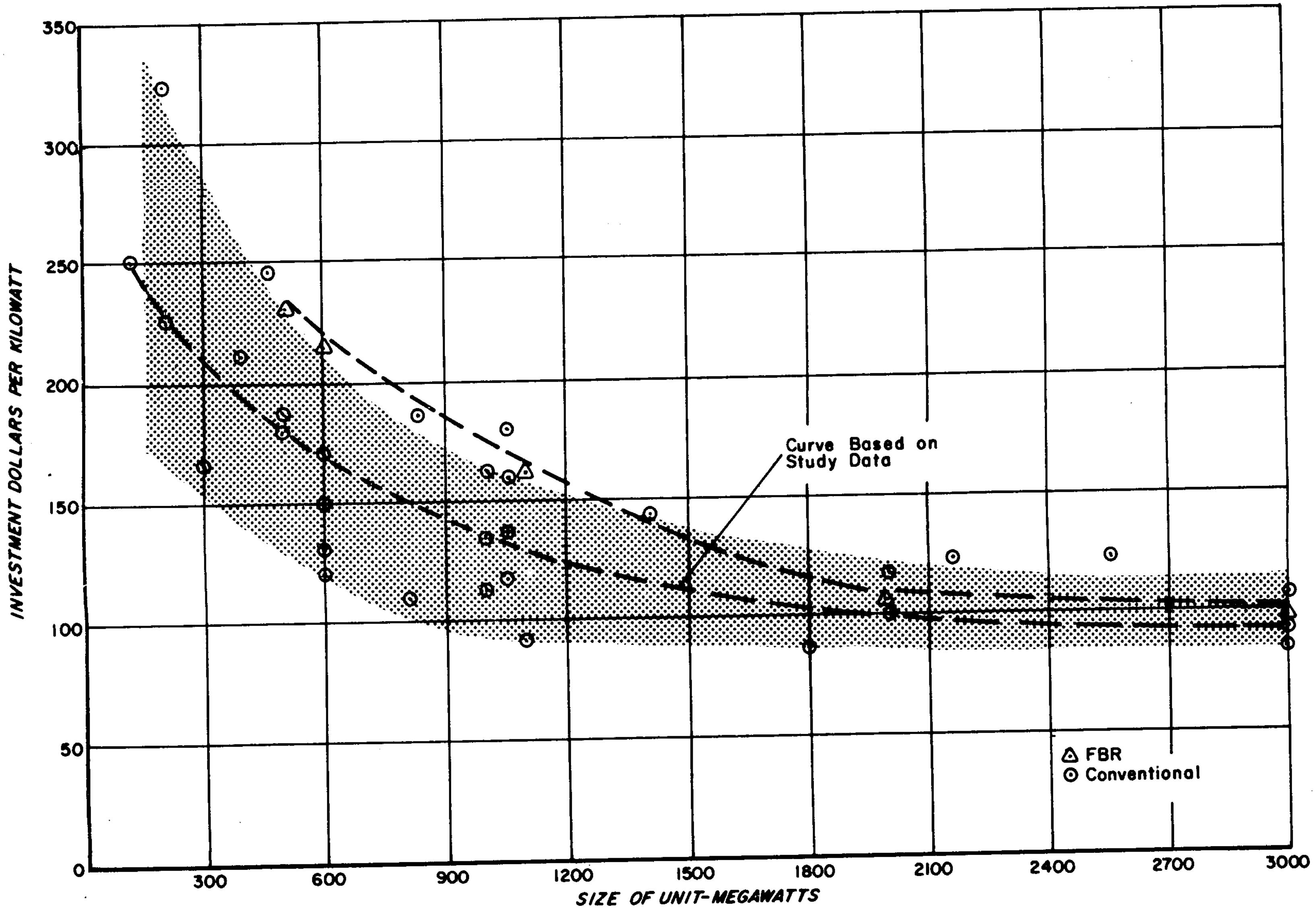

FIGURE 4-2. PLANT INVESTMENT vs PLANT SIZE (NUCLEAR)

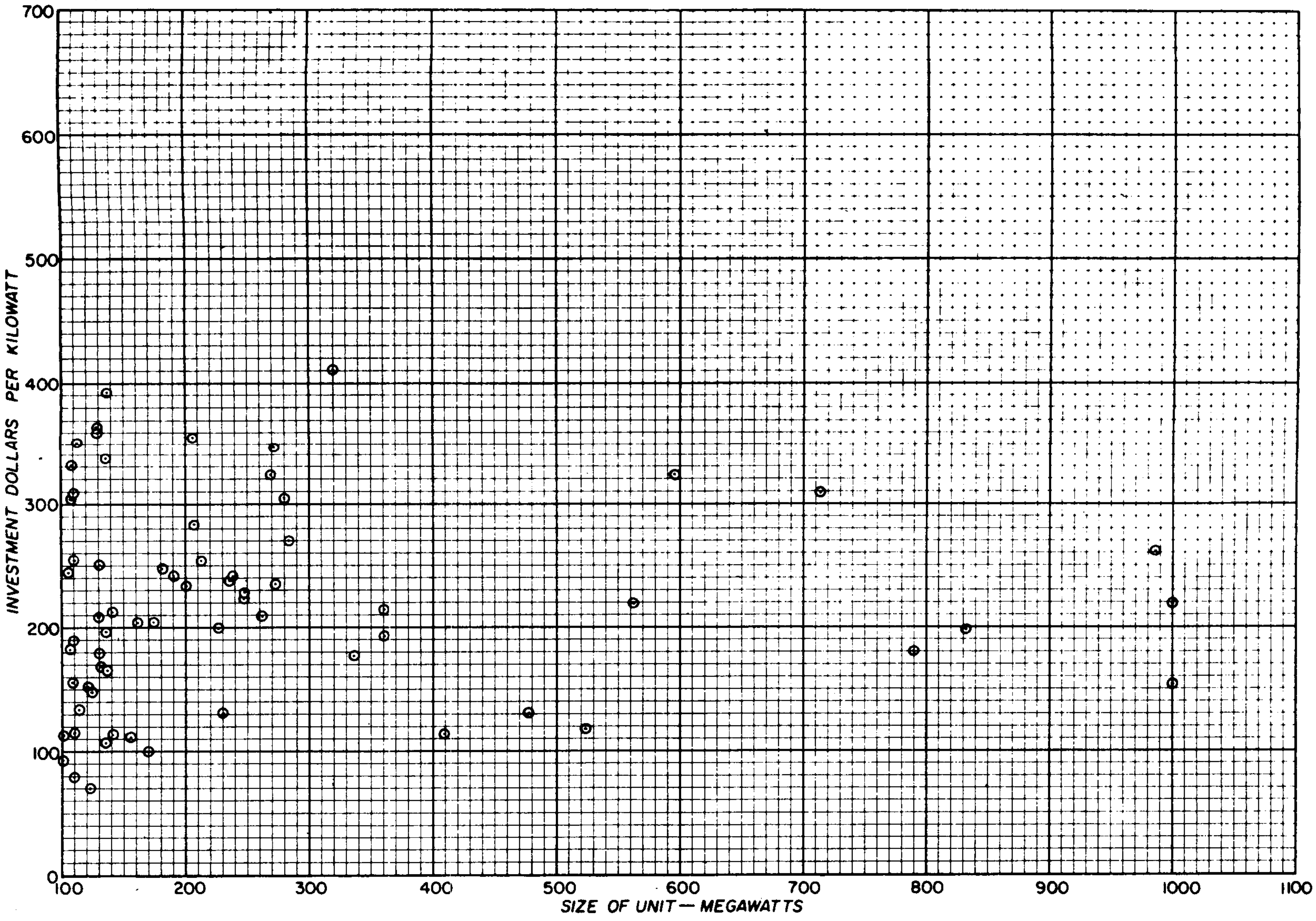

FIGURE 4-3. PLANT INVESTMENT vs PLANT SIZE (HYDROELECTRIC)

factors for each specific location must be applied to these investment cost curves. These factors have been compiled for various areas by the Department of Defense from actual bid experience and have been adjusted relative to a base index of 1.0 for Washington, D.C.

The following formula has been used in the determination of power plant investment at each site.

$$I = (S\ (i \times 1.15) \times P_1 \times A_1) + (S\ (i \times 1.15) \times P_2 \times A_2) = S \times AF\ (i \times 1.15)$$

where I = Power plant investment in dollars
S = Proposed power plant size in kilowatts
i = Base investment cost \$/kw for the power plant (figure 4-1)
1.15 = Base investment of 100% plus reserve investment of 15%
P_1 = Fraction of base investment using local labor, materials, etc.
A_1 = Area adjustment factor for local costs
P_2 = Fraction of base investment using base labor, materials, etc.
A_2 = Area adjustment factor for base (Washington, D.C.) costs
AF = Area adjustment factor*

*No capital investment reserve contingency is incorporated in this analyses. Power plant investment analyses normally contain this contingency, but this is not the case for the hydrogen facility evaluation.

For hydrogen facilities use:

$$I_{location} = I \times AF$$

where $I_{location}$ = capital investment at the use point

I = capital investment based on Section 3 data (Washington, D.C., base)

Factors applicable to each location are listed in table 4-16.

TABLE 4-16. LOCATION FACTORS

Location	A_1	A_2	P_1	P_2	AF
Bangkok	1.5	1.0	25	75	1.125
Bangkok (hydro)	1.5	1.0	50	50	1.250
Johannesburg	1.0	1.0	25	75	1.000
Honolulu	1.3	1.0	25	75	1.075
Beirut	1.2	1.0	25	75	1.050
Sao Paulo	1.5	1.0	50	50	1.250
Sydney	1.1	1.0	50	50	1.050
Tokyo	0.8	1.0	50	50	0.900
Los Angeles	1.1	1.0	100	0	1.100
Paris	1.1	1.0	50	50	1.050
New York City	1.3	1.0	100	0	1.300

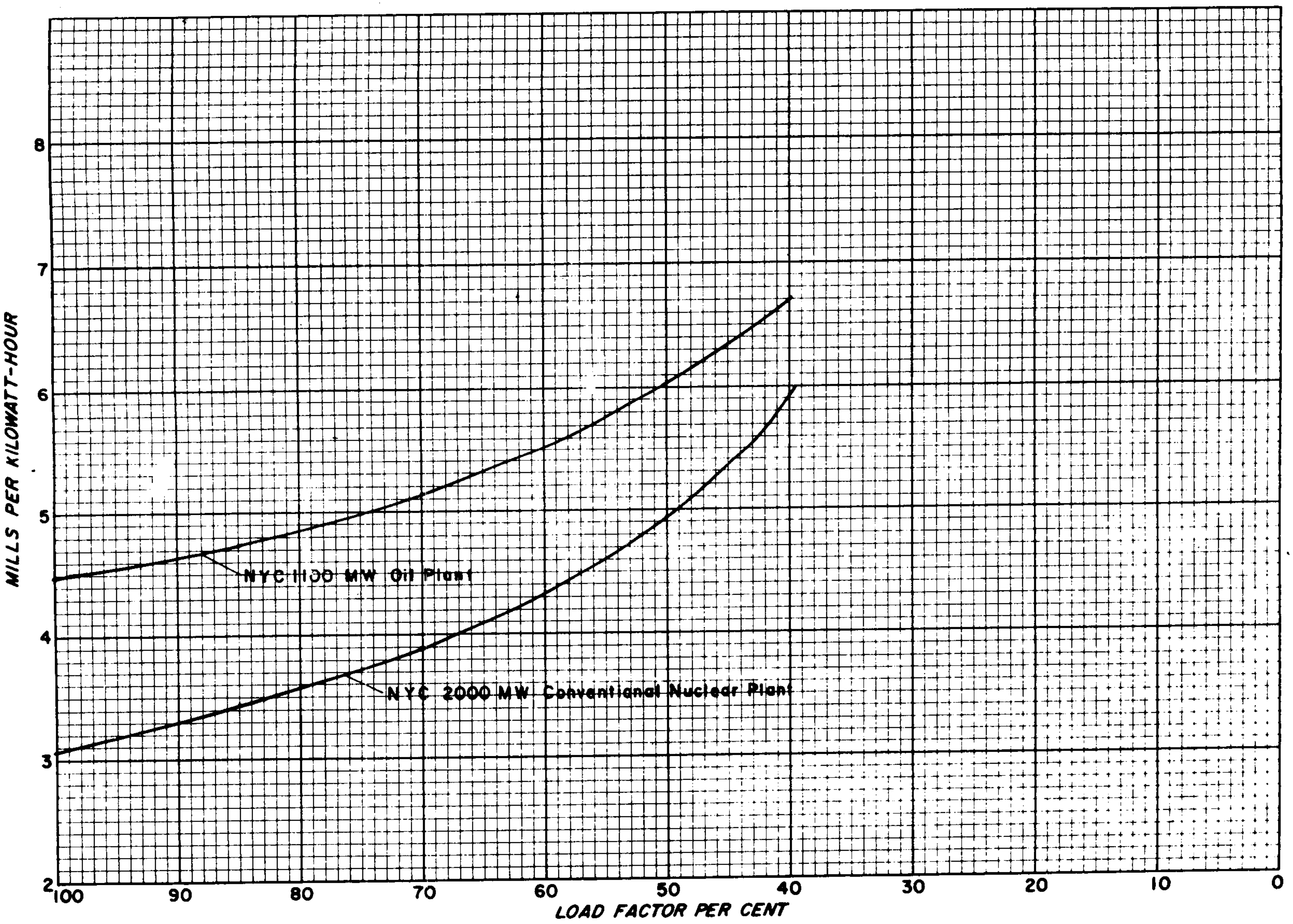

FIGURE 4-4. PLANT LOAD FACTOR vs AVERAGE COST OF POWER

4.9 DETERMINATION OF POWER COST

Many variables will affect the busbar cost of electric power from a new plant. They include plant size, investment, fixed charge rate on the investment, fuel cost, operation and maintenance costs, load factor, and plant availability.

The following assumptions have been made in order to estimate the busbar cost of power. Plant size and investment were determined by using the analysis developed in paragraphs 4-7 and 4-8. Each of the area electric grid systems was assumed to supply electricity to the liquid hydrogen facility continually at 2300 volts at the busbar (or substation in the case of transmitted power). Calculations were for a 30-year contract period, subject to cancellation charges.

4.9.1 LOAD FACTOR

The cost of power, which includes the cost of reserve power, was based on an average production of 0.96 kwh/hr/kw of capacity from the optimum size plant for a 30-year period.

4.9.2 PLANT AVAILABILITY

The optimum size plant is estimated to have 85 percent availability, with reserve power available during the remainder of the year. This hypothetical unit is scheduled to run at 96 percent capacity during availability. The effect of changing this 96 percent load factor on the average cost of power for two different size plants in New York City is shown in figure 4-4.

4.9.3 FIXED-CHARGE RATE

The annual fixed charge rate on the investment has been based on using an annual rate of return on capital investment of 6½ percent per year. This amounts to 4.15 percent levelized rate of return over a 30-year life for fossil and nuclear plants. The income tax effect has been computed based on United States rates, using a capital structure of 50 percent debt and 50 percent equity financing, accelerated depreciation, and investment tax credits. Straight line depreciation, ad valorem taxes at 2.0 percent per year and insurance at 0.15 percent for fossil and 0.40 percent for nuclear plants have been used.[102] A tabulation of this analysis which will be used as the standard for this study for the various type plants is shown in Table 4-17.

TABLE 4-17. FIXED-CHARGE RATE STANDARDS

Item	Fossil	Nuclear	Hydro
Average return on investment	4.15	4.15	4.15
Depreciation	3.35	3.35	2.00
Income-tax effect	1.35	1.35	1.25
Insurance	0.15	0.40	0.10
Property taxes	2.00	2.00	1.00
Total	11.00	11.25	8.50

Variations in fixed charge rates will obviously occur at the different locations. Graphical presentations showing the effect of varying the rate where the slope (M) represents the change in Mills per kwh for each 1 percent change in fixed-charge rate is shown in figures 4-5 through 4-7. Examples of some reported but unconfirmed fixed-charge rates are given in table 4-18.

TABLE 4-18. EXAMPLES OF FIXED-CHARGE RATES

Location	Area Adj. Factor	Fixed-Charge Rate	%/Year
		Nuclear	Fossil
France	1.1	9.4 [105]	8.06[95]
New York City	1.3	12.55[102]	12.05[102]
Montreal	0.9	7.00[39]	7.00[39]
United Kingdom	1.0	10.00[39]	10.00[39]
Tokyo	0.8	14.00[39]	13.00[39]
Los Angeles	1.1	13.00[103]	12.50[95]
Hawaii	1.3	11.30[95]	10.80[95]
TVA	1.0	5.90[97]	5.90[97]

4.9.4 OPERATING AND MAINTENANCE COSTS

The curves for each plant type shown in figure 4-8 have been scaled from a general curve given in data presented.[105] Scaling has been based on analysis of available data and estimated plant complexity.

4.9.5 COST OF FUEL

The cost of fuels presented in table 4-13 is the assumed cost delivered to the air terminal. Storage for oil or coal at the site is included in the power plant investment. Nuclear fuel costs presented in figure 4-9 are based on available data.[99] The fuel costs per kwh generated for each plant size and method have been computed by using 8750 Btu/kwh for coal, oil, and gas, and 10,500 Btu/kwh for nuclear fuel.[95]

4.9.6 RESERVE

A 15 percent reserve is required for each proposed power plant. The investment for this reserve is considered to be comparable in $/kw to the proposed power plant investment. Therefore, the analysis for determination of power cost includes the investment cost of reserve in the formula in paragraph 4.8. Operating and maintenance costs at 115 percent of figure 4-8 data have been used for allocating the costs of maintaining reserves.

Based on the foregoing methodology and analysis, power costs were determined for various size and type plants at each location and are presented in Appendix C. A summary of the power costs anticipated at each air terminal is presented in table 4-19.

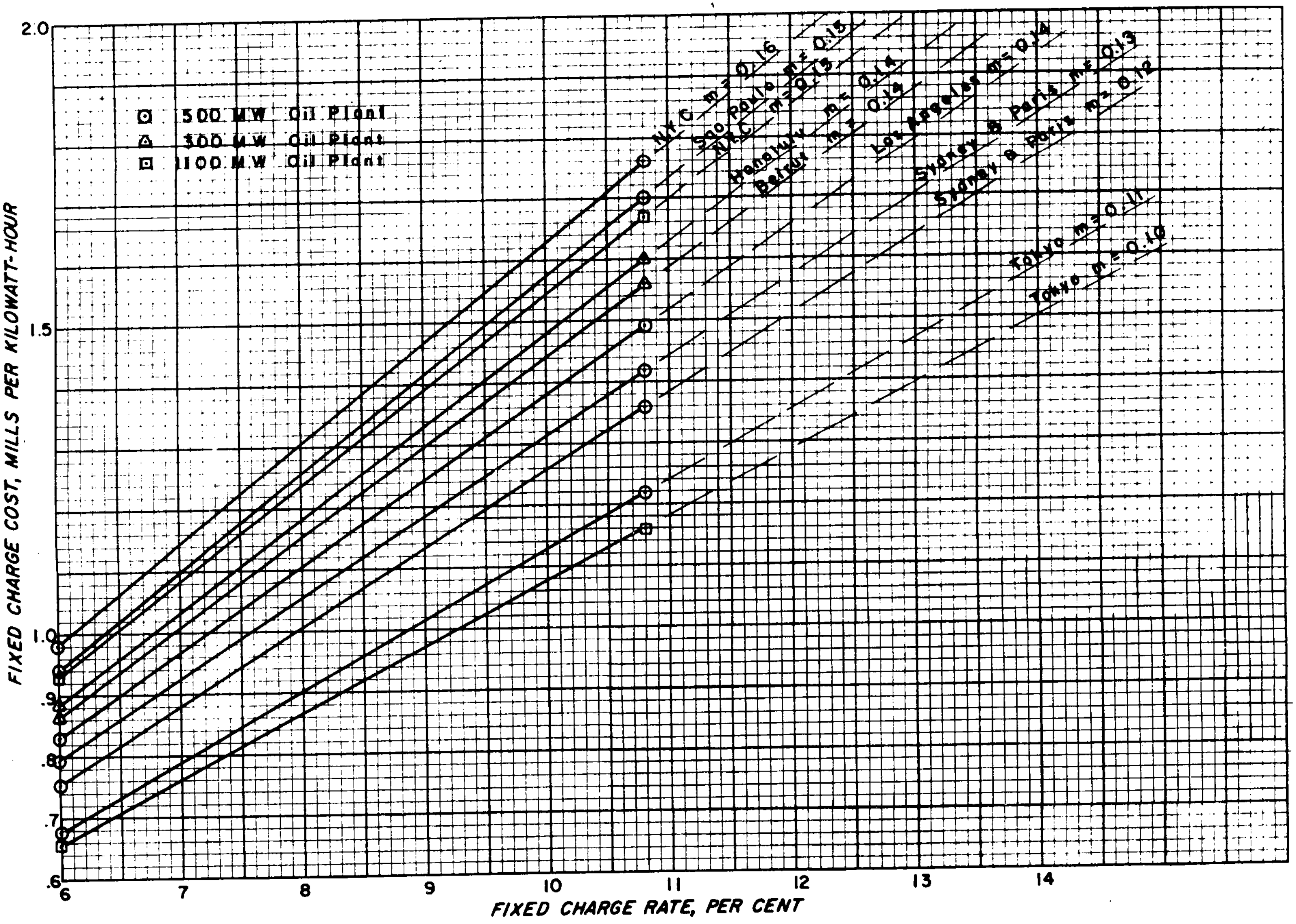

FIGURE 4-5. PLANT FIXED CHARGE RATE vs AVERAGE COST OF POWER (OIL)

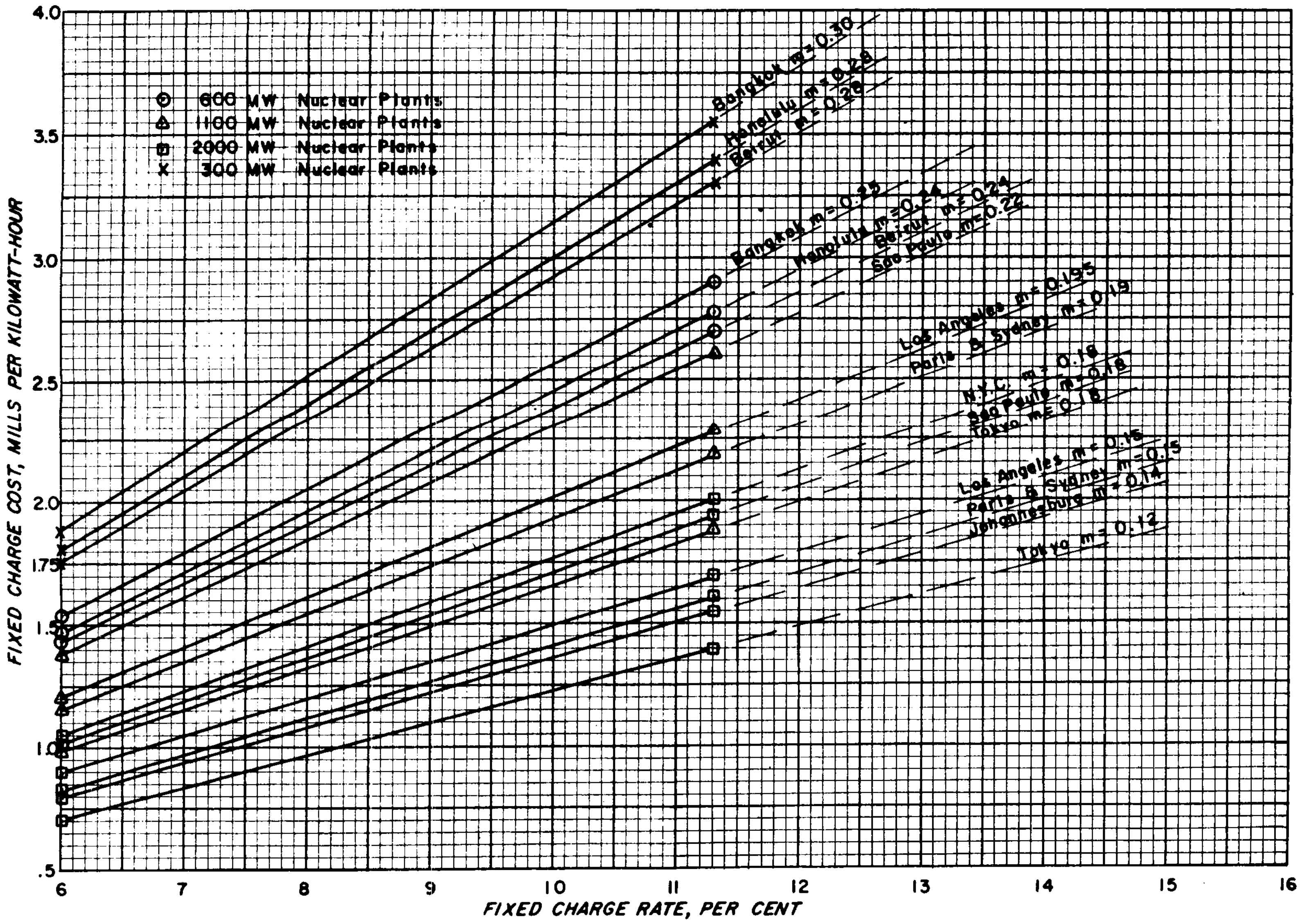

FIGURE 4-6. PLANT FIXED CHARGE RATE vs AVERAGE COST OF POWER (CONVENTIONAL NUCLEAR)

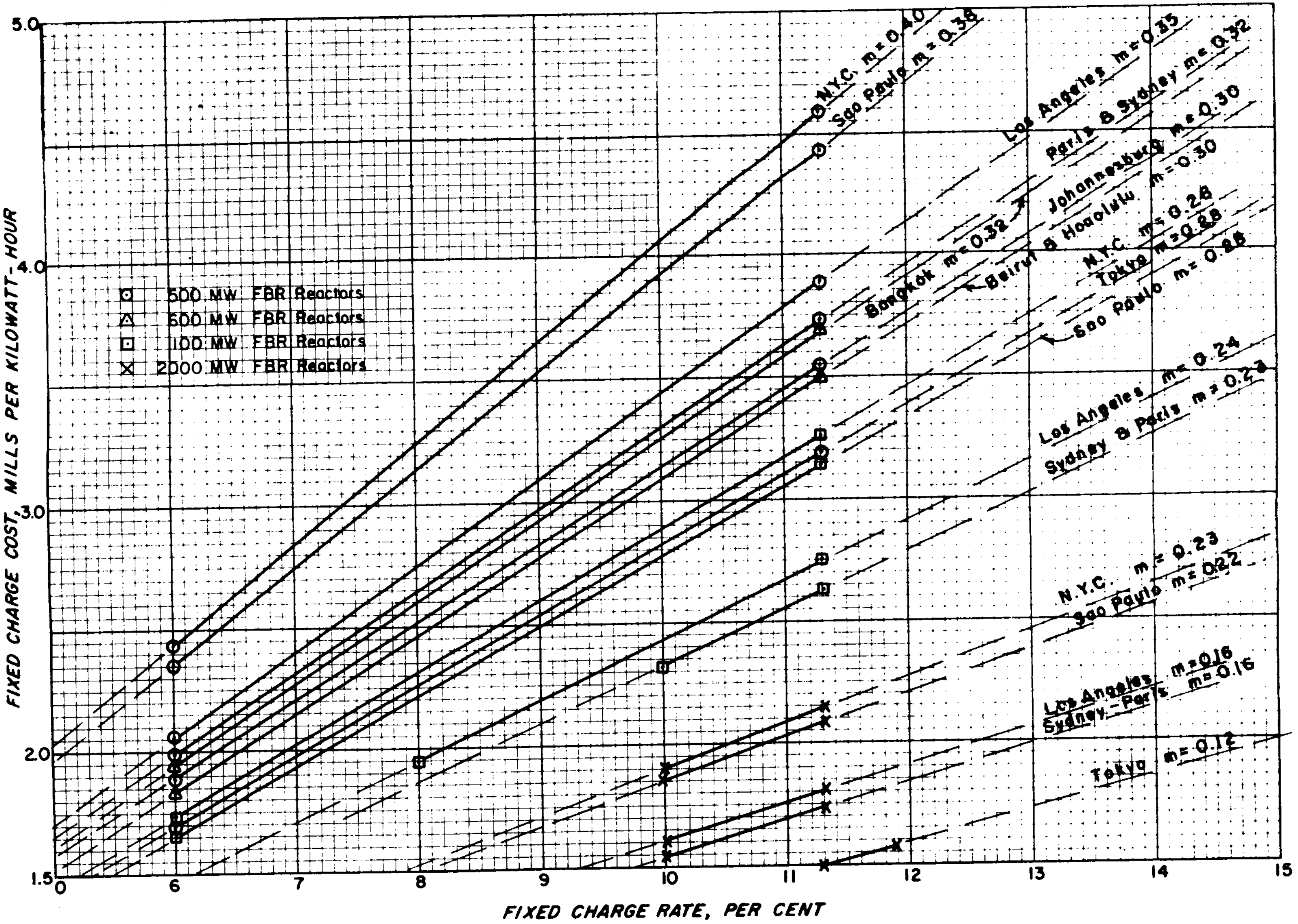

FIGURE 4-7. PLANT FIXED CHARGE RATE vs AVERAGE COST OF POWER (FBR NUCLEAR)

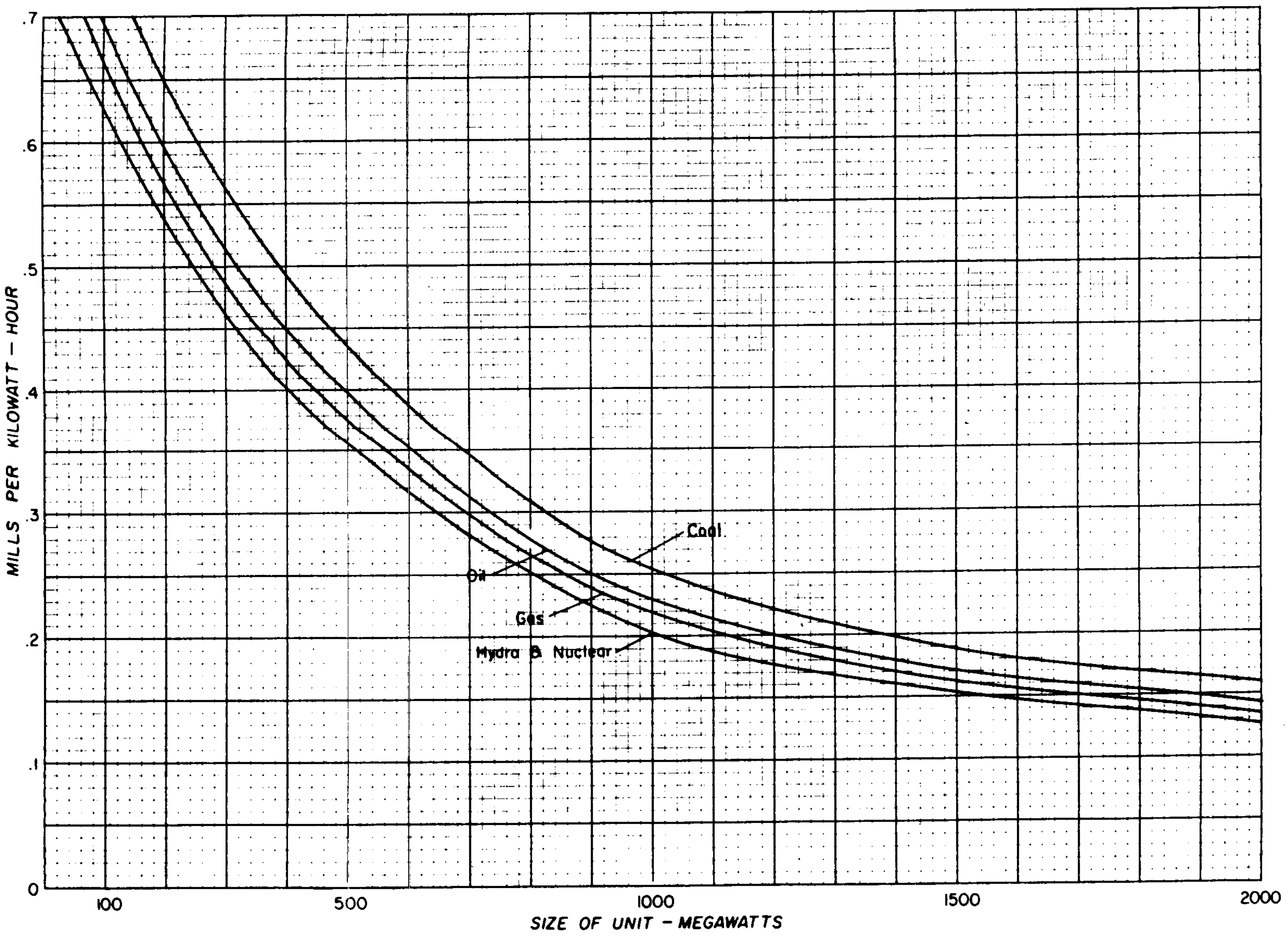

FIGURE 4-8. PLANT OPERATING AND MAINTENANCE COSTS vs PLANT SIZE

TABLE 4-19. UNIT COSTS OF ENERGY BASED ON INTEGRATED SYSTEM DESIGN AT EACH LOCATION

Location	Load mW	Plant Size mW	Fuel	Plant Type	Plant Cost $/kw	Mills Per KWH				
						Fixed Charges	Fuel Cost	Operating	Transmission	Total
HYDROCARBON PROCESS										
Bangkok	113	600	-	Hydroelectric	273	2.76	-	0.35	0.50	3.61
Honolulu	113	150	Oil	Steam	177	2.32	3.02	0.72	-	6.06
Johannesburg	113	1100	Coal	Steam	111	1.45	0.96	0.26	0.13	2.80
Beirut	171	300	Oil	Steam	127	1.66	2.45	0.58	-	4.69
Sao Paulo	226	1100	-	Hydroelectric	252	2.55	-	0.22	0.23	3.00
Sydney	226	1100	Coal	Steam	117	1.53	1.18	0.26	0.15	3.12
Tokyo	272	1100	Nuclear	Fast-breeder	168	2.25	0.60	0.22	-	3.07
Los Angeles	735	2000	Nuclear	Fast-breeder	135	1.80	0.60	0.15	-	2.55
Paris	1135	2000	Nuclear	Fast-breeder	130	1.74	0.60	0.15	-	2.49
New York	1430	2000	Nuclear	Fast-breeder	160	2.13	0.60	0.15	-	2.88
ELECTROLYSIS PROCESS										
Bangkok	530	1500	-	Hydroelectric	268	2.71	-	0.17	0.35	3.23
Honolulu	530	600	Nuclear	Fast-breeder	266	3.57	0.80	0.35	-	4.72
Johannesburg	530	1100	Coal	Steam	111	1.45	0.96	0.26	0.09	2.81
Beirut	990	600	Oil	Steam	109	1.43	2.45	0.38	-	4.26
Sao Paulo	1050	2000	Nuclear	Fast-breeder	154	2.06	0.60	0.15	-	2.81
Sydney	1050	2000	Nuclear	Fast-breeder	130	1.74	0.60	0.15	-	2.49
Tokyo	1300	2000	Nuclear	Fast-breeder	111	1.48	0.60	0.15	-	2.23
Los Angeles	3400	3000	Nuclear	Fast-breeder	127	1.70	0.60	0.11	-	2.41
Paris	5220	3000	Nuclear	Fast-breeder	121	1.62	0.60	0.11	-	2.33
New York	6600	3000	Nuclear	Fast-breeder	150	2.01	0.60	0.11	-	2.72

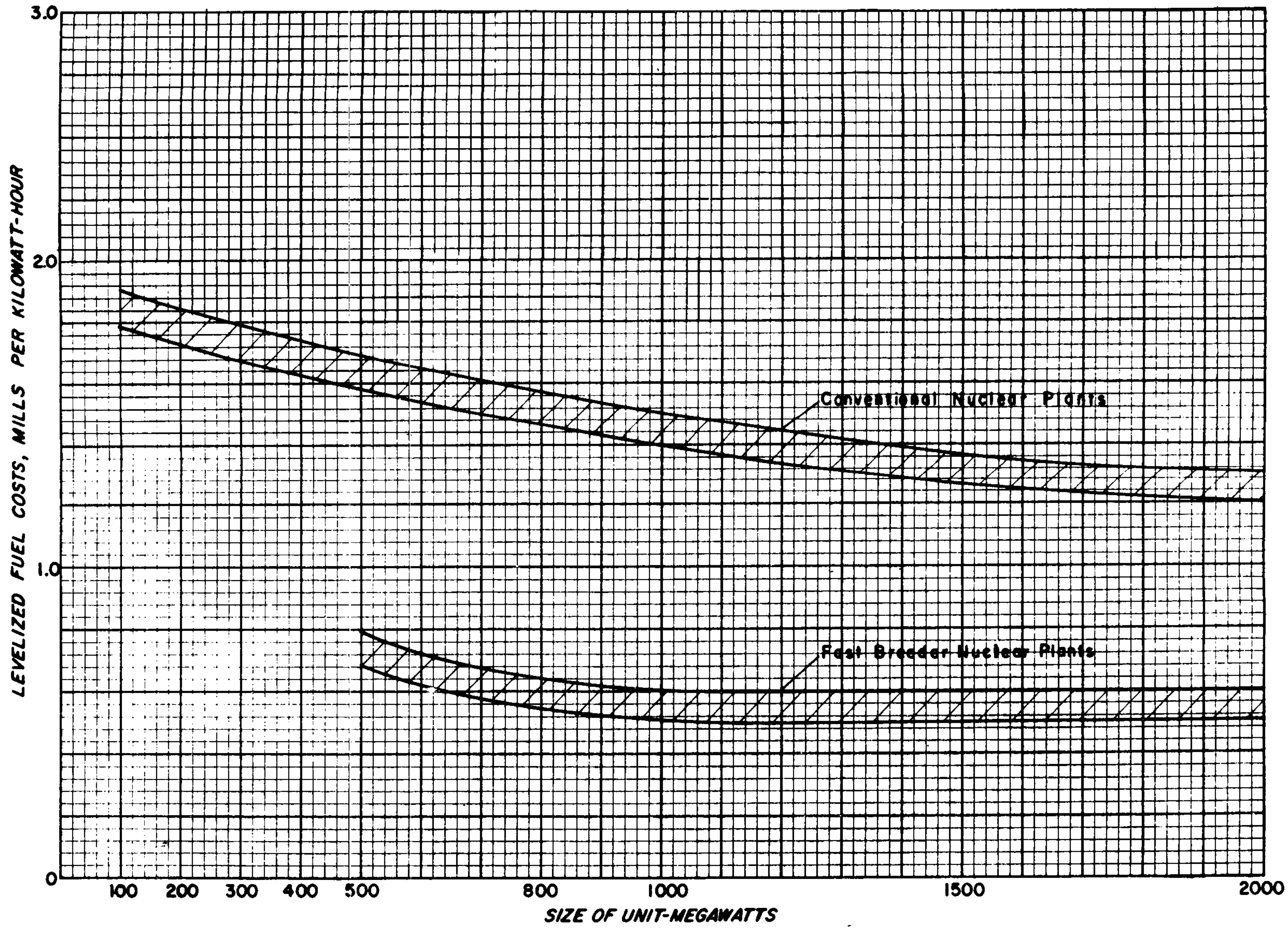

FIGURE 4-9. NUCLEAR FUEL COST vs PLANT SIZE (1980)

It should be noted that fuel costs used in power calculations in some cases are not identical to table 4-19, due to acquisition of new cost data. Most changes are considered insignificant. However, the updated cost data has been utilized in the discussion on hydrogen production in Section 5.

Section 5

COMPOSITE SYSTEM ANALYSES

5.1 GENERAL INFORMATION

Several approaches have been made to satisfy the respective production requirements. Analyses used in this report are not unique in that several other methods could be employed. However, the technique employed does lead to a meaningful development of the relationship between the various use points and their respective liquid hydrogen costs. To lead to this development a number of general assumptions or guidelines were established with respect to the overall analysis technique.

Information is presented for each of the ten geographic locations. Specific considerations lead to the development of hydrogen facilities meeting local production requirements. Costs for producing the liquid hydrogen product are established for each of these locations.

In certain instances more than one composite liquid hydrogen facility is economically feasible. In those cases comparative cost data is developed. Further, combinations of industry with the liquid hydrogen facility can be anticipated. In these cases general discussions relate to the effect of the peripheral industries upon the cost of the liquid hydrogen product.

All of the liquid hydrogen facilities are considered to be adjacent to the product use point. The cost data developed for the raw materials and process energy is related to each use point.

The production from any one facility has been taken as the nominal amount delivered to the aircraft servicing manifold. New York City, for example, requires a delivery of 2500 T/D. This assumption does not account for losses beyond the interface of the manifold. Actual requirements at the interface will be somewhat higher and could be considered in another analyses when more information is developed for liquid hydrogen use at the specific location.

Political considerations have not been incorporated in the analysis. Concerns such as air pollution, were judged to be outside the scope of this report. These considerations can be incorporated in future analyses when more specific evaluations are required.

Safety considerations have been incorporated in the buildup of each facility in terms of its components. These considerations, made with respect to safety in the development of the four respective subsystems, are considered adequate for all the locations.

The supply of energy and raw materials has been taken as "at the liquid hydrogen facility". In the first analysis no benefit can be gained by separating the energy and material source from the liquid hydrogen facility. Minemouth power generation with electric transmission to the liquid hydrogen facility is an example of this.

5.2 ANALYSIS TECHNIQUE

For this study, it was not possible to determine an aircraft schedule that would reflect anything less than continuous full requirement usage rate. The liquid hydrogen facilities are considered to 100 percent net production at all times, except in the case of a catastrophic occurence. Costs calculated, then, incorporate all the features required to provide this reliability. If an alternate method of operation would be desirable, the cost data can easily be adjusted.

Capital investment costs for the liquid hydrogen facilities will be adjusted by suitable area factors. Detail information describing these factors is presented in Section 4, paragraph 4.8.

Fixed charge rates are taken as 11 percent per year. This rate is based upon a 15-year straight line depreciation of money. No taxes or profit are reflected in this capital charge rate. The buildup of the charge rate, reflecting the current practice in the process industry, is presented in table 5-1.

TABLE 5-1. CHARGE RATE BUILDUP

Annual Fixed Item	Rate
Depreciation	6.67%/yr.
Insurance	0.33%/yr.
Average Interest[a]	3.00%/yr.
General & Administrative	1.00%/yr.
TOTAL	11.00%/yr.

[a] 3 percent is taken as the average interest corresponding to a rate on the unpaid balance of principal outstanding of 6 percent.

5.2.1 GAS PRODUCTION AND PURIFICATION INFORMATION

It has been assumed that a representative "on-stream" factor for the modular units of this subsystem is 93 percent. This is, 93 percent of the time, each module will be able to operate at rated capacity. The longest duration for an outage of a module is ten consecutive days.

This assumption reflects the best projections by the manufacturers of steam reforming and partial oxidation equipment. Those projections have been assumed to be representative of the coal processing equipment and the electrolysis equipment. Extension of the projection to coal equipment seems reasonable based upon the similarity of processes. Electrolytic processing is most unlike the other processes in terms of equipment. No real basis for a corresponding on-stream factor was determinable for this processing. Opinions range from electrolytic processing being grossly unreliable to that of being nearly 100 percent reliable. As an aid to analysis, then, all modules of the various processes have been assumed to perform with a 93 percent on-stream factor.

Projection of modular sizes from current technology by equipment and manufacturers indicate that the steam reforming and partial oxidation processes will be limited to not more than 250 T/D in the future. The other processes, less sophisticated in current development, are more difficult to project. Therefore, the 250 T/D module was chosen as a representative size for the purpose of this study.

It should be pointed out that the hydrogen halide process is not considered in the detailed analyses. A preliminary review of the process, from a cost standpoint, does not result in its being competitive. The extreme capital investment burden of this process is the basic cause for the extremely costly operation. As a process, hydrogen halide will remain as one of the future hydrogen producing methods, with the hope that economics will improve with time.

5.2.2 GASEOUS HYDROGEN LIQUEFACTION AND CONVERSION

The guidelines applicable to analysis of this subsystem are those established in Section 3, paragraph 3.3. That is, the high level cascade refrigeration low-level hydrogen recycle refrigeration process will be employed. Further, product orthopara conversion level will be taken as 95 percent para. Losses of hydrogen within the process are taken as 4 percent of the net production. That is, for each pound of hydrogen product, 1.04 pounds of hydrogen feed gas is required. This loss represents the experience of present technology and therefore, permits the sizing of the gas plant based upon the capacity required from the liquefaction units.

Modular sizes of liquefaction units could range from 250 to 2500 T/D. However, from an operational standpoint, it will be more desirable to have multiple units for the larger capacity installations. For purposes of this report, it will be assumed that the liquefaction modules can range to 1000 T/D capacity. The on-stream factor for each module will be taken as 97 percent with the longest outage period being taken as 7 consecutive days.

The on-stream data reflects current operating experience with large tonnage cryogenic process plants. Experience with future large scale plants may, in fact, result in improved on-stream characteristics. Use of the noted values, then, represents a conservative approach at this time.

It should be noted that liquefaction cost determination is, in fact, dependent somewhat upon the specific gasification process supplying the gaseous hydrogen. For example, a partial oxidation process will generate the makeup nitrogen gas required by the liquefaction equipment. Also, electrolytic process energy requirements will be such that lower cost energy should be available to the facility. These dependencies will be borne out in the respective analyses.

5.2.3 STORAGE OF LIQUID HYDROGEN

A variety of storage systems could be used. Preliminary economic analysis indicates that the evacuated perlite, double-wall tank configurations would be the more desirable configurations. This type of storage equipment design has been incorporated in the analysis of the respective facilities.

Facility operating experience requires that at least two storage vessels be built at each facility. Having two vessels will allow the simultaneous production of liquid hydrogen and the loading of aircraft at two different respective rates. Further, a measure of reliability is inherent in the storage facility in the event one tank has a failure.

Storage equipment sizing is primarily dependent upon the modules in other facility subsystems. The major portion of the storage capacity has the function of providing backup inventory for various subsystem module production outage. Determination of this portion of the storage capacity will differ for the various liquid hydrogen facilities. In addition to the production outage backup, an "operational flywheel" backup is provided equivalent to one day's withdrawal. This additional storage capacity provides for backup necessary to account for short duration operational production irregularities that can occur.

It is indicated that two vessels will be used to provide the required storage capacity. However, actual storage capacity must be somewhat greater than that required for outage and operational flywheel. The operation of the two tanks is such that inventory in each tank will cycle about the required level. In order to have the required inventory at all times, the oversize tanks allow for "overfilling" of one while the other is below its required level. Without being able to project the exact operational scheduling for each facility, it is felt that an "oversize factor" of 5 percent will result in satisfactory operation. This results in the actual storage capacity being:

$$\text{Storage Capacity} = 1.05 \times (\text{Subsystem Outage} + \text{One Day's Withdrawal})$$

Storage system losses are based upon the heat leak characteristics of the tanks chosen. Also, a loss due to conversion of ortho hydrogen to para hydrogen is considered. The loss due to conversion is taken as 1.25 percent of the liquid product from the liquefaction units (paragraph 3.3). The two losses, then, must be considered in sizing the amount of liquid production required.

5.2.4 DISTRIBUTION OF LIQUID HYDROGEN

It is judged that liquid hydrogen distribution will be limited to transfer from facility storage direct to aircraft fueling manifold. That is, there will be no extensive distribution of liquid hydrogen. This is in keeping with the premise that the liquid hydrogen facility is located adjacent to the use point. In this instance, the distribution of liquid hydrogen would be accomplished by a facility pipeline. Details of this facility piping are presented as examples in the report detailing pipeline distribution. The standard two mile lengths and attendant 7 percent loss characteristics will be used for the basis of the composite analyses. Other liquid hydrogen distribution systems could be incorporated at any location for distributing incremental product for uses other than aircraft. Such considerations, however, are not incorporated in the analyses.

5.2.5 CALCULATION TECHNIQUE

The calculations are classified into two categories; facility description (sizing), and production cost determination. The facility description is concerned with the calculations detailing the unit plant size and the operating requirements. These calculations relate to each location product usage within the general guidelines as presented. The production cost determination details the actual facility operation cost with respect to the system described. Ultimately, the cost determined, ratioed with the net product, yields a unit cost of product.

In application, the calculations detailing the facility descriptions are a series of trial and error estimates, ultimately defining an acceptable system. The trial and error calculations account for the losses of the product, the unit sizes, etc. The system ultimately described will provide the required product with sufficient reliability. The systems are characterized by the respective subsystem nominal size and daily average production. Based upon these two sets of characteristics investment and operating cost can be developed for a given facility. Competition between various systems for a given situation then can be evaluated in terms of the corresponding total production costs.

The trial and error calculations, detailing facility descriptions, are not presented in this report section. A tabulation of the production costs for the facility is made for each location. Two sets of sample calculations for typical facilities are presented in appendix D of this report. These calculations are typical of the larger and the smaller production requirements. Worksheets used in the production cost determination are included as Appendix E of this report. This information complements the tabulations for this section of the report.

5.3 GEOGRAPHICAL ANALYSIS

The information presented summarizes the calculations made in determining the plant subsystem sizes and product cost for each of the ten geographic locations. Detailed calculations of Bangkok and New York City typically show the method of calculating the summarized data and all production cost determination as presented in Appendix D. Several process subsystem combinations are economically feasible. Specifically this means that several types of gas generating processes, when coupled with the optimum liquefaction cycle, must be evaluated to determine the minimum cost of product. Where a type of process is obviously not competitive, a comment is made in lieu of the calculations.

While it is evident that an integrated facility combining the hydrogen production plant with a broader industrial complex, such as a refinery, chemical or petrochemical plant, or any other operation which could share raw materials or exchange byproducts will yield a lower cost of hydrogen, the definite selection of a particular integration of related operations at any of the proposed sites in the post 1980 period would be highly speculative. For each location, therefore, a set of information is presented for the "stand alone" facility, wherein all input costs relate exclusively to hydrogen production, and the resultant summation of costs yields the hydrogen cost at the location. To this base cost of hydrogen, credits can be cited for the value of salable byproducts of the "stand alone" facility. Potential savings and the resultant reduction to this base cost of hydrogen through integration are described in less definitive terms.

5.3.1 BANGKOK

5.3.1.1 Stand Alone Facility

Calculations were made to determine the annual production costs associated with the various combinations of subsystems and the resultant cost per pound of liquid hydrogen. The results are presented in table 5-2.

TABLE 5-2. PRODUCTION COST DATA (BANGKOK)

Component of Cost	Annual Production Cost in Millions of USA Dollars			
	Steam Reforming Based Plant	Partial Oxidation Based Plant	Coal Gasification Based Plant	Electrolysis Based Plant
Raw Materials	5.52	4.51	14.06	0.23
Energy	2.84	3.34	2.90	13.85
Capital Investment	5.75	5.85	7.37	8.33
Operating Cost	3.11	3.29	4.72	4.45
Total	17.22	16.99	29.05	26.86
	Unit Product Cost			
	11.8 cents/lb LH_2	11.6 cents/lb LH_2	19.9 cents/lb LH_2	18.4 cents/lb LH_2

It is apparent that the liquid hydrogen facilities employing electrolysis or coal gasification subsystems are not economically competitive. Therefore, these two systems do not warrant further evaluation for this location.

The competitive comparison, therefore, centers around the two remaining gas production and purification subsystems, steam reforming and partial oxidation.

Both processes are calculated to produce liquid hydrogen for nearly the same cost. Therefore, the choice of process is dependent upon the availability of raw materials and the byproduct credits that can be applied to cost of product as a result of the choice.

As summarized in paragraph 4.4.3.9.1, Thailand neither produces nor uses natural gas in its present industrial community. Practically all of the energy consumed is derived from imported oil. It could be projected, therefore, that oil will continue to be the predominant raw material. This projection is clouded by the fact that the LNG industry is in its early infancy and by the post-1980 era natural gas in liquid form could be available from Indonesia. Therefore, no clear-cut decision can be made at this time regarding the availability of raw materials.

In evaluating the credits due to byproducts, steam reforming and partial oxidation both produce about the same quantity of carbon dioxide per pound of product produced and are, therefore, a standoff.

Although it would be possible to utilize the carbon dioxide in a large chemical process industry, it is difficult to estimate a credit value for approximately 1400 tons of carbon dioxide per day. Therefore, no credit has been assigned.

In addition to carbon dioxide, the partial oxidation process also produces carbon and steam as a byproduct. Byproduct carbon can be used as a fuel, either in the hydrogen facility or in a brickette for market distribution. The value of the carbon (i.e., 10,000 tons per year) at 3 mills per pound is equivalent to $60,000 per year. Approximately 500,000 tons per day of saturated steam at 100 to 300 psia is available as a byproduct and can be marketed, for heating purposes, at a value of about $350,000 per year.

The two marketable byproducts, therefore, can reduce the annual production costs of the partial oxidation process by $410,000 per year, or $0.003 per pound of LH_2.

Based on this, it can, therefore, be seen that the choice of process is not clear cut and will be dependent upon the economy and industrial growth of the country. However, for the purpose of this report, the partial oxidation process is chosen as the logical process for Bangkok.

5.3.1.2 Integrated Facility

The industrial community in Bangkok will be well developed by the post-1980 period. It can be presumed that an industrial complex could be developed incorporating liquid hydrogen, industrial gas processing, and petrochemical process facilities.

Based on incorporating the hydrogen facility in the industrial complex, byproduct oxygen from the electrolytic process could be marketed. If an electrolytic process is selected for the hydrogen production, about 656,000 tons per year of gaseous oxygen would be available for sale within the complex. Based on an optimistic credit of $5.00 per ton for oxygen, the resultant product hydrogen cost is reduced from $0.184/lb LH_2 to $0.164/lb LH_2. Thus, hydrogen from an electrolytic cell still does not appear attractive in Bangkok for an integrated facility.

Partial oxidation appears to be the most practical method to employ. The integrated facility will constitute a large concentrated electrical energy load. Based on the larger power demand for the integrated load, a power cost of $0.00336/kwh instead of the stand alone cost of $0.00391/kwh is assumed. This lower power cost is equivalent to reducing product hydrogen costs to $0.113/lb LH_2 from $0.116.

A big potential, as well, for reducing the cost of product hydrogen lies in finding an alternate method generating or procuring gaseous hydrogen. For example, refinery of gases or tail gases may contain as much as 80 to 99 percent hydrogen by volume. The amount of such gas, however, normally is not large, relatively speaking. For example, a large refinery of current design can "supply" typically a hydrogen facility approximately 30 T/D production. A large chemical process, electric caustic works, for example, may liberate enough hydrogen to support a 10 T/D hydrogen facility. Of course, these high quality feedstocks are not available at no cost.

It might be projected, for example, that refineries, process industry, etc., could supply approximately 25 percent of the hydrogen required at the liquefier. Further, assume that the gaseous hydrogen costs an equivalent of approximately $0.005/lb LH_2 and the remaining 75 percent of the gas at the liquefier costing \$0.052/lb LH_2 (accounting for low cost energy in the partial oxidation calculation). This yields an average gas-to-liquefier cost of \$0.040/lb LH_2 and a product cost decreases from \$0.053/lb LH_2 to \$0.040/lb LH_2 for the liquefier gas. In turn, the total product cost reduces to \$0.103/lb LH_2 from \$0.116/lb LH_2, without counting low cost energy in the liquefaction plant. Accounting for low cost energy in liquefaction, the total product cost will further reduce to \$0.101/lb LH_2. Net product cost with \$0.003/lb LH_2 byproduct credit then becomes \$0.098/lb LH_2.

The value for the product cost is projected as representative of the integrated facility system. Additional considerations could lead to further cost reduction, but it is not judged that the reductions cited would be significantly greater. In other words, a product cost reduction of 20 percent can be taken as the maximum for the Bangkok location. Considering the 20 percent reduction, then the stand alone production cost of \$0.116/lb LH_2 reduces to \$0.091/lb LH_2 for an integrated facility.

5.3.2 HONOLULU

5.3.2.1 Stand Alone Facility

Calculations were made to determine the annual production costs associated with the various combinations of subsystems and the resultant cost per pound of liquid hydrogen. The results are presented in table 5-3.

TABLE 5-3. PRODUCTION COST DATA (HONOLULU)

Component of Cost	Annual Production Cost in Millions of USA Dollars		
	Steam Reforming Based Plant	Partial Oxidation Based Plant	Electrolysis Based Plant
Raw Materials	5.95	4.99	0.47
Energy	4.77	5.60	20.25
Capital Investment	5.49	5.59	7.96
Operating Cost	3.11	3.29	4.45
Total	19.32	19.47	33.13
	Unit Product Cost		
	13.2 cents/lb LH_2	13.3 cents/lb LH_2	22.7 cents/lb LH_2

Data presented in table 5-3 indicates that liquid hydrogen facilities employing an electrolysis subsystem are not competitive with those systems incorporating either steam reforming or partial oxidation. Therefore, no further attention will be directed toward the electrolytic gas plant facility. The steam reforming and partial oxidation facilities, however, have essentially identical production costs.

If a gaseous hydrogen producing facility were to be built in Honolulu today, in all likelihood the first choice would be the partial oxidation process. As in Bangkok, oil is the principle source of energy. No natural gas is currently used in the industrail community. It is projected, however, that large liquid methane distribution systems will be in operation in the mid-Pacific Ocean by the 1980 period, and that either of the two gas processes could be employed.

Therefore, no firm basis for making a selection between these two types of facilities now exists.

As referenced in the Bangkok study, partial oxidation process byproducts are of minimal value, while steam reforming byproducts are of essentially no value. It would not be anticipated that any significant byproduct credit could be established for the Honolulu plant.

As a result, it is projected that the product cost of liquid hydrogen at the Honolulu facility will be \$0.132/lb LH_2.

5.3.2.2 Integrated Facility

The present industrial community in Hawaii is of rather limited development. No real basis can be seen at this point in time for projecting a large or complex industrial development in the future. As a result, it is projected that an integrated concept of the facility will not be feasible in the Honolulu location.

A single exception does exist in that water desalinization may become a necessary industry in Honolulu. The main effect of the water desalinization would result in a reduction in energy cost to the liquid hydrogen facility. It is projected that the lower cost process energy would be \$0.00472/kwh, or the overall production cost of liquid hydrogen from the steam reforming gas plant facility and partial oxidation gas plant facility would be \$0.126/lb LH_2 and \$0.124/lb LH_2, respectively.

Further reductions in the cost of product hydrogen are not projected. Future developments in the industrial community may of course occur in the post-1980 period. However, no basis for making projections of these developments exist at this time. Thus, further integrated concept analysis of the product hydrogen cost is not feasible.

5.3.3 JOHANNESBURG

5.3.3.1 Stand Alone Facility

Calculations were made to determine the annual production costs associated with the various combinations of subsystems and the resultant cost per pound of liquid hydrogen. The results are presented in table 5-4.

TABLE 5-4. PRODUCTION COST DATA (JOHANNESBURG)

Component of Cost	Annual Production Cost in Millions of USA Dollars			
	Steam Reforming Based Plant	Partial Oxidation Based Plant	Coal Gasification Based Plant	Electrolysis Based Plant
Raw Materials	5.64	4.74	5.06	0.25
Energy	2.21	2.59	2.26	11.84
Capital Investment	5.10	5.19	6.55	7.40
Operating Cost	3.11	3.29	4.72	4.45
Total	16.06	15.81	18.59	23.94
	Unit Product Cost			
	11.0 cents/lb LH_2	10.8 cents/lb LH_2	12.7 cents/lb LH_2	16.4 cents/lb LH_2

The data presented indicates that hydrogen systems employing hydrocarbon processing are economically competitive while systems employing coal gasification or water electrolysis are significantly more costly.

Coal gasification based systems, as evaluated, have calculated production costs approximately 20 percent greater than the cost for the hydrocarbon system. Since the coal gasification system is used on a limited basis in today's industrial world and the scale-up to the projected size is far greater than in the steam reforming and partial oxidation system, the order of accuracy for estimating the cool gasification system could vary by this 20 percent. The electrolysis based plant is eliminated due to its still wider product cost margin.

It is, therefore, reasonable to project that either the hydrocarbon processes or the coal gasification could be employed as the gas generating subsystem in the Johannesburg location.

It should also be pointed out that no consideration is made in the calculation for economic influence that likely will be present in the Johannesburg location. Special credits or import duties undoubtedly could be established to encourage the use of the abundant coal resources available in the area for the liquid hydrogen facility. It would seem that as a minimum, the cost for producing the hydrogen by the coal gasification process could ultimately be equivalent to or less than the cost for the hydrocarbon system. In view of this, it could be projected that Johannesburg will, in fact, employ a coal gasification base system. For this analysis,cost of a partial oxidation based liquid hydrogen system will be used for evaluation.

As noted in the analysis for Bangkok, the value of byproduct carbon dioxide is considered to be negligible. This same consideration is being projected to the Johannesburg location. As a result, the representative cost for the liquid hydrogen from the partial oxidation base facility would be \$0.108/ lb LH_2.

5.3.3.2 Integrated Facility

The industrial economy of Johannesburg currently is a well developed and thriving system. The currently existing industrial community offers a significant potential for an integrated facility for Johannesburg location. In particular, the agricultural and basic metals processing industries provide a large, well diversified industrial gas complex for the post-1980 period. Further, it is expected that petroleum processing, although not an extensive industrial activity today, will be developed within the industrial community in the future.

In view of the foregoing, it would be anticipated that the liquid hydrogen system in Johannesburg would be a part of a large industrial complex generating several industrial gases, such as oxygen, nitrogen, argon, etc. This would result in providing lower cost energy to the liquid hydrogen facility and significantly reducing the cost of gaseous oxygen required in a coal gasification subsystem, below that associated with the stand alone facility.

It is projected that the cost of electrical power would be equivalent to the cost of power associated with the electrolysis based hydrogen system. Oxygen, in the integrated system, could be of such demand that effective cost of oxygen to the hydrogen system would be \$3.50/T. Considering these two factors, the production cost of liquid hydrogen would be reduced to \$0.104/lb LH_2 from \$0.108/ lb LH_2. Further reduction might be anticipated, but a minimum point of \$0.100/lb LH_2 is projected.

5.3.4 BEIRUT

5.3.4.1 Stand Alone Facility

Calculations were made to determine the annual production associated with the various combinations of subsystems and the resultant cost per pound of liquid hydrogen. The results are presented in table 5-5.

Coal gasification data has not been included since coal is not judged to be a readily available or an economically competitive raw material. The electrolysis process energy system was included for comparative purposes only. The data shows that the cost of hydrogen for this system is double that of the hydrocarbon processing systems. Petroleum processing and handling is a well developed industrial activity at the Beirut area. Future activities are likely to increase in the post 1980 period, and liquefaction of natural gas now virtually non-existent, can be anticipated as being one of the future petroleum processing industries.

For this reason no clear choice between the partial oxidation and steam reforming based hydrogen system can be made. For all practical purposes the systems are identically competitive and secondary factors such as raw-material availability and cost will be the determining factor in the ultimate choice.

TABLE 5-5. PRODUCTION COST DATA (BEIRUT)

Component of Cost	Annual Production Cost in Millions of USA Dollars		
	Steam Reforming Based Plant	Partial Oxidation Based Plant	Electrolysis Based Plant
Raw Materials	7.08	5.86	0.69
Energy	5.56	6.53	27.46
Capital Investment	7.05	7.18	10.58
Operating Cost	4.12	4.36	5.92
Total	23.81	23.93	44.65
	Unit Product Cost		
	10.9 cents/lb LH_2	10.9 cents/lb LH_2	20.4 cents/lb LH_2

As noted in the previous presentations, credit due to byproduct revenues for the partial oxidation system are comparatively small. A value of approximately 3 percent of the base product cost was established as the credit for byproducts at the Bangkok location. This credit was based upon byproduct revenues associated with waste materials from the partial oxidation system. By comparison, the steam reforming base system has only a single byproduct, carbon dioxide. As in the Bangkok situation carbon dioxide would be a byproduct of questionable value in Beirut.

In view of this the 3 percent byproduct credit established for a partial oxidation system in Bangkok, is taken as a credit for the partial oxidation system in Beirut. No byproduct revenue credit is established for the steam reforming system. This credit results, then, in a net product cost of \$0.106/lb LH_2.

5.3.4.2 Integrated Facility

As indicated, the Beirut location is projected to be a mid-east center of a petroleum processing industry. Complex industrial development associated with the use of or processing of petroleum products are expected to accompany the basic refinery operations. Systems to produce chemical materials in agricultural supplies can be anticipated. This would provide an opportunity for an integrated facility with respect to the liquid hydrogen processing requirement. As a minimum, it would be expected that concentrated electrical loads would result in a lower cost energy and would be available to the hydrogen facility. Also, the use of petroleum byproducts for major products at lower unit costs would be anticipated.

In the end analysis, the cost reduction resulting from an integrated facility at Beirut should be analogous to the 20 percent value established in the Bangkok location. The choice of the system (steam reforming or partial oxidation) will depend upon the economic competition. It is reasonable to expect, therefore, that a product cost approximating \$0.087/lb LH_2 could be achieved in Beirut with an integrated facility.

5.3.5 SAO PAULO

5.3.5.1 Stand Alone Facility

Calculations were made to determine the annual production costs associated with the various combinations of subsystems and the resultant cost per pound of liquid hydrogen. The results are presented in table 5-6.

TABLE 5-6. PRODUCTION COST DATA (SAO PAULO)

	Annual Production Cost in Millions of USA Dollars		
Component of Cost	**Steam Reforming Based Plant**	**Partial Oxidation Based Plant**	**Electrolysis Based Plant**
Raw Materials	10.27	9.21	0.45
Energy	4.73	5.55	24.06
Capital Investment	10.42	10.60	15.92
Operating Cost	5.00	5.30	7.23
Total	30.42	30.66	47.66
	Unit Product Cost		
	10.4 cents/lb LH_2	10.5 cents/lb LH_2	16.3 cents/lb LH_2

Coal gasification has been determined to be too expensive a process for consideration at the Sao Paulo location. For this reason coal gasification data was not presented.

Data for the electrolysis base hydrogen system indicates that the production cost approximates 150 percent of similar costs for the steam reforming and the partial oxidation base systems. In a preliminary analysis the product cost of the electrolysis base system is not competitive with product from the hydrocarbon system.

However, a rather significant amount of uncertainty exists in evaluating the future economic pattern for the Sao Paulo area. It is conceivable for example, that the electrolysis system will be more competitive in the future. There may be more need for electrical energy in the Sao Paulo area than currently can be forecast and projected. This would result in a reduction in the cost or energy. Also, the cost of producing gaseous hydrogen from the hydrocarbon processes may be more expensive than currently projected. However, based on the data available, the hydrocarbon processing systems are the processes to be used in producing hydrogen in a stand alone facility, and in particular, the steam reforming system. The basis for this choice lies in the fact that the supply of oil required for the partial oxidation base process tends to be questionable in the near future. It would be expected that liquefied natural gas delivered by pipeline from the Bolivian fields to the Sao Paulo area would be developed prior to the expensive oil supply systems.

Since the system choice is based upon a steam reforming gas plant, no significant credit can be established for byproducts. In view of this then the projected price for liquid hydrogen products from the stand alone facility is \$0.104/lb LH_2.

It should be noted, however, that the price of liquid hydrogen in the Sao Paulo area is artificially high in the sense that a more desirable geographical location could be found for hydrogen production in this general area. Specifically, a relocation of hydrogen facilities and its use point to an area closer to Buenos Aires, could prove beneficial to the overall economics of the hydrogen facility. This specific point is not pursued in this analysis.

5.3.5.2 Integrated Facility

The future basic industrial activity near Sao Paulo may be directed more rapidly towards supporting an agricultural economy than in expansion of metallurgical processes. An integrated complex facility could be developed with the agricultural chemical industry in producing gas products as the primary revenue materials. Further, such a facility would use natural gas as a raw material.

As stated previously, this expanding industrial economy and the integration of the hydrogen facility into a chemical complex would result in the lowering of the electrical costs to a value comparable to the cost associated with an electrolytic based hydrogen system, i.e., \$0.00281/kwh. This results in a cost reduction for product hydrogen, from \$0.104/lb LH_2 to \$0.102/lb LH_2, using a steam reforming subsystem.

In addition, if the hydrogen facility is integrated with an ammonia facility the cost of natural gas could be reduced from the \$0.40/1000 scf projected in the evaluation, to \$0.30/1000 scf. This would result in cost of the liquid hydrogen being further reduced from \$0.102/lb LH_2 to \$0.098/lb LH_2. Similar costs reduction could be experienced if fuel oil was used as the fuel instead of natural gas in the integrated hydrogen-ammonia complex.

It can be projected that the lowest cost of hydrogen that can be reasonably anticipated, based on an integrated, expanded industrial economy, is \$0.095/lb LH_2.

5.3.6 SYDNEY

5.3.6.1 Stand Alone Facility

Calculations were made to determine the annual production costs associated with the various combinations of subsystems, and the resultant cost per pound of liquid hydrogen. The calculational results are presented in table 5-7.

TABLE 5-7. PRODUCTION COST DATA (SYDNEY)

Component of Cost	Annual Production Cost in Millions of USA Dollars			
	Steam Reforming Based Plant	Partial Oxidation Based Plant	Coal Gasification Based Plant	Electrolysis Based Plant
Raw Materials	9.93	10.48	9.56	0.49
Energy	4.91	5.77	5.02	21.33
Capital Investment	8.75	8.90	11.21	13.37
Operating Cost	5.00	5.30	7.41	7.23
Total	28.59	30.45	33.20	42.42
	Unit Product Cost			
	9.8 cents/lb LH_2	10.4 cents/lb LH_2	11.4 cents/lb LH_2	14.5 cents/lb LH_2

The basic observation drawn is that the electrolysis based hydrogen system is not competitive with the other three systems. The three systems based on hydrocarbon or coal processing are competitive for all practical purposes. The availability of resources for the steam reforming system incorporates questionable projections with respect to the future. Further, one would expect political and economic influences to favor the use of coal in Sydney, making the coal gasification system more economically competitive in any real situation. All things being considered, it is reasonable to project that the Sydney hydrogen system, in fact, will be based on a coal gasification process, and that the cost of hydrogen to meet competitive forces of the optimum system will approach the average cost of the steam reforming based system as calculated Liquid hydrogen in Sydney would therefore cost \$0.098/ lb LH_2.

5.3.6.2 Integrated Systems - Sydney

The industrial economy in the Sydney area is projected to be fully developed and widely diversified in the post-1980 period. In view of this, it is expected that an integrated facility incorporating a liquid hydrogen system would be feasible. Development of the cost of liquid hydrogen product becomes a complex speculative calculation in view of a seemingly infinite variety of integrated system methods. A basic appreciation for the effect upon the cost of hydrogen can be expected based on reduced raw material costs and energy costs that could result from any integrated system. In this instance, it would be expected that the cost of raw material coal, oxygen, and electrical power would be somewhat reduced from the rates applied in the stand alone calculations. By making reasonable assumptions as to the reduced cost of coal, it could be expected that the volume purchase price of this raw material could become equivalent to the mine mouth price of this commodity. Further, the electrical energy cost might be projected to be that equivalent to electrical energy cost used in the electrolysis based system analysis for the stand alone calculation. Further, the cost of oxygen anticipated for the stand alone basis was based upon the supply of oxygen solely for the coal gasification application. It would be expected that the cost of oxygen would be substantially reduced in the integrated system, since the overall market for this commodity could be comparatively large. It might be projected that, in fact, the cost of the oxygen material would be one-half that for the stand alone evaluations. Based on the foregoing observations, a cost reduction approximating 10 to 15 percent can be projected for the production of hydrogen from the integrated facility. Considering then the cost from the stand alone facility to be \$0.098/lb LH_2, an integrated facility production cost would be reduced to \$0.083/lb LH_2.

Further reduction of costs for product from the integrated facility could be anticipated when a full appreciation of the merchant trade in either gaseous hydrogen or liquid hydrogen is considered. This detailed analysis is not incorporated in this study; however, the effect of such merchant distibution of the hydrogen system products would further reduce the end cost for the hydrogen produced.

5.3.7 TOKYO

5.3.7.1 Stand Alone Facility

Calculations were made to determine the annual production costs associated with the various combinations of subsystems and the resultant cost per pound of liquid hydrogen. The results are presented in table 5-8.

The data presented shows that the steam reforming and the partial oxidation base systems are economically comparable. The choice between the two systems is difficult to make. Both natural gas and oil will be imported into Japan to satisfy the raw material requirements of the future.

The supply of natural gas in Japan is currently inadequate, and, as a result of its short supply, has stimulated the importation from outside sources such as Russia and Alaska. All the oil is currently being imported, primarily from the Middle East. The requirements of a 500 T/D hydrogen plant, while adding to the deficit side of a short supply of fuel, should fall well within the normal annual supply increase of both natural gas and oil. Therefore, although the choice is arbitrary, steam reforming has been chosen as the base system in this report with the resultant cost of \$0.096/lb LH_2.

It should be noted, however, that in the case of Japan, the electrolysis base system is less than 40 percent higher than the hydrocarbon processes. The byproduct oxygen generated as a part of this process will amount to 1.642 million tons per year and at a marketable value of $4.00/T will reduce the cost from $0.138 to $0.120/lb LH_2.

TABLE 5-8. PRODUCTION COST DATA (TOKYO)

Component of Cost	Annual Production Cost in Millions of USA Dollars			
	Steam Reforming Based Plant	Partial Oxidation Based Plant	Coal Gasification Based Plant	Electrolysis Based Plant
Raw Materials	13.49	13.23	37.70	1.17
Energy	6.06	7.11	6.19	23.93
Capital Investment	10.05	10.23	12.87	15.84
Operating Cost	6.51	6.95	9.59	9.53
Total	36.11	37.52	66.35	50.47
	Unit Product Cost			
	9.9 cents/lb LH_2	10.3 cents/lb LH_2	18.2 cents/lb LH_2	13.8 cents/lb LH_2

This byproduct credit could serve as the basis for the electrolysis base system becoming more competitive. Particularly when the costs of this yet to be built system can be more accurately estimated.

It is recommended that if plans for a plant in Tokyo become more definitive, the electrolysis base system be competitively re-evaluated.

5.3.7.2 Integrated Facility

No specific projection is made at this time regarding the nature of such a complex. Rather, it is expected that the value established in previous cost will, in fact, be representative for Tokyo as well, that is, it is expected that the integrated facility would result in cost reduction of 10 to 15 percent.

Assuming that the value of 15 percent is that which would be expected, the projected cost product hydrogen in the Tokyo facility is \$0.84/lb LH_2 from the steam reforming based process.

5.3.8 LOS ANGELES

5.3.8.1 Stand Alone Facilty

Calculations were made to determine the annual production cost associated with the various combinations of subsystem and the resultant cost per pound of liquid hydrogen. The results are presented in table 5-9.

TABLE 5-9. PRODUCTION COST DATA (LOS ANGELES)

Component of Cost	Annual Production Cost in Millions of USA Dollars		
	Steam Reforming Based Plant	Partial Oxidation Based Plant	Electrolysis Based Plant
Raw Materials	30.67	28.14	3.01
Energy	13.05	15.33	67.13
Capital Investment	23.03	23.41	38.54
Operating Cost	11.68	12.42	17.23
Total	78.43	79.30	125.91
	Unit Product Cost		
	8.3 cents/lb LH_2	8.4 cents/lb LH_2	13.3 cents/lb LH_2

The data presented summarizes the calculations from those systems considered feasible for the Los Angeles location. Coal gasification base systems were not considered. The remoteness of supply and resultant transportation costs dictated that any process utilizing coal will be extremely expensive and obviously non-competitive.

The electrolysis system is also non-competitive. In fact, the degree of cost differential, when compared to the hydrocarbon based system, is greater in Los Angeles than any location previously analyzed. Therefore, as has occurred at previous locations, the choice of plant must be made between two economically competitive hydrocarbon gas producing systems.

As shown for the Bangkok location, all else being equal, the byproduct credits may dictate the choice of gas producing system. Therefore, for the purpose of the study, a partial oxidation system is chosen and as noted in the Bangkok study, the cost of hydrogen can be reduced by \$0.003/lb LH_2 to \$0.081/lb LH_2.

The Los Angeles location for the hydrogen system has been considered as adjacent to the existing Los Angeles International Airport. It should be noted that significantly different cost values would be associated by relocating the hydrogen facility and airport to a location away from the presently designated point. In particular, relocating a facility north or south from the present area would permit the construction of a completely independent island in the low waters of the ocean with sufficient area for the hydrogen facility, airport, and other supporting activities. No determination has been made of this feasibility in this report. It stands as one of the avenues of endeavor that shoud be followed if further cost information is required in relation to the overall HST program for the Los Angeles area.

5.3.8.2 Integrated Facility

The economic community will be such that full scale integration of facilities will be feasible in the post 1980 period. However, the size of the liquid hydrogen facility is expected to be relatively large by comparison to other industrial facilities projected to be in the area. Further, the quantities of materials that are either required or produced by the liquid hydrogen systems are likely to be excessive with respect to the prospective market within the immediate area of the facility. In view of this, the cost reduction at the Los Angeles location will be less, on a percentage basis, than those projected for other locations. As a minimum, it would be expected that the electric power required by the hydrogen facility could be obtained at a cost equivalent to the costs projected for the electrolysis base system facility. Including the byproduct credit discussed in the previous subsection, a reduction of \$0.006/lb LH_2 can be expected.

Other reasonable projections that might be expected for an integrated facility could amount to an additional cost reduction approximating \$0.004/lb LH_2. Based on this projection, then, it is projected that the integrated facility could result in an overall cost reduction of the product hydrogen from \$0.084/lb LH_2 to a value of \$0.074/lb LH_2.

As indicated, the relocation of the hydrogen facility in the Los Angeles area should be considered in a further investigation. For example, a large integrated facility incorporating a power plant desalinization process, a hydrogen facility, and other processes could result in reducing the cost of liquid hydrogen significantly below those projected.

5.3.9 PARIS

5.3.9.1 Stand Alone Facility

Calculations were made to determine the annual production costs associated with the various combinations of subsystems and the resultant cost per pound of liquid hydrogen. The results are presented in table 5-10.

The data presented indicates that the steam reforming base system is the system type to be employed at the Paris location. It should be noted, however, that for all practical purposes, the two hydrocarbon systems are economically competitive; while the electrolysis base system is, in fact, 30 percent more costly.

A steam reforming base system would be a choice in Paris based on the projected raw material supply system anticipated in the post-1980 period. Liquefied natural gas will be a rather widely distributed and used commodity in France during this period. It would be expected that the high cost of the liquefied natural gas used for the calculation could be tempered somewhat by a wider scale distribution of the liquefied methane as a commodity than is projected in this report. Petroleum products that would be required to sustain a partial oxidation system will continue to be relatively costly when compared with the liquefied natural gas as a raw material. Therefore, it is logical to anticipate the steam reforming base system for a Paris location.

TABLE 5-10. PRODUCTION COST DATA (PARIS)

	Annual Production Cost in Millions of USA Dollars		
Component of Cost	**Steam Reforming Based Plant**	**Partial Oxidation Based Plant**	**Electrolysis Based Plant**
Raw Materials	60.58	58.49	2.32
Energy	19.60	23.01	99.78
Capital Investment	29.43	29.90	49.98
Operating Cost	15.18	16.15	22.38
Total	124.79	127.55	174.46
	Unit Product Cost		
	8.5 cents/lb LH_2	8.7 cents/lb LH_2	11.9 cents/lb LH_2

Considering the Paris hydrogen facility to be a steam reforming based system, little revenue from byproducts therein generated could be anticipated. The major byproduct of carbon dioxide will be of questionable value in the 1980 period particularly in the quantities anticipated in the steam reforming base system. It is projected, therefore, that the product cost for the Paris facility in the stand alone analysis would be \$0.085/lb LH_2.

5.3.9.2 Integrated Facility

An integrated facility at the Paris location would be a distinct possibility. However, as noted for Los Angeles, the size relationship of the hydrogen facility to other projected industries becomes even more significant. It would be expected that the 2000 T/D LH_2 facility would be an extremely large system when compared with other anticipated industries that could be considered for an integrated facility system. In view of this, a large reduction in the cost of producing the liquid hydrogen would not necessarily be projected, but rather a modest cost reduction could be expected, particularly in the event that the power costs to the hydrogen facility are effectively reduced to the equivalent cost used in the calculations for the electrolysis base system. This cost reduction alone amounts to \$0.002/lb LH_2.

It is projected that a further cost reduction, anticipated from an integrated facility, would amount from 6 to 8 percent, or \$0.003/lb LH_2. Therefore, the total cost reduction projected for a complete system integration is \$0.005/lb LH_2. This reduces the production costs from \$0.085/lb LH_2 to \$0.077/lb LH_2.

5.3.10 NEW YORK CITY

5.3.10.1 Stand Alone Facility

Calculations were made to determine the annual production costs associated with the various combinations of subsystems and the resultant cost per pound of liquid hydrogen. The results are presented in table 5-11.

The data presented indicates that the hydrocarbon based systems will be economically competitive for the New York City location. Electrolysis base systems are not projected to be competitive. The choice of the gas producing based system is, therefore, between steam reforming and partial oxidation and must be based on the availability of raw material and the credits applied to the base.

Natural gas is currently available by pipeline from Texas. Its capacity is currently inadequate, therefore, requiring the installation of LNG peak shavings facilities. It is projected that LNG will be tanked into New York from Venezuela in the future with a resultant higher cost.

Oil is currently imported from the Caribbean and with the expansion of the relatively newer fields, this may have a modest tendency to reduce the price. The increased labor and equipment costs will, however, have a tendency to dilute this savings so that in the end the change in costs may offset each other.

Therefore, based on the evaluation and the relative byproduct credits, the partial oxidation based system is chosen for the New York City location.

The cost of petroleum products vary greatly in the New York City area. A subsequent study should incorporate an evaluation of the hydrogen costs at locations other than that chosen for this study. For example, if the airport and the hydrogen facility were located in Northern New Jersey, as opposed to the Long Island location, a significant savings in the cost of hydrogen product could be anticipated. Further, the value of byproduct developed by a facility in the Northern New Jersey location could be somewhat greater than the value of those byproducts generated at the Long Island location.

Estimating the credit revenue anticipated from marketing byproducts for the New York City location is speculative at best. The quantities of byproducts to be produced by the partial oxidation based system will be very large when related to the potential market. It would be expected, however, that as a minimum all byproduct steam could be satisfactorily marketed. The other byproducts (carbon and carbon dioxide) could be marketed to a lesser degree.

Considering the byproduct steam to have a marketable value of $0.30/1000 lb, a credit amounting to $0.0017/lb LH_2 could be expected. It is expected that a total byproduct credit amounting to $0.002/ lb LH_2 is reasonable for the marketing of the steam and a portion of other byproducts. This reuction then results in a projected cost of hydrogen of $0.079/lb LH_2 for the New York facility.

TABLE 5-11. PRODUCTION COST DATA (NEW YORK CITY)

Component of Cost	Annual Production Cost in Millions of USA Dollars		
	Steam Reforming Based Plant	Partial Oxidation Based Plant	Electrolysis Based Plant
Raw Materials	60.62	54.46	5.82
Energy	28.33	33.27	145.57
Capital Investment	41.56	42.25	74.01
Operating Cost	17.35	18.52	26.08
Total	147.86	148.50	251.48
	Unit Product Cost		
	8.1 cents/lb LH_2	8.1 cents/lb LH_2	13.8 cents/lb LH_2

5.3.10.2 Integrated Systems

As with hydrogen facilities located at other points of high product usage, the prospects for developing an integrated facility incorporating liquid hydrogen systems are not extremely attractive. The hydrogen system itself will be an extremely large undertaking, by comparison, to any of several possible peripheral activities that could be combined to form an integrated facility. As a result, it is increasingly difficult to speculate on the effects of a system integration. It would be expected, as a minimum that the byproduct credit established for the stand alone facility would be available for the hydrogen facility with an integrated system. Further, the cost of electric power for the integrated facility will be comparable to that used in the electrolysis based system analysis. These two factors result in reducing the cost to $0.077/lb LH_2. At most a 10 percent reduction of product cost would be projected for the integrated plant, or $0.073/lb LH_2.

Section 6

SYNOPSIS

6.1 DATA ACCURACY

A relative appreciation or ranking of plant designs has been accomplished. More specific cost determinations and calculations for final production cost accuracy are to be completed in future comprehensive studies. Qualitatively, the cost data accuracy range is ± 15 to 20 percent of the values determined. Many inputs in the data development of this study are accurate within a range of ±10 percent, while others are obviously more speculative. A detailed evaluation of calculation accuracy is not justified at this point. The inaccuracies related to some specific system characteristics result in unfavorable economics. (Especially, unproved systems as water electrolysis, hydrogen halide, and coal gasification.) Raw material and energy requirements for the systems considered reflect high process efficiency, approaching stoichiometric limits perhaps, while operating cost and investment data are more speculative. Development work in the future will then reduce the inaccuracies. Reduction of the inaccuracies should tend to make certain systems more competitive, since a conservative approach was taken in the situations of speculative analysis.

6.2 FUTURE PRODUCTION COSTS

The product costs determined and presented in this report are significant in relation to future large scale production of liquid hydrogen and its use. Future cost structures undoubtedly will differ somewhat from the projections presented in this report. The following paragraphs relate to some of the general future production costs.

Fixed charges based upon the plant investment typically constitute 30 to 32 percent of the production cost for hydrocarbon based and electrolysis based plants. Annual fixed charge calculations in this report were based on 11 percent of the original plant investment. The 11 percent value developed does not reflect what may in fact be future applicable rates. Financing, tax considerations, etc., could yield capital charge rates ranging from 5 to 15 percent. The effect upon the unit product cost is apparent. For example, unit product cost for New York City (partial oxidation based plant) ranges from 6.9 to 8.9 cents/lb LH_2, corresponding to annual capital charge rates ranging from 5 to 15 percent.

In addition to variation in capital charge rates, future plant investments will likely be somewhat different from those presented in this study. A base set of geographic adjustment factors was incorporated in the calculations for this study. Future developments could alter significantly the adjustment factors. Also, as noted, capital investment projections for the various plants could be in error.

Future unit product costs could be lower due to variation in associated capital investment calculations. Similarly, considerations of raw material and energy, due to unit product cost, can be expected.

However, the basic unit product cost of raw material and energy requirements are well defined at the present time and will undergo relatively minor change in the future. Approaches to stoichiometric limits and ideal cycle efficiencies restrict further large improvements. Future reduction consideration of unit product cost for the liquid hydrogen due to raw material and energy at that time will be dependent upon the unit costs for these commodities. For the hydrocarbon and electrolysis-based plants, raw material and energy costs typically represent from 55 to 60 percent of the unit product cost. Future commodity costs would not be lowered by more than 10 to 15 percent. In many cases, respective commodity costs could in fact be greater than projected. Assuming a 10 percent decrease in unit commodity costs, the product cost for New York City (partial oxidation based plant) would be 7.7 cents/lb LH_2 reduced from 8.1 cents/lb LH_2. Reduced commodity costs obviously will affect the unit product costs. Likely cost reduction in capital investment charges appears to be more attractive with respect to decreasing product costs.

Operating costs are the smallest component of unit product cost, typically being 8 to 13 percent of the unit product cost for hydrocarbon based and electrolysis based plants. Large changes in future analyses of the operating costs would be required to significantly affect unit product cost. Referring again to New York City (partial oxidation based plant), reduction of operating costs by 20 percent would result in a reduction of unit product cost of 7.9 cents/lb LH_2 from 8.1 cents/lb LH_2.

In addition to the cost reduction information stated, consideration of future integrated facilities could result in cost reductions beyond the 10 to 20 percent used in this report. No fixed understanding of effect can be developed without a specific detailed study. One major manufacturing firm currently is operating an integrated cryogenic-chemical facility. It has been judged that facility integration resulted in a 15 percent lower unit product cost for liquid hydrogen, compared to the "stand alone" plant.

Future large scale liquid production could involve plant bases significantly different from those used in this study. It could, for example, prove to be advantageous to have fewer production points and transport the product to use points. An example of this would be producing liquid hydrogen in the Venezuelan gas fields and transporting it to New York, Sao Paulo, etc., by massive transoceanic shipping. Natural gas fields in Alaska and other raw material concentration points could be similarly developed.

6.3 FUTURE TECHNOLOGY

Future technological development will be directed toward capital investment reduction in respective systems. Product cost is very sensitive to capital value of systems comprising a given plant. Development of construction material, for example, will hopefully lead to lower cost cells employed in electrolysis processes.

Characteristics of the various processes will also receive attention, but further optimization will be in the nature of incremental steps as opposed to a sweeping revision. Machinery efficiencies, for example, may be improved from 70 to 80 percent through future developmental work. Of more significance, however, would be the cost reduction of the capital goods in the process using the more efficient machine.

It is expected that much effort will be expended to develop coal as a raw material for hydrogen production. Another fertile area for process development is that of several products, including gaseous hydrogen, being issued from integrated process plants. As an example, generation of synthesis gas (nitrogen and hydrogen) is used for production of fertilizer. Presumably "sharing of costs" would reduce the burden allocated to the gaseous hydrogen production. Development work relating to fertilizer, incorporating nuclear energy and water processing, has been undertaken within the past two years. With this understanding, in future years, certain entirely new processes may appear.

Appendix A

FOSSIL FUELS

A.1 The tables contained in this appendix list world-wide summaries of resources. The following is an explanation of letter designations that are applicable to the tables contained in this appendix.

A - Approximate

C - Recorded, but confidential

D - Recorded, but not yet available

E - Estimated

N - Negligible

U - Unavailable, not recorded

X - Number refers to proved reserves only

Z - This estimate includes oil in shale and bituminous rock, other figures of oil reserves do not.

a through e - References are defined at the end of each of the tables.

TABLE A-1. FOSSIL FUEL NATURAL RESOURCES (WORLD WIDE)

Reference	Country or Area	Coal (Metric Tons x 10^6)		Oil (Metric Tons x 10^6)		Natural Gas (Cubic Meters x 10^6)	
		Available for Measurement	Estimated Actual Total	Available for Measurement	Estimated Actual Total	Available for Measurement	Estimated Actual Total
b	<u>NORTH AMERICA,</u> (excl. Central America)		798,000				
a	United States	72,000	1,100,000	14,700	76,600	7,468,900	42,500,000
b	United States		753,000	3,275	26,700		
c	United States	1,470,000	4,070,000	14,530	104,470	7,600,000	58,000,000
d	United States		860,000				
e	United States	1,470,000	4,070,000	49,500	104,400	7,600,000	58,000,000
b	Remainder of North America			183	6,900		
a	Canada	42,000	60,000	530X		864,000	
b	Canada		43,000				
	<u>CENTRAL AMERICA</u>						
a	Mexico		2,600	350X		230,000	
b	Mexico		2,000				
a	Honduras		1				
	<u>WEST INDIES</u>						
a	Trinidad & Tobaga			301	U	24,459	U
a	Cuba			N X			

TABLE A-1. FOSSIL FUEL NATURAL RESOURCES (WORLD WIDE) (Continued)

Reference	Country or Area	Coal (Metric Tons x 10^6)		Oil (Metric Tons x 10^6)		Natural Gas (Cubic Meters x 10^6)	
		Available for Measurement	Estimated Actual Total	Available for Measurement	Estimated Actual Total	Available for Measurement	Estimated Actual Total
b	<u>SOUTH AMERICA,</u>				12,200		
b	(incl. Central America)		10,000				
a	Brazil	388	1,700	48X		10,762	
a	Venezuela	29.6	342	2,485	U	952,630	U
b	Venezuela		2,000	1,471	13,660		
a	Colombia		12,500	105X			
b	Colombia		6,000				
b	Others(excl. Venezuela & Colombia)		2,000	1,471	13,660		
a	Argentina	374	454	1,830	2,190	210,000	237,440
a	Chile	96	180	50	250	80,000	300,000
a	Peru		400	53X			
a	Bolivia			15X			
a	Ecuador			3.7X			
b	<u>WESTERN HEMISPHERE</u>			4,930	40,000		

TABLE A-1. FOSSIL FUEL NATURAL RESOURCES (WORLD WIDE) (Continued)

Reference	Country or Area	Coal (Metric Tons x 10^6)		Oil (Metric Tons x 10^6)		Natural Gas (Cubic Meters x 10^6)	
		Available for Measurement	Estimated Actual Total	Available for Measurement	Estimated Actual Total	Available for Measurement	Estimated Actual Total
b	EUROPE, (excl. U.S.S.R.)		301,000	99	2,080		
a	France		4,400	30	U	200,000	
b	France		6,000				
a	Belgium	1,557	3,425				
b	Belgium		3,000				
a	Netherlands	2,394	2,394	C	C	C	C
b	Netherlands		2,000				
a	West Germany	70,200	229,900	69	88	15,000A	30,000
b	West Germany		143,000				
a	Italy	675	700		130		120,000
a	United Kingdom	127,079	169,379	N	N	N	N
b	United Kingdom		85,000			U	U
a	Ireland	13.5	81.5				
a	Norway	24	1,800A				
a	Sweden	60	90				
a	Denmark					N	N
a	Switzerland	N	N				

TABLE A-1. FOSSIL FUEL NATURAL RESOURCES (WORLD WIDE) (Continued)

Reference	Country or Area	Coal (Metric Tons x 10^6)		Oil (Metric Tons x 10^6)		Natural Gas (Cubic Meters x 10^6)	
		Available for Measurement	Estimated Actual Total	Available for Measurement	Estimated Actual Total	Available for Measurement	Estimated Actual Total
	EUROPE – (Cont'd)						
a	Austria	2	3	33.9X	33.9	23,080	23,080
a	Spain	2,784	2,957			U	U
a	Portugal	18	29	U	U	U	U
a	Greece			N	U	80	U
a	U.S.S.R.	143,688	4,630,050	4,400X		2,000,000	23,700,000
b	U.S.S.R.			762	30,500		
a	East Germany			NX			
a	Hungary	210	210	18X			
a	Bulgaria		33	2.1X			
a	Poland	75,500	135,000				
b	Poland		40,000	3.4X			
a	Czechoslovakia		6,450				
b	Czechoslovakia		10,000	1.7X			
a	Yugoslavia	22	237	27X		4,000	

TABLE A-1. FOSSIL FUEL NATURAL RESOURCES (WORLD WIDE) (Continued)

Reference	Country or Area	Coal		Oil		Natural Gas	
		Available for Measurement (Metric Tons x 10^6)	Estimated Actual Total (Metric Tons x 10^6)	Available for Measurement (Metric Tons x 10^6)	Estimated Actual Total (Metric Tons x 10^6)	Available for Measurement (Cubic Meters x 10^6)	Estimated Actual Total (Cubic Meters x 10^6)
	EUROPE – (Cont'd)						
a	Roumania	78	U	130X			
a	Albania			11X			
b	Other Countries		12,000				
b	MIDDLE EAST, (incl. N. Africa)			4,390	45,700		
a	Turkey	489	1,298	75A	C		
	Syria			14X			
	Israel			4.2E	U		
a	Iraq			3,400X		84,000	
a	Iran			4,700X		88,000	
a	Saudi Arabia			7,000X		260,000	
a	Kuwait			8,400X		280,000	
a	Bahrein			32X			
	Neutral Zone			880X			
	Qatar			370X			

TABLE A-1. FOSSIL FUEL NATURAL RESOURCES (WORLD WIDE) (Continued)

Reference	Country or Area	Coal Available for Measurement (Metric Tons x 10^6)	Coal Estimated Actual Total (Metric Tons x 10^6)	Oil Available for Measurement (Metric Tons x 10^6)	Oil Estimated Actual Total (Metric Tons x 10^6)	Natural Gas Available for Measurement (Cubic Meters x 10^6)	Natural Gas Estimated Actual Total (Cubic Meters x 10^6)
	<u>MIDDLE EAST</u> – (Cont'd)						
a	Southern Arabia			480X			
a	United Arab Rep.			510	U		
b	<u>ASIA</u>, (incl. U.S.S.R.)		1,147,000				
b	Asia - Non-Communist		41,000	15	3,670		
a	Afghanistan						
a	Pakistan	165	365	5.5	U	579,420	634,510
a	India	U	57,832				
b	India		32,000	2	4.5	500	800
a	Burma			6.4X			
a	Thailand			N			
a	Cambodia, Laos, Vietnam	200	1,000				
a	Japan	5,723	19,248	3.9	7.2	142,496	585,497
b	Japan		5,000				
a	Korea		833				
a	Taiwan			N			

TABLE A-1. FOSSIL FUEL NATURAL RESOURCES (WORLD WIDE)(Continued)

Reference	Country or Area	Coal Available for Measurement (Metric Tons x 10^6)	Coal Estimated Actual Total (Metric Tons x 10^6)	Oil Available for Measurement (Metric Tons x 10^6)	Oil Estimated Actual Total (Metric Tons x 10^6)	Natural Gas Available for Measurement (Cubic Meters x 10^6)	Natural Gas Estimated Actual Total (Cubic Meters x 10^6)
b	<u>AFRICA</u>		35,000	7.5	1,230		
a	Union of South Africa	21,443	63,355				
b	Union of South Africa		34,000				
b	Remainder of Africa		1,000				
a	Algeria	9	20	N	U		
a	Basutoland	N	N				
a	Bechuanaland	506	U				
a	Congo, Rep. of		100	N			
a	Madagascar	100	300				
a	Morocco	14.8	95.8	2	7	800	800
a	Nigeria	U	U	45	315	75,000	375,000
a	Rhodesia & Nyassaland	1,760	6,613				
a	Swaziland	2,022	5,022				
a	Tanganykia	300	400				
a	Angola			4.1X			

TABLE A-1. FOSSIL FUEL NATURAL RESOURCES (WORLD WIDE) (Continued)

Reference	Country or Area	Coal		Oil		Natural Gas	
		Available for Measurement (Metric Tons x 10^6)	Estimated Actual Total (Metric Tons x 10^6)	Available for Measurement (Metric Tons x 10^6)	Estimated Actual Total (Metric Tons x 10^6)	Available for Measurement (Cubic Meters x 10^6)	Estimated Actual Total (Cubic Meters x 10^6)
	AFRICA – (Cont'd)						
a	Gabon			21X			
a	Ghana			N	N		
a	Libya			410X			
a	French Sahara			600	U	1,200,000	
b	EASTERN HEMISPHERE			5,400	63,370		
	OCEANIA						
b	(Australasia)		29,000				
a	Australia	1,791	12,844		N		300
b	Australia		29.000		500		
a	New Zealand	45.5	144	U	U	U	U
	Papua - New Guinea			NX			
a	Indonesia	U	U	1,200X		U	U
b	Indonesia			168	4,740		
a	Philippines	2.7	35.4	N			
a	North Bornea	4.9	23.3				

TABLE A-1. FOSSIL FUEL NATURAL RESOURCES (WORLD WIDE) (Concluded)

Reference	Country or Area	Coal (Metric Tons x 10^6) Available for Measurement	Coal (Metric Tons x 10^6) Estimated Actual Total	Oil (Metric Tons x 10^6) Available for Measurement	Oil (Metric Tons x 10^6) Estimated Actual Total	Natural Gas (Cubic Meters x 10^6) Available for Measurement	Natural Gas (Cubic Meters x 10^6) Estimated Actual Total
	OCEANIA – (Cont'd)						
a	Sarawak	0.3	4.8	96X		N	N
	New Caledonia	5	15				
	COMMUNIST COUNTRIES						
a	U.S.S.R.	143,688	4,630,050	3,300X		2,000,000	23,700,000
b	U.S.S.R.		600,000	762	23,630		
a	China		1,011,000	70X			
b	China		506,000				
b	WORLD			10,370	190,500		212,000,000
c	World	771,000	13,744,000	45,700	518,000	51,000,000	589,000,000
c	World			81,600Z	2,896,000Z		

References:

a. Survey of Energy Resources, World Power Conference, 1962
b. Energy Resources, Publication 1000-D, National Research Council, 1962
c. Energy R&D and National Progress, U.S. Office of Science & Technology, 1964, p. 105
d. "Coal", National Coal Association, 1966
e. "Supplies, Costs, and Uses of the Fossil Fuels", U.S. Dept. of the Interior, 1963

TABLE A-2. WATER POWER RESOURCES IN COUNTRIES OF TENTATIVE PLANT LOCATIONS

		Estimated Gross Theoretical Potential							
Ref	Country or Area	Q95[1] MW	Q50[2] MW	Arith Mean[3] MW	Total Capacity MW	Corresponding Average Gen Kwh x 10^6	MW Installed	% of Total Power	10 yr Growth Ratio[4]
a	United States	28,955	63,534	98,221	147,600	643,300	-	-	-
b	United States	34,000	-	121,346	-	-	38,600	17	1.66
c	United States	-	-	-	-	-	42,899	17.8	-
b	Middle Atlantic	3,331	5,867	7,357	-	-	4,407	-	-
b	California,USA	2,120	6,564	13,073	-	-	4,624	-	-
b	Hawaii, USA	67	186	280	-	-	34	-	-
a	Brazil	16,457	U	U	-	-	-	-	-
b	Brazil	15,000	-	180,000	-	-	3,850	80	1.8
c	Brazil	-	-	-	-	-	5,219	73.2	
a	France	U	U	7,700	U	67,600	-	-	-
b	France	4,000	12,000	-	-	-	10,900	47	1.57
c	France	-	-	-	-	-	12,312	48.2	-
c	Lebanon	-	-	-	-	-	126	45.0	-
a	South Africa	U	U	U	U	U	-	-	-
b	South Africa	335	-	10,000	-	-	25	U	U
c	South Africa	-	-	-	-	-	5	0.08	-
a	Japan	9,420	D	17,470	35,370	130,090	-	-	-
b	Japan	9,000	-	18,750	-	-	14,000	53	2.04

TABLE A-2. WATER POWER RESOURCES IN COUNTRIES OF TENTATIVE PLANT LOCATIONS (Concluded)

		Estimated Gross Theoretical Potential							
Ref	Country or Area	Q95[1] MW	Q50[2] MW	Arith Mean[3] MW	Total Capacity MW	Corresponding Average Gen Kwh x 10^6	MW Installed	% of Total Power	10 yr Growth Ratio[4]
a	Australia	1,224	2,928	3,915	7,555	23,200	-	-	-
b	Australia	750	28,500	-	-	-	2,000	2.5	4.84
c	Australia	-	-	-	-	-	2,052	25.8	-
a	Thailand	U	U	1,160	-	-	-	-	-
b	Thailand	3,000	-	15,000	-	-	30	U	U
c	Thailand	-	-	-	-	-	140	25.5	-
b	World	-	-	-	2,724,044	-	180,938	33	2.04
c	World	-	-	-	-	-	210,001	28.6	-

References:

a Survey of Energy, World Power Conference, 1962

b Young, L. L., Summary of Developed and Potential Waterpower of the U.S. and Other Countries of the World, 1955-62, Geological Survey 483, Washington, 1964

c World Power Data, 1964, Federal Power Commission, May 1966

Notes:

[1] Q95 is the natural or modified flow available 95 percent of the time.
[2] Q50 is the natural or modified flow available 50 percent of the time.
[3] Arith Mean is the arithmetical mean flow as measured over a period of time.
[4] Normal 10 yr growth ratio is considered to be 2.00, a doubling of hydroelectric capacity each 10 years.

TABLE A-3. WORLD RESERVES - OXIDE OF URANIUM

Ref	Country or Area	Measured Reserves: More Than 0.1% U308	Measured Reserves: 0.01 to 0.1% U308	Total Reserves	Minimum Or Known	Maximum
a	Canada	D	D	325,000		
b	Canada				450,000	3,000,000
c	Canada				236,000	
a	United States		U	210,000		
b	United States				700,000	3,000,000
c	United States				323,000	
b	Europe & Asia				480,000	18,000,000
c	Europe & Asia				110,000	400,000
a	France	15,000	U	26,947		
c	France				34,000	
a	West Germany	D	D	300		
a	Greece	C	C	C		
a	Italy	1,000	U	4,000		
a	Portugal	5,200	170	U		
a	Spain	552	600	2,662		
a	Sweden	N	9,000	1,100,000		
a	India	8,228	2,908	25,554		
a	Japan	97	229	1,722		

TABLE A-3. WORLD RESERVES - OXIDE OF URANIUM (Concluded)

Ref	Country or Area	Measured Reserves		Total Reserves	Minimum Or Known	Maximum
		More Than 0.1% U308	0.01 to 0.1% U308			
a	Australia	12,200	1,150	15,700		
b	Australia				8,000	3,000,000
c	Australia				13,700	
b	Africa				2,000	8,000,000
a	Angola	U	U	15,000		
a	Congo	7,500				
a	Gabon	6,100	U	U		
a	Madagascar	658	U	U		
a	South Africa	1,190	21,780	212,570		
c	South Africa				127,000	
b	World				1,840,000	45,000,000
c	World				860,000	1,150,000

References:

a Survey of Energy Resources, World Power Conference, 1962

b Energy Resources, Publication 1000-D, National Research Council, 1962

c Energy R&D and National Progress, U.S. Office of Science & Technology, 1964

Appendix B

PROJECTED POWER AVAILABILITY

B.1 REPRESENTATION

The figures contained in this appendix are referenced in Section 4, paragraph 4.5.3.

B.2 INSTALLED CAPACITY

The installed capacity for each location referenced on the figures is as follows:

(1) Actual Installed Capacity 1948 to 1964

(2) Constant Annual Rate of Growth (1955 to 1964 Average)

(3) Constant Annual Incremental Growth (1959 to 1964 Average)

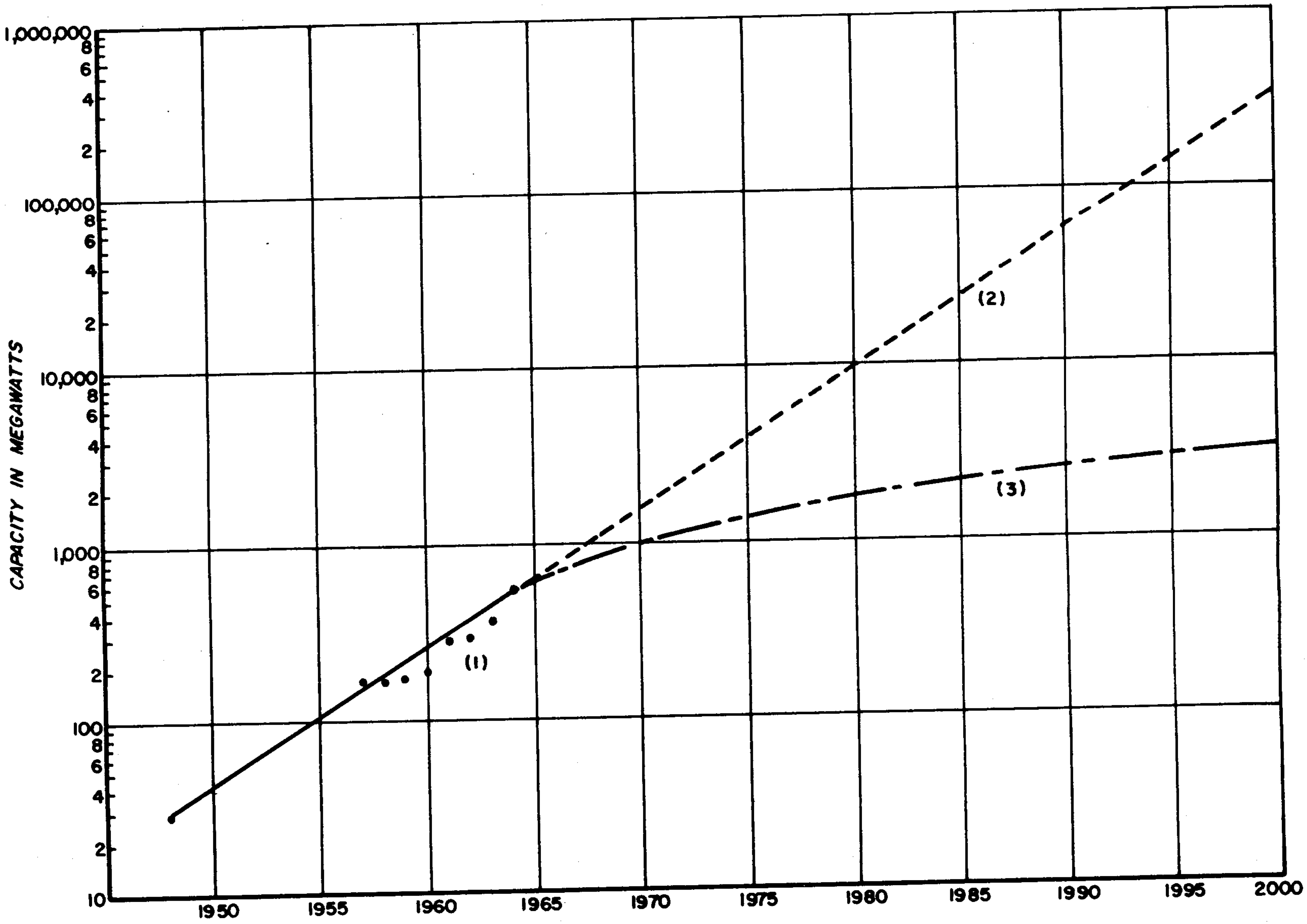

FIGURE B-1. ELECTRIC ENERGY PROJECTIONS - THAILAND

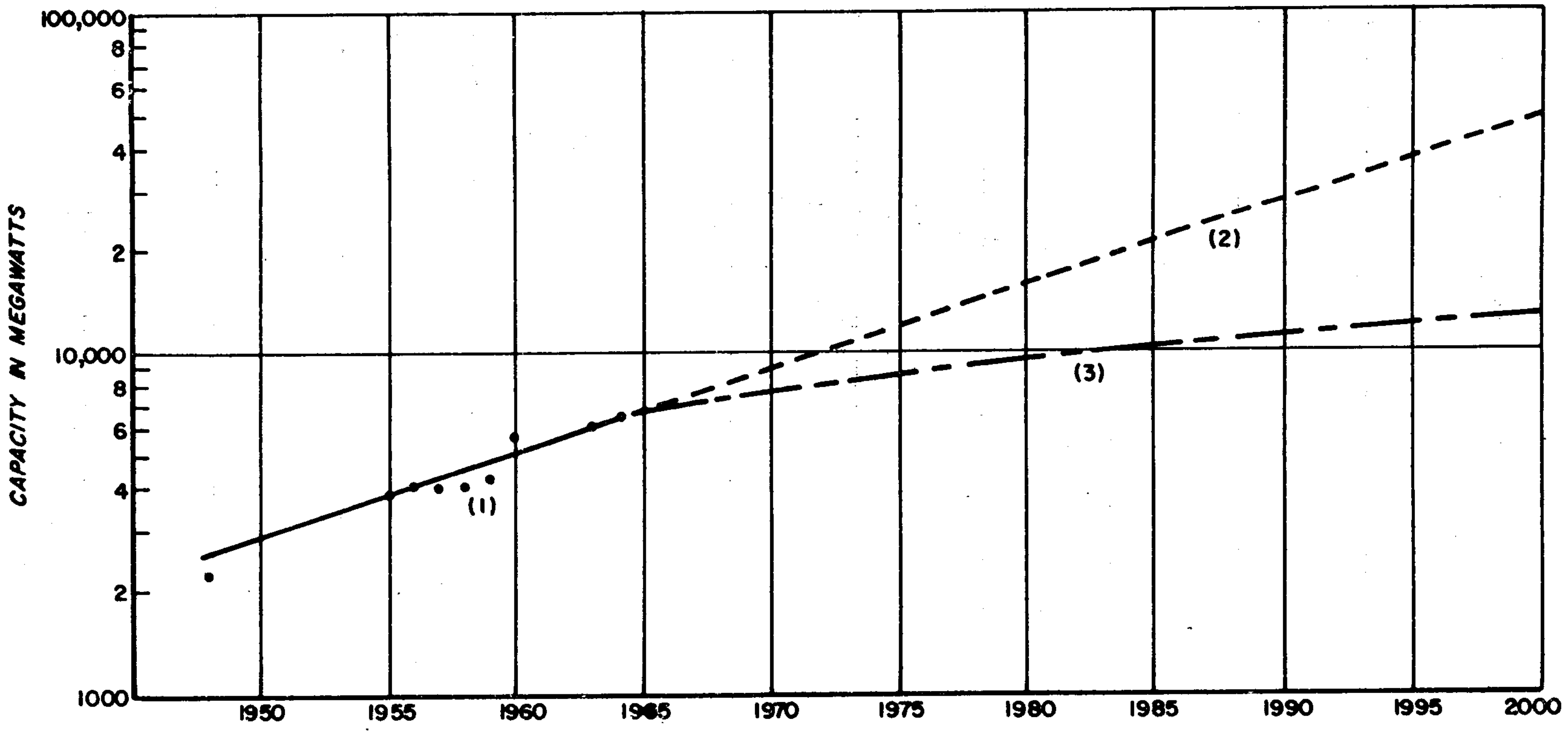

FIGURE B-2. ELECTRIC ENERGY PROJECTIONS - SOUTH AFRICA

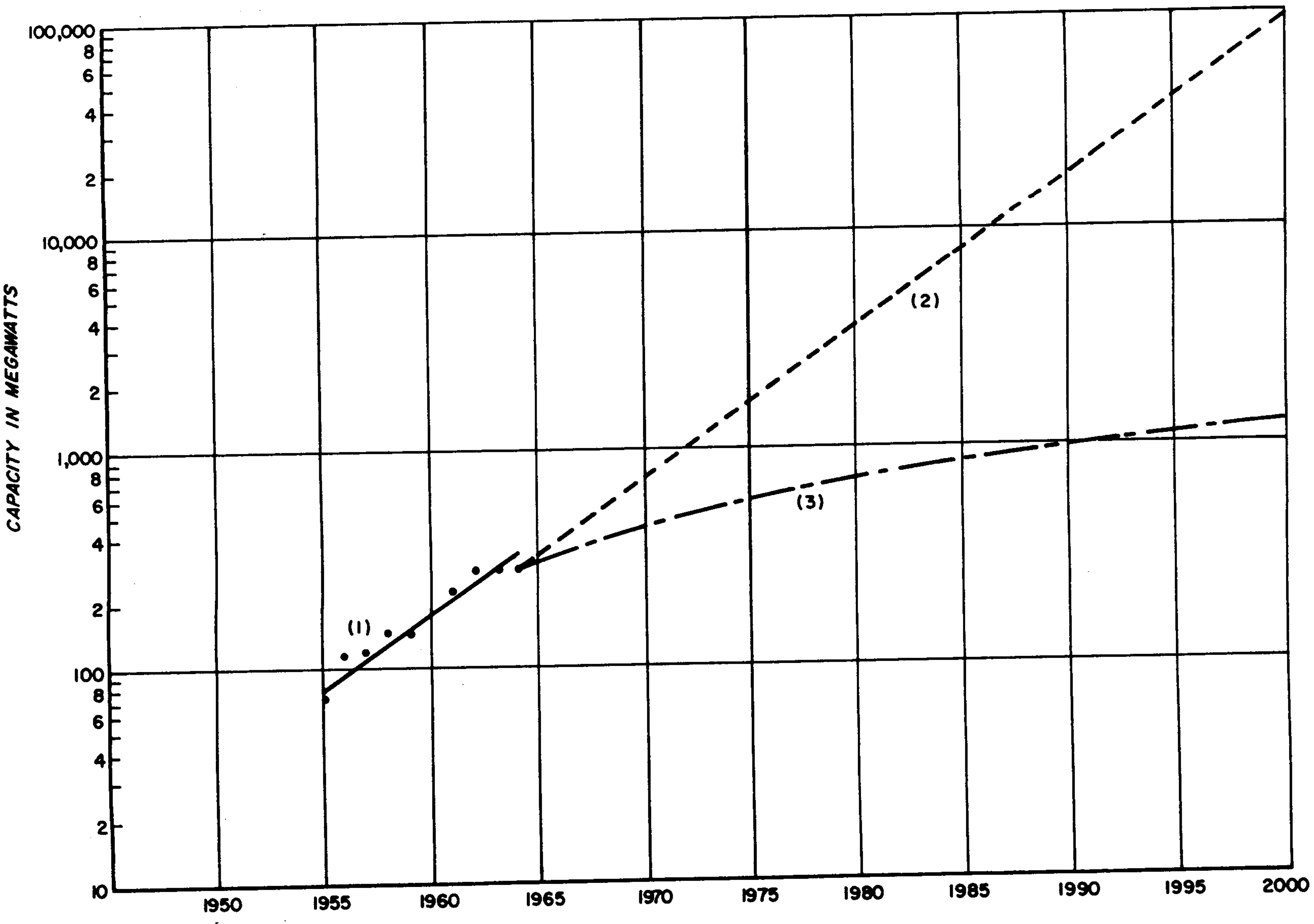

FIGURE B-3. ELECTRIC ENERGY PROJECTIONS - LEBANON

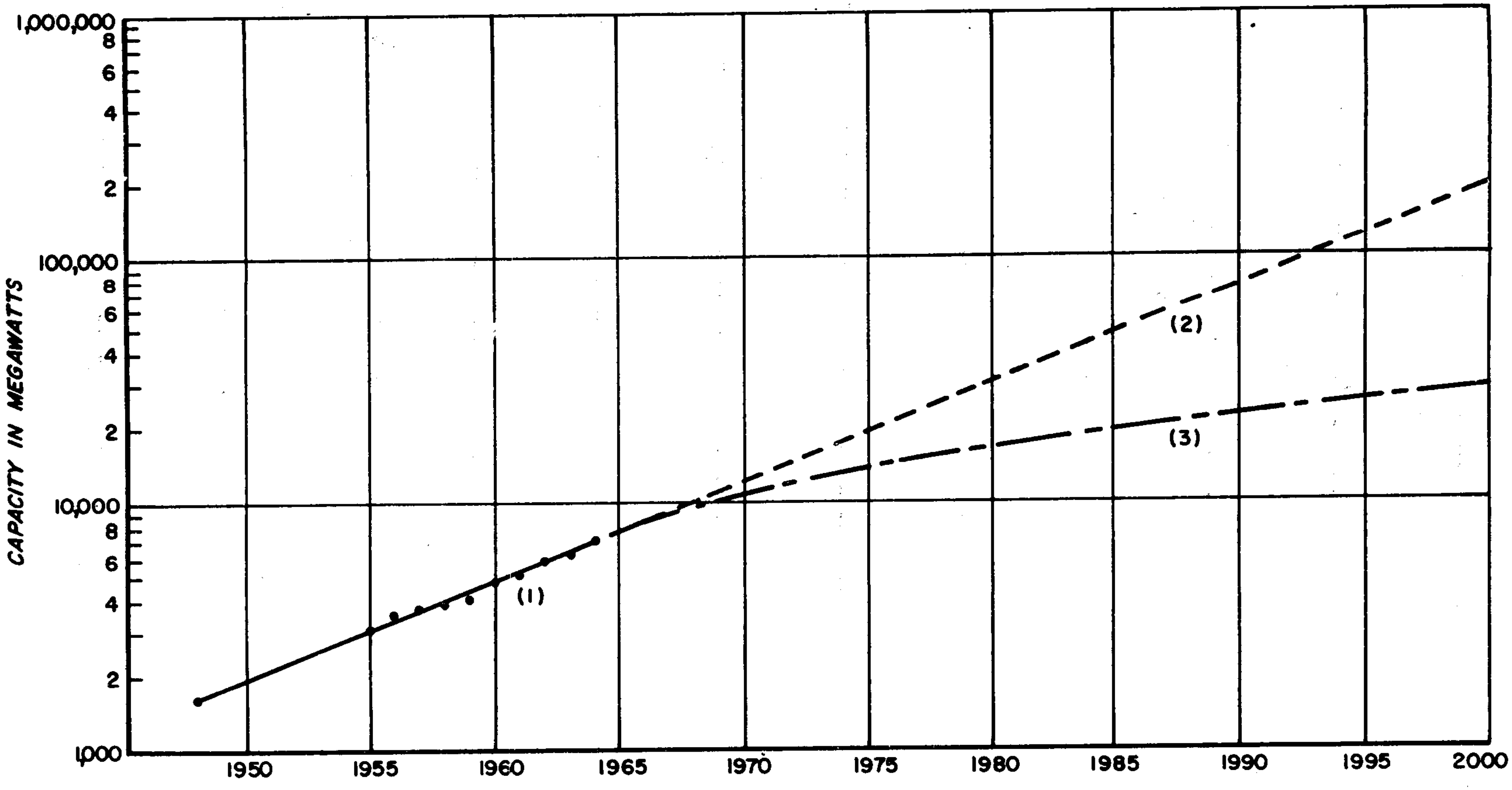

FIGURE B-4. ELECTRIC ENERGY PROJECTIONS - BRAZIL

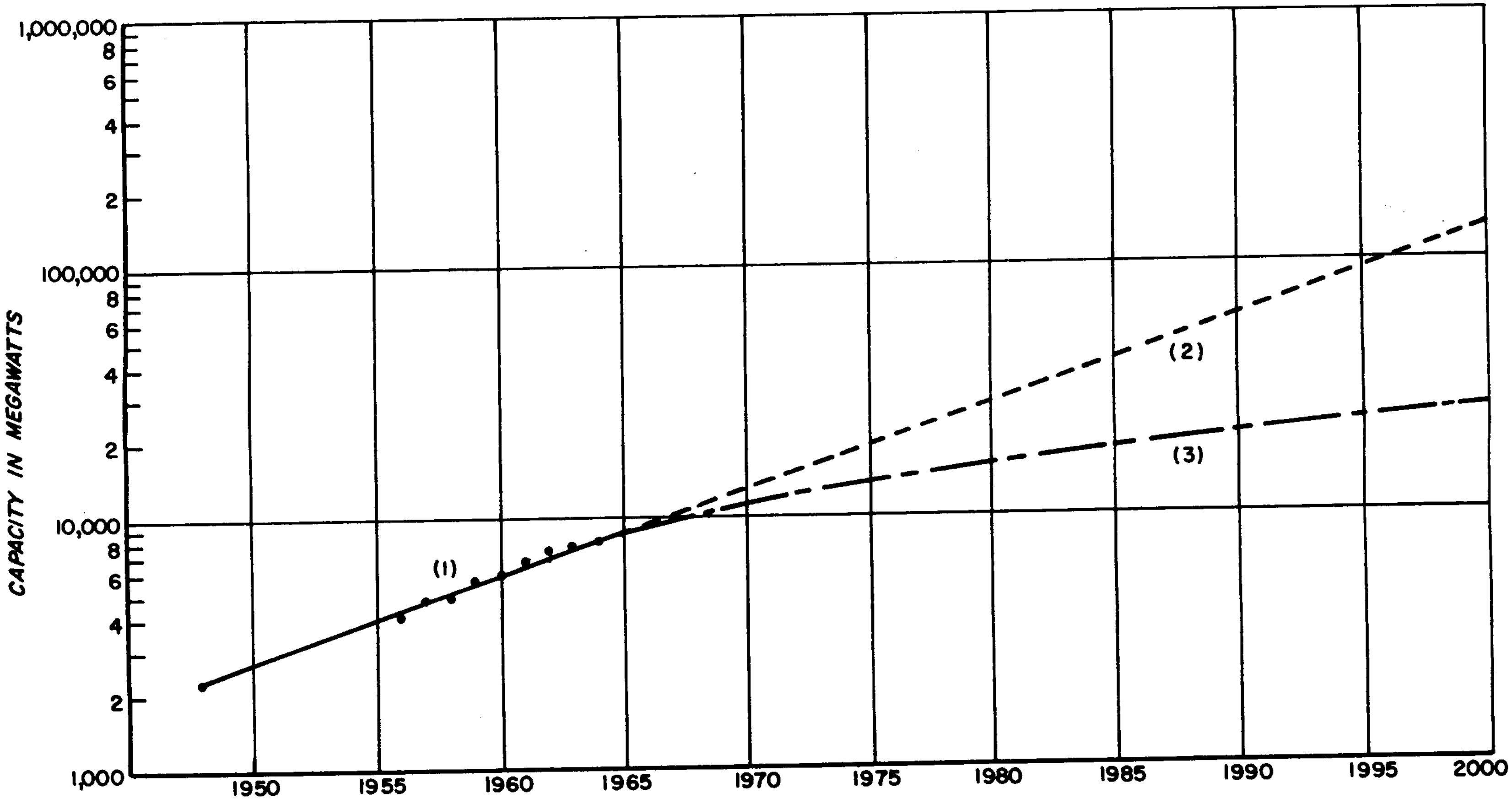

FIGURE B-5. ELECTRIC ENERGY PROJECTIONS - AUSTRALIA

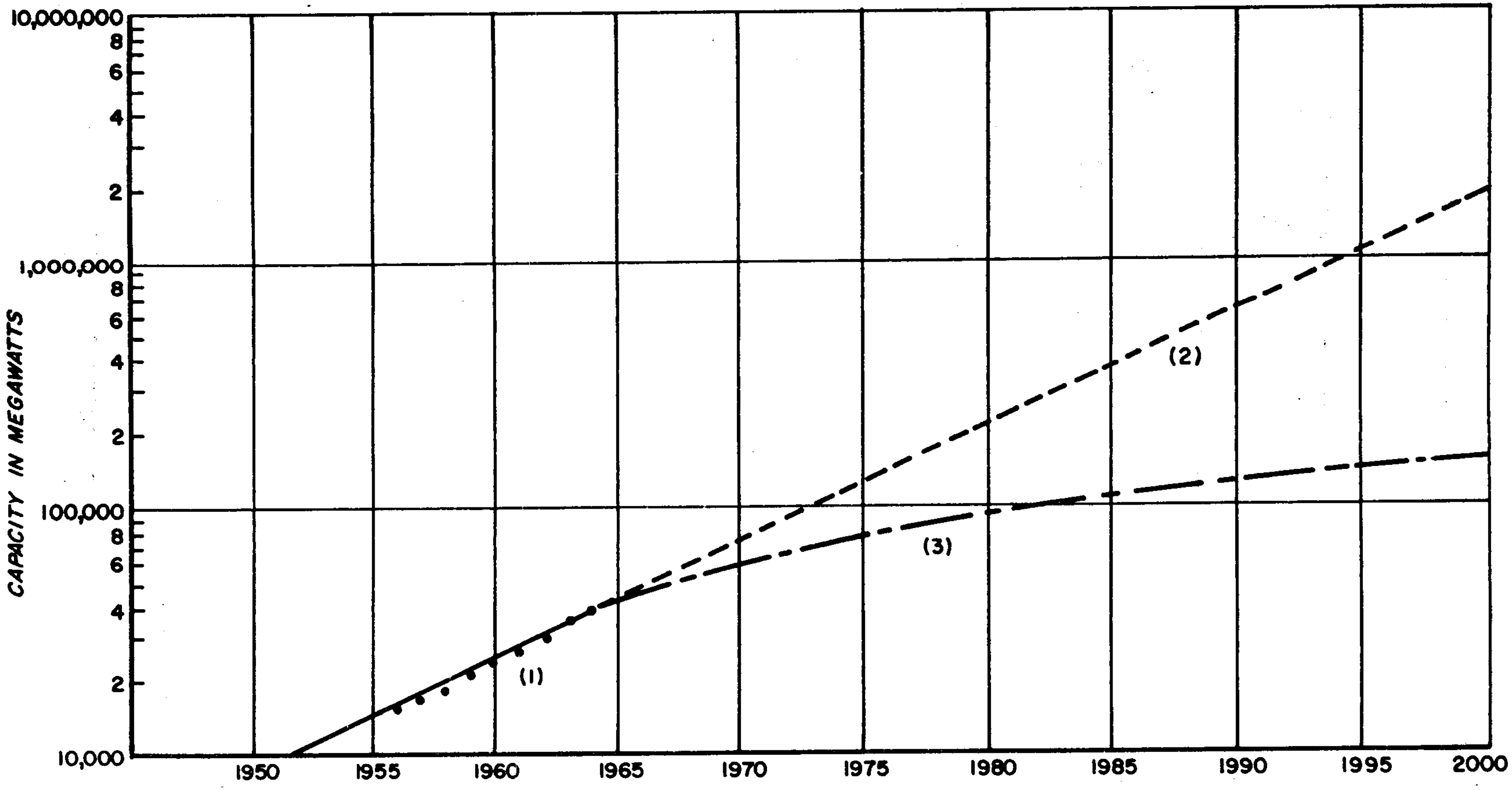

FIGURE B-6. ELECTRIC ENERGY PROJECTIONS - JAPAN

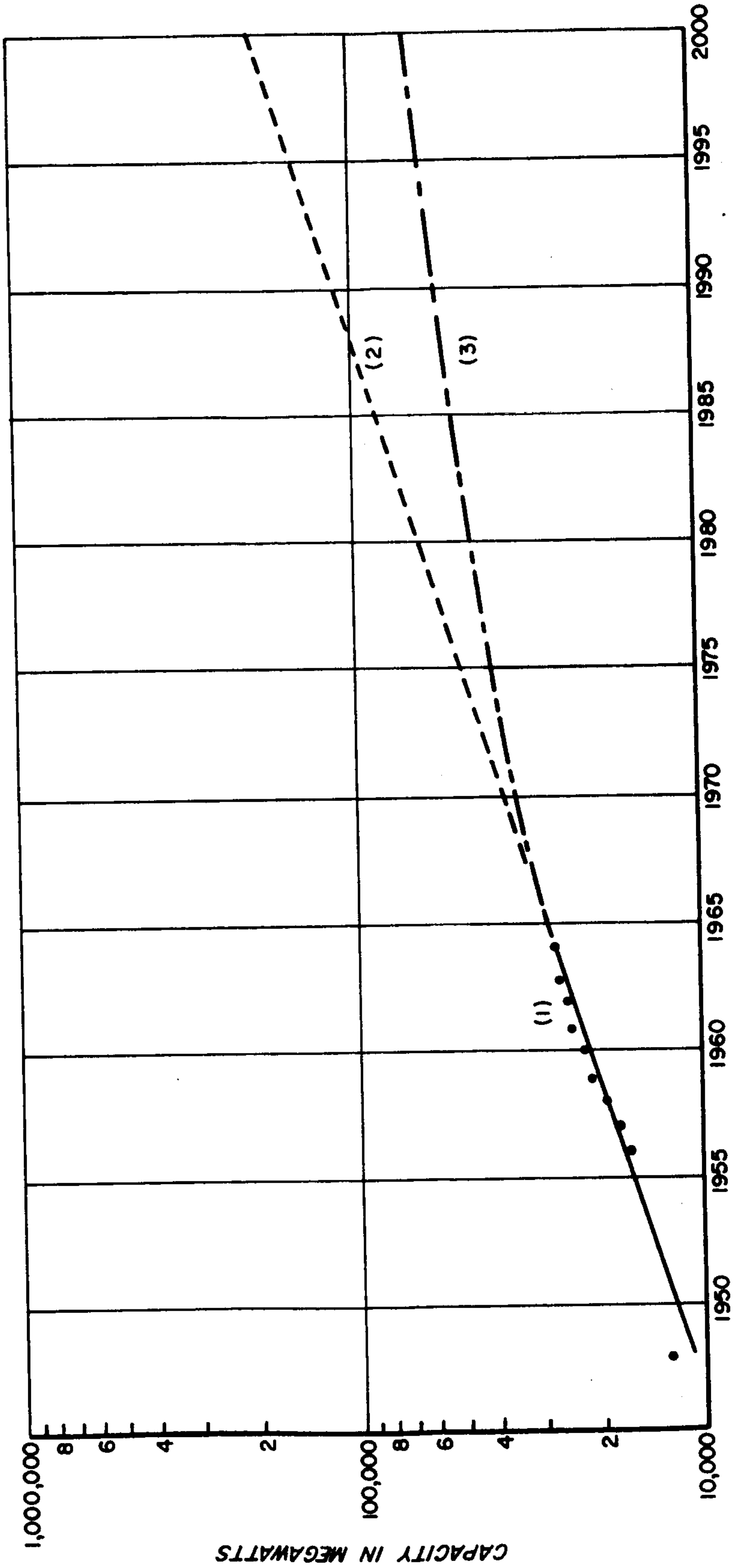

FIGURE B-7. ELECTRIC ENERGY PROJECTIONS - FRANCE

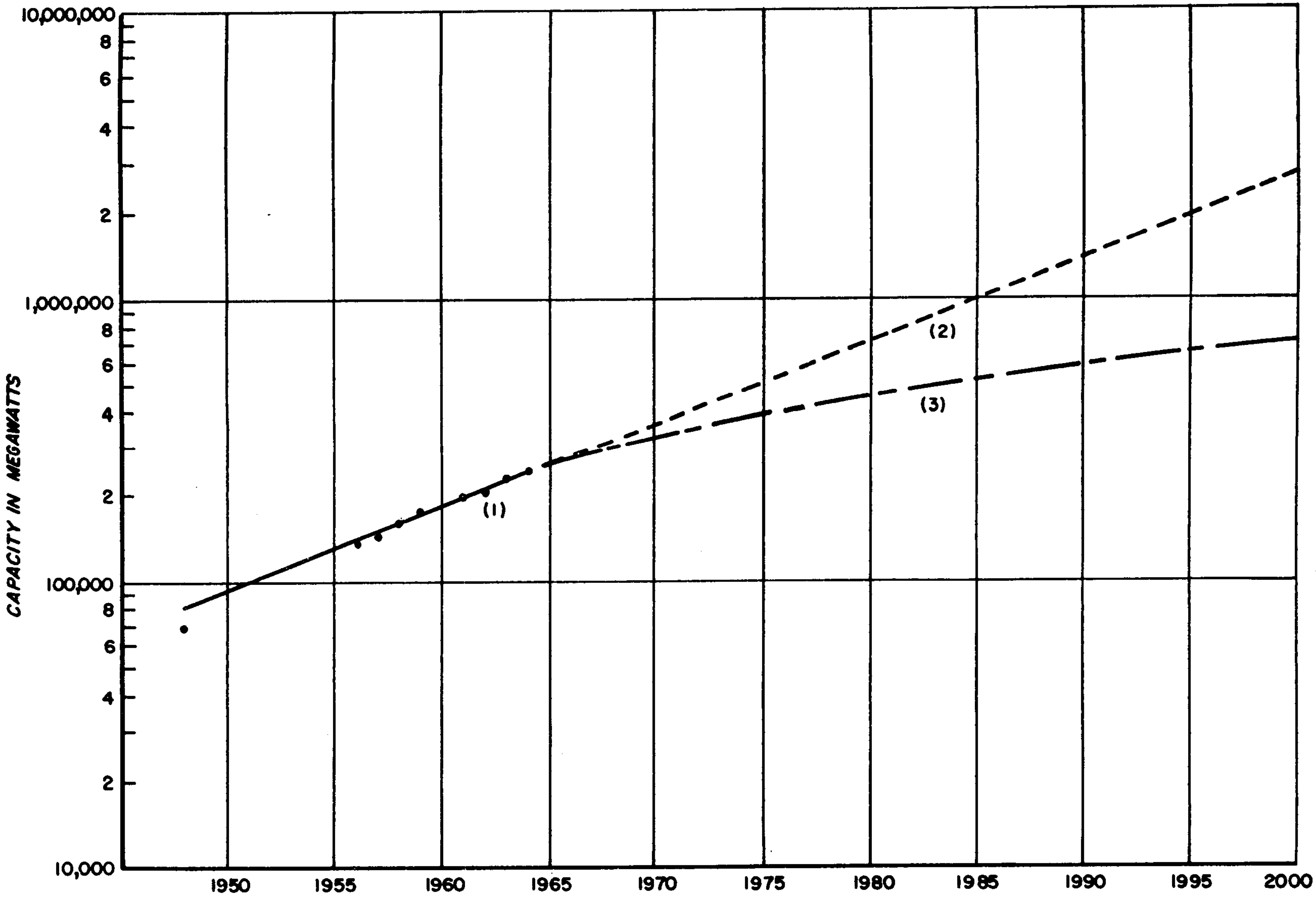

FIGURE B-8. ELECTRIC ENERGY PROJECTIONS - UNITED STATES

Appendix C

ESTIMATED COST OF ELECTRIC POWER

C.1 BANGKOK

The cost of power delivered to the proposed 250 T/D liquid hydrogen plant at the air terminal within the city of Bangkok can be optimized by the purchase of power from the proposed hydroelectric dam at Pa Mong, a distance of 350 miles north on the Thailand-Laotian border. Total installed capacity of this project is estimated at 1500 mW by 1980. Costs to transmit the power to the use-point have been extrapolated as incremental costs using curves developed on AC transmission for 500 and 700 KVA at 50 percent load factor.[92] Results are summarized as follows for use in section 5.

Process	mW Plant Size	Plant Type	Busbar (Mills)	Transmission Mills per Kwh	Total (Mills)
Hydrocarbon	600	Hydroelectric	3.11	0.50	3.61
Electrolysis	1500	Hydroelectric	2.88	0.35	3.23

A graphical presentation of power cost for the various plant types and sizes are shown in figure C-1.

C.1.1 ELECTRIC PLANT INVESTMENT (Paragraph 4.8)

Type	150 mW $/Kw	300 mW $/Kw	600 mW $/Kw
Oil	185	136	117
Nuclear	313	265	216
LNG	173	129	110
Coal	213	164	138
Hydro	567	355	273

C.1.2 ELECTRIC PLANT FIXED COSTS (Paragraph 4.9)

Type	Fixed Charge Rate %/Year	150 mW Mills/Kwh	300 MW Mills/Kwh	600mW Mills/Kwh
Oil	11.0	2.42	1.78	1.52
Nuclear	11.25	4.19	3.55	2.89
LNG	11.0	2.26	1.69	1.44
Coal	11.0	2.79	2.15	1.80
Hydro	8.5	5.73	3.59	2.76

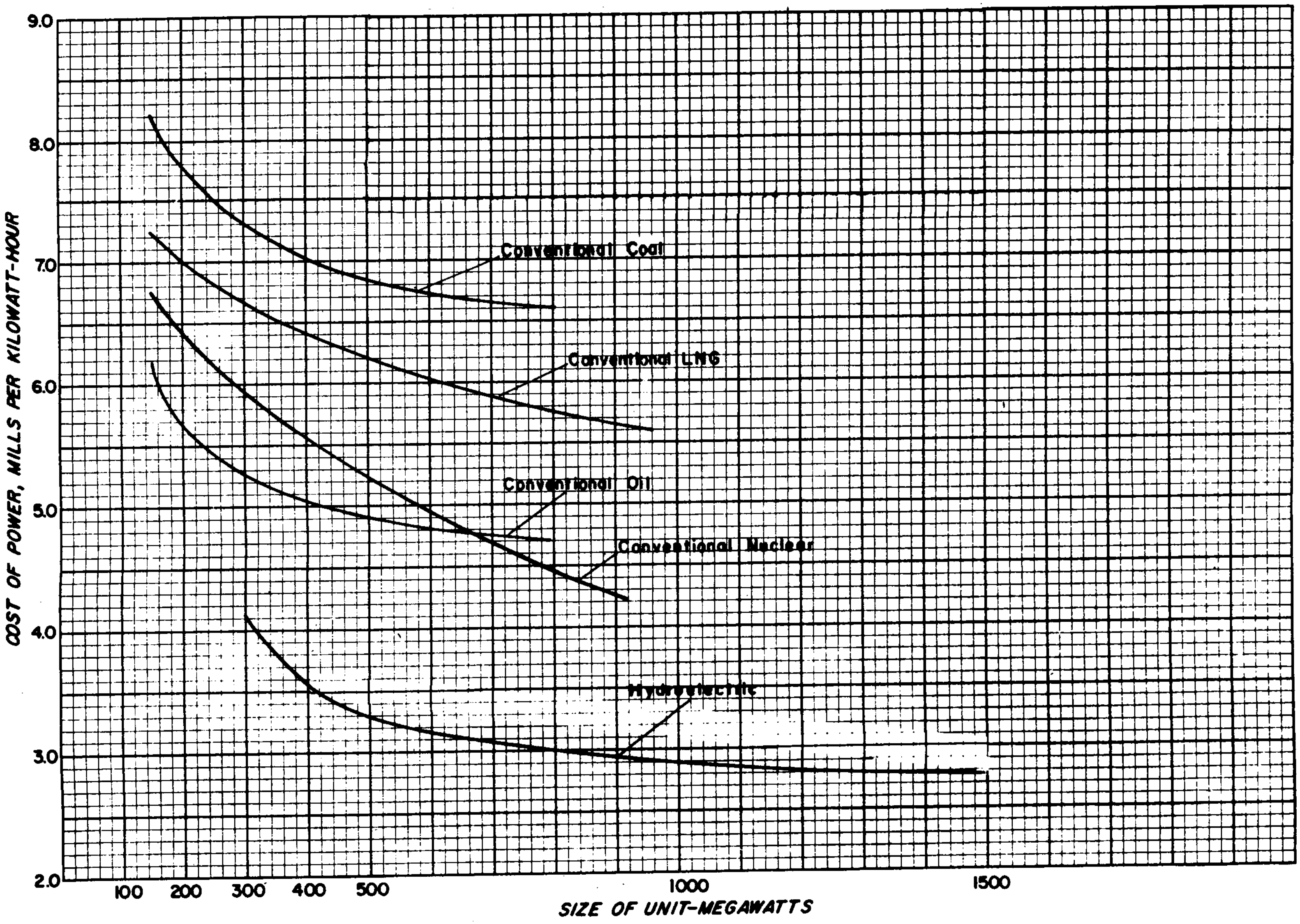

FIGURE C-1. RELATIONSHIP OF POWER COST TO PLANT SIZE (BANGKOK)

C.1.3 ELECTRIC PLANT OPERATING AND MAINTENANCE COSTS (Paragraph 4.9.4)

Type	150 mW Mills/Kwh	300 mW Mills/Kwh	600 mW Mills/Kwh
Oil	0.72	0.58	0.38
Nuclear	0.65	0.53	0.35
LNG	0.69	0.55	0.37
Coal	0.78	0.64	0.44
Hydro	0.65	0.53	0.35

C.1.4 ELECTRIC ENERGY BUSBAR COST

Type	150 mW Mills/Kwh	300 mW Mills/Kwh	600 mW Mills/Kwh
Oil	6.12	5.34	4.88
Nuclear	6.72	5.88	4.89
LNG	7.50	6.79	6.36
Coal	8.12	6.99	6.44
Hydro	6.38	4.12	3.11

C.1.5 ELECTRIC PLANT FUEL COSTS (Paragraph 4.9.5)

	150 mW		300 mW		600 mW	
Type	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh
Oil	34	2.98	34	2.98	34	2.98
Nuclear	17.9	1.88	17.1	1.80	15.7	1.65
LNG	52	4.55	52	4.55	52	4.55
Coal	48	4.20	48	4.20	48	4.20
Hydro	–	–	–	–	–	–

C.2 HONOLULU

An oil-fired power plant located at the LH_2 plant - air terminal is the optimum power-generating method for low-cost power for the hydrocarbon processes in Honolulu. Purchase of power at the busbar with no transmission is presumed. In the electrolytic process, a 600-mW FBR unit at the site is assumed. This disregards established methods in determination of plant size. However, the large load associated with electrolysis makes such a size possible. Reliability is considered the same as at other locations, although obviously the single large plant integrated in the Honolulu system carries more risk of interruptions in power supply. Results are summarized as follows for use in section 5.

Process	**mW Plant Size**	**Plant Type**	**Busbar Mills per Kwh**
Hydrocarbon	150	Oil	6.06
Electrolysis	600	FBR	4.72

A graphical presentation of power cost for the various plant types and sizes is shown in figure C-2.

C.2.1 ELECTRIC PLANT INVESTMENT (Paragraph 4.8)

Type	150 mW $/Kw	300 mW $/Kw	600 mW $/Kw
Oil	177	130	111
Nuclear	298	253	206
FBR	--	--	266

C.2.2 ELECTRIC PLANT FIXED COSTS (Paragraph 4.9)

Type	Fixed Charge Rate %/Year	150 mW Mills/Kwh	300 mW Mills/Kwh	600 mW Mills/Kwh
Oil	11.0	2.32	1.70	1.46
Nuclear	11.25	4.00	3.40	2.76
FBR	11.25	---	---	3.57

C.2.3 ELECTRIC PLANT OPERATING AND MAINTENANCE COSTS (Paragraph 4.9.4)

Type	150 mW Mills/Kwh	300 mW Mills/Kwh	600 mW Mills/Kwh
Oil	0.72	0.58	0.38
Nuclear	0.65	0.53	0.35
FBR	---	---	0.35

C.2.4 ELECTRIC ENERGY TOTAL COST

Type	150 mW Mills/Kwh	300 mW Mills/Kwh	600 mW Mills/Kwh
Oil	6.06	5.30	4.86
Nuclear	6.53	5.73	4.76
FBR	---	---	4.72

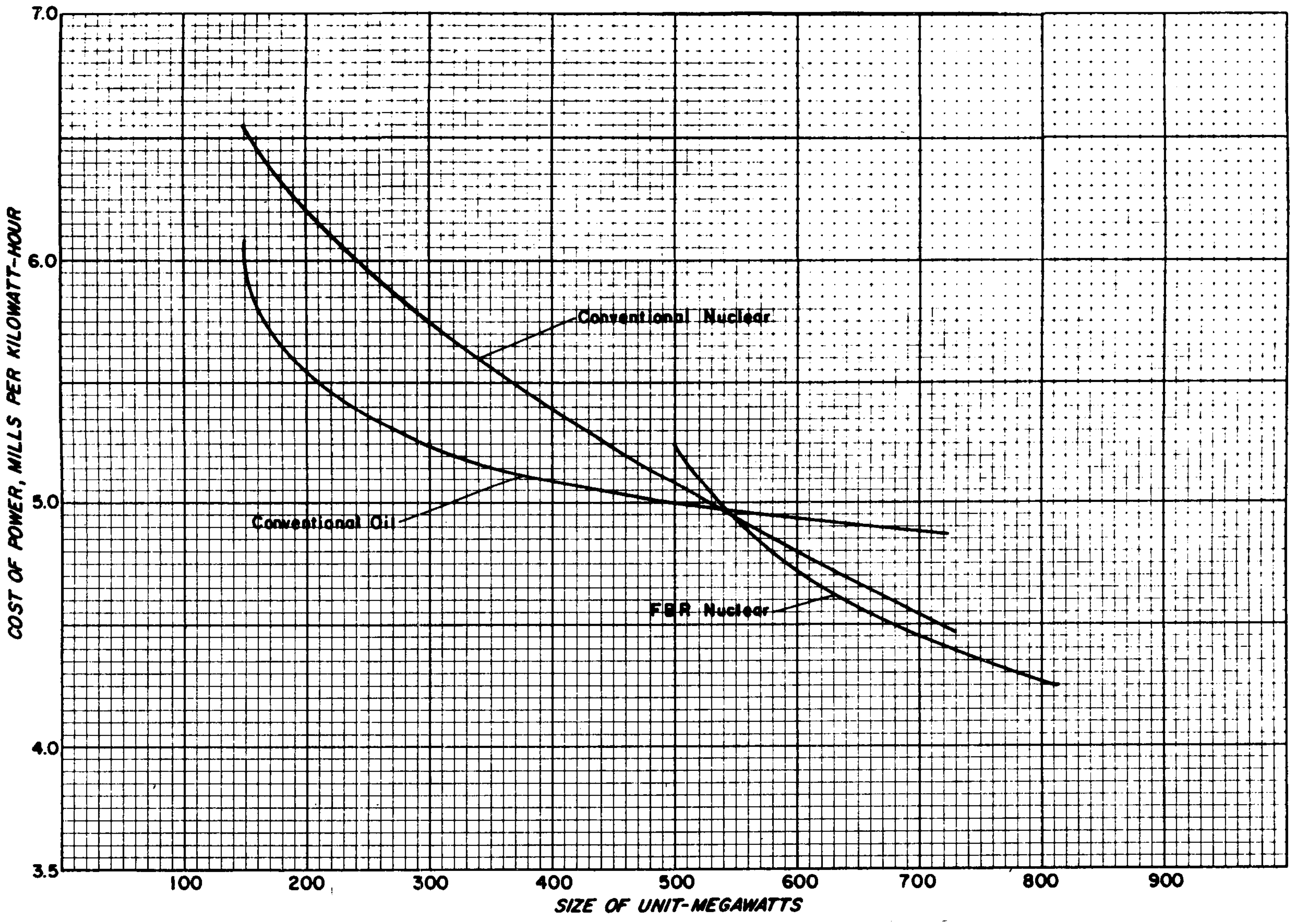

FIGURE C-2. RELATIONSHIP OF POWER COST TO PLANT SIZE (HONOLULU)

C.2.5 ELECTRIC PLANT FUEL COSTS (Paragraph 4.9.5)

	150 mW		300 mW		600 mW	
Type	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh
Oil	34.6	3.02	34.6	3.02	34.6	3.02
Nuclear	17.9	1.88	17.1	1.80	15.7	1.65
FBR	---	---	---	---	7.62	0.80

C.3 JOHANNESBURG

A coal-fired plant located at the mine mouth within 100 miles of the LH_2 plant-air terminal is the optimum system for low-cost power for the hydrocarbon and electrolytic processes. Power transmission costs have been derived from the AC curves in the FPC report.[92] Results are summarized as follows for use in section 5.

Process	mW Plant Size	Plant Type	Busbar	Transmission Mills per Kwh	Total
Hydrocarbon	1100	Coal	2.67	0.13	2.80
Electrolysis	1100	Coal	2.67	0.09	2.76

A graphical presentation of power cost for the various plant types and sizes is shown in figure C-3.

C.3.1 ELECTRIC PLANT INVESTMENT (Paragraph 4.8)

Type	500 mW $/Kw	1100 mW $/Kw	2000 mW $/Kw
Nuclear	206	156	115
FBR	264	187	123
Coal	126	111	--
Coal gas	143	135	125

C.3.2 ELECTRIC PLANT FIXED COSTS (Paragraph 4.9)

Type	Fixed Charge Rate %/Year	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000mW Mills/Kwh
Nuclear	11.25	2.74	2.07	1.54
FBR	11.25	3.54	2.51	1.65
Coal	11.0	1.65	1.45	----
Coal gas	11.0	1.88	1.77	1.63

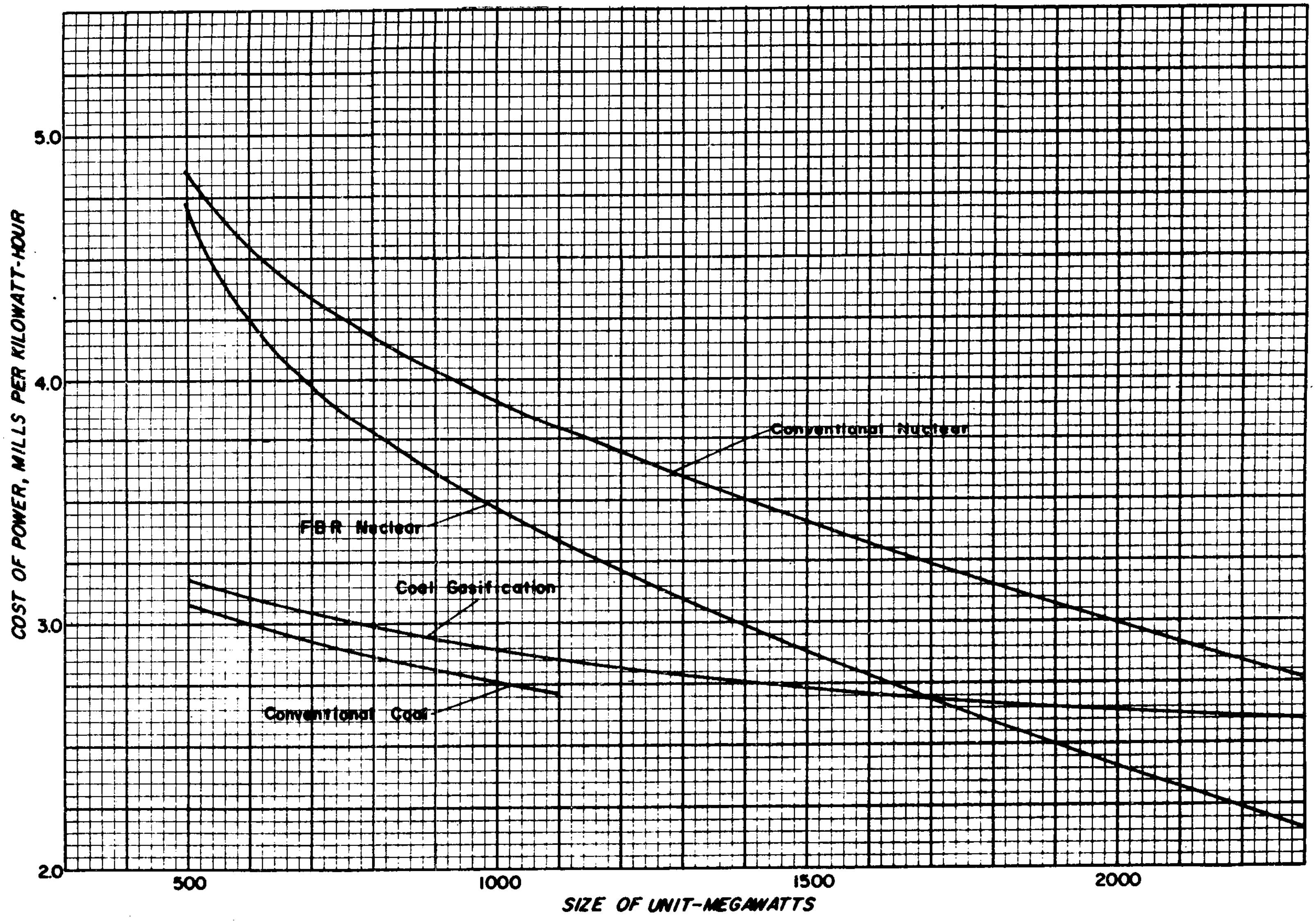

FIGURE C-3. RELATIONSHIP OF POWER COST TO PLANT SIZE (JOHANNESBURG)

C.3.3 ELECTRIC PLANT OPERATING AND MAINTENANCE COSTS (Paragraph 4.9.4)

Type	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh
Nuclear	0.40	0.22	0.15
FBR	0.40	0.22	0.15
Coal	0.49	0.26	---
Coal gas	0.49	0.26	0.18

C.3.4 ELECTRIC ENERGY TOTAL COSTS

Type	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh
Nuclear	4.84	3.78	2.99
FBR	4.74	3.33	2.40
Coal	3.10	2.67	---
Coal gasifier	3.20	2.86	2.64

C.3.5 ELECTRIC PLANT FUEL COSTS (Paragraph 4.9.5)

	500 mW		1100 mW		2000 mW	
Type	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh
Nuclear	16.1	1.70	14.0	1.49	12.4	1.30
FBR	7.62	0.80	5.7	0.60	5.7	0.60
Coal	11	0.96	11	0.96	11	---
Coal gasifier	11	0.83	11	0.83	11	0.83

C.4 BEIRUT

Oil-fired power plants for hydrocarbon and electrolytic processes are the optimum low-cost power systems for Beirut. The plant is presumed located adjacent to the LH_2 plant air terminal. Results are summarized as follows for use in section 5.

Process	mW Plant Size	Plant Type	Busbar Mills per Kwh
Hydrocarbon	300	Oil	4.69
Electrolysis	600	Oil	4.26

A graphical presentation of the various plant types and sizes is shown in figure C-4.

C.4.1 ELECTRIC PLANT INVESTMENT (Paragraph 4.8)

Type	150 mW $/Kw	300 mW $/Kw	600 mW $/Kw
Oil	173	127	109
Nuclear	291	248	202
Nuclear FBR	---	---	260

C.4.2 ELECTRIC PLANT FIXED COSTS (Paragraph 4.9)

Type	Fixed Charge Rate %/Year	150 mW Mills/Kwh	300 mW Mills/Kwh	600 mW Mills/Kwh
Oil	11.0	2.26	1.66	1.43
Nuclear	11.25	3.90	3.31	2.71
Nuclear	11.5	---	---	3.48

C.4.3 ELECTRIC PLANT OPERATING AND MAINTENANCE COSTS (Paragraph 4.9.4)

Type	150 mW Mills/Kwh	300 mW Mills/Kwh	600 mW Mills/Kwh
Oil	0.72	0.58	0.38
Nuclear	0.65	0.53	0.35
Nuclear FBR	---	---	0.35

C.4.4 ELECTRIC ENERGY TOTAL COST

Type	150 mW Mills/Kwh	300 mW Mills/Kwh	600 mW Mills/Kwh
Oil	5.43	4.69	4.26
Nuclear	6.47	5.64	4.71
Nuclear FBR	---	---	4.63

C.4.5 ELECTRIC PLANT FUEL COSTS (Paragraph 4.9.5)

	150 mW		300 mW		600 mW	
Type	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh
Oil	28	2.45	28	2.45	28	2.45
Nuclear	17.9	1.88	17.1	1.80	15.7	1.65
FBR	---	---	---	---	7.62	0.80

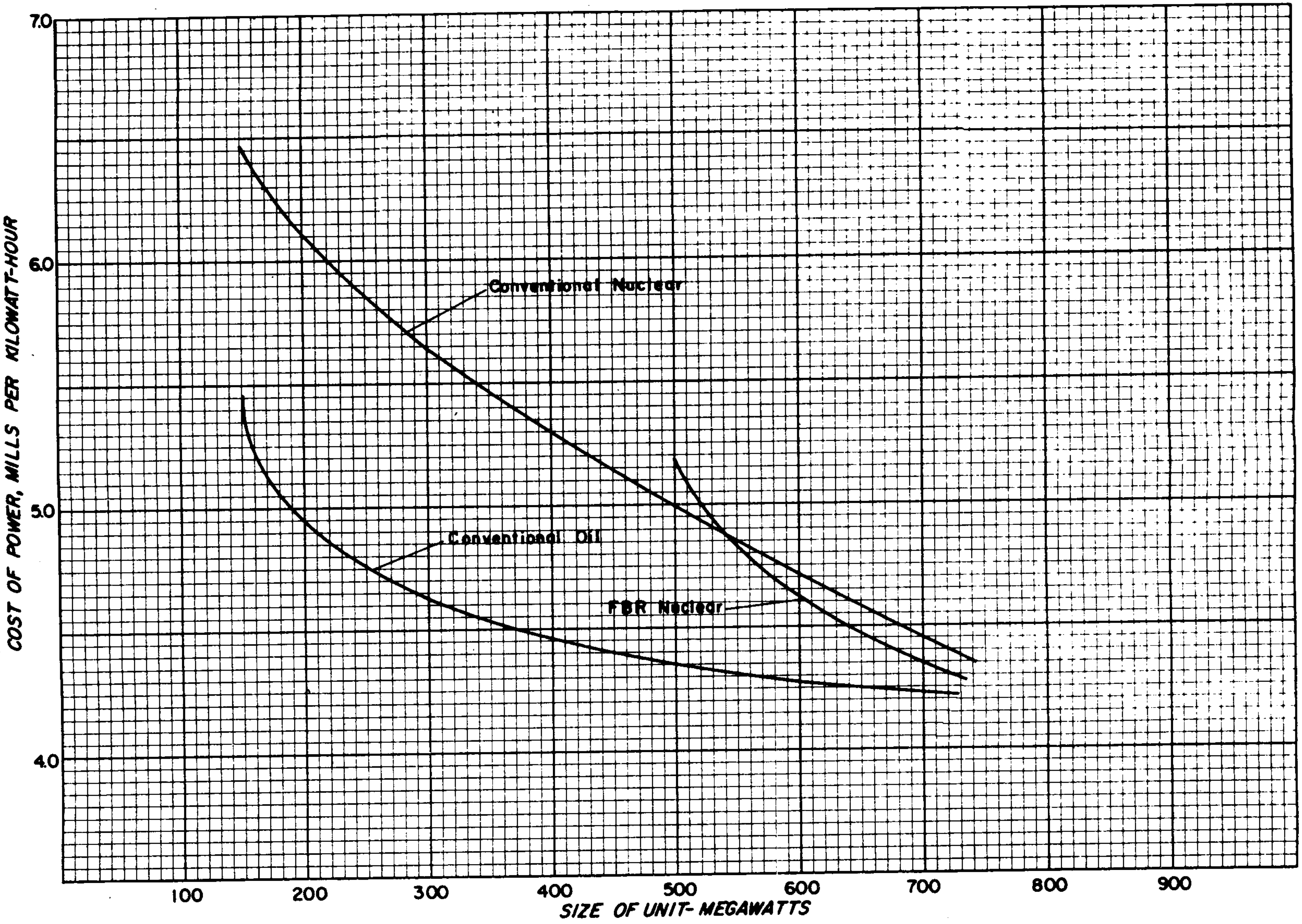

FIGURE C-4. RELATIONSHIP OF POWER COST AND PLANT SIZE (BEIRUT)

C.5 SAO PAULO

Low-cost hydroelectric power from 1980 hydro plants near the Furnase Dam on the Rio Grande appears to be the optimum low-cost source of power for hydrocarbon processes. The large load requirements of the electrolytic process would best be served from a dependable large size nuclear reactor located at the LH_2 plant. The cost to transmit the hydro power 175 miles to the Sao Paulo grid has been extrapolated from the FPC power transmission curves. Results are summarized as follows for use in Section 5.

Process	**mW Plant Size**	**Plant Type**	**Busbar**	**Transmission Mills per Kwh**	**Total**
Hydrocarbon	1100	Hydroelectric	2.77	0.23	3.00
Electrolytic	2000	FBR	2.81	0	2.81

A graphical presentation of power cost for the various plant types and sizes is shown in figure C-5.

C.5.1 ELECTRIC PLANT INVESTMENT (Paragraph 4.8)

Type	500 mW $/Kw	1100 mW $/Kw	2000 mW $/Kw
Oil	132	119	---
Nuclear	256	194	144
Nuclear FBR	328	233	154
Hydro	266	252	---

C.5.2 ELECTRIC PLANT FIXED COSTS (Paragraph 4.9)

Type	Fixed Charge Rate %/Year	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh
Oil	11.0	1.73	1.56	---
Nuclear	11.25	3.43	2.60	1.92
Nuclear FBR	11.25	4.39	3.14	2.06
Hydro	8.5	2.69	2.55	---

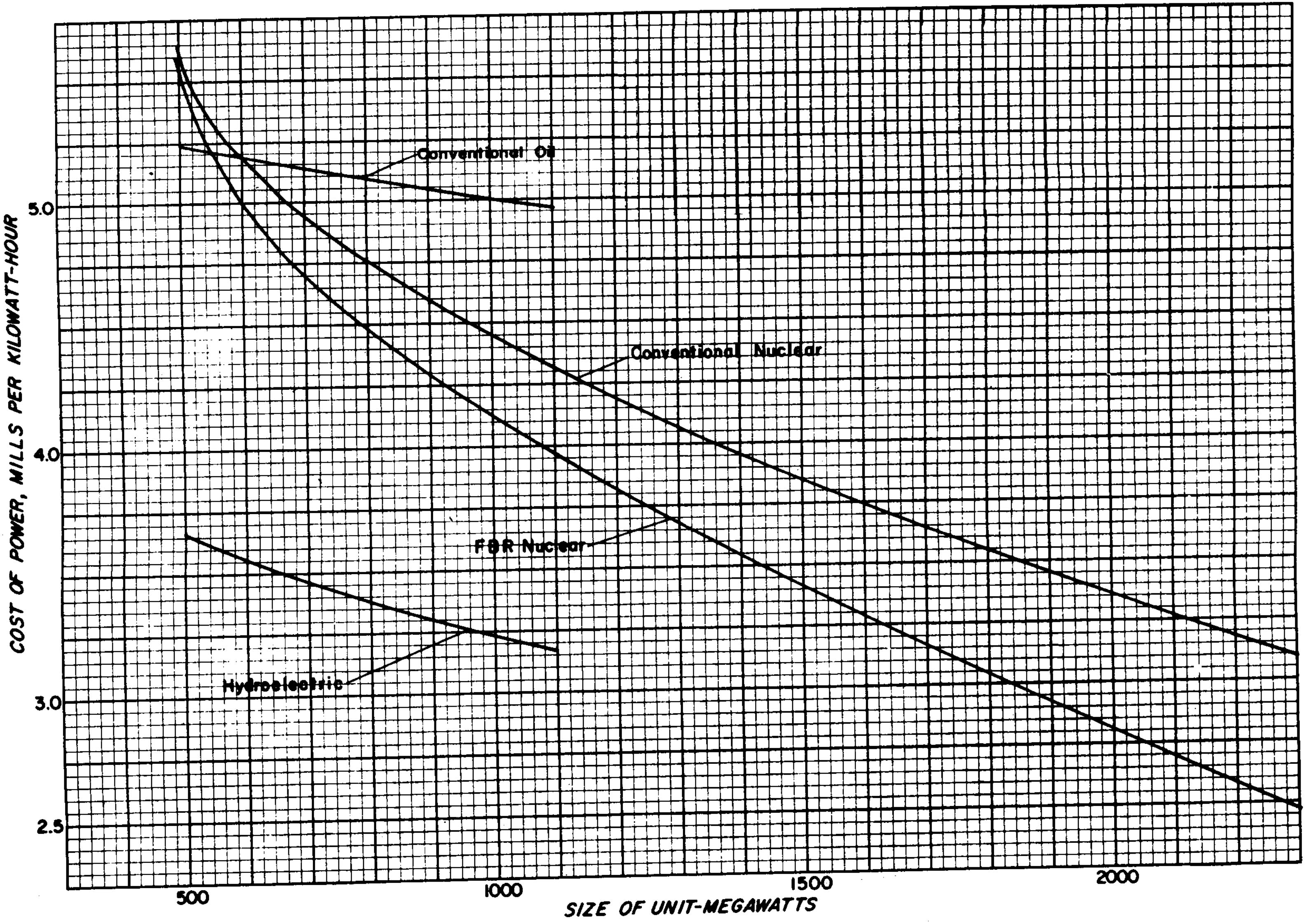

FIGURE C-5. RELATIONSHIP OF POWER COST TO PLANT SIZE (SAO PAULO)

C.5.3 ELECTRIC PLANT OPERATING AND MAINTENANCE COSTS (Paragraph 4.9.4)

Type	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh
Oil	0.46	0.25	---
Nuclear	0.40	0.22	0.15
Nuclear FBR	0.40	0.22	0.15
Hydro	0.40	0.22	---

C.5.4 ELECTRIC ENERGY TOTAL COST

Type	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh
Oil	5.25	4.87	---
Nuclear	5.53	4.31	3.37
Nuclear FBR	5.59	3.96	2.81
Hydro	3.09	2.77	---

C.5.5 ELECTRIC PLANT FUEL COSTS (Paragraph 4.9.5)

	500 mW		1100 mW		2000 mW	
Type	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh
Oil	35	3.06	35	3.06	35	---
Nuclear	16.1	1.70	14.0	1.49	12.4	1.30
FBR	7.62	0.80	5.7	0.60	5.7	0.60
Hydro	---	---	---	---	---	---

C.6 SYDNEY

Minemouth coal plants with transmission lines to serve the hydrocarbon processes and nuclear powered fast-breeder reactors at the LH_2 plant site to serve the electrolytic loads are the optimum low cost power systems available for Sydney. Although hydroelectric is the lowest cost power available, the unreliability of this source eliminates it from consideration for Sydney. Additional cost of transmission would make it comparable to the choice selected as follows:

Process	**mW Plant Size**	**Plant Type**	**Busbar**	**Transmission Mills per Kwh**	**Total**
Hydrocarbon	1100	Coal	2.97	0.15	3.12
Electrolytic	2000	FBR	2.49	0	2.49

A graphical presentation of power cost for the various plant types and sizes is shown in figure C-6.

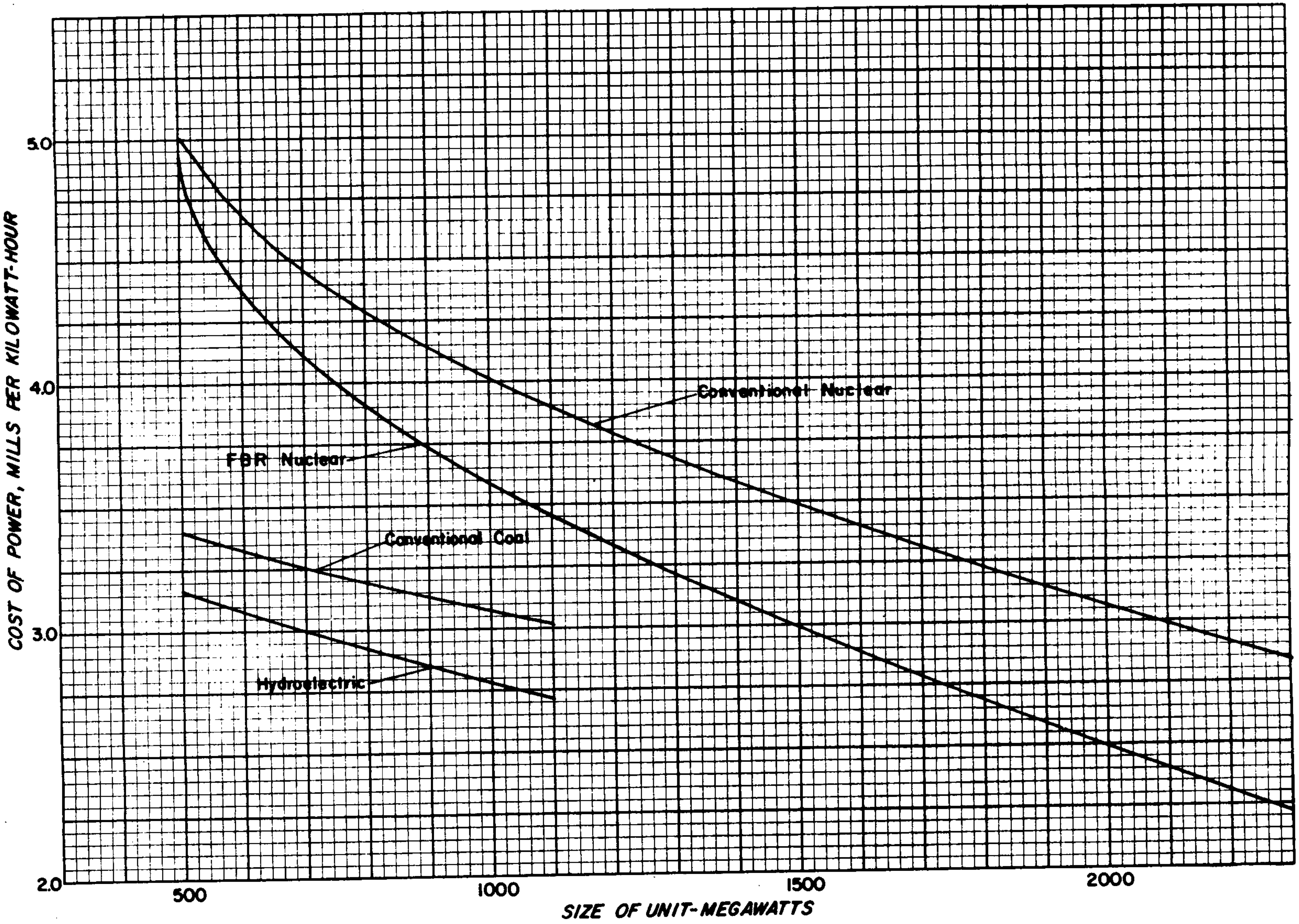

FIGURE C-6. RELATIONSHIP OF POWER COST TO PLANT SIZE (SYDNEY)

C.6.1 ELECTRIC PLANT INVESTMENT (Paragraph 4.8)

Type	500 mW $/Kw	1100 mW $/Kw	2000 mW $/Kw
Nuclear	215	163	120
FBR	275	195	130
Coal	134	117	---
Hydro	223	211	---

C.6.2 ELECTRIC PLANT FIXED COSTS (Paragraph 4.9)

Type	Fixed Charge Rate %/Year	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh
Nuclear	11.25	2.88	2.18	1.61
FBR	11.25	3.68	2.62	1.74
Coal	11.0	1.75	1.53	---
Hydro	8.5	2.25	2.13	---

C.6.3 ELECTRIC PLANT OPERATING AND MAINTENANCE COSTS (Paragraph 4.9.4)

Type	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh
Nuclear	0.40	0.22	0.15
FBR	0.40	0.22	0.15
Coal	0.49	0.26	---
Hydro	0.40	0.22	---

C.6.4 ELECTRIC ENERGY TOTAL COST

Type	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh
Nuclear	4.98	3.89	3.06
FBR	4.88	3.44	2.49
Coal	3.42	2.97	---
Hydro	2.65	2.35	---

C.6.5 ELECTRIC PLANT FUEL COSTS (Paragraph 4.9.5)

	500 mW		1100 mW		2000 mW	
Type	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh
Nuclear	16.1	1.70	14.0	1.49	12.4	1.30
FBR	7.62	0.80	5.7	0.60	5.7	0.60
Coal	13.5	1.18	13.5	1.18	13.5	---
Hydro	---	---	---	---	---	---

C.7 TOKYO

Nuclear-powered fast-breeder reactors will provide the Tokyo LH_2 power requirements in 1980. Due to the large electrolytic load requirement, a larger reactor has been used for this process. The relatively smaller power requirements for the hydrocarbon processes, in themselves, do not appear to warrant a power plant larger than 1100 in this analysis. Results are summarized as follows:

Process	**mW Plant Size**	**Plant Type**	**Busbar**
Hydrocarbon	1100	FBR	3.07
Electrolysis	2000	FBR	2.23

A graphical presentation of power cost for the various plant types and sizes is shown in figure C-7.

C.7.1 ELECTRIC PLANT INVESTMENT (Paragraph 4.8)

Type	500 mW $/Kw	1100 mW $/Kw	2000 mW $/Kw
Oil	95	86	---
Nuclear	184	140	103
FBR	236	168	111

C.7.2 ELECTRIC PLANT FIXED COSTS (Paragraph 4.9)

Type	Fixed Charge Rate %/Year	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh
Oil	11.00	1.25	1.12	---
Nuclear	11.25	2.46	1.87	1.38
FBR	11.25	3.16	2.25	1.48

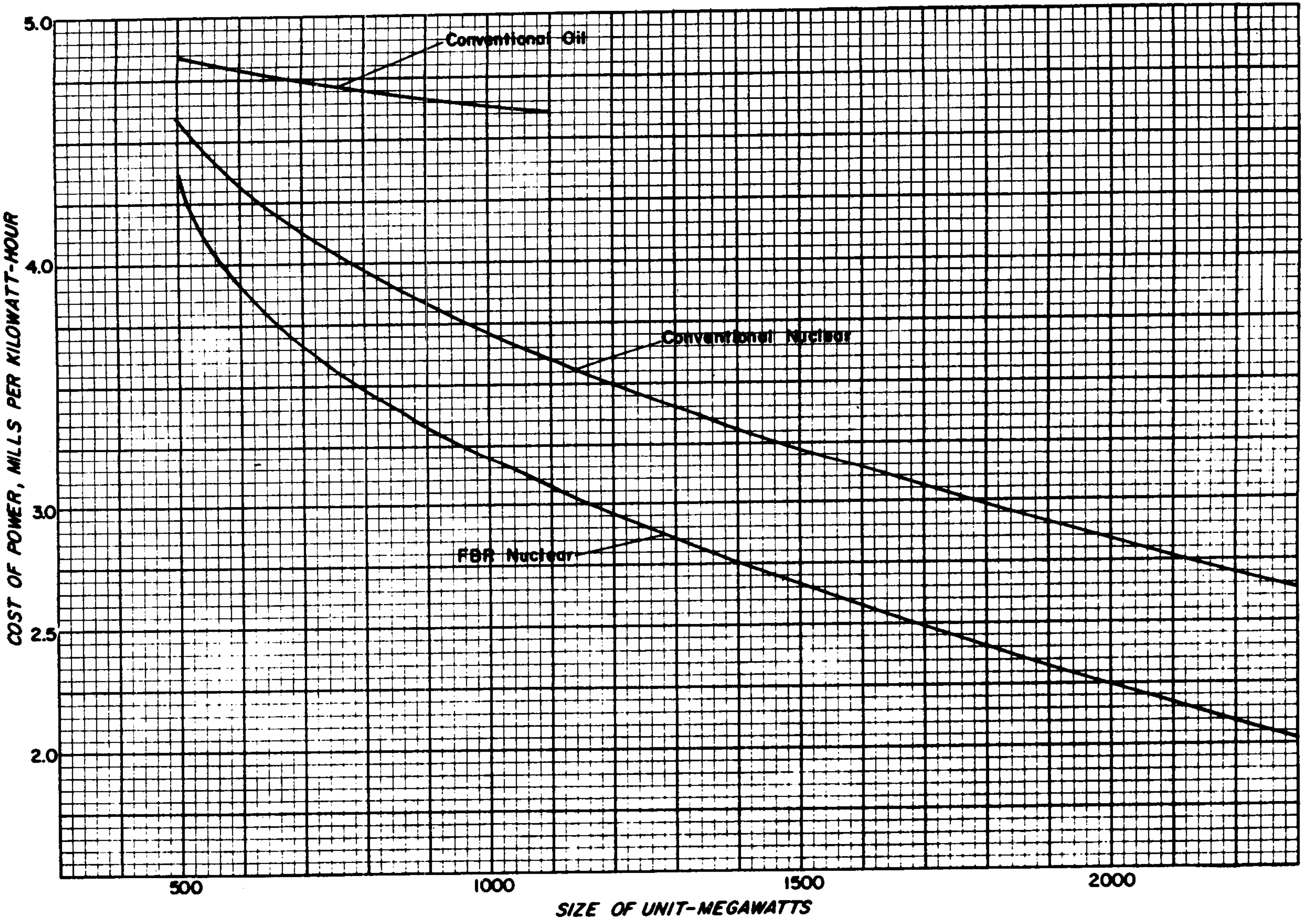

FIGURE C-7. RELATIONSHIP OF POWER COST TO PLANT SIZE (TOKYO)

C.7.3 ELECTRIC PLANT OPERATING AND MAINTENANCE COSTS (Paragraph 4.9.4)

Type	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh
Oil	0.46	0.25	---
Nuclear	0.40	0.22	0.15
FBR	0.40	0.22	0.15

C.7.4 ELECTRIC ENERGY TOTAL COST

Type	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh
Oil	4.86	4.52	---
Nuclear	4.56	3.58	2.83
FBR	4.36	3.07	2.23

C.7.5 ELECTRIC PLANT FUEL COSTS (Paragraph 4.9.5)

Type	500 mW		1100 mW		2000 Mw	
	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh
Oil	36.0	3.15	36.0	3.15	---	---
Nuclear	16.1	1.70	14.0	1.49	12.4	1.30
FBR	7.62	.80	5.7	.60	5.7	.60

C.8 LOS ANGELES

Nuclear-powered fast-breeder reactors will supply the lowest cost power for this LH_2 plant facility. The plants are assumed to be located at the air terminal. Maximum size units are based on recent studies pertaining to reactor sizes. Results are summarized as follows:

Process	mW Plant	Plant Type	Busbar
Hydrocarbon	2000	FBR	2.55
Electrolysis	3000	FBR	2.41

A graphical presentation of power cost for the various plant types and sizes is shown in figure C-8.

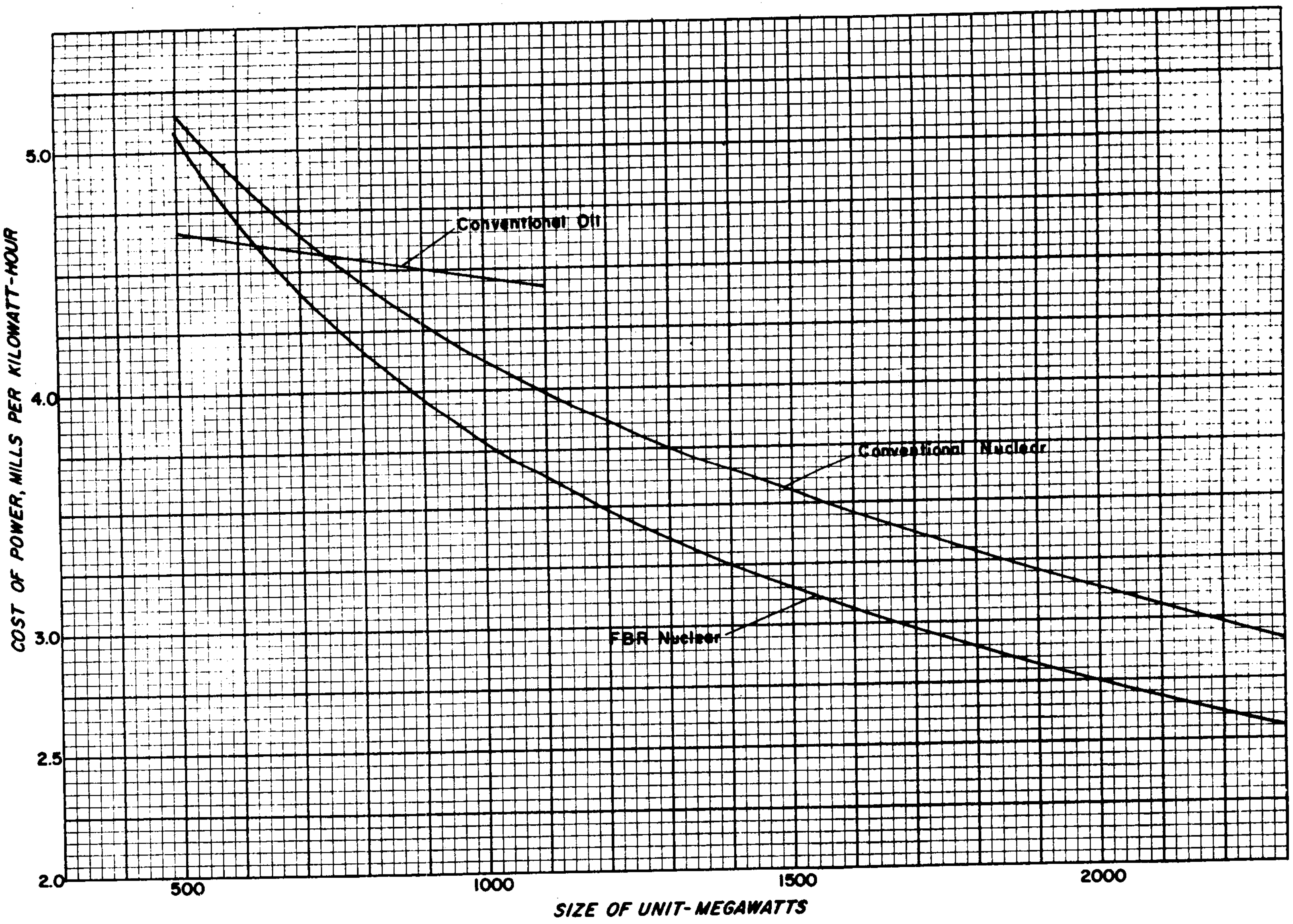

FIGURE C-8. RELATIONSHIP OF POWER COST TO PLANT SIZE (LOS ANGELES)

C.8.1 ELECTRIC PLANT INVESTMENT (Paragraph 4.8)

Type	500 mW $/Kw	1100 mW $/Kw	2000 mW $/Kw	3000 mW $/Kw
Oil	116	105	---	---
Nuclear	225	170	127	119
FBR	288	206	135	127

C.8.2 ELECTRIC PLANT FIXED COSTS (paragraph 4.9)

Type	Fixed Charge Rate %/Year	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh	3000 mW Mills/Kwh
Oil	11.0	1.52	1.37	---	---
Nuclear	11.25	3.01	2.29	1.69	1.59
FBR	11.25	3.85	2.74	1.80	1.70

C.8.3 ELECTRIC PLANT OPERATING AND MAINTENANCE COSTS (Paragraph 4.9.4)

Type	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh	3000 mW Mills/Kwh
Oil	0.46	0.25	---	---
Nuclear	0.40	0.22	0.15	0.11
FBR	0.40	0.22	0.15	0.11

C.8.4 ELECTRIC ENERGY TOTAL COST

Type	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh	3000 mW Mills/Kwh
Oil	4.55	4.19	---	---
Nuclear	5.11	4.00	3.14	3.00
FBR	5.05	3.56	2.55	2.41

C.8.5 ELECTRIC PLANT FUEL COSTS (Paragraph 4.9.5)

	500 mW		1100 mW		2000 mW		3000 mW	
Type	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh
Oil	29.4	2.57	29.4	2.57	---	---	---	---
Nuclear	16.1	1.70	14.0	1.49	12.4	1.30	12.4	1.30
FBR	7.62	0.80	5.7	0.60	5.7	0.60	5.7	0.60

C.9 PARIS

The Paris analysis is similar to that for Los Angeles. The results are presented as follows:

Process	mW Plant Size	Plant Type	Busbar
Hydrocarbon	2000	FBR	2.49
Electrolysis	3000	FBR	2.33

A graphical presentation of power cost for the various plant types and sizes is shown in figure C-9.

C.9.1 ELECTRIC PLANT INVESTMENT (Paragraph 4.8)

Type	500 mW $/Kw	1100 mW $/Kw	2000 mW $/Kw	3000 mW $/Kw
Oil	110	100	---	---
Nuclear	215	163	120	114
FBR	275	195	130	121

C.9.2 ELECTRIC PLANT FIXED COSTS (Paragraph 4.9)

Type	Fixed Charge Rate %/Year	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh	3000 mW Mills/Kwh
Oil	11.0	1.45	1.31	---	---
Nuclear	11.25	2.88	2.18	1.61	1.52
FBR	11.25	3.68	2.62	1.74	1.62

C.9.3 ELECTRIC PLANT OPERATING AND MAINTENANCE COSTS (Paragraph 4.9.4)

Type	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh	3000 mW Mills/Kwh
Oil	0.46	0.25	---	---
Nuclear	0.40	0.22	0.15	0.11
FBR	0.40	0.22	0.15	0.11

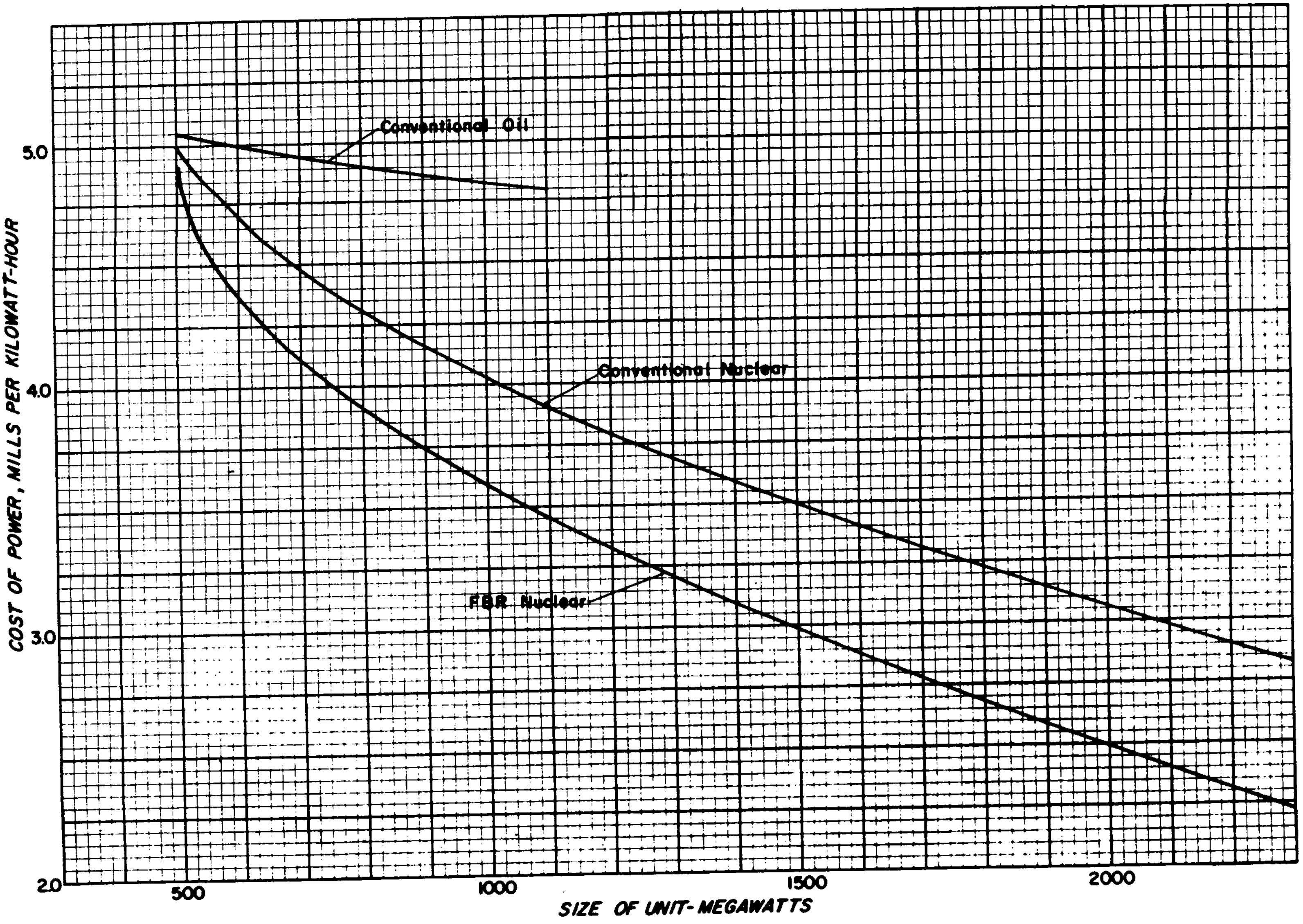

FIGURE C-9. RELATIONSHIP OF POWER COST TO PLANT SIZE (PARIS)

C.9.4 ELECTRIC ENERGY TOTAL COSTS

Type	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh	3000 mW Mills/Kwh
Oil	5.53	5.18	---	---
Nuclear	4.98	3.89	3.06	2.93
FBR	4.88	3.44	2.49	2.33

C.9.5 ELECTRIC PLANT FUEL COSTS (Paragraph 4.9.5)

Type	500 mW cents/MBtu	500 mW Mills/Kwh	1100 mW cents/MBtu	1100 mW Mills/Kwh	2000 mW cents/MBtu	2000 mW Mills/Kwh
Oil	41.4	3.62	41.4	3.62	---	---
Nuclear	16.1	1.70	14.0	1.49	12.4	1.30
FBR	7.62	0.80	5.7	0.60	5.7	0.60

C.10 NEW YORK

The New York City analysis is similar to that for Los Angeles. The results are presented as follows:

Process	**mW Plant Size**	**Plant Type**	**Busbar**
Hydrocarbon	2000	FBR	2.88
Electrolysis	3000	FBR	2.72

A graphical presentation of power cost for the various plant types and sizes is shown in figure C-10.

C.10.1 ELECTRIC PLANT INVESTMENT (Paragraph 4.8)

Type	500 mW $/Kw	1100 mW $/Kw	2000 mW $/Kw	3000 mW $/Kw
Oil	138	124	---	---
Nuclear	266	202	149	141
FBR	342	243	160	150
Coal	166	145	---	---

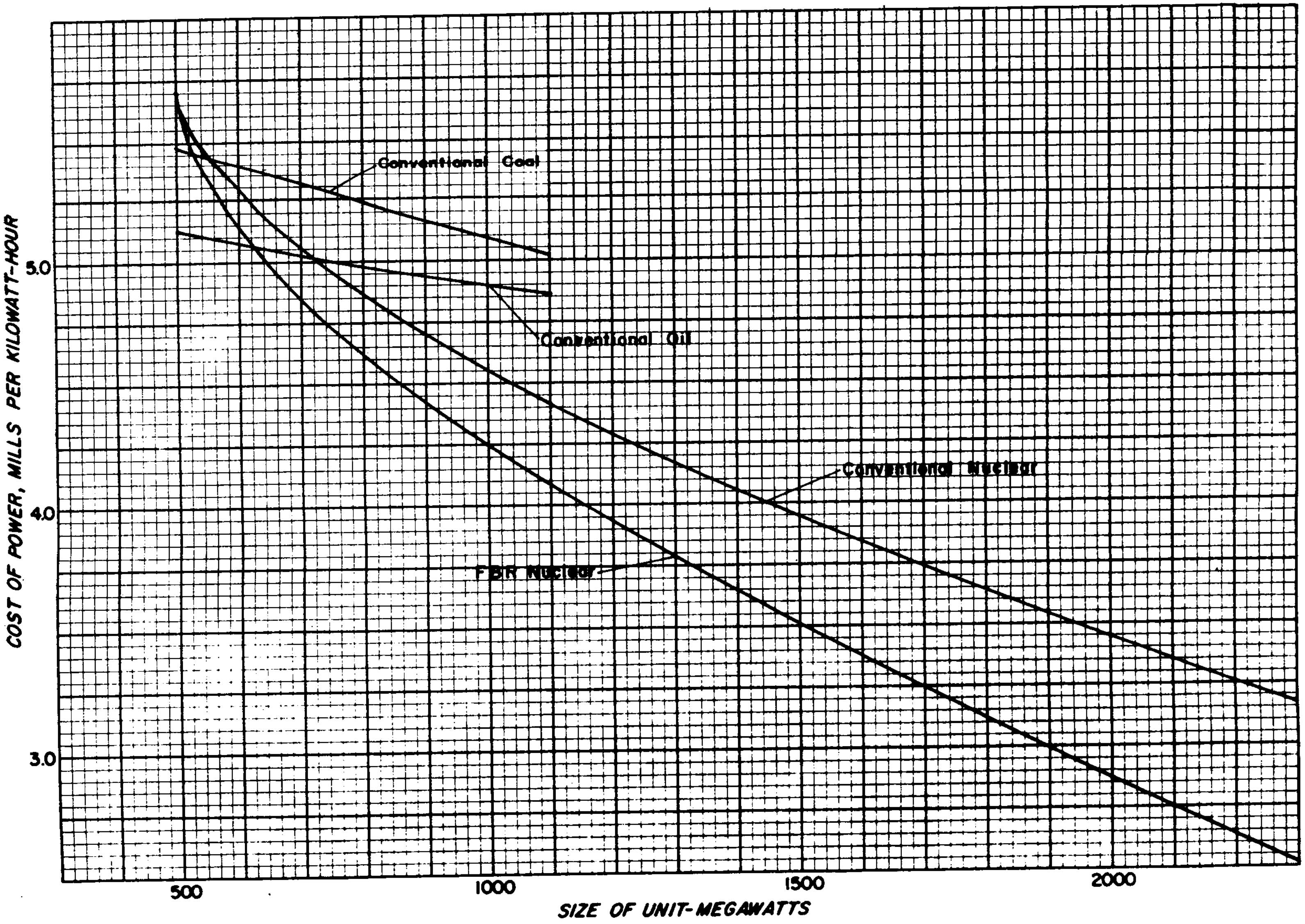

FIGURE C-10. RELATIONSHIP OF POWER COST TO PLANT SIZE (NEW YORK)

C.10.2 ELECTRIC PLANT FIXED COSTS (Paragraph 4.9)

Type	Fixed Charge Rate %/Year	500 mW Mills/Kwh	1100 mw Mills/Kwh	2000 mW Mills/Kwh	3000 mW Mills/Kwh
Oil	11.0	1.85	1.66	---	---
Nuclear	11.25	3.56	2.71	2.00	1.87
FBR	11.25	4.56	3.24	2.13	2.01
Coal	11.0	2.22	1.94	---	---

C.10.3 ELECTRIC PLANT OPERATING AND MAINTENANCE COSTS (Paragraph 4.9.4)

Type	500 mW Mills/Kwh	1100 mw Mills/Kwh	2000 mw Mills/Kwh	3000 mw Mills/Kwh
Oil	0.46	0.25	---	---
Nuclear	0.40	0.22	0.15	0.11
FBR	0.40	0.22	0.15	0.11
Coal	0.49	0.27	---	---

C.10.4 ELECTRIC ENERGY TOTAL COST

Type	500 mW Mills/Kwh	1100 mW Mills/Kwh	2000 mW Mills/Kwh	3000 mW Mills/Kwh
Oil	5.20	4.80	---	---
Nuclear	5.66	4.39	3.45	3.28
FBR	5.75	4.06	2.88	2.72
Coal	5.51	5.01	---	---

C.10.5 ELECTRIC PLANT FUEL COSTS (Paragraph 4.9.5)

	500 mW		1100 mW		2000 mW		3000 mW	
Type	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh	cents/MBtu	Mills/Kwh
Oil	33	2.89	33	2.89	33	---	---	---
Nuclear	16.1	1.70	14.0	1.49	12.4	1.30	12.4	1.30
FBR	7.62	0.80	5.7	0.60	5.7	0.60	5.7	0.60
Coal	32	2.80	32	2.80	32	---	---	---

Appendix D

DETAILED FACILITY DESCRIPTION CALCULATIONS FOR BANGKOK AND NEW YORK CITY

D.1 BANGKOK

D.1.1 FACILITY DESCRIPTION CALCULATIONS

The following paragraphs relate information for the Bangkok facility description calculations.

D.1.1.1 Distribution

Vacuum insulated pipe lines will be used. Pipe line of 18" IPS nominal size, for 2 miles long is considered. This provides the capability of loading one aircraft at a time. The net product to be delivered is 200T/D. A loss of 7 percent is anticipated in this type of transfer system as referenced in paragraph 3.5.4.1.3. Thus, the product transferred into the distribution system must be 200 T/D/0.93, or 215 T/D. This product required for distribution is the new product required from the storage system.

D.1.1.2 Storage

Evacuated, perlite, insulated, double wall tanks will be used. The capacity of the two tanks is dependent upon the reliability of the production system (i.e., the gas production and gas liquefaction systems). Sufficient reserve must be maintained to backup the outages of these systems. Single modules will be employed for the gas and purification plant and the liquefaction conversion plant. The gas plant module will have an on-stream factor of 93 percent, as referenced in paragraph 5.2.1. The on-stream factor for the liquefaction unit is 97 percent, as referenced in paragraph 5.2.2. Minimum storage capacity backup is equivalent to the maximum duration of outage for the normal gas plant production (10 days). Since the modular size of the gas plant is taken as 250 tons, the primary storage requirement is then 2500 tons.

In accord with the detail information referenced in paragraph 5.2.3, additional amount of operational flywheel storage must be provided. This additional amount, equivalent to one day's nominal draw, is 215 tons; an oversize factor of 5 percent must be applied to the storage capacity. Thus, the required storage capacity (V) is calculated as:

$$\begin{aligned} V &= (2500 + 215)\,T \times 1.05, \\ &= 2851\ T \end{aligned}$$

For the facility, two tanks of 1400 T nominal storage capacity will be used.

In order to define the net production from the liquefaction unit to storage, the heat leak and the ortho-para conversion storage losses must be added to the production required for input to the distribution system. Using the storage tank sizes as a base, the storage and conversion losses can be projected. Examining the heat leak data as referenced in paragraph 3.4.1.4, the anticipated storage losses for a 1400 T storage unit are 0.019 percent/day. This yields a heat leak storage loss, then, of 0.5 T/D for the tanks. The conversion loss is taken as 1.25 percent of the product delivered to the storage system as referenced in paragraph 5.2.3. This product delivered (P) is calculated as:

$$P = 215\ \text{T/D} + 0.5\ \text{T/D} + (0.0125)\ (P),$$

$$(0.9875)\ (P), = 215.5\ \text{T/D},$$

$$P = 218\ \text{T/D}$$

This calculated value of 218 T/D represents the average daily production of liquid by the liquefaction plant.

D.1.1.3 Liquefaction and Conversion

Production capability of the liquefaction unit must be that the average daily production is maintained when plant outage is considered. For example, the single liquefaction unit will have an on stream factor of 97 percent. Thus, a production capability of at least 218 T/Dx0.97 or 225 T/D must be attainable. However, the required production capability is also influenced by the reliability of the gas plant which supplies the feedstock to the liquefier. In this case the gas module has an on stream factor of 93 percent. Thus, a production capability of at least 218 T/D/0.93, or 234 T/D is required. The capability of 234 T/D effectively provides for plant outage assuming the 3 percent outage of the liquefier occurs during the 7 percent outage of the gas plant. The "worst case" would be that of assuming the respective outages were not concurrent, but additive. This would result in the production outage being 10 percent or (7% + 3%), with the required production outage being 10 percent or (7% + 3%), with the required production capability of the liquefier 218 T/D x 0.90, or 242 T/D.

A nominal liquefaction unit size of 235 T/D has been selected for Bangkok. Average daily production from the liquefier unit was calculated to be 218 T/D. Gas plant production to support this liquefaction rate must allow for losses within the liquefaction equipment. To account for the losses, 4 percent "overproduction" is necessary from the gas equipment as referenced in paragraph 5.2.2. Thus, the average daily production from the gas plant calculates to 218 T/D x 1.04, or 227 T/D. Gas production at the average daily production rate of 227 T/D, is sufficient to support the overall facility delivery schedule.

Considering the gas plant on-stream factor of 93 percent, the required production capability calculates to 227 T/D/0.93, or 244 T/D. A single unit of 250 T/D nominal size has been selected for Bangkok.

D.1.1.4 Bangkok Summary

These calculations constitute a consistent set of data describing the Bangkok liquid hydrogen facility. Pertinent information is summarized as follows:

(1) Distribution
Two miles long of one 18-inch IPS vacuum insulated pipeline.

(2) Storage
Two evacuated perlite double wall tanks, 1400 T each.

(3) Liquefaction and Conversion
Nominal Plant Size 235 T/D - Average Daily Production 218 T/D

(4) Gas Production and Purification
Nominal Plant Size 250 T/D - Average Daily Production 227 T/D

D.2 NEW YORK CITY

D.2.1 FACILITY DESCRIPTION CALCULATIONS

The following paragraphs relate information for the New York City facility description calculations.

D.2.1.1 Distribution

Vacuum insulated pipe lines will be used. Two 24" IPS nominal size, 2 miles long, of pipelines are considered. These provide the capability of loading four aircraft simultaneously. The net product to be delivered to the aircraft manifolds is 2500 T/D. A loss of 7 percent is anticipated in the use of this type of transfer system as referenced in paragraph 3.5.4.1.3. Thus, the product transferred into the distribution system must be 2500 T/D/0.93, or 2688 T/D. This product required for distribution, in turn, is the product required from the storage system.

D.2.1.2 Storage

Two sets of storage calculations for this location have been made to satisfy actual requirements. This is accomplished in order to reach a suitable facility design to meet the production needs for this location.

D.2.1.2.1 Calculation 1 – Evacuated, perlite insulated, double wall tanks will be used. The capacity of the tanks (two tanks) is dependent upon reliability of the production systems (i.e., the gas producing plant and the liquefaction plant). Sufficient reserve must be maintained to backup the outages of these production subsystems.

It is estimated that 11 or 12 modular gas plant units will be used, each having an on-stream factor or 93 percent, as referenced in paragraph 5.2.1. Assuming a 12 unit modular plant and a 7 percent down-time per unit it can be seen that 84 percent of the time 11 out of the 12 units will be operating and 16 percent of the time all 12 units would be operating. It is, therefore, possible that the eleven unit production would be sufficient to provide the overall facility related gas production. As a minimum, the product storage backup required is for situation when one of the 250 T/D gas units goes down for the expected maximum duration as referenced in paragraph 5.2.2. Storage back up for the gas plant would be nominally 2500 tons.

Considering the liquefaction plant, it is possible that 3 modular units will be used, each having a nominal capacity of 1000 T/D and an on-stream factor of 97 percent, as referenced in paragraph 5.2.1. Assuming, therefore, 3 modular units and a 3 percent down-time per unit, it can be seen that 1 of the 3 units will be off-stream 9 percent of the time or the 3 units will be on-stream 91 percent. The overall facility operation could not proceed normally with only 2 of the 3 liquefaction units operating. Thus, the storage capacity must be such that backup is provided from this type of outage. The maximum duration of such an outage is as referenced in paragraph 5.2.2. Storage capacity, required for the liquefaction plant outage must then have a minimum backup capacity of 7000 tons.

Since the storage capacity required for the liquefaction plant outage will also satisfy the capacity of gas plant outage, the 7000 ton storage capacity becomes the governing factor. In addition to the 7000 ton required for plant outages, an operating flywheel equivalent to one day's nominal draw must be provided, as referenced in paragraph 5.2.3. Thus, an additional storage of 2688 T must be incorporated. Further, a contingent storage of 5 percent must also be provided, as referenced in paragraph 5.1.4. Thus, the required storage capacity (V) is calculated as:

$$\begin{aligned} V &= (7000 + 2688)\,T \times 1.05, \\ &= 10{,}172\,T. \end{aligned}$$

For the facility, two tanks of 5000 T nominal storage capacity will be used. If the assumptions related to the gas and liquefaction plants prove in error, a revised storage system will be required.

In order to define the production required from the liquefaction plant, the heat leak and ortho-para conversion storage losses must be added to the production required for distribution. Using the above storage size as a base, the storage and conversion losses can be projected. Extrapolating the heat leak data as referenced in paragraph 3.4.1.4, the anticipated storage loss for a 5000 T unit is 0.014 percent per day. This yields a heat leak storage loss of 1.4 T/D. The conversion loss is taken as 1.25 percent of the product delivered to the storage system as referenced in paragraph 5.2.3. This product delivered to storage (P) is calculated as:

$$P = 2688\,T/D + 1.4\,T/D + (0.0125)\ (P),$$

$$(0.9875)\ (P), = 2689.44\,T/D$$

$$P = 2723\,T/D$$

This calculated value of 2723 T/D represents the average daily production of liquid by the liquefaction plant.

D.2.1.2.2 <u>Calculation 2</u> – Following the development of calculation 1, the minimum storage back up capacity is 7 x 935/T, or 6545/T. Counting an additional day's equivalent draw (2688/T) and 5 percent oversize, the required storage capacity (V) is calculated as:

$$V = (6545\ T + 2688\ T) \times 1.05$$
$$= 9695\ T$$

For the facility two tanks of 4800 T nominal storage will be used. Anticipated heat leak storage loss for the tanks is 0.014 percent per day, or 1.3 T/D. Accounting for conversion loss, the product delivered to storage (P) is calculated as:

$$P = 2688\ T/D + 1.3\ T/D + 0.0125\ (P),$$
$$P = 2723\ T/D$$

D.2.1.3 Liquefaction and Conversion

Two sets of liquefaction and conversion calculations have been made to satisfy the actual requirements, and in order to justify the storage calculations.

D.2.1.3.1 Calculation 1 – The on-stream factor for the liquefaction equipment is 97 percent per module. Storage sizing for the liquefaction plant operation can be based upon each unit operating at 97 percent of the time, or the liquefaction facility has a composite on-stream factor of 97 percent. The liquefaction plant size, then, is calculated as 2723 T/D/0.97, or 280 T/D.

Three units of nominal 935 T/D size would satisfy the requirements. However, it should be noted that estimates (storage sizing) used in other calculations were based upon 1000 T/D modular sizes. A second set of data must be developed based upon the smaller sized units.

D.2.1.3.2 Calculation 2 – Production required from the liquefaction equipment is essentially the same as determined in calculation 1. The choice of three nominal 935 T/D size units thus remains consistent.

D.2.1.4 Gas Production and Purification

Average daily production from the liquefier unit is 2723 T/D. Production by the gas unit must allow for the 4 percent loss of feedstock in the liquefier. Daily average gas production then is 2723 T/D x 1.04 or 2832 T/D. However, the liquefier equipment, operating with a 97 percent on-stream factor, will require gas production at the rate of 2832 T/D/0.97, or 2920 T/D. The gas equipment then must have a capability of daily production at the rate of 2920 T/D for 97 percent of the time.

To satisfy the production requirements, thirteen gas plant modules of 250 T/D nominal size will be utilized. Each of the units will have a 93 percent on-stream factor. With this number of units, production capability would be nominally 3000 T/D for 91 percent of the time, and 3250 T/D for the remaining 9 percent of the time. In this manner sufficient gas production capability is maintained for the liquefaction equipment supply.

D.2.1.5 New York City Summary

These calculations constitute a consistent set of data describing the New York City liquid hydrogen facility. Pertinent information is summarized as follows:

(1) Distribution

Two miles long of vacuum insulated pipeline; two 29-inch IPS

(2) Storage

Two evacuated perlite double wall tanks, 4800 T each.

(3) Liquefaction and Conversion

Nominal plant size 2805 T/D
Average daily production 2723 T/D

(4) Gas Production and Purification

Nominal Plant Size 3000 T/D
Average Daily Production 2832 T/D

APPENDIX E

CALCULATIONS FOR VARIOUS LOCATIONS

NATIONAL / 42-381 50 SHEETS
42-382 100 SHEETS
Made in U.S.A.

FACILITY DESCRIPTION – BANGKOK

DISTRIBUTION
1 – 18" NPS × 2 MILE, SI VJ LINE (ALLOWS 1 AIRCRAFT LOADING)
PRODUCT DELIVERED IS 200 T/D. PRODUCT RECEIVED IS 215 T/D.

STORAGE
2 – 1400 T NOMINAL SIZE, DOUBLE WALL, EVACUATED PERLITE INSUL. TANKS.
PRODUCT DISCHARGED IS 215 T/D. PRODUCT RECEIVED IS 218 T/D.

LIQUEFACTION AND CONVERSION
1 – 235 T/D NOMINAL SIZE UNIT WITH 97% ON STREAM FACTOR.
DAILY AVERAGE PRODUCTION IS 218 T/D.

GAS PRODUCTION AND PURIFICATION
1 – 250 T/D NOMINAL SIZE UNT WITH 93% ON STREAM FACTOR.
DAILY AVERAGE PRODUCTION IS 227 T/D.

PRODUCTION COST DETERMINATION

PROCESS, ITEM, UNIT REQUIREMENT	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
STEAM REFORMING			
FEED – NATURAL GAS $1.90 \frac{\#NG}{\#GH_2} = 1.90 \frac{\#NG}{\#GH_2} \times \frac{MCF\ NG}{41.6\ \#NG}$	7.57×10^6 MCF (BASED ON 227 T/D GAS PROD., 365 DAY/YR.)	\$0.45/MCF	3.41
$= 0.0457 \frac{MCF\ NG}{\#GH_2}$			
FUEL – NATURAL GAS $1.44 \frac{\#NG}{\#GH_2} - 4.4 \times 10^{-5} \frac{\#NG/\#GH_2}{T/D} \times 250\ T/D$			

NATIONAL 42-381 50 SHEETS
42-382 100 SHEETS
Made in U.S.A.

Process, Item, Unit Requirement — BANGKOK		Annual Requirement	Unit Cost	Annual Cost $\times 10^{-6}$ (\$)
Fuel - Natural Gas, Cont'd.	$1.43 \frac{\#NG}{\#GH_2}$			
Fuel - Oil Equiv of Nat. Gas (Assume oil is same as for Honolulu)	$1.43 \frac{\#NG}{\#GH_2} \times \frac{MCF\ NG / 41.6\ \#NG}{MCF\ NG / 1,000,000\ BTU}$ $\times \frac{BBL / 42\ GAL}{155,200\ BTU / GAL}$ $= 5.27 \times 10^{-3} \frac{BBL}{\#GH_2}$	0.87×10^6 BBL	\$2.15/BBL	1.87
Cooling Water & Boiler Water Make Up	$(22.4 + 16.6) \frac{\#H_2O}{\#GH_2} = 39.0 \frac{\#H_2O}{\#GH_2}$	3.23×10^6 T	\$0.0072/T	0.02
Process Energy	$0.47 \frac{KWH}{\#GH_2}$	77.88×10^6 KWH	\$0.00361/KWH	0.28
Byproduct CO₂	$5.3 \frac{\#CO_2}{\#GH_2}$	0.44×10^6 T	—	—
Capital Investment	$\$7.65 \times 10^6 \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.7} \times 1.125$ $= \$8.606 \times 10^6$	$\$8.606 \times 10^6 \times 0.11$ $= \$0.95 \times 10^6$	—	0.95
Operating Cost	$\$1.28 \times 10^6/YR \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.73} = \$1.28 \times 10^6/YR$	$\$1.28 \times 10^6$	—	1.28
Total Cost				7.81

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PROCESS, ITEM, UNIT REQUIREMENT — BANGKOK	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST × 10^{-6} ($)
PARTIAL OXIDATION			
FEED - OIL OR BUNKER "C" $3.89 \frac{\#OIL}{\#GH_2}$ $= 3.89 \frac{\#OIL}{\#GH_2} \times \frac{BBL/42.0\ GAL}{8.40\#OIL/GAL}$ $= 0.011\ BBL/\#GH_2$	1.82×10^6 BBL	$2.15/BBL	3.91
FUEL - BANGKOK OIL EQUIV OF BUNKER "C" (ASSUME OIL IS THE SAME AS HONOLULU) $0.40 \frac{\#\text{"C"}}{\#GH_2} - 3.1 \times 10^{-5} \frac{\#C/\#GH_2}{T/D} \times 250\ T/D$ $= 0.39 \frac{\#\text{"C"}}{\#GH_2}$ $= 0.39 \frac{\#\text{"C"}/\#GH_2}{8.108\ \#\text{"C"}/GAL\ \text{"C"}} \times \frac{152{,}000\ BTU/GAL\ \text{"C"}}{155{,}200\ BTU/GAL\ OIL}$ $\times\ BBL/42\ GAL$ $= 0.0011 \frac{BBL}{\#GH_2}$	0.18×10^6 BBL	$2.15/BBL	0.39
COOLING WATER & BOILER WATER MAKE UP $(12.5 + 12.5) \frac{\#H_2O}{\#GH_2} = 25.0 \frac{\#H_2O}{\#GH_2}$	2.07×10^6 T	$0.0072/T	0.01
PROCESS ENERGY $1.30 \frac{KWH}{\#GH_2}$	2.15×10^8 KWH	$0.00361/KWH	0.78
BYPRODUCTS $6.0 \frac{\#CO_2}{\#GH_2}$	0.50×10^6 T	—	—
$5.0 \frac{\#STEAM}{\#GH_2}$	0.41×10^6 T	—	—
$0.11 \frac{\#C}{\#GH_2}$	9.1×10^3 T	—	—

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PROCESS, ITEM,	UNIT REQUIREMENT BANGKOK	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
CAPITAL INVESTMENT	$\$8.46 \times 10^6 \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.7} \times 1.125$ $= \$9.518 \times 10^6$	$\$9.518 \times 10^6 \times 0.11$ $= \$1.05 \times 10^6$	—	1.05
OPERATING COST	$\$1.46 \times 10^6/Y \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.73} = \$1.46 \times 10^6/Y$	$\$1.46 \times 10^6$	—	1.46
TOTAL COST				7.60
COAL GASIFICATION (BITUM.)				
FEED - WATER	$20.3\ \frac{\#\ H_2O}{\#\ GH_2}$	1.682×10^6 T	\$0.0072/T	0.01
FUEL - COAL EQUIV OF W. VA. BITUM.	$8.74\ \frac{\#\ W.VA.\ BIT}{\#\ GH_2}$ $= 8.74\ \frac{\#\ W.VA.\ BIT}{\#\ GH_2} \times \frac{14{,}290\ BTU/\#\ W.VA\ BIT}{11{,}100\ BTU/\#\ BAN.\ COAL}$ $= 11.25\ \frac{\#\ BAN\ COAL}{\#\ GH_2}$	0.932×10^6 T	\$12.50/T	11.65
OXYGEN	$5.20\ \frac{\#\ O_2}{\#\ GH_2}$	0.431×10^6 T	\$5.00/T	2.16
COOLING WATER MAKE UP	$26.0\ \frac{\#\ H_2O}{\#\ GH_2}$	2.154×10^6 T	\$0.0072/T	0.02
PROCESS ENERGY	$0.57\ \frac{KWH}{\#\ GH_2}$	94.45×10^6 KWH	\$0.00361/KWH	0.34

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PROCESS, ITEM,	UNIT REQUIREMENT BANGKOK	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
BYPRODUCT CO_2	$24.0 \frac{\# CO_2}{\# GH_2}$	1.99×10^6 T	—	—
CAPITAL INVESTMENT	$\$20.8 \times 10^6 \times \left(\frac{250 \text{ T/D}}{250 \text{ T/D}}\right)^{0.7} \times 1.125$ $= \$23.400 \times 10^6$	$\$23.4 \times 10^6 \times 0.11$ $= \$2.57 \times 10^6$	—	2.57
OPERATING COST	$\$2.89 \times 10^6/\text{Y} \times \left(\frac{250 \text{ T/D}}{250 \text{ T/D}}\right)^{0.65} = \$2.89 \times 10^6$	$\$2.89 \times 10^6$		2.89
TOTAL COST				19.64
ELECTROLYSIS (LO PRESSURE)				
FEED & COOLING WATER	$(9.0 + 11.5) \frac{\# H_2O}{\# GH_2} = 20.5 \frac{\# H_2O}{\# GH_2}$	1.70×10^6 T	\$0.0072/T	0.01
PROCESS ENERGY	$21.6 \frac{\text{KWH}}{\# GH_2}$	3.58×10^9 KWH	\$0.00323/KWH	11.56
BYPRODUCT O_2	$7.92 \frac{\# O_2}{\# GH_2}$	0.66×10^6 T	—	—
CAPITAL INVESTMENT	$\$28.5 \times 10^6 \left(\frac{250 \text{ T/D}}{250 \text{ T/D}}\right)^{0.88} \times 1.125 = \$32.063 \times 10^6$	$\$32.063 \times 10^6 \times 0.11$ $= \$3.53 \times 10^6$	—	3.53
OPERATING COST	$\$2.62 \times 10^6/\text{Y} \times \left(\frac{250 \text{ T/D}}{250 \text{ T/D}}\right)^{0.73} = \$2.62 \times 10^6/\text{Y}$	$\$2.62 \times 10^6$	—	2.62
TOTAL COST				17.72

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PROCESS, ITEM, UNIT REQUIREMENT BANGKOK			ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
LIQUEFACTION					
REFRIGERANTS	N_2 (NOT REQ'D FOR P.O. FEEDSTOCK)	$0.04 \frac{\# N_2}{\# LH_2}$	3.2×10^3 T	\$6.50/T	0.02
	C_1H_4	$0.01 \frac{\# NG}{\# LH_2} \times \frac{MCF}{41.6 \# NG} = 24 \times 10^{-5} \frac{MCF}{\# LH_2}$	38.2×10^3 MCF	\$0.45/MCF	0.02
	C_3H_8	$0.02 \frac{\# C_3}{\# LH_2}$	1.6×10^3 T	\$25.00/T	0.04
	C_2H_4	$0.02 \frac{\# E_2}{\# LH_2}$	1.6×10^3 T	\$80.00/T	0.13
COOLING WATER MAKE UP		$24.7 \frac{\# H_2O}{\# LH_2}$	1.97×10^6 T	\$0.0072/T	0.01
PROCESS ENERGY	(HYDROCARBON, COAL FEEDSTOCK)	$4.46 \frac{KWH}{\# LH_2}$	70.98×10^7 KWH	\$0.00361/KWH	2.56
	(ELECTROLYSIS FEEDSTOCK)	$4.46 \frac{KWH}{\# LH_2}$	70.98×10^7 KWH	\$0.00323/KWH	2.29
CAPITAL INVESTMENT		$\$31.5 \times 10^6 \times \left(\frac{235\ T/D}{250\ T/D}\right)^{0.8} \times 1.125 = \$33.726 \times 10^6$	$\$33.726 \times 10^6 \times 0.11 = \$3.71 \times 10^6$		3.71
OPERATING COST		$\$1.91 \times 10^6/Y \times \left(\frac{235\ T/D}{250\ T/D}\right)^{0.65} = \$1.83 \times 10^6/Y$	$\$1.83 \times 10^6$	—	1.83
TOTAL COST	(STEAM REF.)				8.32
	(PART. OX.)				8.30
	(COAL GAS)				8.32
	(ELECTROL.)				8.05

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PROCESS, ITEM	BANGKOK	ANNUAL REQUIREMENT	ANNUAL COST $\times 10^{-6}$ (\$)
STORAGE			
CAPITAL INVESTMENT	$2 \times \$2.97 \times 10^6 \times 1.125 = \$6.683 \times 10^6$	$\$6.683 \times 10^6 \times 0.11 = \$0.74 \times 10^6$	0.74
DISTRIBUTION			
CAPITAL INVESTMENT	$2\ MI \times \frac{5300\ FT}{MI} \times 18\ IN \times \frac{\$15}{FT\ IN} \times 1.125 = \$3.220 \times 10^6$	$\$3.220 \times 10^6 \times 0.11 = \$0.35 \times 10^6$	0.35

PRODUCTION COST SUMMARY

GAS PLANT TYPE	GAS PLANT ANNUAL COST $\times 10^{-6}$ (\$)	LIQ. PLANT ANNUAL COST $\times 10^{-6}$ (\$)	STORAGE ANNUAL COST $\times 10^{-6}$ (\$)	DISTRIB. ANNUAL COST $\times 10^{-6}$ (\$)	TOTAL ANNUAL COST $\times 10^{-6}$ (\$)	PRODUCT UNIT COST (¢/#LH_2)
STEAM REFORMING	7.81	8.32	0.74	0.35	17.22	11.8
PARTIAL OXIDATION	7.60	8.30	0.74	0.35	16.99	11.6
COAL GASIFICATION	19.64	8.32	0.74	0.35	29.05	19.9
ELECTROLYSIS	17.72	8.05	0.74	0.35	26.86	18.4

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FACILITY DESCRIPTION – HONOLULU

FACILITY DESCRIPTION FOR HONOLULU IS THE SAME AS FOR BANGKOK; INFORMATION IS NOT REPEATED HERE.

PRODUCTION COST DETERMINATION

PROCESS, ITEM, UNIT REQUIREMENT	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST × 10^{-6} (\$)
STEAM REFORMING			
FEED – NATURAL GAS $1.90 \frac{\#NG}{\#GH_2} = 1.90 \frac{\#NG}{\#GH_2} \times \frac{MCF}{41.6\ \#NG} = 0.0457 \frac{MCF}{\#GH_2}$	7.57×10^6 MCF	\$0.45/MCF	3.41
FUEL – NATURAL GAS $1.44 \frac{\#NG}{\#GH_2} - 4.4 \times 10^{-5} \frac{\#NG/\#GH_2}{T/D} \times 250\ T/D$ $= 1.43 \frac{\#NG}{\#GH_2}$			
FUEL – OIL EQUIV OF NAT. GAS $1.43 \frac{\#NG}{\#GH_2} \times \frac{MCF/41.6\ \#NG}{MCF/10^6\ BTU} \times \frac{BBL/42\ GAL}{155{,}200\ BTU/GAL}$ $= 5.27 \times 10^{-3} \frac{BBL}{\#GH_2}$	0.87×10^6 BBL	\$2.25/BBL	1.96
COOLING WATER & BOILER WATER MAKE UP $(22.4 + 16.6) \frac{\#H_2O}{\#GH_2} = 39.0 \frac{\#H_2O}{\#GH_2}$	3.23×10^6 T	\$0.072/T	0.23
PROCESS ENERGY $0.47 \frac{KWH}{\#GH_2}$	77.88×10^6 KWH	\$0.00606/KWH	0.47
BYPRODUCT CO_2 $5.3 \frac{\#CO_2}{\#GH_2}$	0.44×10^6 T	—	—

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PROCESS, ITEM, UNIT REQUIREMENT — HONOLULU	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
CAPITAL INVESTMENT $\$7.65 \times 10^6 \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.7} \times 1.075 = \$8.223 \times 10^6$	$\$8.223 \times 10^6 \times 0.11$	—	0.90
OPERATING COST $\$1.28 \times 10^6 \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.73} = \$1.28 \times 10^6$	$\$1.28 \times 10^6$	—	1.28
TOTAL COST			8.25
PARTIAL OXIDATION			
FEED - OIL OR BUNKER "C" $3.89 \frac{\#\text{"C"}}{\#GH_2} = 3.89 \frac{\#\text{"C"}}{\#GH_2} \times \frac{BBL/42\ GAL}{8.40\ \#\text{"C"}/GAL}$ $= 0.011 \frac{BBL}{\#GH_2}$	1.823×10^6 BBL	$2.25/BBL	4.10
FUEL - HONOLULU OIL EQUIV. OF BUNKER "C" $0.40 \frac{\#\text{"C"}}{\#GH_2} - 3.1 \times 10^{-5} \frac{\#\text{"C"}/\#GH_2}{T/D} \times 250\ T/D$ $= 0.39 \frac{\#\text{"C"}}{\#GH_2}$ $= \frac{0.39\ \#\text{"C"}/\#GH_2}{8.108\ \#\text{"C"}/GAL\ \text{"C"}} \times \frac{152,000\ BTU/GAL\ \text{"C"}}{155,200\ BTU/GAL\ OIL} \times BBL/42\ GAL\ OIL$ $= 0.0011 \frac{BBL}{\#GH_2}$	0.182×10^6 BBL		0.41

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PROCESS, ITEM, UNIT REQUIREMENT HONOLULU	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
COOLING WATER & BOILER WATER MAKE UP $(12.5 + 12.5)\ \frac{\#H_2O}{\#GH_2} = 25.0\ \frac{\#H_2O}{\#GH_2}$	2.07×10^6 T	\$0.072/T	0.15
PROCESS ENERGY $1.30\ \frac{KWH}{\#GH_2}$	2.154×10^8 KWH	\$0.00606/KWH	1.30
BYPRODUCTS CO_2 $6.0\ \frac{\#CO_2}{\#GH_2}$	0.50×10^6 T	—	—
STEAM $5.0\ \frac{\#H_2O}{\#GH_2}$	0.41×10^6 T	—	—
C $0.11\ \frac{\#C}{\#GH_2}$	9.1×10^3 T	—	—
CAPITAL INVESTMENT $\$8.46 \times 10^6 \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.7} \times 1.075 = \$9.095 \times 10^6$	$\$9.095 \times 10^6 \times 0.11$ $= 1.00 \times 10^6$	—	1.00
OPERATING COST $\$1.46 \times 10^6/Y \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.73} = \$1.46 \times 10^6/Y$	$\$1.46 \times 10^6$	—	1.46
TOTAL COST			8.42
ELECTROLYSIS (LOW PRESSURE)			
FEED & COOLING WATER MAKE UP $(9.0 + 11.5)\ \frac{\#H_2O}{\#GH_2} = 20.5\ \frac{\#H_2O}{\#GH_2}$	1.70×10^6 T	\$0.072/T	0.12
PROCESS ENERGY $21.6\ \frac{KWH}{\#GH_2}$	3.58×10^9 KWH	\$0.00472/KWH	16.90

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PROCESS, ITEM, UNIT REQUIREMENTS HONOLULU			ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
BYPRODUCT O_2		$7.92 \frac{\# O_2}{\# GH_2}$	0.66×10^6 T	—	—
CAPITAL INVESTMENT		\$ $28.5 \times 10^6 \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.88} \times 1.075 =$ \$ 30.638×10^6	\$ $30.638 \times 10^6 \times 0.11$ = \$ 3.37×10^6	—	3.37
OPERATING COST		\$ $2.62 \times 10^6/Y \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.73} =$ \$ $2.62 \times 10^6/Y$	\$ 2.62×10^6		2.62
TOTAL COST					23.01
LIQUEFACTION					
REFRIGERANTS	N_2	(NOT REQUIRED FOR PART. OXID. FEEDSTOCK) $0.04 \frac{\# N_2}{\# LH_2}$	3.2×10^3 T	\$6.50 /T	0.02
	C_1H_4	$0.01 \frac{\# C_1}{\# GH_2} = 0.01 \frac{\# C_1}{\# GH_2} \times \frac{MCF}{41.6 \# C_1} = 24 \times 10^{-5} \frac{MCF}{\# LH_2}$	38.2×10^3 MCF	\$0.45/MCF	0.02
	C_3H_8	$0.02 \frac{\# C_3}{\# LH_2}$	1.6×10^3 T	\$25.00 /T	0.04
	C_2H_4	$0.02 \frac{\# E_2}{\# LH_2}$	1.6×10^3 T	\$80.00/T	0.13
COOLING WATER MAKE UP		$24.7 \frac{\# H_2O}{\# GH_2}$	1.97×10^6 T	\$0.072/T	0.14
PROCESS ENERGY	(HYDROCARBON FEEDSTOCK)	$4.46 \frac{KWH}{\# LH_2}$	70.98×10^7 KWH	\$0.00606/KWH	4.30
	(ELECTROLYSIS FEEDSTOCK)	$4.46 \frac{KWH}{\# LH_2}$	70.98×10^7 KWH	\$0.00472/KWH	3.35

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PROCESS, ITEM,	UNIT REQUIREMENT HONOLULU	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
CAPITAL INVESTMENT	$\$31.5 \times 10^6 \times \left(\frac{235\ T/D}{250\ T/D}\right)^{0.80} \times 1.075 = \$32.23 \times 10^6$	$\$32.23 \times 10^6 \times 0.11 = \$3.55 \times 10^6$	—	3.55
OPERATING COST	$\$1.91 \times 10^6/Y \times \left(\frac{235}{250}\right)^{0.65} = \$1.83 \times 10^6/Y$	$\$1.83 \times 10^6$	—	1.83
TOTAL COST	(STEAM REF.)			10.03
	(PART. OX.)			10.01
	(ELECTROLYSIS)			9.08
STORAGE				
CAPITAL INVESTMENT	$2 \times \$2.97 \times 10^6 \times 1.075 = \$6.386 \times 10^6$	$\$6.386 \times 10^6 \times 0.11 = \$0.70 \times 10^6$	—	0.70
DISTRIBUTION				
CAPITAL INVESTMENT	$2\ MI \times \frac{5300\ FT}{MI} \times 18\ IN \times \frac{\$.15}{FT\ IN} \times 1.075 = \$3.077 \times 10^6$	$\$3.077 \times 10^6 \times 0.11 = \$0.34 \times 10^6$	—	0.34

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PRODUCTION COST SUMMARY HONOLULU

GAS PLANT TYPE	GAS PLANT ANNUAL COST $\times 10^{-6}$ (\$)	LIQ. PLANT ANNUAL COST $\times 10^{-6}$ (\$)	STORAGE ANNUAL COST $\times 10^{-6}$ (\$)	DISTRIB. ANNUAL COST $\times 10^{-6}$ (\$)	TOTAL ANNUAL COST $\times 10^{-6}$ (\$)	PRODUCT UNIT COST (¢/#LH$_2$)
STEAM REFORMING	8.25	10.03	0.70	0.34	19.32	13.2
PARTIAL OXIDATION	8.42	10.01	0.70	0.34	19.47	13.3
ELECTROLYSIS	23.01	9.08	0.70	0.34	33.13	22.7

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FACILITY DESCRIPTION – JOHANNESBURG

FACILITY DESCRIPTION FOR JOHANNESBURG IS THE SAME AS FOR BANGKOK; INFORMATION IS NOT REPEATED HERE.

PRODUCTION COST DETERMINATION

PROCESS, ITEM, UNIT REQUIREMENT	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
STEAM REFORMING			
FEED – NATURAL GAS $1.90 \frac{\# NG}{\# GH_2} = 1.90 \frac{\# NG}{\# GH_2} \times \frac{MCF}{41.6 \# NG} = 0.0457 \frac{MCF}{\# GH_2}$	7.57×10^6 MCF	\$0.45/MCF	3.41
FUEL – NATURAL GAS $1.44 \frac{\# NG}{\# GH_2} - 4.4 \times 10^{-5} \frac{\# NG / \# GH_2}{T/D} \times 250\ T/D$ $= 1.43 \frac{\# NG}{\# GH_2}$			
FUEL – JOHAN. OIL EQUIV. OF NAT. GAS (OIL IS SAME AS FOR HONOLULU) $1.43 \frac{\# NG}{\# GH_2} \times \frac{MCF / 41.6 \# NG}{MCF / 10^6\ BTU} \times \frac{BBL / 42\ GAL}{155{,}200\ BTU/GAL}$ $= 5.27 \times 10^{-3} \frac{BBL}{\# GH_2}$	0.87×10^6 BBL	\$2.25/BBL	1.96
COOLING WATER & BOILER WATER MAKE UP $(22.4 + 16.6) \frac{\# H_2O}{\# GH_2} = 39.0 \frac{\# H_2O}{\# GH_2}$	3.23×10^6 T	\$0.012/T	0.04

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PROCESS, ITEM,	UNIT REQUIREMENT JOHNNESBURG	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
PROCESS ENERGY	$0.47 \frac{KWH}{\#GH_2}$	77.88×10^6 KWH	\$0.00280/KWH	0.22
BYPRODUCT CO_2	$5.3 \frac{\#CO_2}{\#GH_2}$	0.44×10^6 T	—	—
CAPITAL INVESTMENT	$\$7.65 \times 10^6 \left(\frac{250\ T/D}{250\ T/D}\right)^{0.7} \times 1.000 = \$7.650 \times 10^6$	$\$7.650 \times 10^6 \times 0.11$ $= \$0.84 \times 10^6$	—	0.84
OPERATING COST	$\$1.28 \times 10^6/Y \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.73} = \$1.28 \times 10^6/Y$	$\$1.28 \times 10^6$	—	1.28
TOTAL COST				7.75
PARTIAL OXIDATION				
FEED - OIL OR BUNKER "C"	$3.89 \frac{\#"C"}{\#GH_2} = 3.89 \frac{\#"C"}{\#GH_2} \times \frac{BBL/42\ GAL}{8.400\ \#"C"/GAL}$ $= 0.011 \frac{BBL}{\#GH_2}$	1.823×10^6 BBL	\$2.25/BBL	4.10
FEED - JOHAN. OIL EQUIV. OF BUNKER "C"	$0.40 \frac{\#"C"}{\#GH_2} - 3.1 \times 10^{-5} \frac{\#"C"/\#GH_2}{T/D} \times 250\ T/D$ $= 0.39 \frac{\#"C"}{\#GH_2}$ $= \frac{0.39\ \#"C"/\#GH_2}{8.108\ \#"C"/GAL"C"} \times \frac{152{,}000\ BTU/GAL"C"}{155{,}200\ BTU/GAL\ OIL} \times BBL/42\ GAL\ OIL$ $= 0.0011 \frac{BBL}{\#GH_2}$	0.182×10^6 BBL	\$2.25/BBL	0.41

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PROCESS, ITEM, UNIT REQUIREMENT JOHANNESBERG	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
COOLING WATER & BOILER WATER MAKE UP $(12.5 + 12.5) \frac{\# H_2O}{\# GH_2} = 25.0 \frac{\# H_2O}{\# GH_2}$	2.07×10^6 T	\$0.012/T	0.02
PROCESS ENERGY $1.30 \frac{KWH}{\# GH_2}$	2.154×10^8 KWH	\$0.00280/KWH	0.60
BYPRODUCTS CO_2 $6.0 \frac{\# CO_2}{\# GH_2}$	0.50×10^6 T	——	——
STEAM $5.0 \frac{\# H_2O}{\# GH_2}$	0.41×10^6 T	——	——
C $0.11 \frac{\# C}{\# GH_2}$	9.1×10^3 T	——	——
CAPITAL INVESTMENT $\$8.46 \times 10^6 \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.70} \times 1.000 = \$8.460 \times 10^6$	$\$8.460 \times 10^6 \times 0.11$ $= \$0.93 \times 10^6$	——	0.93
OPERATING COST $\$1.46 \times 10^6/Y \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.73} = \$1.46 \times 10^6/Y$	$\$1.460 \times 10^6$	——	1.46
TOTAL COST			7.52
COAL GASIFICATION (BITUM.)			
FEED - WATER $20.3 \frac{\# H_2O}{\# GH_2}$	1.682×10^6 T	\$0.012/T	0.02
FUEL - COAL EQUIV OF W. VA. BITUM. $8.74 \frac{\# W.VA.}{\# GH_2} = 8.74 \frac{\# W.VA.}{\# GH_2} \times \frac{14{,}290\ BTU/\# W.VA.}{9{,}500\ BTU/\# COAL}$ $= 13.15 \frac{\# COAL}{\# GH_2}$	1.090×10^6 T	\$2.40/T	2.62

NATIONAL 42-381 50 SHEETS 42-382 100 SHEETS Made in U.S.A.

PROCESS, ITEM, UNIT REQUIREMENT	JOHANNESBURG	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
OXYGEN	$5.20 \frac{\# O_2}{\# 6H_2}$	0.431×10^6 T	\$5.00 /T	2.16
COOLING WATER MAKE UP	$26.0 \frac{\# H_2O}{\# 6H_2}$	2.154×10^6 T	\$0.012 /T	0.03
PROCESS ENERGY	$0.57 \frac{KWH}{\# 6H_2}$	94.95×10^6 KWH	\$0.00281/KWH	0.27
BYPRODUCT CO₂	$24.0 \frac{\# CO_2}{\# 6H_2}$	1.99×10^6 T	—	—
CAPITAL INVESTMENT	$\$20.8 \times 10^6 \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.7} \times 1.000 = \$20.800 \times 10^6$	$\$20800 \times 10^6 \times 0.11$ $= \$2.29 \times 10^6$	—	2.29
OPERATING COST	$\$2.89 \times 10^6/Y \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.69} = \$2.890 \times 10^6/Y$	$\$2.890 \times 10^6$	—	2.89
TOTAL COST				10.28
ELECTROLYSIS (LOW PRESSURE)				
FEED & COOLING WATER MAKE UP	$(9.0 + 11.5) \frac{\# H_2O}{\# 6H_2} = 20.5 \frac{\# H_2O}{\# 6H_2}$	1.70×10^6 T	\$0.012 /T	0.02
PROCESS ENERGY	$21.6 \frac{KWH}{\# 6H_2}$	3.58×10^9 KWH	\$0.00276/KWH	9.88
BYPRODUCT O₂	$7.92 \frac{\# O_2}{\# 6H_2}$	0.66×10^6 T	—	—
CAPITAL INVESTMENT	$\$28.5 \times 10^6 \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.88} \times 1.000 = \$28.500 \times 10^6$	$\$28.500 \times 10^6 \times 0.11$ $= \$3.14 \times 10^6$	—	3.14

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PROCESS, ITEM, UNIT REQUIREMENT JOHANNESBURG	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST × 10^{-6} (\$)
OPERATING COSTS $\$2.62 \times 10^6/Y \times \left(\frac{250\ T/D}{250\ T/D}\right)^{0.73} = \$2.62 \times 10^6/Y$	$\$2.62 \times 10^6$	—	2.62
TOTAL COST			15.66
LIQUEFACTION			
REFRIGERANTS N_2 (NOT REQ'D FOR PART. OXID. FEEDSTOCK) $0.04 \frac{\# N_2}{\# LH_2}$	3.2×10^3 T	\$6.50/T	0.02
C_1H_4 $0.01 \frac{\# C_1}{\# LH_2} = 0.01 \frac{\# C_1}{\# LH_2} \times \frac{MCF}{41.6\ \# C_1} = 2.4 \times 10^{-5} \frac{MCF}{\# LH_2}$	38.2×10^3 MCF	\$0.45/MCF	0.02
C_3H_8 $0.02 \frac{\# C_3}{\# LH_2}$	1.6×10^3 T	\$25.00/T	0.04
C_2H_4 $0.02 \frac{\# E_2}{\# LH_2}$	1.6×10^3 T	\$80.00/T	0.13
COOLING WATER MAKE UP $24.7 \frac{\# H_2O}{\# LH_2}$	1.97×10^6 T	\$0.012/T	0.02
PROCESS ENERGY (HYDROCARBON, COAL GAS FEEDSTOCK) $4.46 \frac{KWH}{\# LH_2}$	70.98×10^7 KWH	\$0.00281/KWH	1.99
(ELECTROLYSIS FEEDSTOCK) $4.46 \frac{KWH}{\# LH_2}$	70.98×10^7 KWH	\$0.00276/KWH	1.96
CAPITAL INVESTMENT $\$31.5 \times 10^6 \times \left(\frac{235\ T/D}{260\ T/D}\right)^{0.80} \times 1.000 = \$29.979 \times 10^6$	$\$29.979 \times 10^6 \times 0.11$ $= 3.30 \times 10^6$	—	3.30
OPERATING COST $\$1.91 \times 10^6/Y \times \left(\frac{235\ T/D}{260\ T/D}\right)^{0.65} = \$1.83 \times 10^6/Y$	$\$1.83 \times 10^6$	—	1.83
TOTAL COST (STEAM REF.)			7.35

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Process, Item,	Unit Requirement JOHANNESBURG	Annual Requirement	Unit Cost	Annual Cost $\times 10^{-6}$ (\$)
Total Cost	(Partial Oxid.)			7.33
	(Coal Gasif.)			7.35
	(Electrolysis)			7.32
Storage				
Capital Investment	$2 \times \$2.97 \times 10^6 \times 1.000 = \$5.940 \times 10^6$	$\$5.940 \times 10^6 \times 0.11$ $= \$0.65 \times 10^6$	—	0.65
Distribution				
Capital Investment	$2 \text{ Mi} \times \frac{5300 \text{ Ft}}{\text{Mi}} \times 18 \text{ In} \times \frac{\$15}{\text{Ft In}} \times 1.000$ $= \$2.862 \times 10^6$	$\$2.862 \times 10^6 \times 0.11$ $= \$0.31 \times 10^6$	—	\$ 0.31

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PRODUCTION COST SUMMARY JOHANNESBURG

GAS PLANT TYPE	GAS PLANT ANNUAL COST $\times 10^{-6}$ (\$)	LIQ. PLANT ANNUAL COST $\times 10^{-6}$ (\$)	STORAGE ANNUAL COST $\times 10^{-6}$ (\$)	DISTRIB. ANNUAL COST $\times 10^{-6}$ (\$)	TOTAL ANNUAL COST $\times 10^{-6}$ (\$)	PRODUCT UNIT COST (¢/#LH_2)
STEAM REFORMING	7.75	7.35	0.65	0.31	16.06	11.0
PARTIAL OXIDATION	7.52	7.33	0.65	0.31	15.81	10.8
COAL GASIFICATION (BITUM.)	10.28	7.35	0.65	0.31	18.59	12.7
ELECTROLYSIS (LOW PRESSURE)	15.66	7.32	0.65	0.31	23.94	16.4

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42-382 100 SHEETS
Made in U.S.A.

FACILITY DESCRIPTION – BEIRUT

DISTRIBUTION.

1 - 18" NPS × 2 MILE, SI VJ LINE (ALLOWS 1 AIRCRAFT LOADING AT A TIME)

PRODUCT DELIVERED TO MANIFOLD IS 300 T/D. PRODUCT RECEIVED FROM STORAGE IS 323 T/D, I.E. DISTRIBUTION LOSSES ARE 23 T/D.

STORAGE

2 - 1500 T NOMINAL SIZE, DOUBLE WALL, EVALUATED PERLITE INSULATED TANKS.

PRODUCT DISCHAGED FOR DISTRIBUTION IS 323 T/D. PRODUCT RECIEVED FROM THE LIQUEFIER IS 328 T/D, I.E. STORAGE LOSSES TOTAL TO 5 T/D.

LIQUEFACTION AND CONVERSION

1 - 355 T/D NOMINAL SIZE UNIT WITH 97% ON STREAM FACTOR (355 T/D SIZE MATCHES REQUIRED CAPACITY NECESSARY TO PROCESS GAS PLANT EFFLUENT)

DAILY AVERAGE PRODUCTION IS 328 T/D.

GAS PRODUCTION AND PURIFICATION

1 - 250 T/D NOMINAL SIZE UNIT WITH 93% ON STREAM FACTOR.

1 - 125 T/D NOMINAL SIZE UNIT WITH 93% ON STREAM FACTOR.

(NOMINAL SIZE OF GAS PLANT IS TAKEN AS 375 T/D.)

DAILY AVERAGE PRODUCTION IS 341 T/D.

PRODUCTION COST DETERMINATION

PROCESS, ITEM, UNIT REQUIREMENT	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
STEAM REFORMING			
FEED - NATURAL GAS $1.90 \frac{\#NG}{\#GH_2} = 1.90 \frac{\#NG}{\#GH_2} \times \frac{MCF}{41.6 \#NG} = 0.0457 \frac{MCF}{\#GH_2}$	11.376×10^6 MCF	$0.35/MCF	3.98

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42-382 100 SHEETS
Made in U.S.A.

PROCESS, ITEM, UNIT REQUIREMENT BEIRUT	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
FUEL - NATURAL GAS $1.44 \frac{\#NG}{\#GH_2} - 4.4 \times 10^{-5} \frac{\#NG/\#GH_2}{T/D} \times 375\ T/D$ $= 1.42 \frac{\#NG}{\#GH_2}$			
FUEL - BEIRUT OIL EQUIV OF NAT. GAS $1.42 \frac{\#NG/\#GH_2}{41.6\ \#NG/MCF} \times \frac{10^6\ BTU/MCF}{157{,}300\ BTU/GAL} \times \frac{BBL}{42\ GAL}$ $= 51.7 \times 10^{-4} \frac{BBL}{\#GH_2}$	1.287×10^6 BBL	$\$1.74$/BBL	2.24
COOLING WATER & BOILER WATER MAKE UP $(22.4 + 16.6) \frac{\#H_2O}{\#GH_2} = 39.0 \frac{\#H_2O}{\#GH_2}$	4.854×10^6 T	$\$0.072$/T	0.35
PROCESS ENERGY $0.47 \frac{KWH}{\#GH_2}$	1.170×10^8 KWH	$\$0.00469$/KWH	0.55
BYPRODUCT CO_2 $5.30 \frac{\#CO_2}{\#GH_2}$	0.66×10^6 T	—	—
CAPITAL INVESTMENT $\$7.65 \times 10^6 \times \left(\frac{375\ T/D}{250\ T/D}\right)^{0.7} \times 1.050 = \$10.669 \times 10^6$	$\$10.669 \times 10^6 \times 0.11$ $= \$1.17 \times 10^6$	—	1.17
OPERATING COST $\$1.28 \times 10^6/Y \times \left(\frac{375\ T/D}{250\ T/D}\right)^{0.73} = \$1.72 \times 10^6/Y$	$\$1.72 \times 10^6$	—	1.72
TOTAL COST			10.01

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42-382 100 SHEETS
Made in U.S.A.

PROCESS, ITEM, UNIT REQUIREMENT BEIRUT		ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
PARTIAL OXIDATION				
FEED - OIL OR BUNKER "C"	$3.89 \frac{\#\,OIL}{\#GH_2} = 3.89 \frac{\#\,OIL}{\#GH_2} \times \frac{BBL / 42\ GAL}{8.570\ \#OIL/GAL}$ $= 0.0108 \frac{BBL}{\#GH_2}$	2.688×10^6 BBL	\$1.74 /BBL	4.68
FUEL - BEIRUT OIL EQUIV OF BUNKER "C"	$0.40 \frac{\#\text{"C"}}{\#GH_2} - 3.1 \times 10^{-5} \frac{\#\text{"C"}/\#GH_2}{T/D} \times 375\ T/D$ $= 0.388 \frac{\#\text{"C"}}{\#GH_2}$ $= \frac{0.388\ \#\text{"C"}/\#GH_2}{8.108\ \#\text{"C"}/GAL\text{"C"}} \times \frac{152{,}000\ BTU/GAL\text{"C"}}{157{,}300\ BTU/GAL}$ $\times \frac{BBL}{42\ GAL} = 0.0011 \frac{BBL}{\#GH_2}$	0.274×10^6 BBL	\$1.74 /BBL	0.48
COOLING WATER & BOILER MAKE UP WATER	$(12.5 + 12.5) \frac{\#H_2O}{\#GH_2} = 25.0 \frac{\#H_2O}{\#GH_2}$	3.112×10^6 T	\$0.072 /T	0.22
PROCESS ENERGY	$1.30 \frac{KWH}{\#GH_2}$	3.236×10^8 KWH	\$0.00469/KWH	1.52
BYPRODUCTS	CO_2 $6.0 \frac{\#CO_2}{\#GH_2}$	0.75×10^6 T	—	—
	STEAM $5.0 \frac{\#H_2O}{\#GH_2}$	0.62×10^6 T	—	—
	C $0.11 \frac{\#C}{\#GH_2}$	13.7×10^3 T	—	—
CAPITAL INVESTMENT	$\$8.46 \times 10^6 \times \left(\frac{375}{250}\right)^{0.7} \times 1.050 = \$11.798 \times 10^6$	$\$11.798 \times 10^6 \times 0.11$ $= \$1.30 \times 10^6$		1.30

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PROCESS, ITEM, UNIT REQUIREMENT BEIRUT	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
OPERATING COST $\$1.46 \times 10^6/Y \times \left(\frac{375\ T/D}{250\ T/D}\right)^{0.73} = \$1.96 \times 10^6/Y$	$\$1.96 \times 10^6$		1.96
TOTAL COST			10.16
ELECTROLYSIS (LO PRESSURE)			
FEED & COOLING WATER $(9.0 + 11.5) \frac{\#H_2O}{\#GH_2} = 20.5 \frac{\#H_2O}{\#GH_2}$	2.551×10^6 T	\$0.072 /T	0.18
PROCESS ENERGY $21.6 \frac{KWH}{\#GH_2}$	5.377×10^9 KWH	\$0.00426/KWH	22.91
BYPRODUCT O_2 $7.92 \frac{\#O_2}{\#GH_2}$	0.99×10^6 T	—	—
CAPITAL INVESTMENT $\$28.5 \times 10^6 \times \left(\frac{375\ T/D}{250\ T/D}\right)^{0.88} \times 1.050 = \$42.763 \times 10^6$	$\$42.763 \times 10^6 \times 0.11$ $= \$4.70 \times 10^6$	—	4.70
OPERATING COST $\$2.62 \times 10^6/Y \times \left(\frac{375\ T/D}{250\ T/D}\right)^{0.73} = \$3.52 \times 10^6/Y$	$\$3.52 \times 10^6$	—	3.52
TOTAL COST			31.31
LIQUEFACTION			
REFRIGERANTS N_2 (NOT REQUIRED FOR PART. OX. FEEDSTOCK) $0.04 \frac{\#N_2}{\#LH_2}$	4.79×10^3 T	\$6.50 /T	0.03
C_1H_4 $0.01 \frac{\#C_1}{\#LH_2} = 0.01 \frac{\#C_1}{\#LH_2} \times \frac{MCF}{41.6\ \#C_1} = 24 \times 10^{-5} \frac{MCF}{\#LH_2}$	57.5×10^3 MCF	\$0.35 /MCF	0.02

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PROCESS, ITEM, UNIT REQUIREMENT **BEIRUT**	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
REFRIGERANTS (CONT'D.) C_2H_6 0.02 $\frac{\# C_2}{\# LH_2}$	2.39×10^3 T	\$25.00/T	0.06
C_2H_4 0.02 $\frac{\# E_L}{\# LH_2}$	2.39×10^3 T	\$80.00/T	0.19
COOLING WATER MAKE UP 24.7 $\frac{\# H_2O}{\# LH_2}$	2.957×10^6 T	\$0.072/T	0.21
PROCESS ENERGY (HYDROCARBON FEEDSTOCK) 4.46 $\frac{KWH}{\# LH_2}$	1.068×10^9 KWH	\$0.00469/KWH	5.01
(ELECTROLYSIS FEEDSTOCK) 4.46 $\frac{KWH}{\# LH_2}$	1.068×10^9 KWH	\$0.00426/KWH	4.55
CAPITAL INVESTMENT $\$31.5 \times 10^6 \left(\frac{355\ T/D}{250\ T/D}\right)^{0.8} \times 1.050 = \$43.791 \times 10^6$	$\$43.791 \times 10^6 \times 0.11 = \$4.82 \times 10^6$	——	4.82
OPERATING COST $\$\ 1.91 \times 10^6/Y \left(\frac{355\ T/D}{250\ T/D}\right)^{0.65} = \$2.40 \times 10^6/Y$	$\$2.40 \times 10^6$		2.40
TOTAL COST (STEAM REF.)			12.74
(PART. OXID.)			12.71
(ELECTROLYSIS)			12.28
STORAGE			
CAPITAL INVESTMENT $2 \times \$3.15 \times 10^6 \times 1.050 = \$6.615 \times 10^6$	$\$6.615 \times 10^6 \times 0.11 = \$0.73 \times 10^6$	——	0.73

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PROCESS, ITEM	BEIRUT	ANNUAL REQUIREMENT	ANNUAL COST $\times 10^{-6}$ ($)
DISTRIBUTION			
CAPITAL INVESTMENT	$2\ \text{MI} \times \frac{5300\ \text{FT}}{\text{MI}} \times 18\ \text{IN} \times \frac{\$15}{\text{FT IN}} \times 1.050 = \$3.005 \times 10^6$	$\$3.005 \times 10^6 \times 0.11 = \$0.33 \times 10^6$	0.33

DUCTION COST SUMMARY

AS PLANT TYPE	GAS PLANT ANNUAL COST $\times 10^{-6}$ ($)	LIQ. PLANT ANNUAL COST $\times 10^{-6}$ ($)	STORAGE ANNUAL COST $\times 10^{-6}$ ($)	DISTRIB. ANNUAL COST $\times 10^{-6}$ ($)	TOTAL ANNUAL COST $\times 10^{-6}$ ($)	PRODUCT UNIT COST (¢/#LH$_2$)
TEAM REFORMING	10.01	12.74	0.73	0.33	23.81	10.9
ARTIAL OXIDATION	10.16	12.71	0.73	0.33	23.93	10.9
LECTROLYSIS	31.31	12.28	0.73	0.33	44.65	20.4

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42-382 100 SHEETS
Made in U.S.A.

FACILITY DESCRIPTION – SAO PAULO

DISTRIBUTION

1 - 18" NPS x 2 MILE, SI - VJ LINE (ALLOWS 1 AIRCRAFT LOADING AT A TIME)
PRODUCT DELIVERED TO MANIFOLD IS 400 T/D. PRODUCT RECEIVED FROM STORAGE IS 430 T/D,
I.E. DISTRIBUTION LOSSES ARE 30 T/D.

STORAGE

2 - 2000 T NOMINAL SIZE, DOUBLE WALL, EVACUATED PERLITE INSULATED TANKS.
PRODUCT DISCHARGED FOR DISTRIBUTION IS 430 T/D. PRODUCT RECEIVED FROM THE LIQUEFIER IS 436 T/D, I.E. STORAGE LOSSES TOTAL TO 6 T/D.

LIQUEFACTION AND CONVERSION

1 - 470 T/D NOMINAL SIZE UNIT WITH 97% ON STREAM FACTOR (470 T/D SIZE MATCHES) MATCHES REQUIRED CAPACITY TO PROCESS GAS PLANT EFFLUENT)
DAILY AVERAGE PRODUCTION IS 436 T/D.

GAS PRODUCTION AND PURIFICATION

2 - 250 T/D NOMINAL SIZE UNITS WITH 93% ON STREAM FACTOR. (NOMINAL SIZE OF GAS PLANT IS TAKEN AS 500 T/D)
DAILY AVERAGE PRODUCTION IS 453 T/D.

PRODUCTION COST DETERMINATION

PROCESS, ITEM, UNIT REQUIREMENT	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
STEAM REFORMING			
FEED - NATURAL GAS $1.90 \frac{\#NG}{\#GH_2} = 1.90 \frac{\#NG}{\#GH_2} \times \frac{MCF}{41.6\ \#NG} = 0.0457 \frac{MCF}{\#GH_2}$	15.113×10^6 MCF	$0.40 /MCF	6.04

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PROCESS, ITEM, UNIT REQUIREMENT SAO PAULO	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
FUEL - NATURAL GAS $1.44 \frac{\# NG}{\# GH_2} - 4.4 \times 10^{-5} \frac{\# NG/\# GH_2}{T/D} \times 500\ T/D$ $= 1.42 \frac{\# NG}{\# GH_2}$			
FUEL - SAO PAULO OIL EQUIV. OF NAT. GAS. $1.42 \frac{\# NG/\# GH_2}{41.6\ \# NG/MCF} \times \frac{10^6\ BTU/MCF}{151{,}300\ BTU/GAL} \times \frac{BBL}{42\ GAL}$ $= 53.7 \times 10^{-4}\ BBL / \# GH_2$	1.776×10^6 BBL	\$2.11 / BBL	3.75
COOLING WATER & BOILER WATER MAKE UP $(22.4 + 16.6) \frac{\# H_2O}{\# GH_2} = 39.0 \frac{\# H_2O}{\# GH_2}$	6.448×10^6 T	\$0.0072 /T	0.05
PROCESS ENERGY $0.47 \frac{KWH}{\# GH_2}$	1.554×10^8 KWH	\$0.00300 /KWH	0.47
BY PRODUCT CO_2 $5.30 \frac{\# CO_2}{\# GH_2}$	0.88×10^6 T	———	———
CAPITAL INVESTMENT $\$7.65 \times 10^6 \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.7} \times 1.250 = \$15.534 \times 10^6$	$\$15.534 \times 10^6 \times 0.11$ $= \$1.71 \times 10^6$	———	1.71
OPERATING COST $\$1.28 \times 10^6/Y \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.73} = \$2.12 \times 10^6/Y$	$\$2.12 \times 10^6$		2.12
TOTAL COST			14.14

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PROCESS, ITEM, UNIT REQUIREMENT SAO PAULO	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
FEED – OIL OR BUNKER "C" $3.89 \frac{\#\text{OIL}}{\#GH_2} = 3.89 \frac{\#\text{OIL}}{\#GH_2} \times \frac{BBL/42\,GAL}{8.055\ \#\text{OIL}/GAL}$ $= 0.0115 \frac{BBL}{\#GH_2}$	3.803×10^6 BBL	\$2.11 /BBL	8.02
FUEL – SAO PAULO OIL EQUIV OF BUNKER "C" $0.40 \frac{\#\text{"C"}}{\#GH_2} - 3.1 \times 10^{-5} \frac{\#\text{"C"}/\#GH_2}{T/D} \times 500\ T/D = 0.385 \frac{\#\text{"C"}}{\#GH_2}$ $= 0.385 \frac{\#\text{"C"}/\#GAL}{8.100\ \#\text{"C"}/GAL\ \text{"C"}} \times \frac{152{,}000\ BTU/GAL\ \text{"C"}}{151{,}300\ BTU/GAL} \times \frac{BBL}{42\ GAL}$ $= 0.0011 \frac{BBL}{\#GH_2}$	0.364×10^6 BBL	\$2.11 /BBL	0.77
COOLING WATER & BOILER MAKE UP WATER $(12.5 + 12.5) \frac{\#H_2O}{\#GH_2} = 25.0 \frac{\#H_2O}{\#GH_2}$	4.134×10^6 T	\$0.0072 /T	0.03
PROCESS ENERGY $1.30 \frac{KWH}{\#GH_2}$	4.299×10^8 KWH	\$0.00300 /KWH	1.29
BYPRODUCTS CO₂ $6.0 \frac{\#CO_2}{\#GH_2}$	0.99×10^6 T	—	—
STEAM $5.0 \frac{\#H_2O}{\#GH_2}$	0.83×10^6 T	—	—
C $0.11 \frac{\#C}{\#GH_2}$	18.2×10^3 T	—	—
CAPITAL INVESTMENT $\$8.46 \times 10^6 \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.7} \times 1.250 = \$17.179 \times 10^6$	$\$17.179 \times 10^6 \times 0.11$ $= \$1.89 \times 10^6$	—	1.89
OPERATING COST $\$1.46 \times 10^6\ Y / \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.73} = \$2.42 \times 10^6/Y$	$\$2.42 \times 10^6$		2.42
TOTAL COST			14.42

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PROCESS, ITEM, UNIT REQUIREMENT SAO PAULO	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
ELECTROYSIS (LOW PRESSURE)			
FEED & COOLING WATER $(9.0 + 11.5)\frac{\#H_2O}{\#GH_2} = 20.5 \frac{\#H_2O}{\#GH_2}$	3.390×10^6 T	\$0.0072 /T	0.02
PROCESS ENERGY $21.6 \frac{KWH}{\#G\ H_2}$	7.143×10^9 KWH	\$0.00281/KWH	20.07
BYPRODUCT O_2 $7.92 \frac{\#O_2}{\#GH_2}$	1.31×10^6 T	——	——
CAPITAL INVESTMENT $\$28.5 \times 10^6 \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.88} \times 1.250 = \$65.586 \times 10^6$	$\$65.586 \times 10^6 \times 0.11 = \$7.21 \times 10^6$	——	7.21
OPERATING COST $\$2.62 \times 10^6/Y \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.73} = \$4.35 \times 10^6/Y$	$\$4.35 \times 10^6$	——	4.35
TOTAL COST			31.65
LIQUEFACTION			
REFRIGERANTS N_2 (NOT REQ'D FOR PART. OXID. FEEDSTOCK) $0.04 \frac{\#N_2}{\#LH_2}$	6.37×10^3 T	\$6.50/T	0.04
C_1H_4 $0.01 \frac{\#C_1}{\#GH_2} = 0.01 \frac{\#C_1}{\#GH_2} \times \frac{MCF}{41.6\ \#C_1} = 24 \times 10^{-5} \frac{MCF}{\#LH_2}$	76.4×10^3 MCF	\$0.40 /MCF	0.03
C_3H_8 $0.02 \frac{\#C_3}{\#LH_2}$	3.18×10^3 T	\$25.00 /T	0.08
C_2H_4 $0.02 \frac{\#E_2}{\#LH_2}$	3.18×10^3 T	\$80.00 /T	0.25
COOLING WATER MAKE UP $24.7 \frac{\#H_2O}{\#LH_2}$	3.931×10^6 T	\$0.0072 /T	0.03

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PROCESS, ITEM, UNIT CONSUMPTION	SAO PAULO	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
PROCESS ENERGY	(HYDROCARBON FEEDSTOCK) $4.46 \frac{KWH}{\# LH_2}$	1.420×10^9 KWH	\$0.00300/KWH	4.26
	(ELECTROLYSIS FEEDSTOCK) $4.46 \frac{KWH}{\# LH_2}$	1.420×10^9 KWH	\$0.00281/KWH	3.99
CAPITAL INVESTMENT	$\$31.5 \times 10^6 \times \left(\frac{470\ T/D}{250\ T/D}\right)^{0.8} \times 1.250 = \$65.244 \times 10^6$	$\$67.449 \times 10^6 \times 0.11 = \$7.18 \times 10^6$	—	7.18
OPERATING COST	$\$1.91 \times 10^6/Y \times \left(\frac{470\ T/D}{250\ T/D}\right)^{0.65} = \$2.88 \times 10^6/Y$	$\$2.88 \times 10^6$	—	2.88
TOTAL COST	(STEAM REF.)			14.75
	(PART. OXID)			14.71
	(ELECTROLYSIS)			14.48
STORAGE				
CAPITAL INVESTMENT	$2 \times 4.150 \times 10^6 \times 1.250 = \$10.375 \times 10^6$	$\$10.375 \times 10^6 \times 0.11 = \$1.14 \times 10^6$	—	1.14
DISTRIBUTION				
CAPITAL INVESTMENT	$2\ MI \times \frac{5300\ FT}{MI} \times 18\ IN \times \frac{\$15}{FT\ IN} \times 1.250 = \$3.578 \times 10^6$	$\$3.578 \times 10^6 \times 0.11 = \$0.39 \times 10^6$	—	0.39

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PRODUCTION COST SUMMARY SAO PAULO

GAS PLANT TYPE	GAS PLANT ANNUAL COST x 10^{-6} ($)	LIQ. PLANT ANNUAL COST x 10^{-6} ($)	STORAGE ANNUAL COST x 10^{-6} ($)	DISTRIB. ANNUAL COST x 10^{-6} ($)	TOTAL ANNUAL COST x 10^{-6} ($)	PRODUCT UNIT COST (¢/lb LH$_2$)
STEAM REFORMING	14.14	14.75	1.14	0.39	30.42	10.4
PARTIAL OXIDATION	14.42	14.71	1.14	0.39	30.66	10.5
ELECTROLYSIS	31.65	14.48	1.14	0.39	47.66	16.3

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FACILITY DESCRIPTION – SYDNEY

FACILITY DESCRIPTION FOR SYDNEY IS THE SAME AS FOR SAO PAULO; INFORMATION IS NOT REPEATED HERE.

PRODUCTION COST DETERMINATION

PROCESS, ITEM, UNIT REQUIREMENT	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
STEAM REFORMING			
FEED – NATURAL GAS $1.90 \frac{\#NG}{\#GH_2} = 1.90 \frac{\#NG}{\#GH_2} \times \frac{MCF}{41.6\ \#NG} = 0.0457 \frac{MCF}{\#GH_2}$	15.113×10^6 MCF	\$0.35/MCF	5.29
FUEL – NATURAL GAS $1.44 \frac{\#NG}{\#GH_2} - 1.4 \times 10^{-5} \frac{\#NG/\#GH_2}{T/D} \times 500\ T/D$ $= 1.42 \frac{\#NG}{\#GH_2}$			
FUEL – SYDNEY OIL EQUIV OF NAT. GAS $1.42 \frac{\#NG/\#GH_2}{41.6\ \#NG/MCF} \times \frac{10^6\ BTU/MCF}{138{,}800\ BTU/GAL} \times \frac{BBL}{42\ GAL}$ $= 58.6 \times 10^{-4} \frac{BBL}{\#GH_2}$	1.938×10^6 BBL	\$2.12/BBL	4.11
COOLING WATER & BOILER MAKE UP WATER $(22.4 + 16.6) \frac{\#H_2O}{\#GH_2} = 39.0 \frac{\#H_2O}{\#GH_2}$	6.448×10^6 T	\$0.012/T	0.08
PROCESS ENERGY $0.47 \frac{KWH}{\#GH_2}$	1.554×10^8 KWH	\$0.00312/KWH	0.48
BYPRODUCT CO_2 $5.30 \frac{\#CO_2}{\#GH_2}$	0.88×10^6 T	———	———

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Process, Item, Unit Requirement SYDNEY	Annual Requirement	Unit Cost	Annual Cost $\times 10^{-6}$ (\$)
Capital Investment $\$7.65 \times 10^6 \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.7} \times 1.050 = \$13.049 \times 10^6$	$\$13.049 \times 10^6 \times 0.11 = \$1.44 \times 10^6$	——	1.44
Operating Cost $\$1.28 \times 10^6 / Y \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.73} = \$2.12 \times 10^6 / Y$	$\$2.12 \times 10^6$		2.12
Total Cost			13.52
Partial Oxidation			
Feed - Oil or Bunker "C" $3.89 \frac{\#OIL}{\#GH_2} = 3.89 \frac{\#OIL}{\#GH_2} \times \frac{BBL/42\ GAL}{7.085\ \#OIL/GAL}$ $= 0.0131 \frac{BBL}{\#GH_2}$	4.332×10^6 BBL	\$2.12/BBL	9.18
Fuel - Sydney oil equiv of Bunker "C" $0.40 \frac{\#\text{"C"}}{\#GH_2} - 3.1 \times 10^{-5} \frac{\#\text{"C"}/\#GH_2}{T/D} \times 500\ T/D = 0.385 \frac{\#\text{"C"}}{\#GH_2}$ $0.385 \frac{\#\text{"C"}/\#GH_2}{8.108\ \#\text{"C"}/GAL} \times \frac{152{,}000\ BTU/GAL\ \text{"C"}}{138{,}800\ BTU/GAL} \times \frac{BBL}{42\ GAL}$ $= 0.0012 \frac{BBL}{\#GH_2}$	0.397×10^6 BBL	\$2.12/BBL	0.84
Cooling Water & Boiler Make Up Water $(12.5 + 12.5) \frac{\#H_2O}{\#GH_2} = 25.0 \frac{\#H_2O}{\#GH_2}$	4.134×10^6 T	\$0.012/T	0.05
Byproducts CO_2 $6.0 \frac{\#CO_2}{\#GH_2}$	0.99×10^6 T	——	——
Steam $6.0 \frac{\#H_2O}{\#GH_2}$	0.83×10^6 T	——	——
C $0.11 \frac{\#C}{\#GH_2}$	18.2×10^3 T	——	——

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PROCESS, ITEM, UNIT REQUIREMENT	SYDNEY	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
CAPITAL INVESTMENT	$\$8.46 \times 10^6 \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.7} \times 1.050 = \$14.430 \times 10^6$	$\$14.430 \times 10^6 \times 0.11 = \$1.59 \times 10^6$	——	1.59
OPERATING COST	$\$1.46 \times 10^6/Y \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.73} = \$2.42 \times 10^6/Y$	$\$2.42 \times 10^6$	——	2.42
PROCESS ENERGY	$1.30 \frac{KWH}{\# GH_2}$	4.299×10^8 KWH	\$0.00312/KWH	1.34
TOTAL COST				15.42
COAL GASIFICATION (BITUM.)				
FEED WATER & COOLING WATER MAKE UP	$(20.3 + 26.0) \frac{\# H_2O}{\# GH_2} = 46.3 \frac{\# H_2O}{\# GH_2}$	7.655×10^6 T	\$0.012/T	0.09
FUEL - SYDNEY COAL EQUIV OF W. VA. BITUM.	$8.74 \frac{\# W.VA}{\# GH_2} = 8.74 \frac{\# W.VA.}{\# GH_2} \times \frac{14{,}290\ BTU/\# W.VA}{11{,}100\ BTU/\# COAL} = 11.25 \frac{\# COAL}{\# GH_2}$	1.860×10^6 T	\$3.00/T	5.58
OXYGEN	$5.20 \frac{\# O_2}{\# GH_2}$	0.860×10^6 T	\$4.00/T	3.44
PROCESS ENERGY	$0.57 \frac{KWH}{\# GH_2}$	1.885×10^8 KWH	\$0.00312/KWH	0.59
BYPRODUCT CO_2	$24.0 \frac{\# CO_2}{\# GH_2}$	3.97×10^6 T	——	——
CAPITAL INVESTMENT	$\$20.8 \times 10^6 \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.7} \times 1.050 = \$35.479 \times 10^6$	$\$35.479 \times 10^6 \times 0.11 = \$3.90 \times 10^6$	——	3.90

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PROCESS, ITEM, UNIT PROCUREMENT	SYDNEY	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST x10^{-6} (\$)
OPERATING COST	$\$2.89 \times 10^6/Y \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.65} = \$4.53 \times 10^6/Y$	$\$4.53 \times 10^6$	—	4.53
TOTAL COST				18.13
ELECTROLYSIS (LOW PRESSURE)				
FEED & COOLING WATER	$(9.0 + 11.5)\frac{\#H_2O}{\#GH_2} = 20.5\ \frac{\#H_2O}{\#GH_2}$	$3.390 \times 10^6\ T$	\$0.012 /T	0.04
PROCESS ENERGY	$21.6\ \frac{KWH}{\#GH_2}$	7.143×10^9 KWH	\$0.00249 /KWH	17.79
BYPRODUCT O_2	$7.92\ \frac{\#O_2}{\#GH_2}$	$1.31 \times 10^6\ T$	—	—
CAPITAL INVESTMENT	$\$28.5 \times 10^6 \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.88} \times 1.050 = \$55.092 \times 10^6$	$\$55.092 \times 10^6 \times 0.11 = \$6.06 \times 10^6$	—	6.06
OPERATING COST	$\$2.62 \times 10^6/Y \times \left(\frac{500\ T/D}{250\ T/D}\right)^{0.73} = \$4.35 \times 10^6/Y$	$\$4.35 \times 10^6$	—	4.35
TOTAL COST				28.24
LIQUEFACTION				
REFRIGERANTS N_2	(NOT REQ'D FOR PART. OXID. FEEDSTOCK) $0.04\ \frac{\#N_2}{\#LH_2}$	$6.37 \times 10^6\ T$	\$6.50 /T	0.04
C_1H_4	$0.01\ \frac{\#C_1}{\#LH_2} = 0.01\ \frac{\#C_1}{\#LH_2} \times \frac{MCF}{41.6\ \#C_1} = 2.4 \times 10^{-5}\ \frac{MCF}{\#LH_2}$	76.4×10^3 MCF	\$0.35/MCF	0.03
C_3H_8	$0.02\ \frac{\#C_3}{\#LH_2}$	$3.18 \times 10^3\ T$	\$25.00 /T	0.08

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PROCESS, ITEM, UNIT REQUIREMENT	SYDNEY	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
REFRIGERANTS CONT'D.	C_2H_4 $0.02 \frac{\#E_2}{\#LH_2}$	3.18×10^3 T	\$80.00	0.25
COOLING WATER MAKE UP	$24.7 \frac{\#H_2O}{\#LH_2}$	3.931×10^6 T	\$0.012 /T	0.05
CAPITAL INVESTMENT	$\$31.5 \times 10^6 \times \left(\frac{470\ T/D}{250\ T/D}\right)^{0.8} \times 1.050 = \$54.772 \times 10^6$	$\$54.772 \times 10^6 \times 0.11 = \$6.02 \times 10^6$	——	6.02
OPERATING COST	$\$1.91 \times 10^6/Y \times \left(\frac{470\ T/D}{250\ T/D}\right)^{0.65} = \$2.88 \times 10^6/Y$	$\$2.88 \times 10^6$	——	2.88
PROCESS ENERGY	(HYDROCARBON, COAL. GASIF. FEEDSTOCK) $4.46 \frac{KWH}{\#LH_2}$	1.420×10^9 KWH	\$0.00312/KWH	4.43
	(ELECTROLYSIS FEEDSTOCK) $4.46 \frac{KWH}{\#LH_2}$	1.420×10^9 KWH	\$0.00249/KWH	3.54
TOTAL COST	(STEAM REF)			13.78
	(PART. OXID)			13.74
	(COAL GASIF.)			13.78
	(ELECTROL.)			12.89
STORAGE				
CAPITAL INVESTMENT	$2 \times \$4.150 \times 10^6 \times 1.050 = \$8.715 \times 10^6$	$\$8.715 \times 10^6 \times 0.11 = \$0.96 \times 10^6$	——	0.96

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PROCESS, ITEM, UNIT REQUIREMENT SYDNEY	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
DISTRIBUTION			
CAPITAL INVESTMENT 2 MI × $\frac{5300 \text{ FT}}{\text{MI}}$ × 18 IN × $\frac{\$15}{\text{FT IN}}$ × 1.050 = \$3.005×10⁶	\$3.005×10⁶ × 0.11 = \$0.33×10⁶	—	0.33

PRODUCTION COST SUMMARY

GAS PLANT TYPE	GAS PLANT ANNUAL COST $\times 10^{-6}$ ($)	LIQ. PLANT ANNUAL COST $\times 10^{-6}$ ($)	STORAGE ANNUAL COST $\times 10^{-6}$ ($)	DISTRIB. ANNUAL COST $\times 10^{-6}$ ($)	TOTAL ANNUAL COST $\times 10^{-6}$ ($)	PRODUCT UNIT COST (¢/#LH₂)
STEAM REFORMING	13.52	13.78	0.96	0.33	28.59	9.8
PARTIAL OXIDATION	15.42	13.74	0.96	0.33	30.45	10.4
COAL GASIFICATION	18.13	13.78	0.96	0.33	33.20	11.4
ELECTROLYSIS	28.24	12.89	0.96	0.33	42.42	14.5

FACILITY DESCRIPTION – TOKYO

DISTRIBUTION

1 - 24" NPS × 2 MILE, SI-VJ LINE (ALLOWS 2 SIMULTANEOUS AIRCRAFT LOADINGS.) PRODUCT DELIVERED TO MANIFOLD IS 500 T/D. PRODUCT RECEIVED FROM STORAGE IS 538 T/D, I.E. DISTRIBUTION LOSSES ARE 38 T/D.

STORAGE

2 - 2500 T NOMINAL SIZE, DOUBLE WALL, EVACUATED PERLITE INSULATED TANKS. PRODUCT DISCHARGED FOR DISTRIBUTION IS 538 T/D. PRODUCT RECEIVED FROM THE LIQUEFIER IS 546 T/D, I.E. STORAGE LOSSES TOTAL TO 8 T/D.

LIQUEFACTION AND CONVERSION

1 - 690 T/D NOMINAL SIZE UNIT WITH 97% ON STREAM FACTOR (690 T/D SIZE MATCHES REQUIRED CAPACITY TO PROCESS GAS PLANT EFFLUENT.)

DAILY AVERAGE PRODUCTION IS 546 T/D.

GAS PRODUCTION AND PURIFICATION

3 - 250 T/D NOMINAL SIZE UNITS EACH WITH A 97% ON STREAM FACTOR

(NOMINAL SIZE OF GAS PLANT IS TAKEN AS 750 T/D)

DAILY AVERAGE PRODUCTION IS 568 T/D.

PRODUCTION COST DETERMINATION

PROCESS, ITEM, UNIT REQUIREMENT	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST ×10^{-6} ($)
STEAM REFORMING			
FEED - NATURAL GAS $1.90 \frac{\#NG}{\#GH_2} = 1.90 \frac{\#NG}{\#GH_2} \times \frac{MCF}{[illegible]1.6\ \#NG} = 0.0957 \frac{MCF}{\#GH_2}$	18.949×10^6 MCF	\$0.368/MCF	6.97

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PROCESS, ITEM, UNIT REQUIREMENT Tokyo	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
FUEL – NATURAL GAS $1.44 \frac{\#NG}{\#GH_2} - 4.4 \times 10^{-5} \frac{\#NG/\#GH_2}{T/D} \times 750 T/D = 1.41 \frac{\#NG}{\#GH_2}$			
FUEL – TOKYO OIL EQUIV. OF NAT. GAS $1.41 \frac{\#NG/\#GH_2}{41.6 \#NG/MCF} \times \frac{10^6 BTU/MCF}{149,400 BTU/GAL} \times \frac{BBL}{42 GAL} = 54.0 \times 10^{-4} \frac{BBL}{\#GH_2}$	2.239×10^6 BBL	$2.27 /BBL	5.08
COOLING WATER & BOILER WATER MAKE UP $(22.4 + 16.6) \frac{\#H_2O}{\#GH_2} = 39.0 \frac{\#H_2O}{\#GH_2}$	0.085×10^6 T	$0.072 /T	0.58
PROCESS ENERGY $0.47 \frac{KWH}{\#GH_2}$	1.949×10^8 KWH	$0.00307 /KWH	0.60
BYPRODUCT CO_2 $5.30 \frac{\#CO_2}{\#GH_2}$	0.11×10^6 T	—	—
CAPITAL INVESTMENT $\$7.65 \times 10^6 \left(\frac{750 T/D}{250 T/D}\right)^{0.7} \times 0.900 = \$14.858 \times 10^6$	$\$14.858 \times 10^6 \times 0.11$ $= \$1.63 \times 10^6$	—	1.63
OPERATING COST $\$1.28 \times 10^6/Y \left(\frac{750 T/D}{250 T/D}\right)^{0.73} = \$2.82 \times 10^6$	$\$2.82 \times 10^6$		2.82
TOTAL COST			17.68
PARTIAL OXIDATION			
FEED – OIL OR BUNKER "C" $3.89 \frac{\#OIL}{\#GH_2} = 3.89 \frac{\#OIL}{\#GH_2} \times \frac{BBL/42 GAL}{7.8950 \#OIL/GAL}$ $= 0.0117 \frac{BBL}{\#GH_2}$	4.851×10^6 BBL	$2.27/BBL	11.01

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PROCESS, ITEM, UNIT REQUIREMENT Tokyo	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
FUEL - TOKYO OIL EQUIV. OF BUNKER "C" $0.40 \frac{\#\text{"C"}}{\#GH_2} - 3.1 \times 10^{-5} \frac{\#\text{"C"}/\#GH_2}{T/D} \times 750\ T/D = 0.377 \frac{\#\text{"C"}}{\#GH_2}$ $\frac{0.377\ \#\text{"C"}/\#GH_2}{8.108\ \#\text{"C"}/GAL\ \text{"C"}} \times \frac{152{,}000\ BTU/GAL\ \text{"C"}}{149{,}400\ BTU/GAL\ OIL} \times \frac{BBL}{42\ GAL} = 0.0011 \frac{BBL}{\#GH_2}$	0.456×10^6 BBL	\$2.27/BBL	1.04
COOLING WATER & BOILER MAKE UP WATER $(12.5 + 12.5) \frac{\#H_2O}{\#GH_2} = 25.0 \frac{\#H_2O}{\#GH_2}$	5.183×10^6 T	\$0.072/T	0.37
PROCESS ENERGY $1.30 \frac{KWH}{\#GH_2}$	5.390×10^8 KWH	\$0.00307/KWH	1.65
BYPRODUCTS CO₂ $6.0 \frac{\#CO_2}{\#GH_2}$	1.24×10^6 T	—	—
STEAM $5.0 \frac{\#H_2O}{\#GH_2}$	1.04×10^6 T	—	—
C $0.11 \frac{\#C}{\#GH_2}$	22.8×10^3 T	—	—
CAPITAL INVESTMENT $\$8.46 \times 10^6 \times \left(\frac{750\ T/D}{250\ T/D}\right)^{0.7} \times 0.900 = \$16.431 \times 10^6$	$\$16.431 \times 10^6 \times 0.11$ $= \$1.81 \times 10^6$	—	1.81
OPERATING COST $\$1.46 \times 10^6/Y \times \left(\frac{750\ T/D}{250\ T/D}\right)^{0.73} = \$3.26 \times 10^6/Y$	$\$3.26 \times 10^6$	—	3.26
TOTAL COST			19.14
COAL GASIFICATION (BITUM.)			
FEED WATER & COOLING WATER MAKE UP $(20.3 + 26.0) \frac{\#H_2O}{\#GH_2} = 46.3 \frac{\#H_2O}{\#G}$	9.599×10^6 T	\$0.072/T	0.69

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PROCESS, ITEM, UNIT REQUIREMENT Tokyo	ANNUAL REQUIREMENTS	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
FUEL - COAL EQUIV. OF W. VA BITUM. $8.74 \frac{\#W.Va.}{\#6H_2} \cdot 8.74 \frac{\#W.Va.}{\#6H_2} \times \frac{14,290 \text{ BTU}/\#W Va}{11,100 \text{ BTU}/\# \text{COAL}} = 11.25 \frac{\#\text{COAL}}{\#6H_2}$	2.332×10^6 T	$13.42 /T	31.30
OXYGEN $5.20 \frac{\#O_2}{\#6H_2}$	1.078×10^6 T	$ 4.50 /T	4.85
PROCESS ENERGY $0.57 \frac{KWH}{\#6H_2}$	2.363×10^8 T	$0.00307/KWH	0.73
BYPRODUCT CO₂ $24.0 \frac{\#CO_2}{\#6H_2}$	4.98×10^6 T	—	—
CAPITAL INVESTMENT $\$20.8 \times 10^6 \left(\frac{750 \text{ T/D}}{250 \text{ T/D}}\right)^{0.70} \times 0.900 = \$40.416 \times 10^6$	$\$40.416\ 10^6 \times 0.11 = \$4.45 \times 10^6$	—	4.45
OPERATING COST $\$2.89 \times 10^6/Y \times \left(\frac{750 \text{ T/D}}{250 \text{ T/D}}\right)^{0.65} = \$5.90 \times 10^6/Y$	$\$5.90 \times 10^6$	—	5.90
TOTAL COST			47.92
ELECTROLYSIS (LOW PRESSURE)			
FEED & COOLING WATER $(9.0 + 11.5) \frac{\#H_2O}{\#6H_2} = 20.5 \frac{\#H_2O}{\#6H_2}$	4.250×10^6 T	$0.072/T	0.31
PROCESS ENERGY $21.6 \frac{KWH}{\#6H_2}$	8.956×10^9 KWH	$0.00223/KWH	19.97
BY PRODUCT O₂ $7.92 \frac{\#O_2}{\#6H_2}$	1.64×10^6 T	—	—
CAPITAL INVESTMENT $\$28.5 \times 10^6 \times \left(\frac{750 \text{ T/D}}{250 \text{ T/D}}\right)^{0.88} \times 0.900 = \$67.434 \times 10^6$	$\$67.434 \times 10^6 \times 0.11 = \$7.42 \times 10^6$	—	7.42

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PROCESS, ITEM,	UNIT REQUIREMENT TOKYO	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
OPERATING COST	$\$2.62 \times 10^6/Y \times \left(\frac{750\ T/D}{250\ T/D}\right)^{0.73} = \$5.84 \times 10^6/Y$	$\$5.84 \times 10^6$	——	5.84
TOTAL COST				33.54
LIQUEFACTION				
REFRIGERANTS	N_2 (NOT REQ'D. FOR PART. OXID. FEEDSTOCK) $0.04 \frac{\# N_2}{\# LH_2}$	7.97×10^3 T	\$6.50/T	0.05
	C_1H_4 $0.01 \frac{\# NG}{\# LH_2} = 0.01 \frac{\# NG}{\# LH_2} \times \frac{MCF}{41.6\ \# NG} = 24 \times 10^{-5} \frac{MCF}{\# LH_2}$	95.7×10^3 MCF	\$0.368/MCF	0.04
	C_3H_8 $0.02 \frac{\# C_3}{\# LH_2}$	3.99×10^3 T	\$25.00/T	0.10
	C_2H_4 $0.02 \frac{\# E_t}{\# LH_2}$	3.99×10^3 T	\$80.00/T	0.32
COOLING WATER MAKE UP	$24.7 \frac{\# H_2O}{\# LH_2}$	4.922×10^6 T	\$0.072/T	0.35
PROCESS ENERGY	(HYDROCARBON, COAL GASIF. FEEDSTOCK) $4.46 \frac{KWH}{\# LH_2}$	1.778×10^9 KWH	\$0.00307/KWH	5.46
	(ELECTROLYSIS FEEDSTOCK) $4.46 \frac{KWH}{\# LH_2}$	1.778×10^9 KWH	\$0.00223/KWH	3.96
CAPITAL INVESTMENT	$\$31.5 \times 10^6 \times \left(\frac{690\ T/D}{250\ T/D}\right)^{0.8} \times 0.900 = \$63.873 \times 10^6$	$\$63.873 \times 10^6 \times 0.11 = \$7.03 \times 10^6$	——	7.03
OPERATING COST	$\$1.91 \times 10^6/Y \left(\frac{690\ T/D}{250\ T/D}\right)^{0.65} = \$3.69 \times 10^6/Y$	$\$3.69 \times 10^6$	——	3.69
TOTAL COST	(STEAM REFORMING, COAL GASIFICATION)			17.04

PROCESS, ITEM, UNIT REQUIREMENT Tokyo	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
TOTAL COST CONT'D. (PARTIAL OXIDATION)			16.99
(ELECTROLYSIS)			15.54
STORAGE			
CAPITAL INVESTMENT $2 \times \$5.10 \times 10^6 \times 0.900 = \$9.180 \times 10^6$	$\$9.180 \times 10^6 \times 0.11$	—	\$1.01
DISTRIBUTION			
CAPITAL INVESTMENT $2\,\text{MI} \times 5300 \frac{\text{FT}}{\text{MILE}} \times 24\,\text{IN} \times \frac{\$15}{\text{FT IN}} \times 0.900 = \$3.434 \times 10^6$	$\$3.434 \times 10^6 \times 0.11$		\$0.38

NATIONAL / 42-381 50 SHEETS
42-382 100 SHEETS
Made in U.S.A.

PRODUCTION COST SUMMARY TOKYO

GAS PLANT TYPE	GAS PLANT ANNUAL COST $\times 10^{-6}$ (\$)	LIQ. PLANT ANNUAL COST $\times 10^{-6}$ (\$)	STORAGE ANNUAL COST $\times 10^{-6}$ (\$)	DISTRIB. ANNUAL COST $\times 10^{-6}$ (\$)	TOTAL ANNUAL COST $\times 10^{-6}$ (\$)	PRODUCT UNIT COST (¢/# LH_2)
STEAM REFORMING	17.68	17.04	1.01	0.38	36.11	9.9
PARTIAL OXIDATION	19.14	16.99	1.01	0.38	37.52	10.3
COAL GASIFICATION	47.92	17.04	1.01	0.38	66.35	18.2
ELECTROLYSIS	33.54	15.54	1.01	0.38	50.47	13.8

NATIONAL / 42-381 50 SHEETS
42-382 100 SHEETS
Made in U.S.A.

FACILITY DESCRIPTION — LOS ANGELES

DISTRIBUTION

2 - 24" NPS × 2 MILE, SI-VJ LINE (ALLOWS 4 AIRCRAFT LOADING AT A TIME)
PRODUCT DELIVERED TO MANIFOLDS IS 1300 T/D. PRODUCT RECEIVED FROM STORAGE IS 1398 T/D, I.E. DISTRIBUTION LOSSES ARE 98 T/D.

STORAGE

2 - 3600 T NOMINAL SIZE, DOUBLE WALL, EVACUATED PERLITE INSULATED TANKS.
PRODUCT DISCHARGED FOR DISTRIBUTION IS 1398 T/D. PRODUCT RECEIVED FROM THE LIQUEFIER IS 1417 T/D, I.E. STORAGE LOSSES TOTAL TO 19 T/D.

LIQUEFACTION AND CONVERSION

2 - 800 T/D NOMINAL SIZE UNITS WITH 97% ON STREAM FACTOR
DAILY AVERAGE PRODUCTION IS 1417 T/D.

GAS PRODUCTION AND PURIFICATION

7 - 250 T/D NOMINAL SIZE UNITS WITH 93% ON STREAM FACTOR
DAILY AVERAGE PRODUCTION IS 1474 T/D.

PRODUCTION COST DETERMINATION

PROCESS, ITEM, UNIT DESCRIPTION	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
STEAM REFORMING			
FEED - NATURAL GAS $1.90 \frac{\#NG}{\#GH_2} = 1.90 \frac{\#NG}{\#GH_2} \times \frac{MCF}{41.6 \#NG} = 0.0457 \frac{MCF}{\#GH_2}$	49.17×10^6 MCF	\$0.338 /MCF	16.62
FUEL - NAT. GAS $1.44 \frac{\#NG}{\#GH_2} - 4.4 \times 10^{-5} \frac{\#NG/\#GH_2}{T/D} \times 1750\ T/D = 1.36 \frac{\#NG}{\#GH_2}$			

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PROCESS, ITEM, UNIT REQUIREMENT LOS ANGELES	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
FUEL - LOS ANGELOS OIL EQUIV OF NAT. GAS $1.36 \frac{\#\text{NG}}{\#\text{GH}_2} \times \frac{10^6 \text{ BTU/MCF}}{91.6 \#\text{NG/MCF}} \times \frac{\text{BBL}/42 \text{ GAL}}{152,500 \text{ BTU/GAL}} = 51.0 \times 10^{-4} \frac{\text{BBL}}{\#\text{GH}_2}$	5.488×10^6 BBL	\$1.88/BBL	10.32
COOLING WATER & BOILER WATER MAKE UP $(22.4 + 16.6) \frac{\#\text{H}_2\text{O}}{\#\text{GH}_2} = 39.0 \frac{\#\text{H}_2\text{O}}{\#\text{GH}_2}$	20.98×10^6 T	\$0.072/T	1.51
PROCESS ENERGY $0.47 \frac{\text{KWH}}{\#\text{GH}_2}$	5.057×10^8 KWH	\$0.00255/KWH	1.29
BYPRODUCT CO_2 $5.30 \frac{\#\text{CO}_2}{\#\text{GH}_2}$	2.85×10^6 T	—	—
CAPITAL INVESTMENT $\$7.65 \times 10^6 \times \left(\frac{1750 \text{ T/D}}{250 \text{ T/D}}\right)^{0.7} \times 1.10 = \$32.819 \times 10^6$	$\$32.819 \times 10^6 \times 0.11$ $= \$3.61 \times 10^6$	—	3.61
OPERATING COST $\$1.28 \times 10^6/\text{Y} \times \left(\frac{1750 \text{ T/D}}{250 \text{ T/D}}\right)^{0.73} = \$5.30 \times 10^6/\text{Y}$	$\$5.30 \times 10^6$	—	5.30
TOTAL COST			38.65
PARTIAL OXIDATION			
FEED - OIL OR BUNKER "C" $3.89 \frac{\#\text{OIL}}{\#\text{GH}_2} = \frac{3.89 \#\text{OIL}/\#\text{GH}_2}{8.155 \#\text{OIL/GAL}} \times \frac{\text{BBL}}{42 \text{GAL}} = 0.0114 \frac{\text{BBL}}{\#\text{GH}_2}$	12.267×10^6 BBL	\$1.88/BBL	23.06
FUEL - LOS ANGELOS OIL EQUIV OF BUNKER "C" $0.40 \frac{\#\text{"C"}}{\#\text{GH}_2} - 3.1 \times 10^{-5} \frac{\#\text{"C"}/\#\text{GH}_2}{\text{T/D}} \times 1750 \text{ T/D} = 0.346 \frac{\#\text{"C"}}{\#\text{GH}_2}$ $\frac{0.346 \#\text{"C"}/\#\text{GH}_2}{8.108 \#\text{"C"}/\text{GAL "C"}} \times \frac{152,000 \text{ BTU/GAL "C"}}{152,500 \text{ BTU/GAL OIL}} \times \frac{\text{BBL}}{42 \text{GAL OIL}} = 0.0010 \frac{\text{BBL}}{\#\text{GH}_2}$	1.076×10^6 BBL	\$1.88/BBL	2.02

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PROCESS, ITEM, UNIT REQUIREMENT LOS ANGELES	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
COOLING WATER & BOILER WATER MAKE UP $(12.5 + 12.5)\frac{\#H_2O}{\#GH_2} = 25.0 \frac{\#H_2O}{\#GH_2}$	13.45×10^6 T	\$0.072/T	0.97
PROCESS ENERGY $1.30 \frac{KWH}{\#GH_2}$	1.399×10^9 KWH	\$0.00255/KWH	3.57
BYPRODUCT CO_2 $6.0 \frac{\#CO_2}{\#GH_2}$	3.23×10^6 T	——	——
STEAM $5.0 \frac{\#H_2O}{\#GH_2}$	2.69×10^6 T	——	——
C $0.11 \frac{\#C}{\#GH_2}$	59.2×10^3 T	——	——
CAPITAL INVESTMENT $\$8.46 \times 10^6 \times \left(\frac{1750\ T/D}{250\ T/D}\right)^{0.7} \times 1.10 = \$36.293 \times 10^6$	$\$36.293 \times 10^6 \times 0.11$ $= \$3.99 \times 10^6$	——	3.99
OPERATING COST $\$1.46 \times 10^6/Y \times \left(\frac{1750\ T/D}{250\ T/D}\right)^{0.73} = \$6.04 \times 10^6/Y$	$\$6.04 \times 10^6$	——	6.04
TOTAL COST			39.65
ELECTROLYSIS (LOW PRESSURE)			
FEED & COOLING WATER $(9.0 + 11.5)\frac{\#H_2O}{\#GH_2} = 20.5 \frac{\#H_2O}{\#GH_2}$	11.03×10^6 T	\$0.072/T	0.79
PROCESS ENERGY $21.6 \frac{KWH}{\#GH_2}$	23.24×10^9 KWH	\$0.00241/KWH	56.01
BYPRODUCT O_2 $7.92 \frac{\#O_2}{\#GH_2}$	4.26×10^6 T	——	——
CAPITAL INVESTMENT $\$28.5 \times 10^6 \times \left(\frac{1750\ T/D}{750\ T/D}\right)^{0.88} \times 1.100 = \$173.84 \times 10^6$	$\$173.84 \times 10^6 \times 0.11$ $= \$19.12 \times 10^6$	——	19.12

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PROCESS, ITEMS	UNIT REQUIREMENT LOS ANGELES	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
OPERATING COST	$\$2.62 \times 10^6/Y \times \left(\frac{1750\ T/D}{250\ T/D}\right)^{0.73} = \$10.85 \times 10^6/Y$	$\$10.85 \times 10^6$	—	10.85
TOTAL COST				86.77
LIQUEFACTION AND CONVERSION				
REFRIGERANTS N_2	(NOT REQ'D FOR PART. OXID. FEED STOCK) $0.04 \frac{\#N_2}{\#LH_2}$	20.69×10^3 T	$\$6.50/T$	0.13
C_1H_4	$0.01 \frac{\#NG}{\#LH_2} = 0.01 \frac{\#NG}{\#LH_2} \times \frac{MCF}{41.6\ \#NG} = 2.4 \times 10^{-5} \frac{MCF}{\#LH_2}$	298.3×10^3 MCF	$\$0.338/MCF$	0.08
C_2H_6	$0.02 \frac{\#C_2}{\#LH_2}$	10.34×10^3 T	$\$25.00/T$	0.26
C_2H_4	$0.02 \frac{\#E_2}{\#LH_2}$	10.34×10^3 T	$\$80.00/T$	0.83
COOLING WATER MAKE UP	$24.7 \frac{\#H_2O}{\#LH_2}$	12.77×10^6 T	$\$0.072/T$	0.92
PROCESS ENERGY (HYDROCARBON FEEDSTOCK)	$4.46 \frac{KWH}{\#LH_2}$	4.613×10^9 KWH	$\$0.00255/KWH$	11.76
(ELECTROLYSIS FEEDSTOCK)	$4.46 \frac{KWH}{\#LH_2}$	4.613×10^9 KWH	$\$0.00241/KWH$	11.12
CAPITAL INVESTMENT	$\$31.5 \times 10^6 \times \left(\frac{1600\ T/D}{250\ T/D}\right)^{0.8} \times 1.100 = \$152.81 \times 10^6$	$\$152.81 \times 10^6 \times 0.11 = \$16.81 \times 10^6$	—	16.81
OPERATING COST	$\$1.91 \times 10^6/Y \times \left(\frac{1600\ T/D}{250\ T/D}\right)^{0.65} = \$6.38 \times 10^6/Y$	$\$6.38 \times 10^6$		6.38

NATIONAL / 42-381 50 SHEETS 42-382 100 SHEETS Made in U.S.A.

PROCESS, ITEM, UNIT REQUIREMENT — LOS ANGELES	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
TOTAL COST (STEAM REFORMING)			37.17
(PART. OXID.)			37.04
(ELECTROL.)			36.53
STORAGE			
CAPITAL INVESTMENT $2 \times \$7.00 \times 10^{6} \times 1.10 = \$15.400 \times 10^{6}$	$\$15.400 \times 10^{6} \times 0.11$		1.69
DISTRIBUTION			
CAPITAL INVESTMENT $2 \times 2\ \text{MI} \times \frac{5300\ \text{FT}}{\text{MI}} \times 24\ \text{IN} \times \frac{\$15}{\text{FT-IN}} \times 1.100 = \$8.395 \times 10^{6}$	$\$8.395 \times 10^{6} \times 0.11$		0.92

PRODUCTION COST SUMMARY

GAS PLANT TYPE	GAS. PLANT ANNUAL COST $\times 10^{-6}$ (\$)	LIQ. PLANT ANNUAL COST $\times 10^{-6}$ (\$)	STORAGE ANNUAL COST $\times 10^{-6}$ (\$)	DISTRIB. ANNUAL COST $\times 10^{-6}$ (\$)	TOTAL ANNUAL COST $\times 10^{-6}$ (\$)	PRODUCT UNIT COST (¢/MLH₂)
STEAM REFORMING	38.65	37.17	1.69	0.92	78.43	8.3
PARTIAL OXIDATION	39.65	37.04	1.69	0.92	79.30	8.4
ELECTROLYSIS	86.77	36.53	1.69	0.92	125.91	13.3

FACILITY DESCRIPTION — PARIS

DISTRIBUTION

2 - 24" NPS × 2 MILE, SI - VJ LINES. (ALLOWS 4 AIRCRAFT LOADING AT A TIME.) PRODUCT DELIVERED TO MANIFOLDS IS 2000 T/D. PRODUCT RECEIVED FROM STORAGE IS 2151 T/D, I.E. DISTRIBUTION LOSSES TOTAL TO 151 T/D.

STORAGE

2 - 4100 T NOMINAL SIZE, DOUBLE WALL, EVACUATED PERLITE INSULATED TANKS. PRODUCT DISCHARGED FOR DISTRIBUTION IS 2151 T/D. PRODUCT RECEIVED FROM THE LIQUEFIER IS 2179 T/D, I.E. STORAGE LOSSES TOTAL TO 28 T/D.

LIQUEFACTION AND CONVERSION

3 - 800 T/D NOMINAL SIZE UNITS WITH 97% ON STREAM FACTOR. DAILY AVG PRODUCTION IS 2179 T/D.

GAS PRODUCTION AND PURIFICATION

10 - 260 T/D NOMINAL SIZE UNITS WITH 93% ON STREAM FACTOR. DAILY AVG PRODUCTION IS 2266 T/D.

PRODUCTION COST DETERMINATION

PROCESS, ITEM, UNIT DESCRIPTION	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
STEAM REFORMING			
FEED - NATURAL GAS $1.90 \frac{\#NG}{\#GH_2} = 1.90 \frac{\#NG}{\#GH_2} \times \frac{MCF}{41.6} = 0.0457 \frac{MCF}{\#GH_2}$	75.60×10^6 MCF	$0.48/MCF	36.29
FUEL - NATURAL GAS $1.44 \frac{\#NG}{\#GH_2} - 9.4 \times 10^{-5} \frac{\#NG/\#GH_2}{T/D} \times 2500 T/D = 1.33 \frac{\#NG}{\#GH_2}$			

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42-382 100 SHEETS
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PROCESS, ITEM, UNIT REQUIREMENT PARIS	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
FUEL - PARIS OIL EQUIV. TO NAT. GAS	9.198×10^6 BBL	\$2.39/BBL	21.86
$1.33 \frac{\#NG}{\#GH_2} \times \frac{MCF/41.6\ \#NG}{MCF/10^6\ BTU} \times \frac{GAL/137{,}600\ BTU}{42\ GAL/BBL} = 55.3 \times 10^{-4} \frac{BBL}{\#GH_2}$			
COOLING WATER & BOILER WATER MAKE UP $(22.4 + 16.6) \frac{\#H_2O}{\#GH_2} = 39.0 \frac{\#H_2O}{\#GH_2}$	32.26×10^6 T	\$0.0072/T	0.23
PROCESS ENERGY $0.47 \frac{KWH}{\#GH_2}$	7.775×10^8 KWH	\$0.00249/KWH	1.94
BYPRODUCT $5.30 \frac{\#CO_2}{\#GH_2}$		———	———
CAPITAL INVESTMENT $\$7.65 \times 10^6 \times \left(\frac{2500\ T/D}{250\ T/D}\right)^{0.7} \times 1.050 = \$40.243 \times 10^6$	$\$40.243 \times 10^6 \times 0.11$	———	4.43
OPERATING COST $\$1.28 \times 10^6/Y \times \left(\frac{2500\ T/D}{250\ T/D}\right)^{0.73} = \$6.89 \times 10^6/Y$	$\$6.89 \times 10^6$	———	6.87
TOTAL COST			71.62
PARTIAL OXIDATION			
FEED - OIL OR BUNKER "C" $3.89 \frac{\#OIL}{\#GH_2} = \frac{3.89\ \#OIL/\#GH_2}{7.000\ \#OIL/GAL} \times \frac{BBL}{42\ GAL} = 0.0132 \frac{BBL}{GAL}$	21.84×10^6 BBL	\$2.39/BBL	52.20
FUEL - PARIS OIL EQUIV TO BUNKER "C"	1.737×10^6 BBL	\$2.39/BBL	4.15
$0.40 \frac{\#\text{"C"}}{\#GH_2} - \frac{3.1 \times 10^{-5}\ \#\text{"C"}/\#GH_2}{T/D} \times 2500\ T/D = 0.323 \frac{\#\text{"C"}}{\#GH_2}$			

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PROCESS, ITEM, UNIT REQUIREMENT PARIS	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
FUEL - CONT'D. $\frac{0.323\ \#\text{"C"}/\#GH_2}{8.108\ \#\text{"C"}/GAL\text{"C"}} \times \frac{152{,}000\ BTU/GAL\text{"C"}}{137{,}600\ BTU/GAL\ OIL} \times \frac{BBL}{42\ GAL\ OIL} = 10.5 \times 10^{-4} \frac{BBL}{\#GH_2}$			
COOLING WATER & BOILER WATER MAKE UP $(12.5 + 12.5) \frac{\#H_2O}{\#GH_2} = 25.0 \frac{\#H_2O}{\#GH_2}$	20.68×10^6 T	$0.0072 /T	0.15
PROCESS ENERGY $1.30 \frac{KWH}{\#GH_2}$	2.150×10^9 KWH	$0.00249/KWH	5.35
BYPRODUCT CO₂ $6.0 \frac{\#CO_2}{\#GH_2}$	4.96×10^6 T	—	—
STEAM $5.0 \frac{\#H_2O}{\#GH_2}$	4.14×10^6 T	—	—
C $0.11 \frac{\#C}{\#GH_2}$	91.0×10^3 T	—	—
CAPITAL INVESTMENT $\$8.46 \times 10^6 \times \left(\frac{2500\ T/D}{250\ T/D}\right)^{0.7} \times 1.050 = \$44.504 \times 10^6$	$\$44.504 \times 10^6 \times 0.11$	—	4.90
OPERATING COST $\$1.46 \times 10^5/Y \times \left(\frac{2500\ T/D}{250\ T/D}\right)^{0.73} = 7.84 \times 10^6/Y$		—	7.84
TOTAL COST			74.59
ELECTROLYSIS			
FEED & COOLING WATER $(9.0 + 11.5) \frac{\#H_2O}{\#GH_2} = 20.5 \frac{\#H_2O}{\#GH_2}$	16.96×10^6 T	$0.0072 /T	0.12
PROCESS ENERGY $21.6 \frac{KWH}{\#GH_2}$	35.73×10^9 T	$0.00233/KWH	83.25

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PROCESS, ITEM, UNIT REQUIREMENT PARIS	ANNUAL REQUISITION	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
BYPRODUCT O_2 $7.92\ \frac{\#O_2}{\#GH_2}$	6.55×10^6 T	——	——
CAPITAL INVESTMENT $\$28.5 \times 10^6 \left(\frac{2500\ T/D}{250\ T/D}\right)^{0.88} \times 1.050$	$227.13 \times 10^6 \times 0.11$	——	24.98
OPERATING COST $\$2.62 \times 10^6 / Y \times \left(\frac{2500\ T/D}{250\ T/D}\right)^{0.73} =$ $\times 10^6/Y$	$\$14.07 \times 10^6$	——	14.07
TOTAL COST			122.42
LIQUEFACTION			
REFRIGERANTS N_2 (NOT REQ'D FOR PART. OXID. FEEDSTOCK) $0.04\ \frac{\#N_2}{\#LH_2}$	31.81×10^3 T	\$6.50/T	0.21
C_1H_4 $0.01\ \frac{\#NG}{\#LH_2} = 0.01\ \frac{\#NG}{\#LH_2} \times \frac{MCF}{41.6\ \#NG} = 24 \times 10^{-5}\ \frac{MCF}{\#LH_2}$	381.8×10^3 MCF	\$0.48/MCF	0.18
C_2H_6 $0.02\ \frac{\#C_2}{\#LH_2}$	15.91×10^3 T	\$25.00/T	0.40
C_2H_4 $0.02\ \frac{\#E_2}{\#LH_2}$	15.91×10^3 T	\$80.00/T	1.27
COOLING WATER MAKE UP $24.7\ \frac{\#H_2O}{\#LH_2}$	19.64×10^6 T	\$0.0072/T	0.14
PROCESS ENERGY (HYDROCARBON FEEDSTOCK) $4.46\ \frac{kWH}{\#LH_2}$	7.094×10^9 KWH	\$0.00249/KWH	17.66
(ELECTROLYSIS FEEDSTOCK) $4.46\ \frac{KWH}{\#LH_2}$	7.094×10^9 KWA	\$0.00233/KWH	16.53
CAPITAL INVESTMENT $\$31.5 \times 10^6 \times \left(\frac{2400\ T/D}{250\ T/D}\right)^{0.8} \times 1.050 = \$202.09 \times 10^6$	$\$202.09\ 10^6 \times 0.11$	——	22.23

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42-382 100 SHEETS
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PROCESS, ITEM,	UNIT REQUIREMENT PARIS	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
OPERATING COST	$\$1.91 \times 10^6/Y \times \left(\frac{2400\ T/D}{250\ T/D}\right)^{0.65} = \$8.31 \times 10^6/Y$	$\$8.31 \times 10^6$	———	8.31
TOTAL COST	(STEAM REFORMING)			50.40
	(PART. OXID.)			50.19
	(ELECTROLYSIS)			49.27
STORAGE				
CAPITAL INVESTMENT	$2 \times \$8.20 \times 10^6 \times 1.050 = \$17.220 \times 10^6$	$\$17.220 \times 10^6 \times 0.11$		1.89
DISTRIBUTION				
CAPITAL INVESTMENT	$2 \times 2\ MI \times \frac{5300\ F}{MI} \times 24\ IN \times \frac{\$15}{FT\ IN} \times 1.050 = \$8.014 \times 10^6$	$\$8.014 \times 10^6 \times 0.11$		0.88

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Production Cost Summary Paris

Gas Plant Type	Gas Plant Annual Cost $\times 10^{-6}$ (\$)	Liq. Plant Annual Cost $\times 10^{-6}$ (\$)	Storage Annual Cost $\times 10^{-6}$ (\$)	Distrib. Annual Cost $\times 10^{-6}$ (\$)	Total Annual Cost $\times 10^{-6}$ (\$)	Unit Product Cost (¢/#LH_2)
Steam Reforming	71.62	50.40	1.89	0.88	124.79	8.5
Partial Oxidation	74.59	50.19	1.89	0.88	127.55	8.7
Electrolysis	122.42	49.27	1.89	0.88	174.46	11.9

NATIONAL 42-381 50 SHEETS
42-382 100 SHEETS
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FACILITY DESCRIPTION — NEW YORK

DISTRIBUTION
2 - 24" NPS × 2 MILE, SI-VJ LINES. (ALLOWS 4 AIRCRAFT LOADING AT A TIME.)
PRODUCT DELIVERED TO MANIFOLDS IS 2500 T/D. PRODUCT RECEIVED FROM STORAGE IS 2688 T/D, I.E. DISTRIBUTION LOSSES TOTAL TO 188 T/D.

STORAGE
2 - 4800 T NOMINAL SIZE, DOUBLE WALL, EVACUATED PERLITE INSULATED TANKS.
PRODUCT DISCHARGED FOR DISTRIBUTION IS 2688 T/D. PRODUCT RECEIVED FROM THE LIQUEFIER IS 2723 T/D, I.E. STORAGE LOSSES TOTAL TO 35 T/D.

LIQUEFACTION AND CONVERSION
3 - 935 T/D NOMINAL SIZE UNITS WITH 97% ON STREAM FACTORS
DAILY AVG. PRODUCTION IS 2723 T/D.

GAS PRODUCTION AND PURIFICATION
13 - 250 T/D NOMINAL SIZE UNITS WITH 93% ONSTREAM FACTORS
DAILY AVG. PRODUCTION IS 2832 T/D.

PRODUCTION COST DETERMINATION

PROCESS, ITEM, UNIT REQUIREMENT	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
STEAM REFORMING			
FEED - NATURAL GAS $1.90 \frac{\#NG}{\#GH_2} = 1.90 \frac{\#NG}{\#GH_2} \times \frac{MCF}{41.6\ \#NG} = 0.0457 \frac{MCF}{\#GH_2}$	94.48×10^6 MCF	\$0.363 /MCF	34.30
FUEL - NATURAL GAS $1.44 \frac{\#NG}{\#GH_2} - 4.4 \times 10^{-5} \frac{\#NG/\#GH_2}{T/D} \times 3250\ T/D = 1.30 \frac{\#NG}{\#GH_2}$			

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42-382 100 SHEETS
Made in U.S.A.

PROCESS, ITEM, UNIT REQUIREMENT — NEW YORK	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
FUEL - NEW YORK OIL EQUIV OF NAT. GAS. $\frac{1.30\ \#NG/\#GH_2}{41.6\ \#NG/MCF} \times \frac{10^6\ BTU/MCF}{148,900\ BTU/GAL} \times \frac{BBL}{42\ GAL} = 0.00500\ \frac{BBL}{\#GH_2}$	10.34×10^6 BBL	\$1.85/BBL	19.13
COOLING WATER & BOILER WATER MAKE UP $(22.4 + 16.6)\ \frac{\#H_2O}{\#GH_2} = 39.0\ \frac{\#H_2O}{\#GH_2}$	40.31×10^6 T	\$0.072/T	2.90
PROCESS ENERGY $0.47\ \frac{KWH}{\#GH_2}$	9.717×10^8 KWH	\$0.00288/KWH	2.80
BYPRODUCT CO_2 $5.30\ \frac{\#CO_2}{\#GH_2}$	5.48×10^6 T	—	—
CAPITAL INVESTMENT $\$7.65 \times 10^6 \times \left(\frac{3250\ T/D}{250\ T/D}\right)^{0.7} \times 1.300 = \$59.869 \times 10^6$	$\$59.869 \times 10^6 \times 0.11$	—	6.59
OPERATING COST $\$1.28 \times 10^6/Y \times \left(\frac{3250\ T/D}{250\ T/D}\right)^{0.73} \times 1.300 = \$8.33 \times 10^6/Y$	$\$8.33 \times 10^6$		8.33
TOTAL COST			74.05
PARTIAL OXIDATION			
FEED - OIL OR BUNKER "C" $3.89\ \frac{\#OIL}{\#GH_2} = \frac{3.84\ \#OIL/\#GAL}{7.850\ \#OIL/GAL} \times \frac{BBL}{42\ GAL} = 0.0118\ \frac{BBL}{\#GH_2}$	24.395×10^6 BBL	\$1.85/BBL	45.13
FUEL - NEW YORK OIL EQUIV OF BUNKER "C" $0.40\ \frac{\#\text{"C"}}{\#GH_2} - 3.1 \times 10^{-5}\ \frac{\#\text{"C"}/\#GH_2}{T/D} \times 3250\ T/D = 0.299\ \frac{\#\text{"C"}}{\#GH_2}$ $\frac{0.299\ \#\text{"C"}/\#GH_2}{8.168\ \#\text{"C"}/GAL\ \text{"C"}} \times \frac{152,000\ BTU/GAL\ \text{"C"}}{148,900\ BTU/GAL\ OIL} \times \frac{BBL}{42\ GAL} = 0.00090\ \frac{BBL}{\#GH_2}$	1.860×10^6 BBL	\$1.85/BBL	3.44

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PROCESS, ITEM, UNIT REQUIREMENT NEW YORK	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
COOLING WATER & BOILER WATER MAKE UP $(12.5 + 12.5)\ \frac{\#H_2O}{\#GH_2} = 25.0\ \frac{\#H_2O}{\#GH_2}$	25.84×10^6 T	\$0.072/T	1.86
PROCESS ENERGY $1.30\ \frac{KWH}{\#GH_2}$	2.688×10^9 KWH	\$0.00288/KWH	7.74
BYPRODUCT CO_2 $6.0\ \frac{\#CO_2}{\#GH_2}$	6.20×10^6 T	—	—
STEAM $5.0\ \frac{\#H_2O}{\#GH_2}$	5.17×10^6 T	—	—
C $0.11\ \frac{\#C}{\#GH_2}$	113.7×10^3 T	—	—
CAPITAL INVESTMENT $\$8.46 \times 10^6 \times \left(\frac{3250\ T/D}{250\ T/D}\right)^{0.7} \times 1.300 = \$66.208 \times 10^6$	$\$66.208 \times 10^6 \times 0.11$	—	7.28
OPERATING COST $\$1.46 \times 10^6/Y \times \left(\frac{3250\ T/D}{250\ T/D}\right)^{0.73} = \$9.50 \times 10^6/Y$	$\$9.50 \times 10^6$	—	9.50
TOTAL COST			74.95
ELECTROLYSIS			
FEED & COOLING WATER $(9.0 + 11.5)\ \frac{\#H_2O}{\#GH_2} = 20.5\ \frac{\#H_2O}{\#GH_2}$	21.19×10^6 T	\$0.072/T	1.53
PROCESS ENERGY $21.6\ \frac{KWH}{\#GH_2}$	44.65×10^9 KWH	\$0.00272/KWH	121.45
BYPRODUCT O_2 $7.92\ \frac{\#O_2}{\#GH_2}$	8.17×10^6 T	—	—

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PROCESS, ITEM, UNIT REQUIREMENT NEW YORK	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ ($)
CAPITAL INVESTMENT. $\$28.5 \times 10^6 \times \left(\frac{3250\ T/D}{250\ T/D}\right)^{0.88} \times 1.300 = \$354.94 \times 10^6$	$\$354.94 \times 10^6 \times 0.11$	———	39.04
OPERATING COST $\$2.62 \times 10^6/Y \times \left(\frac{3250\ T/D}{250\ T/D}\right)^{0.73} = \$17.06 \times 10^6/Y$	$\$17.06 \times 10^6$		17.06
TOTAL COST			179.08
LIQUEFACTION			
REFRIGERANTS N_2 (NOT REQ'D FOR PART. OXID. FEEDSTOCK) $0.04 \frac{\# N_2}{\# LH_2}$	$39.76 \times 10^3\ T$	\$6.50/T	0.26
C_1H_4 $0.01 \frac{\# C_1}{\# LH_2} = \frac{0.01\ \# C_1/\# LH_2}{41.6\ \# C_1/MCF} = 24 \times 10^{-5} \frac{MCF}{\# LH_2}$	0.477×10^6 MCF	\$0.363/MCF	0.17
C_2H_6 $0.02 \frac{\# C_2}{\# LH_2}$	$19.88 \times 10^3\ T$	\$25.00/T	0.50
C_2H_4 $0.02 \frac{\# E_2}{\# LH_2}$	$19.88 \times 10^3\ T$	\$80.00/T	1.59
COOLING WATER MAKE UP $24.7 \frac{\# H_2O}{\# LH_2}$	$24.55 \times 10^6\ T$	\$0.072/T	1.77
PROCESS ENERGY (HYDROCARBON FEEDSTOCK) $4.46 \frac{KWH}{\# LH_2}$	8.866×10^9 KWH	\$0.00288/KWH	25.53
(ELECTROLYSIS FEEDSTOCK) $4.46 \frac{KWH}{\# LH_2}$	8.866×10^9 KWH	\$0.00272/KWH	24.12
CAPITAL INVESTMENT $\$31.5 \times 10^6 \left(\frac{2805\ T/D}{250\ T/D}\right)^{0.8} \times 1.30 = \$282.96 \times 10^6$	$\$282.96 \times 10^6 \times 0.11$	———	31.13
OPERATING COST $\$1.91 \times 10^6/Y \left(\frac{2805\ T/D}{250\ T/D}\right)^{0.65} = \$9.02 \times 10^6/Y$	$\$9.02 \times 10^6$	———	9.02

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PROCESS, ITEM, UNIT REQUIREMENT New York	ANNUAL REQUIREMENT	UNIT COST	ANNUAL COST $\times 10^{-6}$ (\$)
TOTAL COST (STEAM REFORMING)			69.97
(PART OXID.)			69.71
(ELECTROLYSIS)			68.56
STORAGE			
CAPITAL INVESTMENT $2 \times \$9.6 \times 10^6 \times 1.300 = \$24.960 \times 10^6$	$\$24.960 \times 10^6 \times 0.11$	—	2.75
DISTRIBUTION			
CAPITAL INVESTMENT $2 \times 2\,\text{MI} \times \frac{5300\,\text{FT}}{\text{MI}} \times 24\,\text{IN} \times \frac{\$15}{\text{FT IN}} \times 1.300 = \$9.922 \times 10^6$	$\$9.922 \times 10^6 \times 0.11$		1.09

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PRODUCTION COST SUMMARY NEW YORK

GAS PLANT TYPE	GAS PLANT ANNUAL COST $\times 10^{-6}$ ($)	LIQ. PLANT ANNUAL COST $\times 10^{-6}$ ($)	STORAGE ANNUAL COST $\times 10^{-6}$ ($)	DISTRIB. ANNUAL COST $\times 10^{-6}$ ($)	TOTAL ANNUAL COST $\times 10^{-6}$ ($)	PRODUCT UNIT COST (¢/# LH_2)
STEAM REFORMING	74.05	69.97	2.75	1.09	147.86	8.1
PARTIAL OXIDATION	74.95	69.71	2.75	1.09	148.50	8.1
ELECTROLYSIS	179.08	68.56	2.75	1.09	251.48	13.8

Appendix F

LIST OF REFERENCES

1. Wellman, Paul and Katell, Sidney, U.S. Bureau of Mines, Dept. of Interior, Morganstown, W. Va., "How Pressure and Temperature Affect Steam-Methane Reforming," *Hydrocarbon Processing & Petroleum Refiner,* June 1963, Vol. 42, No. 6, p. 135.

2. M. W. Kellogg Company, New York, New York

3. Girdler Corporation, Louisville, Kentucky

4. Chemical Construction Corporation, New York, New York

5. Schweyer, Herbert E., *Process Engineering Economics,* McGraw-Hill Book Company Inc., 1955, p. 58.

6. Zimmerman, O. T. and Lavine, I., Industrial Research Service, Dover, New Hampshire, *Chemical Engineering Costs,* 1950, p.10.

7. Schweyer, Herbert E., *Process Engineering Economics,* McGraw-Hill Book Company Inc., 1955, p. 59.

8. Aries, Robert S. and Newton, Robert D., *Chemical Engineering Cost Estimation,* McGraw-Hill Book Company Inc., 1955, p. 15-16.

9. *The Oil and Gas Journal,* "Steam-Methane Reforming," Foster Wheeler Corporation, April 5, 1965, p. 111.

10. Nelson, W. L., *Guide to Refinery Operating Costs,* The Petroleum Publishing Co., Tulsa, Oklahoma, 1966.

11. *The Oil and Gas Journal,* "Four Improvements Feature Texaco Partial-Oxidation Process," October 7, 1963, p. 127.

12. *The Oil and Gas Journal,* "Four Improvements Feature Texaco Partial-Oxidation Process," October 7, 1963, p. 128.

13. Texaco, Inc., New York, New York

14. *Design Study of Hydrogen Production by Electrolysis,* Allis Chalmers, Research Div., Space and Defense Sciences, Milwaukee, Wis., Final Technial Report, Subcontract No. 2676 (Under Prime Contract No. W-7405-eng-26) prepared for Union Carbide Corporation, Nuclear Div., Oak Ridge, Tenn., Vol. I & II, October, 1966.

15. Costa, R. L. and Grimes, P. G., *Electrolysis as a Source of Hydrogen and Oxygen,* Chemical Engineering Progress Symposium Series, Nuclear Engineering - Part XVII, No. 71, Vol. 63, pp. 45-57.

16. Pieroni, L. J., et al, "Pipeline Gas from Coal Using Nuclear Heat," *Chemical Engineering Progress 60,* No. 6, June (1964)

17. Katell, S. & Faber, J. H., "What H_2 from Coal Costs," *Hydrocarbon Process & Petroleum Ref.,* 43, No. 3, March (1964)

18. Willmott, L. F., et al, *Gasification of Bituminous Coal with O_2 in a Pilot Plant Equipped with Slurry Feeding,* U.S. Bureau of Mines, R1 5996 (1962)

19. Gall, R. L., et al, *Gasification of Coal by Hot Recycle He in a Laboratory Scale Exchanger - Type Gasifier,* U.S. Bureau of Mines, R1 5996 (1962)

20. Curran, C. P., et al, "Carbon Dioxide Acceptor Process," *Chemical Engineering Progress,* 62, No. 2, (February, 1966)

21. Curran, C. P., et al, "CO_2 Acceptor Gasification Process," paper presented to A.C.S., Div. of Fuel Chem., Philadelphia, April 5-10, 1964

22. Lichtig, I., *Conversion of Coal to Liquid Fuels,* U.S. Bureau of Mines, IC 8065 (1961)

23. "Coal Gasification Approaches Pilot Plant Stage," *Chemical Engineering News,* April 18, 1966, pp. 68-72.

24. Linde, H. R., "Coal Gasification . . . .and Natural Gas," *American Gas Journal,* April, 1967, pp. 19-25.

25. *Ammonia Production Feasibility Study,* Contract DA-44-009-AMC-828(X) for U.S.Army Engineer Reactors Group, Army Nuclear Power Group, Fort Belvoir, Va., Dept. of Army Task No. 4A643318D88704 by Air Products and Chemicals, Inc., July, 1965.

26. *Technical Aspects of Ortho-Parahydrogen Conversion,* Contract AF33(616)-7506 and AF33 (657)-10057 for Aeronautical Systems Division of the United States Air Force, Wright-Patterson Air Force Base, Ohio, by Air Products and Chemicals, Inc. 1963.

27. Scott, R. B., *Cryogenic Engineering,* D. Van Nostrand Company, Inc., 1963, p. 290

28. Chicago Bridge and Iron Company, Oak Brook, Illinois

29. *Advances in Cryogenic Engineering,* "Dynamic Response of Fluid and Wall Temperatures during Pressurized Discharge of Liquid from a Container," V. S. Arpaci, J. A. Clark, and W. D. Winer, University of Michigan, Ann Arbor, Michigan, 1961.

30. Institute of Gas Technology, Lowell Gas Co., Lowell, Massachusetts

31. *Advances in Cryogenic Engineering,* "Gas Requirements in Pressurized Transfer of Liquid Hydrogen," D. F. Gluck and J. F. Kline, Lewis Research Center, National Aeronautics and Space Administration, Cleveland, Ohio, 1962.

32. *Advances in Cryogenic Engineering,* "Analysis of the Pressurizing Gas Requirements for an Evaporated Propellant Pressurization System," E. F. Coxe and J. W. Tatom, Lockheed-Georgia Company, Division of Lockheed Aircraft Corp., Marietta, Georgia, 1962.

33. Vance, R. W., and Duke, W. M., *Applied Cryogenic Engineering,* J. Wiley & Sons, New York, 1962.

34. "The Petroleum Situation," Chase Manhattan Bank, New York, October, 1966.

35. Netschert, B. C. and Schurr, S. H., "Energy in the American Economy, Its History and Prospects, 1850-1975," Baltimore, 1960.

36. Netschert, B. C., "The Future Supply of Oil and Gas," Baltimore, 1958

37. "Our Future Gas Supply," *Gas Age,* July, 1966.

38. "Coal," National Coal Policy Conference, Inc., Washington, 1966.

39. No. S-38-66, Seaborg, Glenn T., Dr., Chairman, U.S. Atomic Energy Commission, October 24, 1966.

40. "World Energy Supplies, 1961-1964," Statistical Papers, Series J., No. 9, United Nations, New York, 1966

41. "Mineral Fuels," Minerals Yearbook, Vol II, 1964, U.S. Bureau of Mines, Washington, 1965.

42. "World Wide Oil Report," *Oil and Gas Journal,* December 26,1966.

43. "World Petroleum Report," *Oil and Gas Journal,* Vol. XII, March 15, 1966.

44. "World Petroleum Report," *Oil and Gas Journal,* Vol. XIII, 1967.

45. "Refinery Survey Issue," *Petroleum Times,* January 7, 1966.

46. "World Wide Oil," *Oil and Gas Journal,* December 27, 1965.

47. "Annual Refining Number," *Oil and Gas Journal,* April 3, 1967.

48. "World Crude Production Survey," *Petroleum Times,* July 22, 1966.

49. World Petroleum, Vol. 38, No. 2, February, 1967.

50. "Third Annual Natural Gas in Western Eruope Number," *Petroleum Times,* Nov. 11, 1966.

51. "Summary of Developed and Potential Waterpower of the United States and Other Countries of the World, 1955-62," *Geological Survey Circular 483,* Washington, 1964.

52. "Facts and Figures," *An Investor's Guide to Thailand,* Bangkok Bank Ltd., Bangkok, 1966.

53. "Monthly Review," Bangkok Bank Limited, Bangkok, July, 1966.

54. "European Chemical Prices," *European Chemical News,* Vol. 12, No. 287, July 28, 1967.

55. "Atlas of Crude Oil Pipelines," *Oil and Gas Journal,* 1966.

56. *Pocket Data Book, U.S.A.,* 1967, U.S. Dept. of Commerce, Washington, D.C., December, 1966.

57. "Supplies, Costs, and Uses of the Fossil Fuels," Energy Policy Staff, U.S. Department of the Interior, Washington, D.C., February, 1963.

58. F.P.C. S-171 Steam-Electric Plant Construction Cost and Annual Production Expenses, *Seventeenth Annual Supplement,* 1964, p. x-xv.

59. "Area Reports: International," *Mineral Yearbook, Vol IV,* 1963, U.S. Bureau of Mines, Washington, D.C., 1964.

60. "Area Reports: International," *Mineral Yearbook, Vol. IV,* 1964, U.S. Bureau of Mines, Washington, D.C., 1966.

61. "Nouvelles Réflexions sur les Perspectives Energétiques à Long Terme de la Communauté Européennee," *Bulletin de la Communauté Europeenne du Charbon et de l'Acier, No. 61,* Luxembourg, 1966.

62. "Memorandum on Energy Policy," The Commission of the European Economic Community, Luxembourg, August, 1962.

63. "Gas Supplies for France," American Embassy, Paris, March 15, 1966.

64. "Petroleum in Western Europe Today," *Oil and Gas Journal,* April 18, 1966.

65. "Shipping and Ocean Freight Rates," United Nations, Regional Economic Cooperation Series No. 3, (Asia and the Far East), New York, 1965.

66. "Japanese Economic Statistics," *Bulletin 95,* Economic Planning Agency, Japanese Government, August, 1966.

67. "Monthly Statistics of Japan," *Bulletin 61,* Economic Planning Agency, Japanese Government, July, 1966.

68. "Basic Data on the Economy of Japan," Part 1, No. 62-61, U.S. Dept. of Commerce, November, 1965.

69. "Background Notes - Japan," *Publication 7770,* U.S. Department of State, Washington, D.C., December, 1964.

[illegible]0. "Economic Survey of Asia and the Far East, 1965," United Nations, New York, 1966.

71. "Nineteenth Annual Report, 1965-66," Joint Coal Board, New South Wales, Sydney, October, 1966.

72. "Establishment of Industry in Australia," Australia and New Zealand Bank, Ltd., Melbourne, 1966.

73. "Costs, etc. - Coal Oil, and Natural Gas," New South Wales Department of Decentralization and Development, Sydney, February, 1967.

74. "Natural Gas in Australia," New South Wales Department of Decentralization and Development, Sydney, November, 1965.

75. Oil and Australia, 1966, Petroleum Information Bureau, Melbourne, 1966.

76. Oil and Australia, 1965, Petroleum Information Bureau, Melbourne, 1965.

77. "Petroleum Search in Australia," Petroleum Information Bureau, Melbourne, January, 1967.

78. "Basic Data on the Economy of Australia," *Overseas Business Report, OBR63-38,* U.S. Department of Commerce, Washington, D.C., March, 1963.

79. "Establishing a Business in Australia," *Overseas Business Report, OBR63-108,* U.S. Department of Commerce, Washington, D.C., July, 1963.

80. "El Petroleo de la America Latina," (Map), Petroleum Publishing Co., Tulsa, April, 1957.

81. "El Petroleo de Mexico, Centro y Sudamerica y las Antillas en 1963," (Map), Petroleum Publishing Co., Tulsa, June, 1963.

82. "Basic Data on the Economy of Lebanon," *Overseas Business Report, OBR65-79,* U.S. Department of Commerce, Washington, D.C., December, 1965

83. "Guide to Eastern Oil," (Map), *Oil and Gas International, Vol. 3, No. 6,* June, 1963.

84. "Market Profile - Republic of South Africa," *Overseas Business Report, OBR66-36,* U.S. Department of Commerce, Washington, D.C., June, 1966.

85. "Basic Data on the Economy of South Africa," *Overseas Business Report, OBR66-71,* U.S. Department of Commerce, Washington, D.C., October, 1966.

86. "South Africa - An Economic Survey," Barclay's Bank, D.C.O., London, March, 1966

87. "Review of Long-Term Economic Projections for Selected Countries in the ECAFE Region," Development Programming Technique Series No. 5, United Nations Economic Commission for Asia and the Far East, Bangkok, 1964.

88. "Basic Data of the Economy of Thailand," *Overseas Business Report, OBR66-60,* U.S. Department of Commerce, Washington, D.C., September, 1966.

89. "Private Enterprise Investment Opportunities in Thailand," AID, American Embassy, Bangkok, February, 1966.

90. "Investment in Southeast Asia," Agency for International Development, Washington, D.C., July,1966.

91. *World Power Data, 1964,* Federal Power Commission, Washington, D.C., May, 1966

92. *National Power Survey,* The Federal Power Commission, Parts I and II, 1964.

93. *Monthly Bulletin of Statistics,* United Nations, New York, March, 1967.

94. *Statistical Yearbook, 1964,* United Nations, New York, 1965.

95. *Investigation of Energy Cost,* Gilbert Associates, June 14, 1967.

96. *Annual Report, 1966,* The Electricity Commission of New South Wales, Sydney, November, 1966.

97. Comparison of Coal-Fired and Nuclear Power Plants for the TVA System, Office of Power, Tennessee Valley Authority, Chattanooga, Tennessee, June, 1966.

98. *Only Breeder Can Make Reactor Industry Viable, (preliminary draft),* March 13, 1967, Wright, J. H., Technical Director, Advanced Reactors Division, Westinghouse Electric Corporation.

99. *Cost Analysis of Fossil Fueled Electric Power Generating Systems,* Westinghouse Electric Corporation, Pittsburgh, Pennsylvania, January, 1966.

100. *Nuclear Systems Cost,* Stinson, W. H., Advisory Engineer, Public Utility Headquarters Division, Westinghouse Electric Corporation.

101. *F.P.C. S-180 Hydroelectric Plant Construction Cost and Annual Production Expenses,* Ninth Annual Supplement, 1965.

102. *Consolidated Edison Analysis,* May 31, 1967, Duncan, T.C., Vice President.

103. *Southern California Edison Analysis,* April 17, 1967, Drewry, W.C., Vice President.