Economics of Seawater Desalting in Combination with Ammonia and Power Production

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Economics of Seawater Desalting in Combination with Ammonia and Power Production

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with

Annex A: "Seawater Desalting by Distillation Processes, Including Vapor Compression, As Applied to Integral Chemical Complexes", by S. J. Senatore, Oak Ridge Gaseous Diffusion Plant, Oak Ridge, Tennessee, for Chung-ming Wong, et al, Office of Saline Water.

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UNITED STATES DEPARTMENT OF THE INTERIOR • Walter J. Hickel, Secretary
Carl L. Klein, Assistant Secretary for Water Quality and Research

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FOREWORD

This is one of a continuing series of reports designed to present accounts of progress in saline water conversion and the economics of its application. Such data are expected to contribute to the long-range development of economical processes applicable to low-cost demineralization of sea and other saline water.

Except for minor editing, the data herein are as contained in a report submitted by the contractor. The data and conclusions given in the report are essentially those of the contractor and are not necessarily endorsed by the Department of the Interior.

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I. INTRODUCTION

A. OBJECTIVE

The purpose of this study is to examine the feasibility of lowering the operating cost of producing desalted water by combining its production with that of ammonia (one of the three basic fertilizer materials), and also of electricity, through:

1. The recovery and utilization of the low temperature heat which, in conventional ammonia manufacturing processes, is dissipated to cooling water and thence to the atmosphere.

2. The employment of a common energy center for the production of the high temperature heat needed both for the reforming of natural gas to produce an ammonia synthesis gas, and for the production of high pressure steam. (High pressure steam is used to drive the large ammonia plant compressors. It may also be used to drive the vapor compressors of a VC-VTE plant for the production of desalted water, or to drive turbo-generators.)

As described below, a number of alternate process flowsheets are evaluated, all comprising modern, large-scale fossil-fueled plants. In addition, one case of nuclear energy supply is included in the number of cases studied.

The processes assumed for water desalination are the MSF (multi-stage flash), VC-VTE (vapor recompression utilizing long tube evaporators), and a combination of the two. These are the distillation processes now under most active consideration by the Office of Saline Water for the desalination of ocean water.

Production of ammonia is accomplished by the reforming of natural gas. (There is no significant change in process when naphtha is the raw material .) Most of the ammonia produced in the world today is manufactured in this manner.

To determine the effect of plant scale on unit cost of production, a range of ammonia and water plant capacities are studied.

Preliminary estimates of the capital, operating and unit cost economics of water desalting are derived for each of the alternates investigated.

B. SCOPE

1. Ammonia

A 500 T/D plant is perhaps the smallest that can utilize centrifugal synthesis gas compressors, and thus the smallest of the highly efficient, modern plants. Most of the ammonia plants built during the past few years have capacities ranging from 600-1000 T/D. The nominal rating of the largest single-train plant built to date is 1500 T/D, although it is understood that higher rates have been achieved in actual production. Modern multi-train units at a single plant site have been rated at a total of 3000 T/D.

To encompass the range of plant capacities of industrial importance today and for which technology is currently available four ammonia plant sizes were selected for evaluation : 500, 1000, 1500 and 2000 T/D.

Although architect-engineering firms are prepared to offer single-train 2000 T/D units, it was decided that the more conservative approach of determining unit costs on the basis of two 1000 T/D trains would be utilized in this study. Capital and operating costs for the smaller ammonia plants would be based on a single-train concept.

2. Desalted Water

In the first four alternates studied water production was limited to that amount which could be produced by the recovery of the low temperature heat normally discarded in an ammonia plant. This energy is utilized in horizontal, multi-stage flash plants. The resulting water production rates are 12.5, 21.6, 30.8 and 43.2 MGD at ammonia plant capacities of 500, 1000, 1500 and 2000 T/D respectively.

The remaining alternates were selected to evaluate the effect of employing supplemental energy sources to increase the production rate of water beyond that obtainable merely by recovery of the available low temperature heat in the ammonia plant.

In Alternate 4A, gas turbines are used to drive vapor compressors in a VC-VTE cycle, thereby increasing the production of water in conjunction with a 2000 T/D ammonia plant from 43.2 to 100 MGD.

Alternate 4B employs the same design concept as Alternate 4A, and the same capacity of desalted water and ammonia. However, Alternate 4B envisages the ducting of hot exhaust gases from the gas turbines to the ammonia plant, the gases being used as combustion air in the reforming furnace.

In Alternate 5 a study was made of the effect of increasing the amount of water co-produced with a 1000 T/D ammonia plant to 100 MGD. As in the previous two alternate cases, vapor compressors are used. However, in Alternate 5 the vapor compressors are driven by steam turbines.

Since energy at a high temperature level would be required both for the production of high pressure steam and for the reforming of natural gas, a study of the feasibility of combining both services in a fossil-fired common energy center was undertaken.

3. Electricity

An earlier study (1) examined the feasibility of co-producing desalted water, ammonia and electrical power in a single plant complex containing a fossil-fired energy center. The results of that investigation indicate that the proposed method will produce water at a relatively low unit cost - although requiring a larger capital investment - when compared with other methods of desalinating sea water. Alternate 6, studied herein, is a follow-on to that study. The capacity of the power generating unit is assumed to be 150 MW(e) gross output and the ammonia plant capacity is assumed to be 1000 T/D. The amount of water production is that which may be obtained by utilizing all of the low pressure steam produced in the ammonia plant and the power plant, resulting in a complex which is thermodynamically balanced to preclude the necessity of condensing the exhaust from a steam turbine. Under these conditions it is possible to produce 101.4 MGD of desalted water in the complex.

4. Nuclear Energy

Alternate 7 examines the feasibility of utilizing a nuclear reactor as a high temperature energy center for the co-production of desalted water and ammonia. An 840 MW(th) high-temperature gas-cooled reactor is chosen for the prototype. The amount of ammonia produced is that which may be obtained by utilizing the high temperature portion of the circulating helium stream to reform methane. This calculated to be approximately 1500 T/D. The balance of the heat generated in the reactor is used to preheat process streams for the ammonia process and to generate high pressure steam.

This steam is used to drive both ammonia plant compressors and VC-VTE vapor compressors, as in previous Alternates, and exhausts into the low pressure steam header of an MSF system. The results show that a total of 179 MGD of desalted water can be produced.

5. Summary

Table I presents a summary of each of the Alternates which were investigated.

C. ACKNOWLEDGMENTS

Mr. E. F. Miller, of the Office of Saline Water, U. S. Department of the Interior, Washington, D. C. was responsible for overall coordination and administration of this project. His assistance on many technical aspects of this study is gratefully acknowledged.

Mr. S. J. Senatore, of the Oak Ridge Gaseous Diffusion Plant, Oak Ridge, Tennessee prepared the basic capital and operating cost data for all the desalination plants, and provided consulting assistance and direction in the selection of the desalination process most appropriate for each of the Alternates.

Dr. Maynard Born and Dr. Robert Simon, of Gulf-General Atomic Inc.,helped explore the feasibility of adopting their HTGR design to meet the requirements of a nuclear energy center; also, they provided the basic investment and operating data for the reactor described in Alternate 7.

TABLE I. ALTERNATES INVESTIGATED

Alternate No	Source of Energy	Method of Producing Water	Plant Capacity			Comment
			Water MGD	Ammonia T/D	Power MW(e)	
1	Natural Gas	MSF	12.5	500	nil	
2	Natural Gas	MSF	21.6	1000	nil	
3	Natural Gas	MSF	30.8	1500	nil	
4A	Natural Gas	VC-MSF	100.0	2000	nil	VC driven by gas turbines
4B	Natural Gas	VC-MSF	100.0	2000	nil	Same as 4A, but turbine exhaust used in Primary Reformer
5	Natural Gas	VC-MSF	100.0	1000	nil	VC driven by steam turbines
6	Natural Gas	MSF	100.0	1000	150	Triple Purpose Case
7	Nuclear	VC-MSF	180.0	1500	nil	VC driven by steam turbines

Many more individuals and companies provided assistance in this study - more than can be noted here. However, particular recognition is given to the assistance of the Riley Stoker Corporation in developing the concept of the fossil-fired energy center; to Dr. David Hall of the Los Alamos Scientific Laboratory for technical information pertaining to the "UTHREX" experiment (Ultra High Temperature Reactor Experiment); to the Benfield Corporation for data pertaining to the removal of carbon dioxide from synthesis gas; to Bros Incorporated for data pertaining to the ammonia plant Convection Section; to the Cooper-Bessemer Corporation and the Clark Bros. Company for data pertaining to ammonia plant compressors; and to the General Electric Company and the DeLaval Company for data pertaining to steam turbines.

II. SUMMARY AND CONCLUSIONS

A. GENERAL CONCLUSIONS

The conclusions of this study may be stated briefly as follows:

1. The cost of producing desalted water from sea water in a single purpose MSF plant can be reduced almost 30% by combining its production with that of ammonia.

2. The unit cost of desalted water co-produced with ammonia (and electricity) varies from 27 to 35 cents per thousand gallons in plants having capacities ranging from 12.5 to 180 MGD. (These costs are based on an assumed capital charge of 7% of investment and a natural gas cost of 25¢ per 10^6 Btu.)

3. In such plants the net investment for water desalination varies from \$550,000 to \$925,000 per MGD, being dependent upon plant capacity and process.

4. Between 2.0 and 2.5 MGD of water per 100 T/D of ammonia can be recovered from low temperature heat in a modern ammonia plant. When larger amounts of water are required, a vapor compression cycle can be readily added to the multi-stage flash plant. Alternatively, if there is a demand for electrical power, the plant can be designed to co-produce water, ammonia and electricity.

5. Vapor compressors can be driven by completely independent gas turbines. However, production costs are reduced if the turbine exhaust gases are used as combustion air in the gas reforming furnace of the ammonia plant.

6. A further reduction of unit water costs will result from the employment of a common energy center to produce the high temperature energy utilized both in the gas reforming process and for the generation of steam to drive the vapor compressors. This common energy center may use fossil fuel or nuclear fuel.

7. Use of the concept of a common energy center enables one to design plants having a wide range of relative capacities of ammonia, desalted water and/or electric power.

8. It is entirely feasible, employing current technology, to produce an ammonia synthesis gas within the containment shell of a high temperature gas-cooled reactor, using nuclear heat directly to supply the endothermic heat of reaction of the gas reforming step.

9. The use of nuclear energy, as described above, starts to become economically attractive when the cost of natural gas or other hydrocarbon fuel exceeds 25 cents per 10^6 Btu.

10. On the basis of a preliminary examination it appears that there may be regions of the world in which the economy could be improved significantly if both fresh water and fertilizer could be made available in large quantities. In these regions the employment of the common energy center

concept of the type evaluated in this study might prove attractive.

B. COST COMPARISONS

1. Table II presents a summary of investment and operating costs for each of the multiple-product plants studied, and production costs for equivalent standard single-purpose MSF plants. The data are shown graphically in Fig. 1.

2. Table III lists the total investment required for each Alternate, and the unit cost of each co-product.

C. RECOVERY AND UTILIZATION OF LOW TEMPERATURE HEAT (ALTERNATES 1, 2, 3 and 4)

1. It is relatively simple to modify the design of a modern ammonia plant to incorporate features which enable low temperature heat to be recovered for use in a desalination plant. This energy can be obtained in the form of low pressure steam (from non-condensing steam turbines and low pressure waste heat boilers) which can be used in the brine heaters of an MSF plant; and in the form of process heat which can be imparted to a circulating brine stream.

2. For ammonia plant capacities in the range of 500-2000 T/D, between 2.0 and 2.5 MGD of desalted water can be produced for each 100 T/D of ammonia. Because compressors

TABLE II. INVESTMENT AND PRODUCTION COSTS OF WATER DESALTING PLANTS

Plant Capacity-MGD	12.5	22.0	30.8	43.2	100.0	100.0	100.0	100.0	180.0
Alternate No.	1	2	3	4	4A	4B	5	6	7
Method of Desalting Water	MSF	MSF	MSF	MSF	VC-MSF	VC-MSF	VC-MSF	MSF	VC-MSF
Net Investment ($000)	11,580	17,980	25,850	33,570	71,370	71,630	71,330	61,430	97,800
Net Investment per MGD ($000)	925	815	840	777	714	716	713	607	545
Unit Cost of Water-¢/Kgal. (1)	34.7	31.9	31.1	30.4	31.4	30.4	29.2	26.8	27.3
Comparable Cost of Water from Single-Purpose MSF Plants-¢/Kgal (1) (2)	51	49	46	41	38	38	38	38	37

(1) Capital charges equal 7% of investment; cost of natural gas is 25¢/10^6 Btu (L.H.V.)

(2) Ref. 2

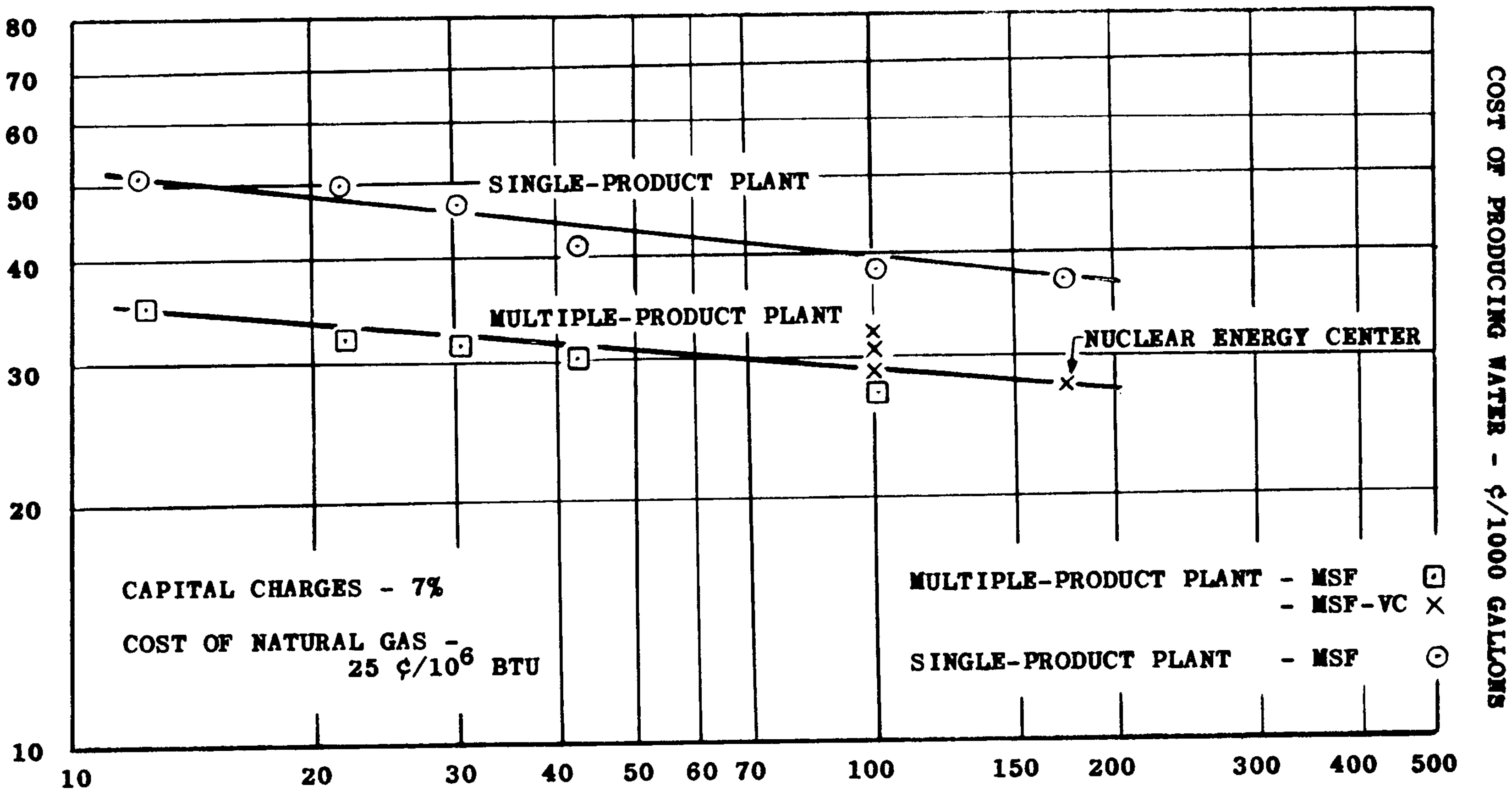


Figure 1-Unit Cost of Desalting Water As A Function of Plant Size

TABLE III. TOTAL INVESTMENT AND PRODUCTION COSTS OF EACH ALTERNATE

Alternate No.	1	2	3	4	4A	4B	5	6	7
<u>Plant Capacity</u>									
Desalted Water-MGD	12.5	21.6	30.8	43.2	100	100	100	100	180
Ammonia-T/D	500	1000	1500	2000	2000	2000	1000	1000	1500
Electricity-MW (e)	-	-	-	-	-	-	-	150	-
Nuclear Reactor-MW (th)	-	-	-	-	-	-	-	-	840
<u>Plant Cost ($000)</u>									
Desalted Water	11,580	17,980	25,850	33,570	71,370	71,630	71,330	61,430	97,800
Ammonia	9,550	16,200	21,700	31,000	31,000	31,000	16,200	16,200	21,300
Electricity	-	-	-	-	-	-	-	16,500	-
Nuclear Reactor	-	-	-	-	-	-	-	-	45,200
Total	21,130	34,180	47,550	64,570	102,370	102,630	87,530	94,130	164,300
<u>Unit Cost of Production</u>									
Desalted Water-¢/Kgal (1)	34.7	31.9	31.1	30.4	31.4	30.4	29.2	26.8	27.3
Ammonia-$/Ton (2)	22.10	18.60	17.15	17.50	17.50	17.50	18.60	18.50	17.50
Electricity-mils/Kwh (1)	-	-	-	-	-	-	-	4.2	-

(1) Capital charges equal 7% of investment; cost of natural gas is 25¢/10^6 Btu (L.H.V.)

(2) Capital charges equal 14% of investment; cost of natural gas is 25¢/10^6 Btu (L.H.V.)

and turbines are more efficient as their sizes increase, the relative amount of recoverable heat - and hence of water production - decreases as the size of a single train ammonia plant increases. Alternatively stated, inefficiencies of energy utilization in small ammonia plants may be recovered by the production of water.

3. The cost of water co-produced with ammonia varies from 30¢ to 35¢ per Kgal., decreasing as the plant size is increased.

4. Because the capital investment is so large, the cost of water production is not too greatly influenced by a change in the cost of natural gas. For example, a 60% change in the price of gas, from 25¢ to either 10¢ or to 40¢, will change the cost of water approximately 5¢/Kgal. (Fig. 2). This is equal to a 15 1/2% change in the cost of water produced in the 43 MGD plant (Alternate 4) and a 13 1/2% change in the cost of water produced in the 12.5 MGD plant (Alternate 1).

5. However, the value of water is sensitive to the value which has been assigned to the ammonia. At the design point, a $1 per ton change in ammonia value, approximately 5%, will be reflected by a 10 to 15% change in the value of water (Fig. 12). The basis for assigning production costs is discussed in Section IV D.

D. <u>INCREASED WATER PRODUCTION BY THE ADDITION OF A VC-VTE CYCLE AND A FOSSIL-FIRED COMMON ENERGY CENTER (ALTERNATES 4A, 4B and 5)</u>

1. Employing a vapor compression cycle to increase the

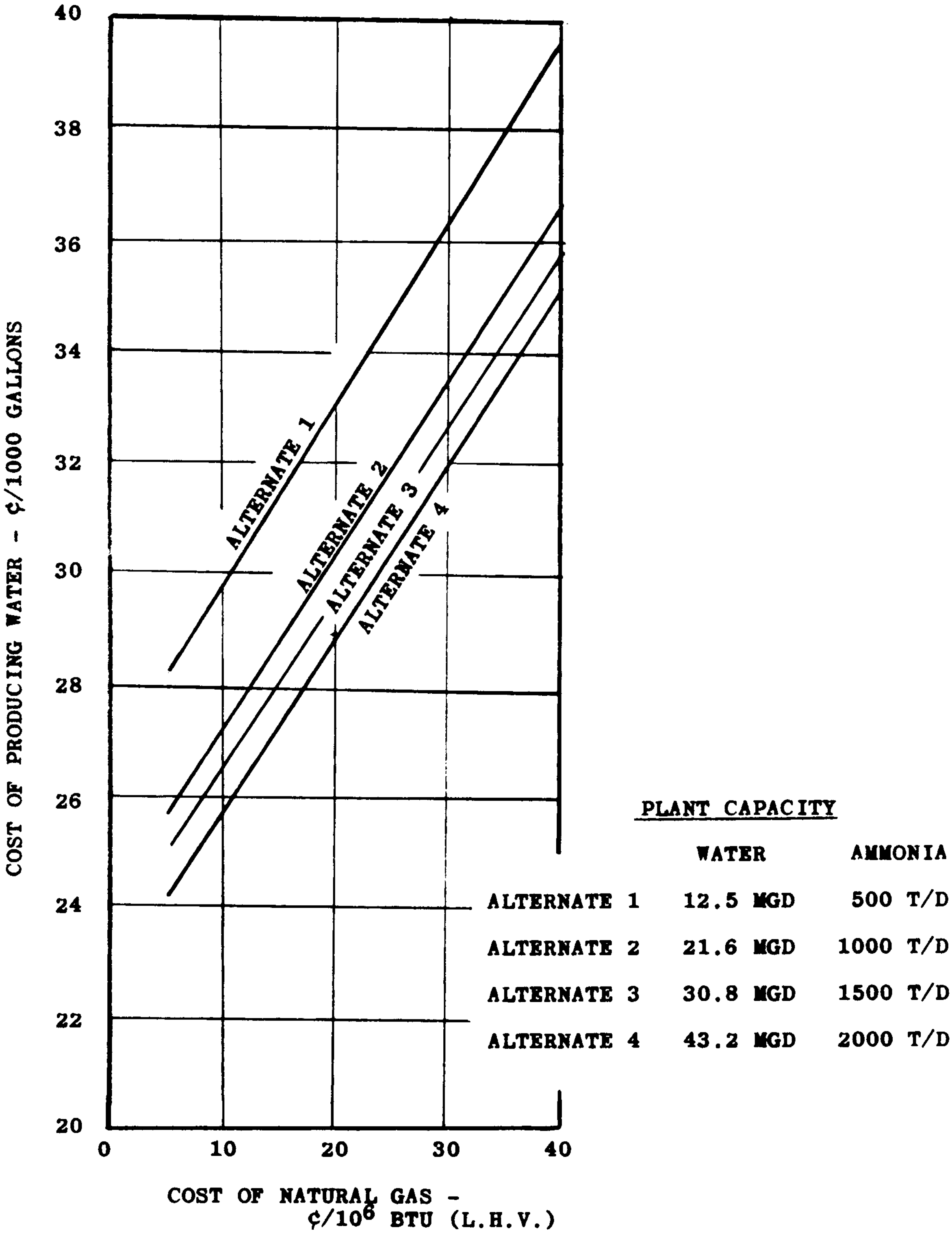


Figure 2-Unit cost of Desalting Water As A Function of Natural Gas Cost (Small Multiple-Product Plants - Alternates 1, 2, 3, & 4)

water desalination capacity of a complex causes a slight increase in the average cost of water production. However, the cost of water is still much less than if it had been manufactured in a single-purpose MSF plant (Fig 1).

2. If gas turbines are used to drive the vapor compressors, overall economy can be achieved if a portion of the exhaust gas is used to replace combustion air in the ammonia plant reforming furnace. This saving will be 0.5 to 1.5¢ per Kgal., depending upon the cost of natural gas (Fig 3).

3. An additional saving of approximately 1¢ per Kgal can be achieved if steam turbines, rather than gas turbines, are used to drive the vapor compressors; and if the ammonia plant gas reforming duty and the high pressure steam boiler duty are combined in a common energy center (Fig 3).

4. Since the ratio of production of water to ammonia has increased, the value of water is not as sensitive to that assigned to ammonia as it had been in the first four Alternates (Fig. 17).

E. CO-PRODUCTION OF DESALTED WATER, AMMONIA AND ELECTRICAL POWER (ALTERNATE 6)

1. The design based on this Alternate produced the lowest unit cost of desalted water - 27¢ per Kgal. This is almost 2 1/2 ¢ lower than the cost of water produced in a plant using steam driven vapor compressors, although the capacity of both plants is 100 MGD (Fig 4).

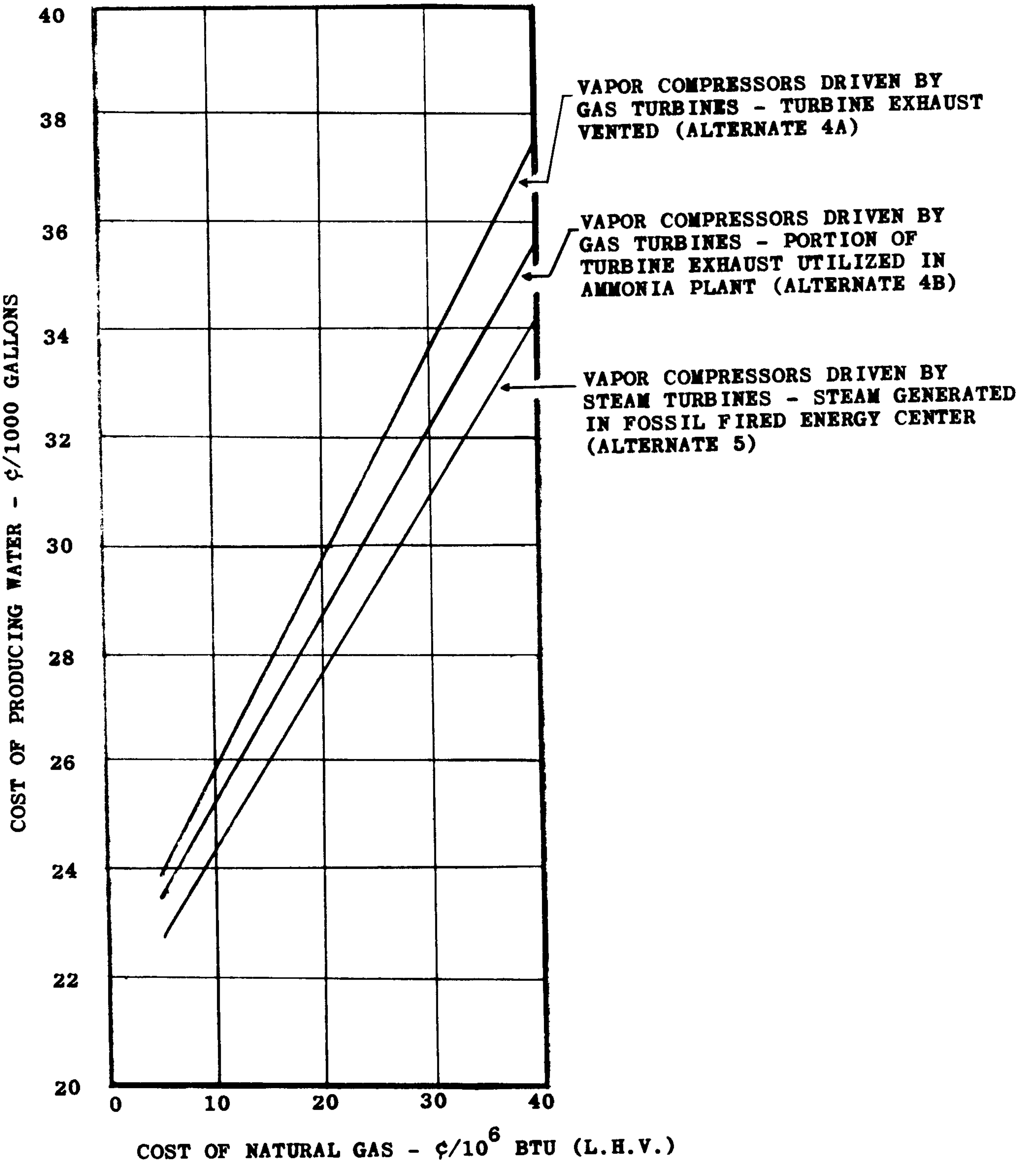


Figure 3-Unit Cost of Desalting Water As A Function of Natural Gas Cost (100 MGD Multiple-Product Plants - Alternates 4A, 4B & 5)

2. As in the case of Alternate 5, a change in the value assigned to ammonia, when expressed in terms of dollars per ton, is reflected by a substantially equal change in the cost of water expressed in terms of cents per Kgal. when the value of electric power is maintained constant (Fig 17d). The value assigned to electric power is that calculated for each assumed gas cost using the Fruth method of exergy proration.

Alternatively, the value assigned to electrical power may be varied, with the value of ammonia held constant at each gas cost (Fig 17e).

F. <u>NUCLEAR FUELED COMMON ENERGY CENTER (ALTERNATE 7)</u>

1. The cost of producing water in a plant based on this Alternate is independent of the cost of natural gas, since in this case natural gas serves only in the role of being an essential material for the reforming reaction and is not also used as a heat source.

2. Fig (4) illustrates the effect of gas cost for Alternates 5,6 and 7. If the production rate of Alternates 5 and 6 were increased to 180 MGD, one would expect a 2¢/Kgal decrease in the cost of producing water. Therefore, if a complex were to produce ammonia and water, the nuclear energy center would be competitive with fossil fuel at a natural gas cost of about 28¢/10^6 Btu. However, if the plant were to produce ammonia, water and electric power, then nuclear energy would not be competitive unless the gas cost was about 33¢/10^6 Btu.

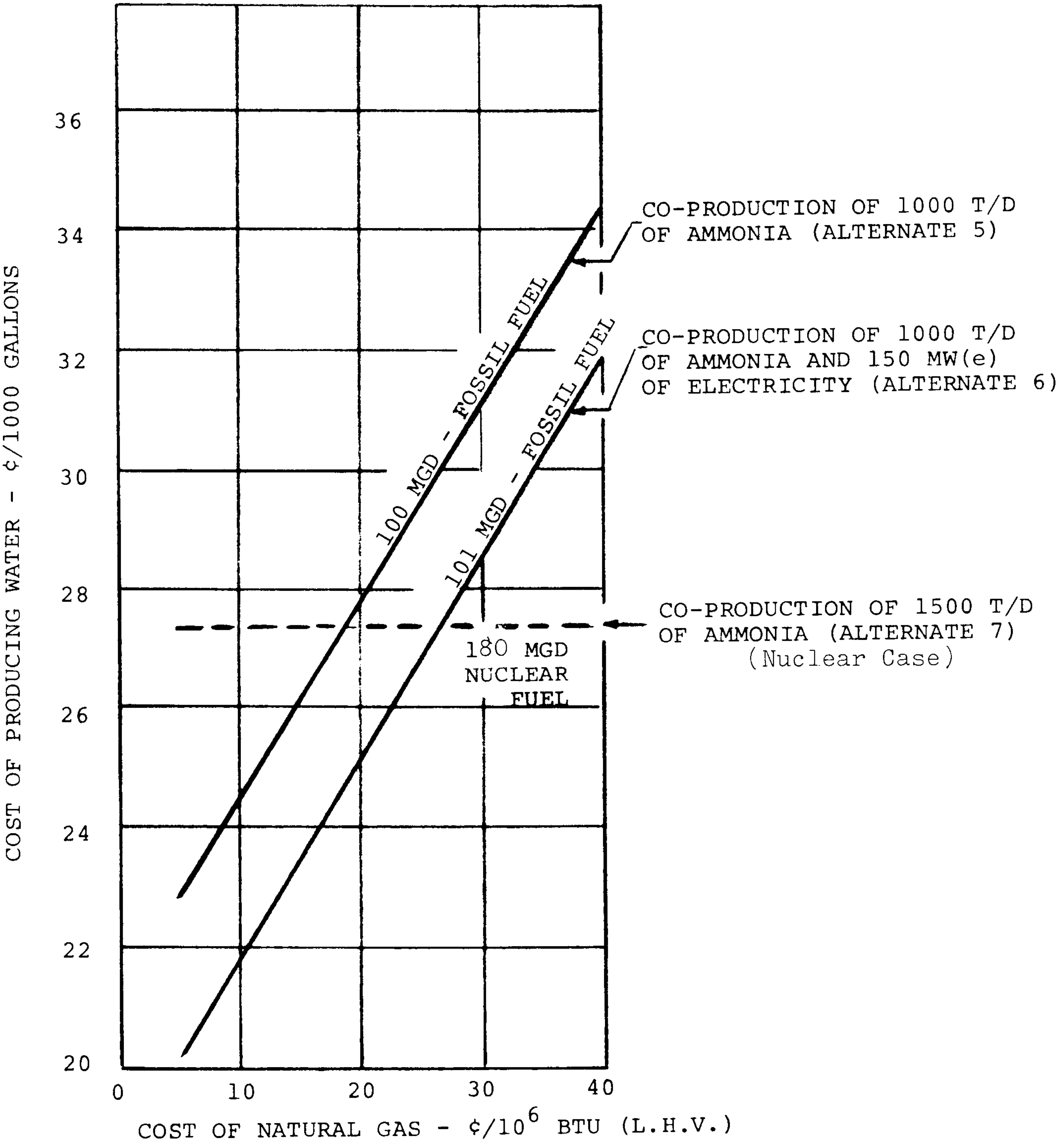


Figure 4-Unit Cost of Desalting Water As A Function of Natural Gas Cost (Large Multiple-Product Plants-Alternates 5, 6 & 7)

3. As shown in Fig 5, the costs of using either nuclear energy or fossil energy to produce ammonia are equal when natural gas costs about 28¢/10^6Btu. Therefore, in areas of higher gas cost, savings in ammonia production costs via the nuclear heat route may be used to decrease the value assigned to desalted water.

4. The effect of arbitrary changes in the assigned production of ammonia and water is shown in Fig. 17f.

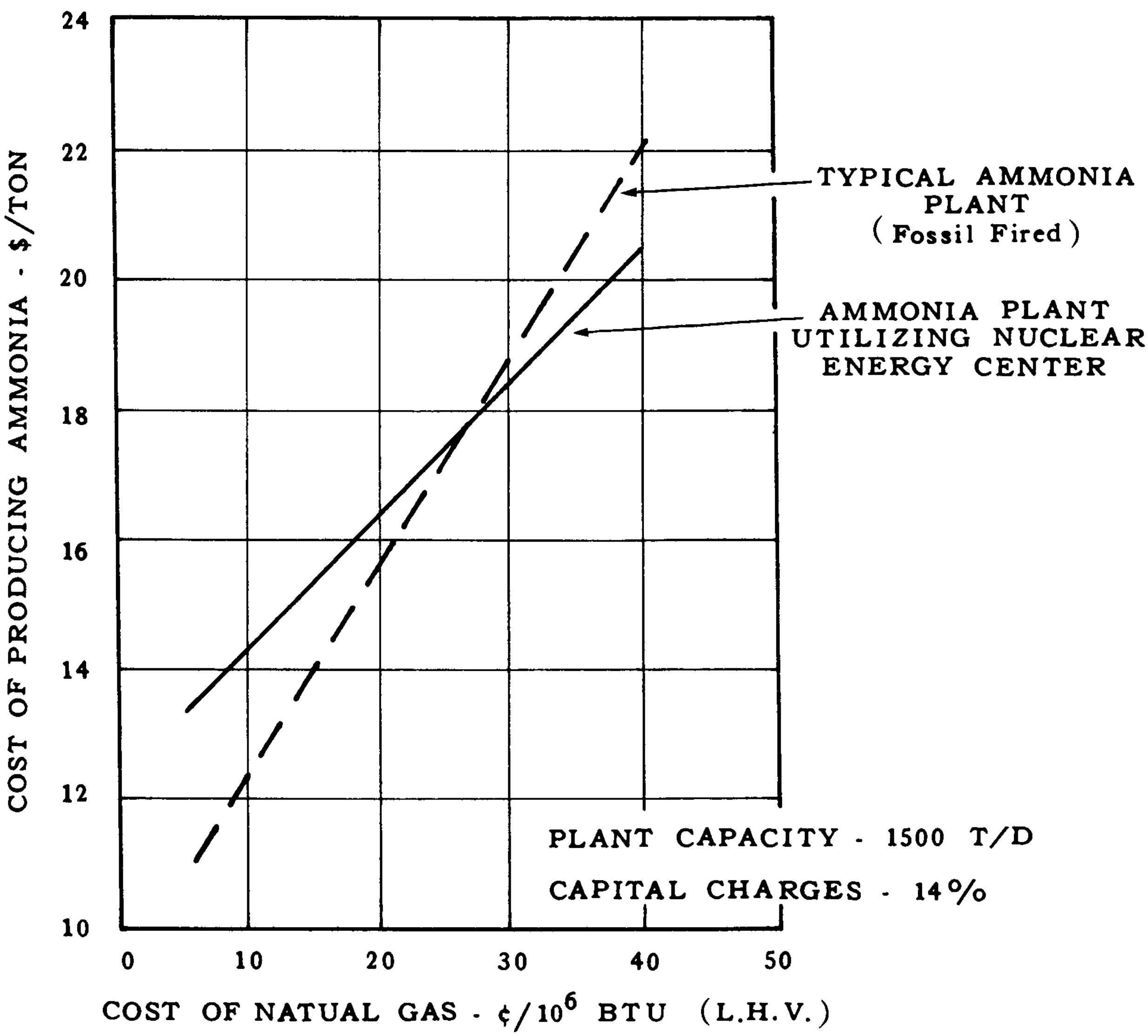


Figure 5 - Unit Cost of Ammonia as a Function of Natural Gas Cost

III. IMPLICATIONS OF THE STUDY: RECOMMENDATIONS

A. OBJECTIVES AND CONCLUSIONS

In very general terms the objectives of this investigation are to study the feasibility of recovering low temperature heat from a petrochemical (ammonia) plant, utilizing this heat for the desalination of water; to investigate the possibility of combining the desalination of reasonably large quantities of water (approx. 100 MGD) with other processes, such as the production of ammonia and/or electricity, in order to reduce costs by sharing certain common facilities; and finally, to study the feasibility of utilizing a nuclear reactor as a direct source of high temperature heat for these processes.

The results of the study indicate that the technology required by all of these concepts is available at the present time, and that their utilization would reduce the cost of producing fresh water significantly below that attainable in a single purpose plant.

B. PROCESS IMPROVEMENTS

As these studies were to be of a fundamental character, and a starting point was required, the basic characteristics of the desalination plant were fixed by referral to previous studies (3) (4). It was not important at that time, nor was it possible before the completion of this study, to optimize the desalination design in accordance with the overall multi-product process.

However, if a particular site were to be selected now, and the cost of energy and of capital established at that particular location, it would probably be possible to decrease the cost of desalting water about 10% below the amounts here calculated. This would be done by determining the optimum performance ratio for the local conditions and, in addition, changing the drives on the large brine and product pumps from motors to steam turbines.

C. NUCLEAR ENERGY CENTER

In areas of relatively high natural gas price, where the use of nuclear energy as the heat source is attractive, an additional reduction of cost can be achieved by using a larger reactor. An 840 MW(th) reactor, used in this study, is considered fairly small and expensive by today's standards. Doubling its size, for example, would decrease the cost of producing water to 24¢/Kgal, and would decrease the cost of ammonia to $16.50.

However, an increase in the size of the nuclear reactor would not necessarily require an equivalent increase in the production rate of either ammonia or water. In this study, nuclear heat is used to produce a synthesis gas mixture of hydrogen plus carbon oxides, and also high pressure steam. By changing the operating conditions in the reforming tubes it should be possible to obtain a synthesis gas which is suitable for the production of methanol. Synthetic methanol is used in the production of a wide variety of chemical and plastic materials, and may also be used as an automotive fuel. Further, as the technology of fuel cells improves it may be possible to utilize hot hydrogen-rich gases for the direct, and very efficient, production of electric power.

The high pressure steam produced in the nuclear reactor can be used to drive vapor compressors, to drive ammonia or methanol synthesis compressors, or to drive turbogenerators.

The electric power which is produced can be used for the production of phosphoric acid, another of the three basic fertilizers, or for other industrial purposes.

Thus, a nuclear energy center, of the type investigated in the study of Alternate 7, may form the basis of a complete agritechnical complex.

A diagram illustrating this concept is presented in Figure 8 of Appendix A.

D. SITING

Regardless of the technical and economic evaluation of the processes which were studied, this investigation can be meaningful only if there are areas of the world which are in need of both desalted water and nitrogenous (and other) fertilizers.

A preliminary investigation indicates that there are a number of regions where the installation of a complex of this type would prove most beneficial.

The plant capacity may be fairly small, supporting the needs of a local region for potable water (plus electricity) and ammonia, or may be large enough to supply the entire basic needs of a developing agriculture in an arid coastal region. Such potential sites exist in the United States, Mexico, the Mid-East, the southern shore of the Mediterranean, Australia, the Persian Gulf, the Arabian peninsula, the desert regions of the coasts of Chile and Peru, and elsewhere. A preliminary analysis of a number of specific locations may be found in Appendix B.

E. RECOMMENDATIONS

A preliminary study of a number of different processes having been completed, the results of which indicate that it is possible to obtain a significant reduction in the cost of desalting water by combining this operation with the production of fixed nitrogen (and electricity) in a multi-product chemical complex, it would be desirable at this time to select a specific plant site and to prepare a design optimized for that location.

Among the factors to be reviewed in the selection of a site are such items as the demand for water and the cost at which its production becomes economically desirable; the need for fertilizer or electricity (or other products which can be co-produced with desalted water); the type, cost and availability of a source of energy; the quality and availability of ocean water or brackish water; and the amount and cost of capital which might be allocated to this project.

The value of desalted water will be influenced by whether it is to be used for municipal and industrial purposes or for agricultural purposes; the value of crops which may be raised at the site; and the cost of alternative sources of water, fertilizer and food.

Undoubtedly, it will prove impossible to obtain completely satisfactory data pertaining to all of these critical factors. However, as a result of an investigation of the marketing and financial aspects of the problem, it would be possible to develop guide lines within which the best technical solution can be found.

The engineering portion of a more complete study would be devoted to the preparation of a reasonably detailed mechanical design and cost estimate of the common energy center (fossil or nuclear fueled) and to the optimization of the process design .

In view of the world wide need of both desalted water and nitrogenous fertilizer, and the potential benefits of a multi-product plant as indicated by this report, it is recommended that this investigation be continued and that a detailed feasibility study be undertaken for one of the aforementioned sites.

IV. GROUND RULES AND ASSUMPTIONS

The ground rules for this study are to be similar to those established for two previous investigations (1) (3). The essentials of these ground rules, plus modifications and additional assumptions which were to be used, are summarized below.

A. RAW MATERIALS AND PRODUCTS

1. Water

Water is obtained from the sea by a long submarine pipe line. Suitable screens and intake pumps are provided. The sea water temperature is 80°F. Its total solids content is 33,600 ppm. Concentrated brine is returned to the sea.

The product water contains 25 ppm of total solids. It has been treated for pacification, but has not been chlorinated.

2. Natural Gas

Gas is available at a pressure of 550 PSIA and a temperature of 80°F.

Its composition is:

Methane	95% (by vol.)
Ethane	3% (by vol.)
Nitrogen	2% (by vol.)
Sulfur	nil

3. Ammonia

Ammonia is the usual commercial grade, having a minimum purity of 99.5% (by weight) and a maximum water content of 0.5%.

B. PROCESSES

1. General Basis

All processes are based on technology and equipment currently available.

2. Desalination

The multi-stage flash process is similar to that detailed by the Foster Wheeler Corporation in a previous study for the Office of Saline Water (3).

The vapor compression process is similar to that detailed by the Oak Ridge National Laboratory in a previous study for the Office of Saline Water (4).

3. Ammonia Production

The ammonia process is similar to those used for the design of modern, large plants based on the reforming of natural gas.

4. Electrical Generation

The power generating station is similar to modern outdoor stations designed to burn natural gas.The station is complete with transformer and switchgear.

C. COST ESTIMATE

1. Plant Limits

Plant costs are battery limits only, and do not include facilities for storing or distributing desalted water, anhydrous ammonia, or electrical power.

However, all off-site facilities normally required by process plants, such as control rooms, administration building, yard piping, internal roads, fencing, etc., are included.

The cost of land, valued at $500 per acre, is included in the cost of each plant.

2. Construction Costs

Equipment costs are based on 1968 prices. Construction costs are equivalent to those prevailing in the Gulf Coast area of the United States.

The plant site is level. There are no unusual grading or other land improvement difficulties. Piling is not required.

D. UNIT COST DERIVATION

1. Plant Life

The desalination plants, the electric power generating plant and the common energy centers are assumed to have useful lives of 30 years.

The ammonia plants are assumed to have useful lives of 15 years.

2. <u>On-Stream Time</u>

It is assumed that each plant will produce 90% of its daily rated capacity during the course of a year.

3. <u>Capital Recovery Rate</u>

A fixed charge rate of 7% is used for desalting systems and for electrical power generating plants. This rate includes both interest and plant amortization, based on a 30-year capital recovery period.

A fixed charge rate of 14% is used for ammonia plants. This rate includes interest and plant amortization, based on a 15-year capital recovery period.

A fixed charge rate of 7% (interest plus plant amortization) also is used for the common energy centers as the bulk of the heat released in these centers is used for the production of desalted water (or electricity).

4. <u>Cost Allocation</u>

a. <u>Desalted Water and Electricity</u>

Capital costs of the steam generating facilities plus the sea water intake pipe lines and pumping station equivalent to that required for condensing steam exhaust from the turbo-generator, and boiler fuel costs , are allocated between the power generating facilities and the water desalting facilities in proportion to the exergy of the steam utilized This is the Fruth Method (5). Calculations are based on the steam extraction cycle and turbine efficiency used in determining the overall steam balance.

b. Desalted Water and Ammonia

Because of the intricacy of energy exchange and energy recovery from ammonia plant process streams, meaningful calculations of investment and energy allocations derived from exergy utilization become quite complex.

Therefore, to simplify the energy relationships, in fossil fired multi-plant complexes the desalination plant is charged with all energy and capital costs which exceed those of an equivalent ammonia-only plant. For similar reasons of simplification, the costs of the nuclear reactor are allocated on the basis of the relative amounts of high temperature energy utilized by each plant.

V. DESCRIPTION OF BASIC PROCESSES

A. WATER DESALINATION

1. Multi-stage Flash

The MSF plant contains sixty-eight stages of heat recovery and four of heat rejection.

Steam, at a temperature of 260°F, is used in the brine heater. In addition, a side stream of brine is pumped from the tube side of stage 53 to the ammonia plant, where it is heated from 132°F to 143°F, and then returned to the flashing brine stream at the same stage. The plant is designed to have a performance ratio (lbs. of desalted water per 1000 Btu of heat) of 16.25 when utilizing steam at 260°F and 5.0 when utilizing heated brine at 143°F.

Electrical power requirements for sea water pumps, product pumps, etc., are 6 Kwh per 1000 gallons of desalted water.

A more complete description of the process may be found in Annex A.

2. Vapor Compression

The VC plant contains two vertical evaporation effects. Steam from the second effect is recycled to the steam heat of the first effect after being compressed from 17.5 to 22 PSIA.

A multi-stage flash plant, of the type described above, is used for feed heating.

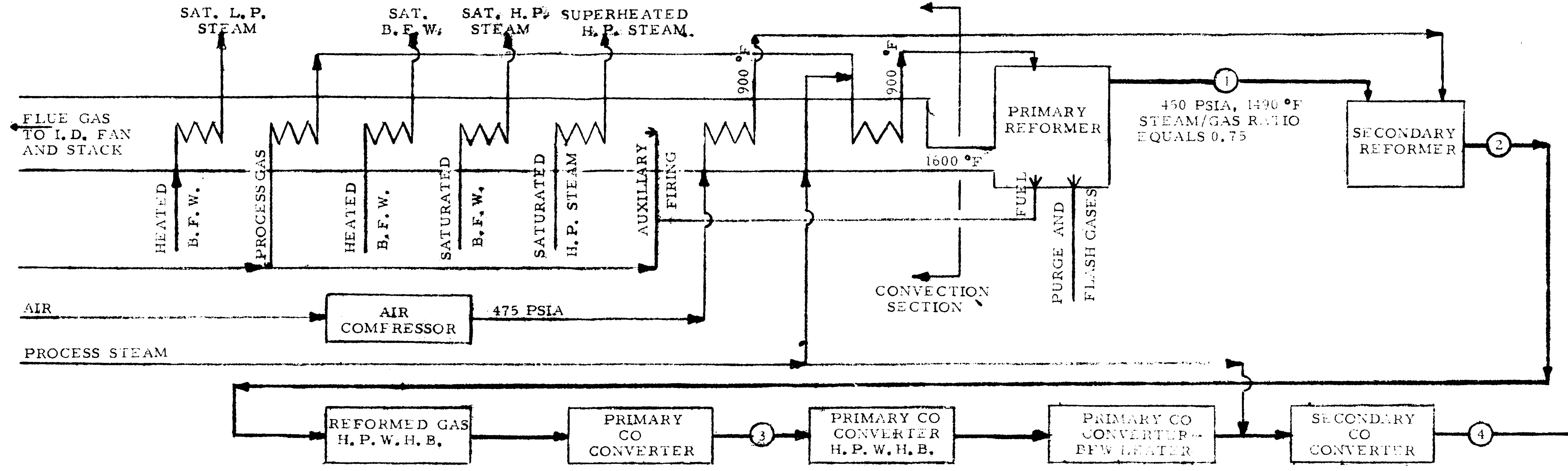
Shaft horsepower requirements for the vapor compressor are 1570 HP per MGD. An additional 10%, equal to 4.2 Kwh per 1000 gallons of desalted water, is required to drive the sea water pumps, product pumps, etc., associated with the VC plant.

A more complete description of the process may be found in Annex A.

B. AMMONIA PRODUCTION

1. Normal Design

The process flow diagram shown in Figure 6 exemplifies the design used for most modern ammonia plants, although certain aspects may be changed by engineering firms in order to optimize their design and to meet local requirements. (This flow diagram also includes two low pressure waste heat boilers, not normally used, which generate steam for the desalination plant.)



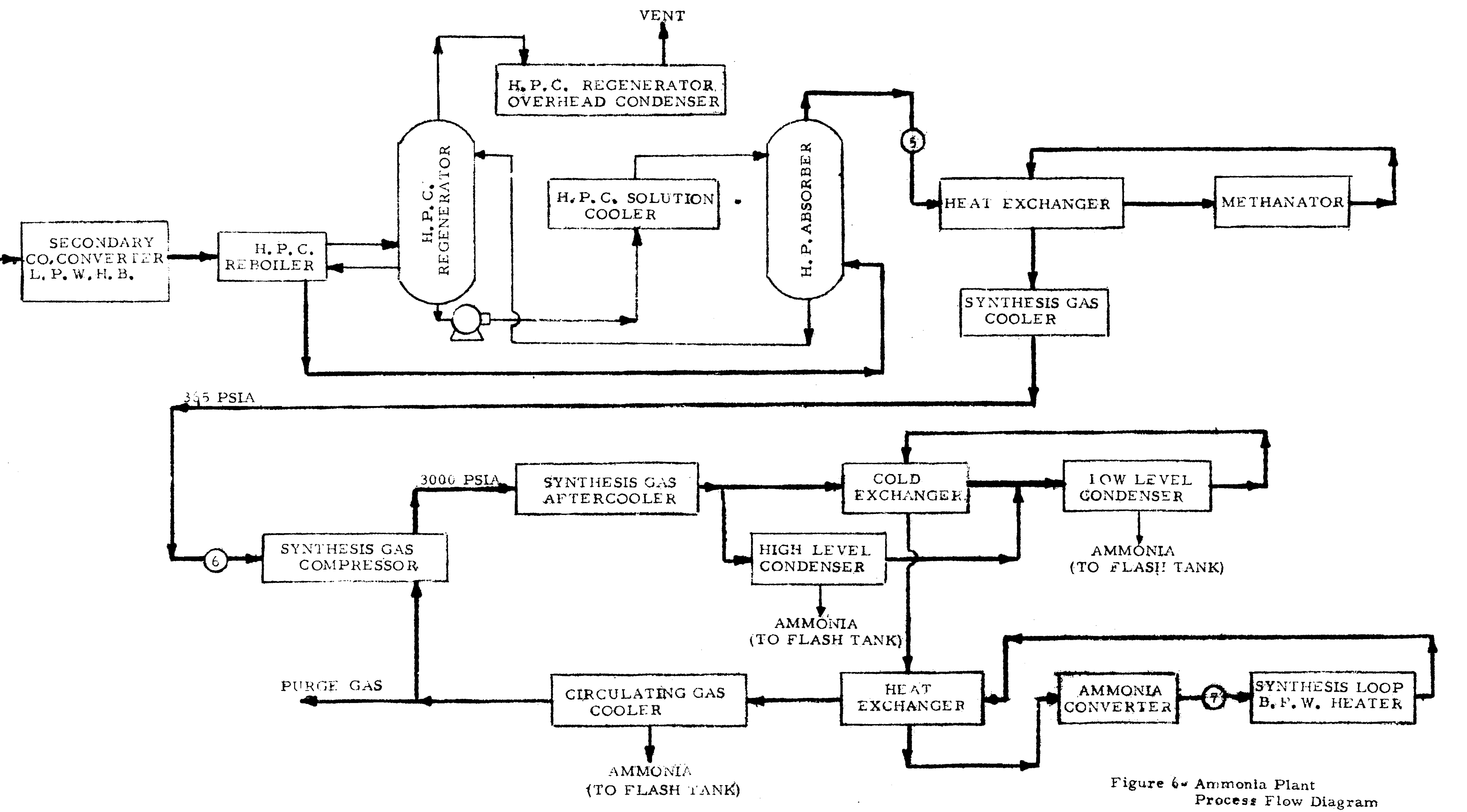


Figure 6. Ammonia Plant Process Flow Diagram

GAS ANALYSIS - MOL PER CENT							
COMPONENT	1	2	3	4	5	6	7
H_2	6 .4	56.5	60.2	61.4	74.6	74.1	55.3
N_2	0.6	22.4	20.5	19.9	24.2	24.7	13.4
CO	9.5	12.9	3.1	0.3	0.4	--	--
CO_2	10.2	7.6	15.6	17.8	0.1	--	--
CH_4	11.3	0.3	0.3	0.3	0.4	0.9	10.0
A	--	0.3	0.3	0.3	0.3	0.3	4.0
NH_3	--	--	--	--	--	--	12.3
TOTAL	100.0	100.0	100.0	100.0	100.0	100.0	100.0

The usual raw material is natural gas (or naphtha). Natural gas and steam are mixed, preheated, and passed through catalyst-filled tubes which are hung in a Primary Reforming Furnace. The energy required to bring the raw materials to the required temperature and to supply the heat of reaction, is obtained by burning additional amounts of natural gas (or other satisfactory hydrocarbon fuel). The following two reactions occur:

(1) $CH_4 + H_2O \rightleftharpoons CO + 3\ H_2$

(2) $CH_4 + 2H_2O \rightleftharpoons CO_2 + 4\ H_2$

The hot flue gas leaving the Reforming Furnace is passed over a series of coils in a Convection Section. As the flue gas is cooled, its sensible heat is used to preheat the natural gas and steam mixture, preheat compressed process air, generate and superheat high pressure steam, preheat boiler feedwater , and preheat natural gas. As the sensible heat of the flue gas is not sufficient to supply all of the required heat, additional amounts of fuel are burned in the Convection Section.

If the ammonia plant is designed in conjunction with a water desalination plant, a coil for the generation of low pressure steam is placed after the natural gas preheat coil in order to recover low temperature heat which would otherwise be vented to the atmosphere.

The process gas mixture leaving the furnace consists of carbon oxides, unconverted methane, hydrogen and steam. It is piped to a catalyst-filled Secondary Reformer. Compressed air is added, providing the nitrogen necessary for the synthesis of ammonia and reducing the methane content of the reformed gas. The reactions which take place are a continuation of the reforming steps indicated above, and the combustion of part of the hydrogen with the oxygen of the air:

$$(3)\quad 2H_2 + O_2 \rightleftharpoons 2H_2O$$

Leaving the Secondary Reformer, the reformed gas passes through a H.P. Reformed Gas Waste Heat Boiler and then to the Primary CO Converter, which contains an adiabatic catalyst bed. Here additional amounts of hydrogen are produced, in accordance with the following shift reaction:

$$(4)\quad CO + H_2O \rightleftharpoons CO_2 + H_2$$

The converted gas is further cooled in a Boiler Feedwater Heater, additional amounts of steam are introduced to improve the equilibrium, and the gas passed over another catalyst in the Secondary CO Converter. The lower temperature, and increased steam quantity, tend to force equation (4) to the right.

The CO_2 must be removed from the converted gas. An activated solution of potassium carbonate is used for this purpose (Benfield process). The hot solution is circulated between an Absorption Tower and a Regeneration Tower. The solution enters the top of the Regenerator Tower. As it flows down the tower, it is scrubbed by a counter-current stream of rising water vapor generated in a H.P.C. Reboiler. Carbon dioxide and water vapor leaving the top of the tower are cooled and condensed in an H.P.C. Overhead Condenser, and the carbon dioxide vented to the atmosphere.

The regenerated carbonate solution is withdrawn from the base of the Regenerator, cooled, and pumped back to the top of the Absorption Tower.

The process gas leaving the Secondary CO Converter is piped to a Secondary CO Converter L.P. Waste Heat Boiler, where steam is generated for use in a desalination plant. (If the ammonia plant is not being designed in conjunction with a desalination plant, heat equivalent to that exchanged in the Boiler may be used to heat boiler feedwater, as indicated in Figure 11.) Leaving the Waste Heat Boiler, the gases enter the H.P.C. Reboiler. A portion of the steam is condensed, and the total of sensible and latent heats provides the energy needed for solution regeneration. The gas mixture leaving the Reboiler is cooled, and piped to the Absorber.

As carbon oxides will poison the synthesis catalyst, all remaining traces of carbon dioxide and carbon monoxide must be removed. This is done by passing the process gas through an adiabatic bed of methanation catalyst, where reactions (1) and (2) are reversed. A suitable heat exchanger is employed to bring the converted gas to the required temperature. The gas mixture is then cooled.

The purified process gas, now designated make-up gas, is compressed to 3000 psi, mixed with a recirculating gas stream, and piped to the ammonia synthesis loop.

The compressed gas leaving the circulator is cooled with water. Leaving the Synthesis Gas Aftercooler, the stream is split. Part of the gas passes through a Cold Exchanger, and the remaining portion is chilled in a refrigerated High Level Condenser. The gas streams are combined and further refrigerated in a Low Level Condenser to condense and remove most of the ammonia which has been formed in the converter. The gas stream, which now contains a very small amount of ammonia, is piped to a Heat Exchanger and then into the Ammonia Converter. Hydrogen and nitrogen are converted into ammonia in accordance with the following reaction:

$$(5)\quad 3H_2 + N_2 \rightleftharpoons 2\ NH_3$$

The reaction is exothermic. Ammonia-rich gases leaving the converter are cooled first by being passed through a Synthesis Loop Boiler Feedwater Heater and then through the Heat Exchanger and a water cooled Circulating Gas Cooler. A small amount of the circulating gas is purged from the system, to remove the methane which had entered with the synthesis make-up gas, and the rest of the circulating gas is returned to the Compressor.

The liquid ammonia product is cooled to -20°F before it leaves the plant. The flash gas, and the purge gas, are refrigerated to condense out the ammonia in these gas streams. The non-condensed gases are piped back to the Primary Reformer, where they are used to supplement the natural gas fuel.

Three large compressors are required. One is used to compress synthesis gas, one to compress air for the Secondary Reformer, and one to compress ammonia to provide the refrigeration needed in the synthesis loop. All of these are driven by condensing steam turbines. Condensing steam turbines are also used to drive the large pumps and fans- the Boiler Feedwater Pumps, the Cooling Water Pumps, and the Induced Draft Fan of the Primary Reformer. Small pumps are driven by electric motors.

The plant is assumed to be located at a place where fresh water is available for make-up to a cooling tower and for boiler feedwater. Standard materials of construction are used for the heat exchangers i.e., 304 stainless steel for the CO_2 Regenerator Overhead Condenser, Admiralty for the steam condenser tubing, and carbon steel for the remaining water coolers.

2. Heat Recovery

a. Energy Sources

It is apparent that, rather than using condensing steam turbines, it would be possible to provide non-condensing steam turbines and exhaust low pressure steam to the brine heaters of a water desalination plant. The benefits to be realized would be equivalent to those obtained in a dual purpose plant producing electrical power and desalted water.

There are, in addition, other sources of low level heat which cannot be utilized in an ammonia plant and are normally discarded either to cooling water or to the atmosphere. A portion of this heat also can be used in an MSF desalting plant.

The amounts of heat which are recoverable at various places in the process are shown in Figure 7. The quantities shown apply to a 1000 T/D plant. The curve includes only those sources from which it is economically feasible to recover energy. Other potential sources have been omitted either because the temperature at which the energy is available is too low, as in the case of the Ammonia Refrigeration Condenser; or because the cost of energy recovery is too high, as in the case of the Synthesis Gas Aftercooler.

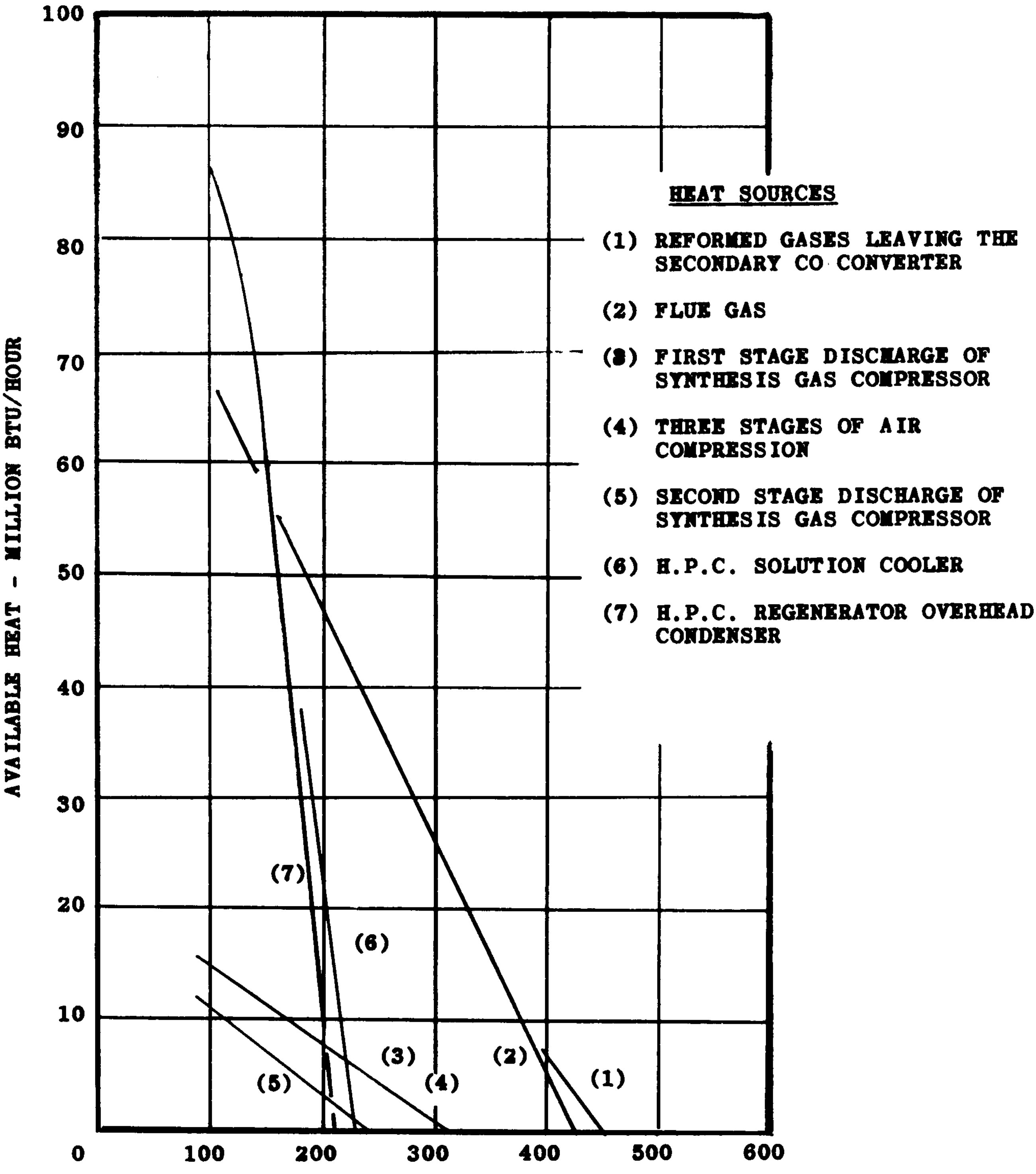


Figure 7-Available Low Temperature Heat Sources in a 1000 T/D Ammonia Plant

The total amount of such available heat is shown in Figure 8.

The desalination plant is able to utilize energy at 260°F, in the form of low pressure steam; and also at lower temperatures, in the form of heat imparted to a brine steam. However, the value of the heat becomes less as the temperature decreases.

To simplify the exchange of energy between the ammonia plant and the desalination plant, heat is assumed to be recovered at two levels only.

Studies were made of the amounts of heat which could be recovered in the form of low pressure steam or hot brine by using differing temperature approaches to process streams. The temperature approach in each exchanger was narrowed until the incremental cost of the increased exchanger surface and associated materials was equal to the incremental value of the recovered heat. In these preliminary optimization studies the cost of equipment was evaluated using a 7% annual fixed charge rate. Low pressure steam was considered to be worth 15¢ per 10^6 Btu and heat transferred to brine was considered to be worth 4.7¢ per 10^6 Btu. (It is recognized that in a more detailed optimization study the value of recovered heat would be a function of the cost of

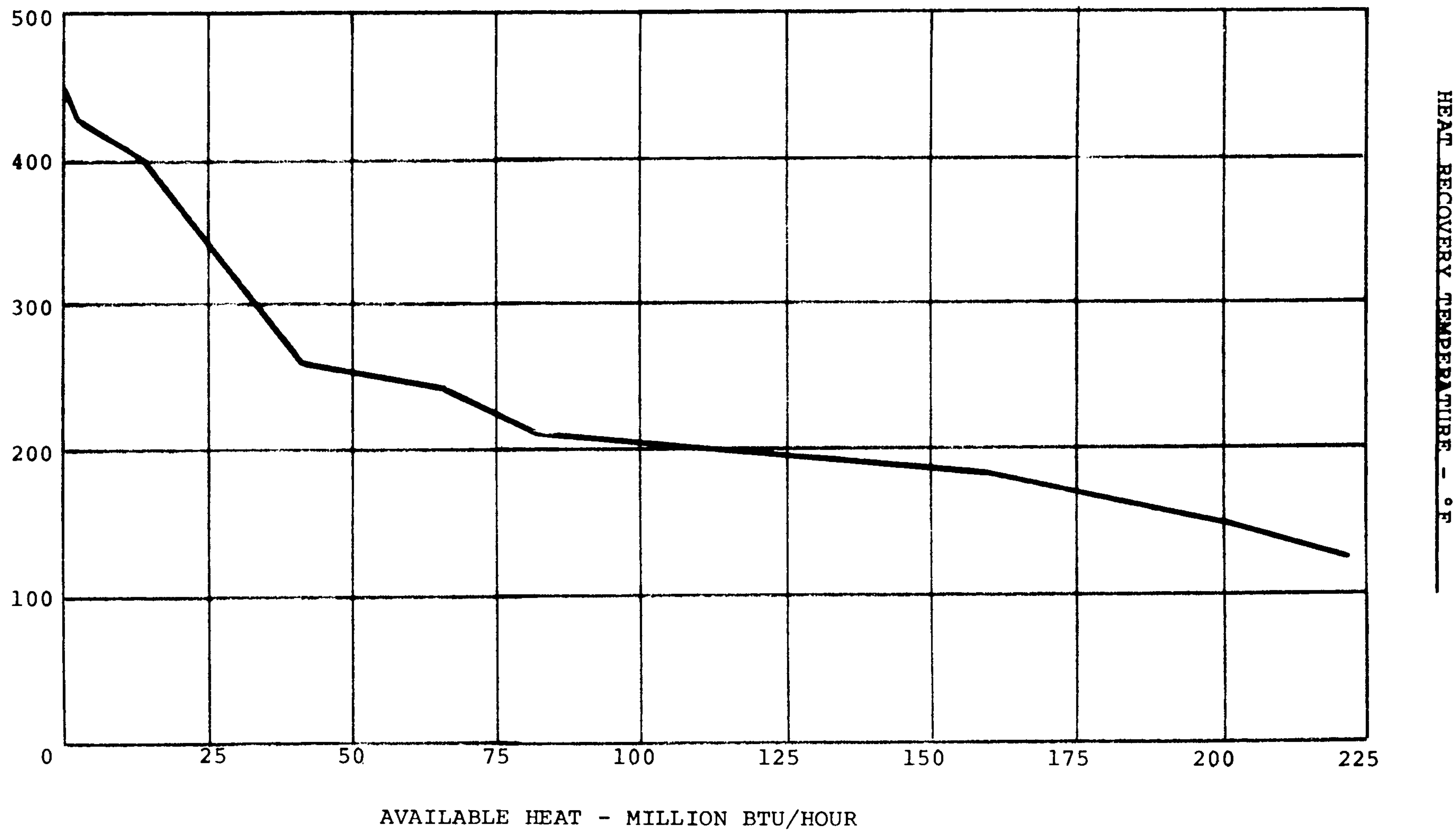


Figure 8 -Total Available Low Temperature Heat in a 1000 T/D Ammonia Plant

natural gas or other prime fuel, and the plant configuration would be changed accordingly.)

These studies indicate that, in general, heating a circulating brine stream from 132°F to 143°F is the optimum range for all of the Alternates which were investigated.

The amount of heat recovered, and the points in the process at which heat recovery was assumed, varied with the size of the plant (and thus the characteristics of the compressors and drives) and, in the more complex plant designs (Alternates 5,6 and 7), with the overall steam and boiler feedwater balance.

In general, low pressure steam boilers are added in the Convection Section of the Gas Reforming Furnace and after the Secondary CO Converter. These boilers generate steam at 37 PSIA, this steam being added to the quantity exhausted from the steam turbines. A pressure drop of 1 1/2 PSI is required to convey the low pressure steam to the desalination plant. The arrangement of the two plants is shown in Figure 9.

b. Heat Exchanger and Pump Materials

Ammonia plants designed in conjunction with desalination plants are assumed to be remote from any source of fresh water.

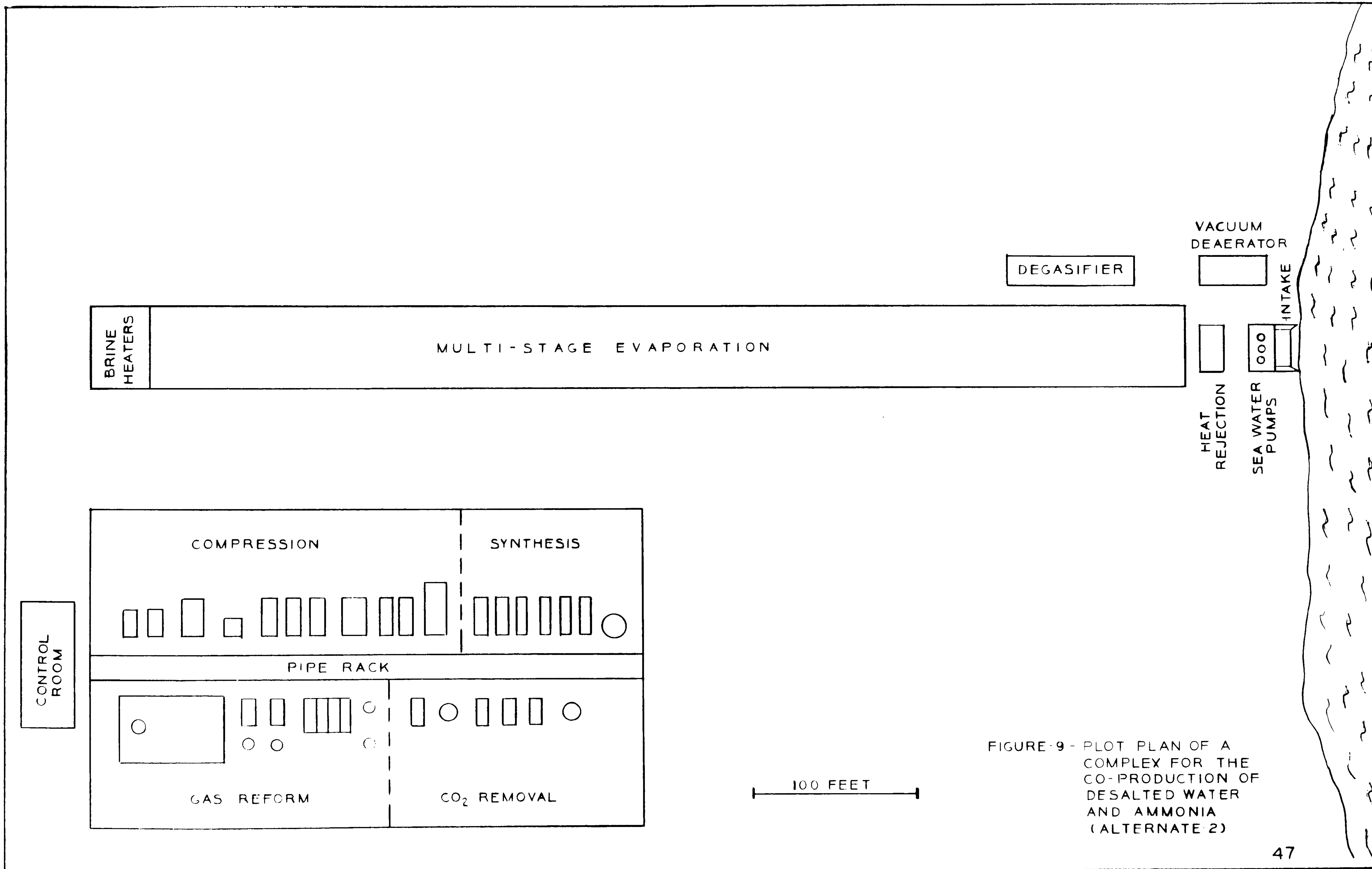


FIGURE 9 - PLOT PLAN OF A COMPLEX FOR THE CO-PRODUCTION OF DESALTED WATER AND AMMONIA (ALTERNATE 2)

Sea water is used for water cooling, and desalted water for boiler feedwater make-up.

Coolers in contact with salt water contain 90-10 cupro-nickel tubes and carbon steel shells, with the salt water on the tube side and the process fluid on the shell side.(6) When such construction cannot be used, either because the process gas pressure is too high to be placed on the shell side (Synthesis Gas Aftercooler) or because the process fluid should not be placed in contact with cupro-nickel (H.P.C. Solution Cooler, H.P.C. Regenerator Overhead Condenser), bimetallic tubes are used. These tubes contain an inner layer of 304 S.S., which is in contact with the process fluid, and an outer layer of 90-10 cupro-nickel in contact with sea water.

Brine pumps have Ni-Resist casings and Ni-Al-Bronze impellers. Specifications and costs of such pumps have been published recently (7)

3. Cost Allocation

The complexity of energy exchange in an ammonia plant makes it very difficult to utilize established methods of cost allocation, such as the Fruth method (5).

The procedure for determining the total investment to be allocated is not straightforward. In addition to the boilers in the Convection Section, each Alternate contains a number of process waste heat boilers and feedwater heaters.

The estimate of total fuel cost also is not direct, as part of the energy used for the generation of steam is derived from heats of reaction rather than from the combustion of natural gas. In addition, as noted, much of the heating of boiler feedwater is accomplished by heat exchange with process streams.

For these reasons it is believed that the purpose of this study is best served and the economic presentation more clearly defined if all changes in energy and capital charges of the ammonia plant, as contrasted with that of a typical single-purpose plant, are reflected by equivalent changes in the cost of desalted water and/or power.

Therefore, two comparable heat balances are calculated for each ammonia plant size. One is based on a typical conventional plant design, and the other is based on the design which had been modified to recover low level heat.

Compressor, pump and turbine efficiencies are corrected for each heat balance, and are representative of those obtainable in equipment available at the present time.

Natural gas and investment requirements are estimated for each case, so that the effect of design modification to accomodate water desalination may be indicated.

C. POWER GENERATION

The power generation cycle is typical of that used in modern plants.

Steam throttle conditions of 1465 PSIA, 825°F are used, retaining the same high pressure steam level for all Alternates. This pressure is very close to the optimum for a 150 MW turbo-generator. (The current IEEE-ASME recommended throttle conditions for a condensing 100 MW station, the highest for which a standard has been established (8), is 1450 PSIA, 1000 °F.) Two uncontrolled extraction stages, at 315 PSIA and 105 PSIA, are used for boiler feedwater heating. The turbine is non-condensing, steam being exhausted at 37 PSIA for use in the desalination plant. A pressure drop of 1.5 PSI is allowed in the piping from the power plant to the desalination plant.

D. COST OF UTILITIES

1. Natural Gas

Calculations are based on a natural gas cost of 25¢ per 10^6 Btu (LHV).

Industrial natural gas costs in the United States range from perhaps 17 to 20¢ in the Southeast to 35 to 40¢ in the West. Therefore 25¢ was selected as being at about the midpoint of the natural gas cost range in those regions of the coastal USA in which large-scale combined ammonia and desalination plants might be sited.

However, as the variable cost of production is primarily a function of the cost of natural gas, curves are presented showing the effect of natural gas cost on product cost.

2. Electricity

Power costs are calculated, assuming electricity to be purchased from a utility which operates a gas-fired boiler. (Except for Alternates 6 and 7.) With a capital charge of 7%, it is assumed that power can be generated and transmitted to the agritechnical complex at the following rates:

Price of Natural Gas ¢/10^6 Btu	Price of Electricity mils/Kwh
15	4.0
25	5.2
40	7.4

In Alternate 6, generated power costs are calculated to be 4.2 mils when gas is 25¢, and this value is used in calculating the cost of water production.

In Alternate 7, it is assumed that electricity , generated in a nuclear reactor, will be available at 4.5 mils/Kwh (9).

VI. ALTERNATES INVESTIGATED

A. ALTERNATE 1 - 12.5 MGD Of DESALTED WATER AND 500 T/D Of AMMONIA

1. Process

The ammonia process flow diagram is similar to that indicated on Figure 6.

The energy balance of the multi-product complex is shown on Figure 10.

The energy balance of a typical 500 T/D ammonia plant is shown on Figure 11.

The multi-product complex has been designed so that the ammonia plant can be run at substantially full capacity regardless of operation of the desalting plant. This has been done by providing condensing sections for the turbines driving the Refrigeration and Synthesis Gas compressors.

During normal operation the only steam passing through these sections is that required to keep the low pressure turbine blades from overheating. However, if some of the modules in the desalting plant are shut down (either for repair or because the water is not required), steam flow to the condensers can be increased. As these turbines contain

Figure 10 - Energy Balance Alternate 1

NOTE : ALL FLOWS IN THOUSANDS OF POUNDS PER HOUR

Figure 11- Energy Balance
Typical 500 T/D Ammonia Plant

extraction - re-entry nozzles, it is possible for the turbines to accept a significant portion of the 37 PSIA steam generated by the Boiler Feedwater, H.P.C., and Brine Cooling Pumps and by the I.D. Fan. This results in a decreased flow of high pressure steam to the turbines and, therefore, a decrease in steam generation in the Convection Section and a reduction in auxiliary firing requirements.

Any remaining 37 PSIA steam is condensed in the serviceable brine heater modules, whether or not desalted water is being produced, and the condensate returned for use as boiler feedwater.

Although the design is penalized to a small degree, both in capital cost and efficiency, by the provision of such steam turbines, the increased cost is warranted by the greater degree of flexibility which is thereby achieved.

The desalination process is described in Annex A.

2. Heat Recovery

a. Brine Circulation

As noted previously, the assumption is made that fresh water (from a cooling tower) will be used in a standard plant, whereas the dual purpose plant will use salt water. Furthermore, the required heat exchange surface will have to be increased when cooling is done by 132°F brine rather than by 80°F sea water.

Table IV lists the major salt water coolers in the ammonia plant, for Alternate 1, and the specified materials and cooling fluid.

The cost of these modifications to the plant design so that process heat can be exchanged with brine or sea water is determined by estimating the required surface and cost of each of these exchangers, deducting the cost of water coolers and condensers required in a standard plant, deducting the installed cost of the cooling tower which would otherwise be required, and providing for the following:

Added cost of erecting the additional heat exchange surface.

Added cost of material and installation for cooling water pipe and valves of cast iron (for sea water) rather than of carbon steel.

Incremental cost of salt water intake station, including submarine pipe line, and effluent system.

Added cost of intake station pumps and drives, as contrasted with cooling water pumps and drive.

Cost of pipe line (cast iron) used to circulate hot brine between the desalting water plant and the ammonia plant.

Increased cost of the brine pump because of the pressure drop in the circulating brine pipe line.

TABLE IV. MAJOR SALT WATER COOLERS
ALTERNATE 1

EQUIPMENT	COOLING MEDIUM	MATERIAL
H.P.C. Regenerator Overhead Condenser (205 to 150°F)	Hot Brine	Bimetallic
H.P.C. Regenerator Overhead Condenser (150 to 100°F)	Sea water	Bimetallic
H.P.C. Solution Cooler	Hot Brine	Bimetallic
Circulating Gas Cooler	Sea water	Bimetallic
1st Stage Synthesis Gas Intercooler (220 to 150°F)	Hot Brine	Bimetallic
1st Stage Synthesis Gas Intercooler (150 to 90°F)	Sea water	Bimetallic
2nd Stage Synthesis Gas Intercooler (220 to 150°F)	Hot Brine	Bimetallic
2nd Stage Synthesis Gas Intercooler (150 to 90°F)	Sea water	Bimetallic
Synthesis Gas Aftercooler	Sea water	Bimetallic
1st, 2nd & 3rd Stage Inter-coolers Air Compressor	Sea water	Cupro-nickel

EQUIPMENT	COOLING MEDIUM	MATERIAL
Ammonia Refrigerator Intercooler	Sea water	Bimetallic
Ammonia Refrigerator Condenser	Sea water	Bimetallic
Synthesis Gas and Refrigeration Steam Turbine Condensers	Sea water	Cupro-nickel

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Cupro-nickel tubes contain 90% copper and 10% nickel
Bimetallic tubes contain a layer of 304SS bonded to a layer of 90-10 cupro-nickel

Engineering and drafting of all of the above.

The estimated increase in plant cost to effect these changes is $225,000.

b. Other Modifications

Other changes in standard ammonia plant design are required. These consisted of:

Increase in duty of the Convection Section, to produce more high pressure steam.

Increase in size of the Boiler Feedwater pumps.

Elimination, or decrease in size, of steam condensers for the compressor and pump drives.

The estimated increase plant cost to effect these changes is $425,000.

c. Total Cost

Therefore, it costs a total of $650,000 to modify a 500 T/D plant so that it can use sea water in its cooling water system and can export useable energy, at low temperature levels, to the desalination plant. This is slightly less than 7% of the installed cost of a conventional 500 T/D ammonia plant. As noted previously, it is assumed that such additional costs would be borne by the desalination plant.

3. Cost of Water and Ammonia

Under the Ground Rules and Assumptions noted in Section IV, water is produced for 35¢ per 1000 gallons and ammonia for $22 per ton. A tabulation of investment and production costs for this Alternate are shown in Tables V and VI.

As the plant produces two distinct products, it should be possible to assign an arbitrary value to either one of the two products, with the second absorbing the difference in cost. The relation of dollar values is shown in Figure 12a. To generalize the results, the same chart has been used to illustrate the effect of locating the complex in an area in which the gas price is 15¢, or an area where the gas price is 40¢.

B. ALTERNATE 2 - 21.6 MGD OF DESALTED WATER AND 1000 T/D OF AMMONIA

1. Process

The plant design is similar to that described for Alternate 1.

The basic concept of heat recovery, utilizing both low pressure steam boilers and brine recovery, is similar to that described for the first Alternate. However, when the ammonia plant is larger it is economically feasible to recover heat from the air compressor intercoolers. Therefore, cupro-nickel intercoolers, heating brine from 132°F to 143°F,

TABLE V. INVESTMENT AND ANNUAL FIXED CHARGES

ALTERNATE 1

PRODUCT	WATER	AMMONIA
NOMINAL RATING	12.5 MGD	500 T/D
ANNUAL PRODUCTION	4,100 x 10^6 Gal.	165,000 Tons
INVESTMENT ($000)		
Basic plant cost	11,250	9,550
Heat recovery, incremental cost	650	
Common off-site facilities	(200)	
Common erection costs	(120)	
Net investment	11,580	9,550
ANNUAL FIXED CHARGES ($000)		
Capital charges	810	1,340
Maintenance materials, catalyst and chemicals	123	320
Maintenance labor	56	150
Operating labor, supervision and administration	127	450
Common maintenance and administration costs	(80)	
Total fixed charges	1,036	2,260

TABLE VI. UTILITY REQUIREMENTS AND TOTAL COST OF PRODUCTION

ALTERNATE 1

PLANT	WATER		AMMONIA
NOMINAL RATING	12.5 MGD		500 T/D
ANNUAL PRODUCTION	4,100 x10^6 Gal.		165,000 Tons
<u>NET UTILITY REQUIREMENTS/YR.</u>			
Natural gas-10^{12}Btu	1.01	(1)	5.34
Power -10^6 Kwh	24.6		6.6
Make-up B.F.W.- 10^6Gal.	16.0	(1)	77.2
<u>ANNUAL VARIABLE CHARGES ($000)</u>			
Natural gas @ $0.25/$10^6$ Btu	252		1,330
Power @ 5.2 mils/Kwh	128		34
Water @ $0.35/$10^3$ Gal.	5		27
Total variable costs	385		1,391
<u>COST OF PRODUCTION</u>			
Fixed charges ($000)	1036		2260
Variable charges ($000)	385		1391
Total cost of production $000)	1421		3651
Cost per 1000 gallons of water	34.7¢		
Cost per ton of ammonia			$22.10

(1) Fuel and boiler feedwater required for supplemental steam generation.

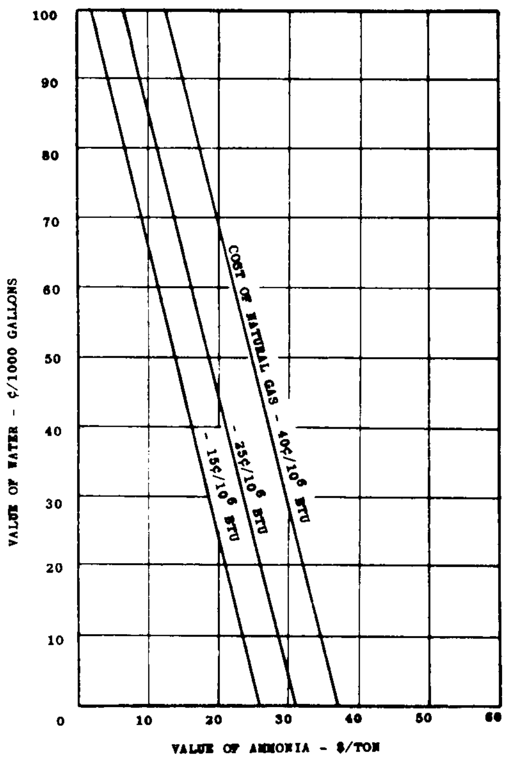


Figure 12a-Dollar Amounts Required to Recover Production Costs, Alternate 1

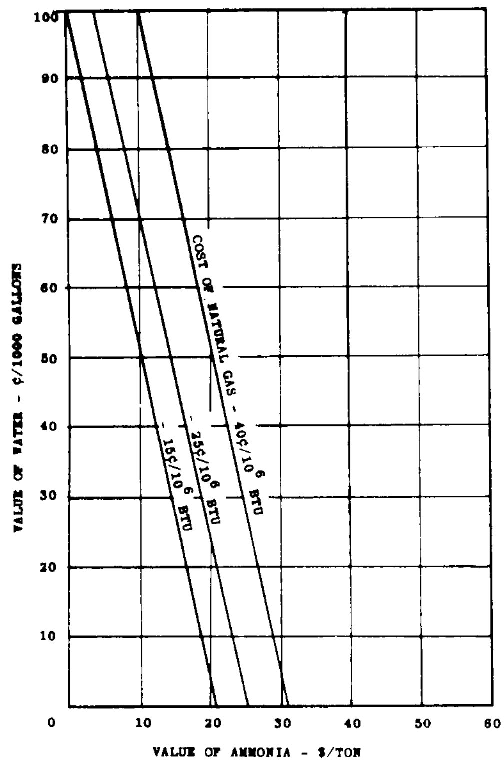


Figure 12b-Dollar Amounts Required to Recover Production Costs, Alternate 2

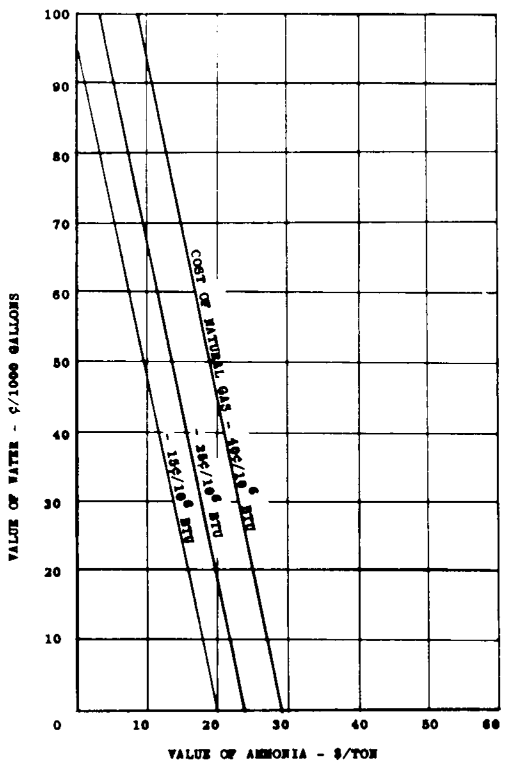


Figure 12c-Dollar Amounts Required to Recover Production Costs, Alternate 3

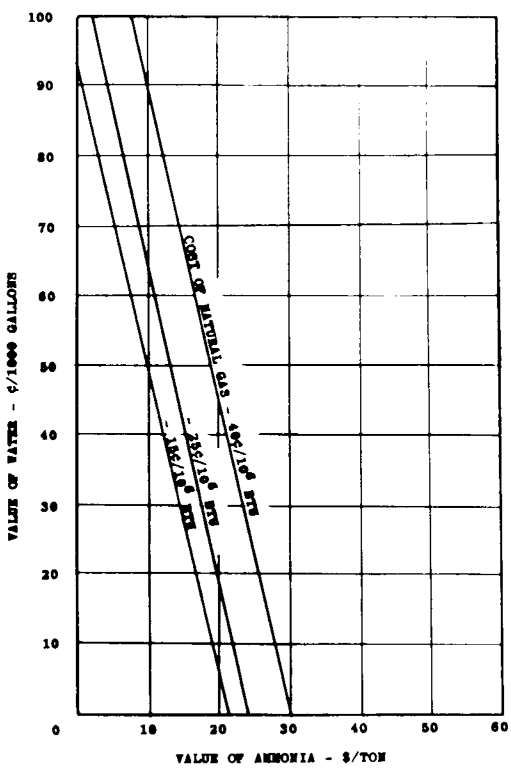


Figure 12d-Dollar Amounts Required to Recover Production Costs, Alternate 4 are used to reduce the compressed air temperature to 150°F. The remaining cooling to 100°F is done using cupro-nickel coolers and 80°F sea water.

At the increased plant size, all compressors and pumps are more efficient. Therefore the quantity of low level heat available for desalination does not increase in direct proportion to the larger ammonia plant capacity.

The energy balances of the multi-product complex is shown in Figure 13. A plot plan of the complex is shown in Figure 9.

The energy balance of a typical 1000 T/D ammonia plant is shown in Figure 14.

2. Cost of Water and Ammonia

Tables VII and VIII show total investment and production costs with the price of natural gas at 25¢ per 10^6 Btu. Under these conditions water can be produced for 32¢/Kgal and ammonia for $18.50/ton.

The effect of utilizing a different allocation of production costs, or of a different natural gas price, is demonstrated in Figure 12b.

C. ALTERNATE 3 30.8 MGD OF DESALTED WATER AND 1500 T/D OF AMMONIA

The process is similar to that described for Alternate 2. Again, an increase in efficiency of mechanical drives in the ammonia plant results in a somewhat lower amount of low-temperature heat available for utilization in the MSF plant.

Figure 13 - Energy Balance
Alternate 2

Figure 14- Energy Balance
Typical 1000T/D Ammonia Plant

NOTE: ALL FLOWS IN THOUSANDS OF POUNDS PER HOUR

TABLE VII. INVESTMENT AND ANNUAL FIXED CHARGES

ALTERNATE 2

PRODUCT	WATER	AMMONIA
NOMINAL RATING	21.6 MGD	1000 T/D
ANNUAL PRODUCTION	$7,100 \times 10^6$ Gal.	370,000 Tons
INVESTMENT ($000)		
Basic plant cost	17,580	16,200
Heat recovery, incremental cost	1,000	
Common off-site facilities	(400)	
Common erection costs	(200)	
Net investment	17,980	16,200
ANNUAL FIXED CHARGES ($000)		
Capital charges	1,255	2,260
Maintenance materials, catalyst and chemicals	203	590
Maintenance labor	87	150
Operating labor, supervision and Administration	165	450
Common maintenance and administration Costs	(80)	
Total fixed charges	1,630	3,450

TABLE VIII. UTILITY REQUIREMENTS AND TOTAL COST OF PRODUCTION

ALTERNATE 2

PLANT	WATER		AMMONIA
NOMINAL RATING	21.6 MGD		1000 T/D
ANNUAL PRODUCTION	7,100 x 10^6Gal.		330,000 Tons
NET UTILITY REQUIREMENTS/YR.			
Natural gas-10^{12}Btu	1.62	(1)	10.28
Power -10^6 Kwh	42.6		13.2
Make-up B.F.W.-10^6 Gal.	20	(1)	168
ANNUAL FIXED CHARGES ($000)			
Natural gas @ $0.25/$10^6$Btu	405		2,560
Power @ 5.2 mils/Kwh	222		69
Water @ $0.32/$10^3$ Gal.	6		54
Total variable costs	633		2,683
COST OF PRODUCTION			
Fixed charges ($000)	1,630		3,450
Variable charges ($000)	633		2,683
Total cost of production ($000)	**2,263**		**6,133**
Cost per 1000 gallons of water	31.9¢		
Cost per ton of ammonia			$18.60

(1) Fuel and boiler feedwater required for supplemental steam generation.

The energy balance of the multi-product complex is shown in Figure 15.

The energy balance of a 1500 T/D ammonia plant is shown in Figure 16.

Tables IX and X show total investment and production costs with the price of natural gas at 25¢ per 10^6 Btu. Under these conditions water can be produced for 31¢/Kgal and ammonia for $17/ton.

Figure 12c indicates the effect of differing allocations of production costs and of natural gas price.

D. ALTERNATE 4 - 43.2 MGD OF DESALTED WATER AND 2000 T/D OF AMMONIA

Although modern fabricating techniques may permit the building of equipment which can be used to produce 2000 T/D, such plants have not yet been built. Therefore, a more conservative approach of utilizing two 1000 T/D ammonia trains was adopted.

The energy balance is the same as that indicated for the 1000 T/D plant (Figure 13) except that all quantities must be doubled.

The required investment and cost of production is shown in Tables XI and XII, and the effect of varying cost allocations and gas price is shown in Figure 12d.

With natural gas valued at 25¢ per 10^6 Btu, water can be produced at 30 1/2¢/Kgal and ammonia at $17.50/ton.

NOTE: ALL FLOWS IN THOUSANDS OF POUNDS PER HOUR

Figure 15 - Energy Balance Alternate 3

NOTE: ALL FLOWS IN THOUSANDS OF POUNDS PER HOUR

Figure 16- Energy Balance
Typical 1500 T/D Ammonia Plant

TABLE IX. INVESTMENT AND ANNUAL FIXED CHARGES

ALTERNATE 3

PRODUCT	WATER	AMMONIA
NOMINAL RATING	30.8	1500 T/D
ANNUAL PRODUCTION	10,400 x 10^6 Gal.	490,000 Tons
INVESTMENT ($000)		
Basic plant cost	25,000	21,700
Heat recovery, incremental cost	1,600	
Common off-site facilities	(450)	
Common erection costs	(300)	
Net investment	25,850	21,700
ANNUAL FIXED CHARGES ($000)		
Capital charges	1,810	3,040
Maintenance materials, catalyst and chemicals	289	830
Maintenance labor	125	150
Operating labor, supervision and Administration	198	450
Common maintenance and administration costs	(80)	
Total fixed charges	2,342	4,470

TABLE X. UTILITY REQUIREMENTS AND TOTAL COST OF PRODUCTION

ALTERNATE 3

PLANT	WATER		AMMONIA
NOMINAL RATING	30.8		1500 T/D
ANNUAL PRODUCTION	$10,400 \times 10^6$ Gal.		490,000 Tons
<u>NET UTILITY REQUIREMENTS/YR.</u>			
Natural gas - 10^{12} Btu	2.25	(1)	15.0
Power - 10^6 Kwh	61.2		19.6
Make-up B.F.W.- 10^6 Gal.	21	(1)	268
<u>ANNUAL VARIABLE CHARGES ($000)</u>			
Natural gas @ $0.25/10 Btu	565		3,750
Power @ 5.2 mils/Kwh	318		102
Water @ $0.31 /$10^3$ Gal.	7		83
Total variable costs	890		3,935
<u>COST OF PRODUCTION</u>			
Fixed charges ($000)	2,342		4,470
Variable charges ($000)	890		3,935
Total cost of production ($000)	3,232		8,405
Cost per 1000 gallons of water	31.1¢		
Cost per ton of ammonia			$17.15

(1) Fuel and boiler feedwater required for supplemental steam generation.

TABLE XI. INVESTMENT AND ANNUAL FIXED CHARGES

ALTERNATE 4

PRODUCT	WATER	AMMONIA
NOMINAL RATING	43.2 MGD	2000 T/D
ANNUAL PRODUCTION	14,200 x 10^6 Gal.	660,000 Tons
INVESTMENT ($000)		
Basic plant cost	32,270	31,000
Heat recovery, incremental cost	1,900	
Common off-site facilities	(400)	
Common erection costs	(200)	
Net investment	33,570	31,000
ANNUAL FIXED CHARGES ($000)		
Capital charges	2,360	4,340
Maintenance materials, catalyst and chemicals	392	1,150
Maintenance labor	161	220
Operating labor, supervision and administration	235	520
Common maintenance and administration costs	(110)	
Total fixed charges	3,038	6,230

TABLE XII. UTILITY REQUIREMENTS AND TOTAL COST OF PRODUCTION

ALTERNATE 4

PLANT	WATER		AMMONIA
NOMINAL RATING	43.2 MGD		2,000 T/D
ANNUAL PRODUCTION	14,200 x 10^6 Gal.		660,000 Tons
NET UTILITY REQUIREMENTS/YR.			
Natural gas - 10^{12} Btu	3.24	(1)	20.56
Power - 10^6 Kwh	85		26.4
Make-up B.F.W. - 10^6 Gal.	40	(1)	336
ANNUAL VARIABLE CHARGES ($000)			
Natural gas @ $0.25/$10^6$ Btu	810		5,120
Power @ 5.2 mils/Kwh	444		138
Water @ $0.30/$10^3$ Gal.	12		100
Total variable costs	1,266		5,358
COST OF PRODUCTION			
Fixed charges ($000)	3,038		6,230
Variable charges ($000)	1,266		5,358
Total cost of production ($000)	4,304		11,588
Cost per 1000 gallons of water	30.4¢		
Cost per ton of ammonia			$17.50

(1) Fuel and boiler feedwater required for supplemental steam generation.

E. ALTERNATE 4A - 100 MGD OF WATER AND 2000 T/D OF AMMONIA

1. Process

The ratio of water to ammonia, as derived in the first four alternates, can be increased by adding a VC (vapor compression) cycle to the plant complex.

The simplest method of incorporating a VC cycle is to have it driven by an entirely independent power source, such as an electric motor or a gas turbine. A gas turbine is used in these studies. Generally a gas turbine is the more desirable drive when fixed charge rates are relatively low.

In this design studied in this Alternate, the exhaust gases from a turbine pass through a fired waste heat boiler and then discharge into the atmosphere at 300°F. The steam which is generated passes through non-condensing helper steam turbines, and exhausts to the 37 PSIA header.The steam turbines drive electric generators which produce enough power to meet all of the pumping requirements for water produced in the VC system. However, power for the pumping needs of the MSF portion of the complex is purchased.

A more complete description of the VC process may be found in Annex A.

Excluding the VC portion of the complex, which is completely independent, the energy balance is the same as that indicated for Alternate 2 (Figure 13), except that all quantities must be doubled.

2. Cost of Water and Ammonia

Tables XIII and XIV show total investment and production costs.

With gas at 25¢ per 10^6 Btu, water is produced for 31.50¢/Kgal. The ammonia price remains $17.50/ton.

Comparing production costs with those derived in Alternate 4, it appears that the addition of the VC cycle has slightly increased the cost of water (Figure 3), even though the size of the water producing facilities has increased by a factor somewhat greater than two Annual fixed charges are almost equal for the two plant designs. The process utilized in Alternate 4 has a somewhat higher overall thermal efficiency, while the process utilized in Alternate 4A permits more water to be produced without requiring a proportionate increase in the size of the ammonia plant.

The effect of utilizing a different allocation of production costs, or of a different natural gas price, is shown in Figure 17a.

F. ALTERNATE 4B - 100 MGD OF WATER AND 2000 T/D OF AMMONIA

1. Process

The thermal efficiency of the complex is improved by utilizing the gas turbine exhaust gases as combustion air in the Primary Reforming Furnace.

The turbine exhaust gases contain 16.1% oxygen by volume, wet basis (18.3% by weight). Used in the ammonia plant, they reduce its fuel requirements by the amount of energy otherwise needed to heat combustion air from its ambient temperature to the stack temperature. The construction of ducting to convey 300°F gases to the Primary Reforming Furnace burners is fairly simple - particularly at the relatively low temperature involved. Similar ducting is used in operating plants to preheat combustion air in a Convection Section, and to utilize hot gas turbine exhaust gases.

Thus, the process studied in this Alternate is the same as that in 4A, except that instead of venting all of the gas turbine exhaust gases to the atmosphere, 41,000 Mols/hr (approximately 25%) are used to supply the oxygen needed by the Primary Reforming Furnace burners and another 22,000 Mols/hr (approximately 14%) used for auxiliary firing in the Convection Section.

The two plants can function independently of each other. If it should become necessary to shut down a gas turbine, or reduce its output, the deficiency of heat in the Reforming Section can be made up by auxiliary firing. Similarly, if it becomes necessary to shut down the ammonia plant or reduce its output, additional amounts of turbine exhaust gas can be vented to the atmosphere.

TABLE XIII. INVESTMENT AND ANNUAL FIXED CHARGES

ALTERNATE 4A

PRODUCT	WATER	AMMONIA
NOMINAL RATING	100 MGD	2000 T/D
ANNUAL PRODUCTION	$33,000 \times 10^6$ Gal.	660,000 Tons
INVESTMENT ($000)		
Basic plant cost	70,070	31,000
Heat recovery, incremental cost	1,900	
Common off-site facilities	(400)	
Common erection costs	(200)	
Net investment	71,370	31,000
ANNUAL FIXED CHARGES ($000)		
Capital charges	4,980	4,340
Maintenance materials, catalyst and chemicals	1,234	1,150
Maintenance labor	351	220
Operating labor, supervision and administration	507	520
Common maintenance and administration costs	(80)	
Total fixed charges	6,992	6,230

TABLE XIV. UTILITY REQUIREMENTS AND TOTAL COST OF PRODUCTION

ALTERNATE 4A

PLANT	WATER		AMMONIA
NOMINAL RATING	100 MGD		2000 T/D
ANNUAL PRODUCTION	33,000 x 10^6 Gal.		660,000 Tons
NET UTILITY REQUIREMENTS/YR.			
Natural gas - 10^{12} Btu	10.62	(1)	20.56
Power - 10^6 Kwh	84		26.4
Make-up B.F.W. - 10^6 Gal.	40	(1)	336
ANNUAL VARIABLE CHARGES			
Natural gas @ $0.25/$10^6$ Btu	2,910		5,120
Power @ 5.2 mils/Kwh	437		138
Water @ $0.31/$10^3$ Gal.	12		107
Total variable costs	3,359		5,365
COST OF PRODUCTION			
Fixed charges ($000)	6,992		6,230
Variable charges ($000)	3,359		5,365
Total cost of production ($000)	10,351		11,595
Cost per 1000 gallons of water	31.4¢		
Cost per ton of ammonia			$17.50

(1) Fuel (3.24 x 10^{12} Btu) and boiler feedwater required for supplemental firing in the ammonia plant, plus fuel (7.38 x 10^{12} Btu) for the gas turbines.

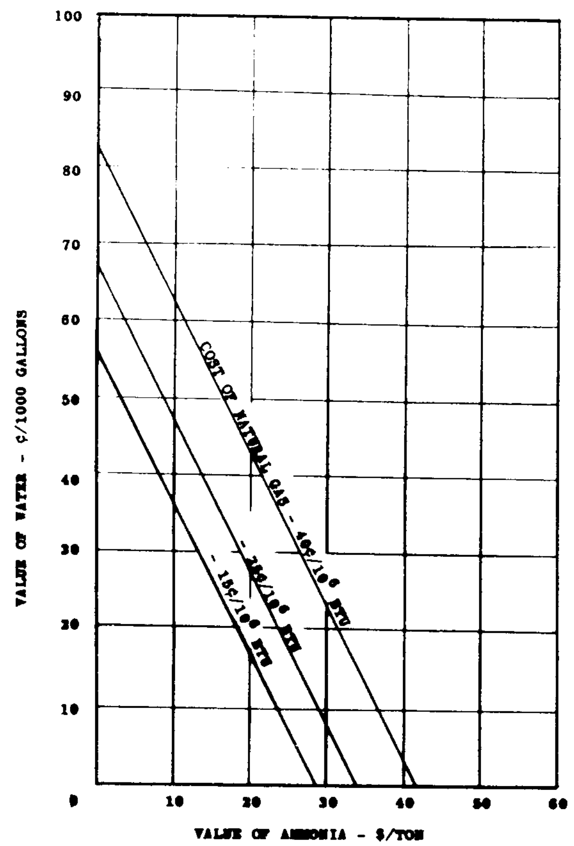


Figure 17a-Dollar Amounts Required to Recover Production Costs, Alternate 4A

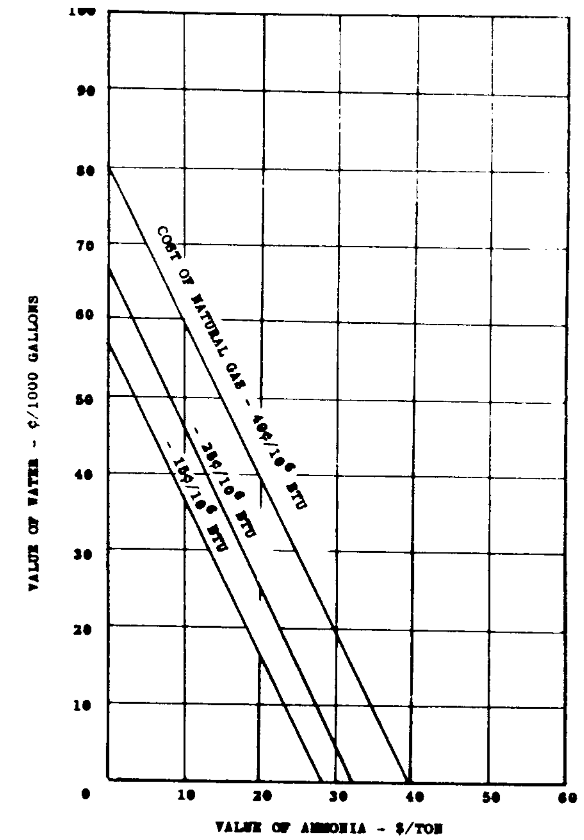


Figure 17b-Dollar Amounts Required to Recover Production Costs, Alternate 4B

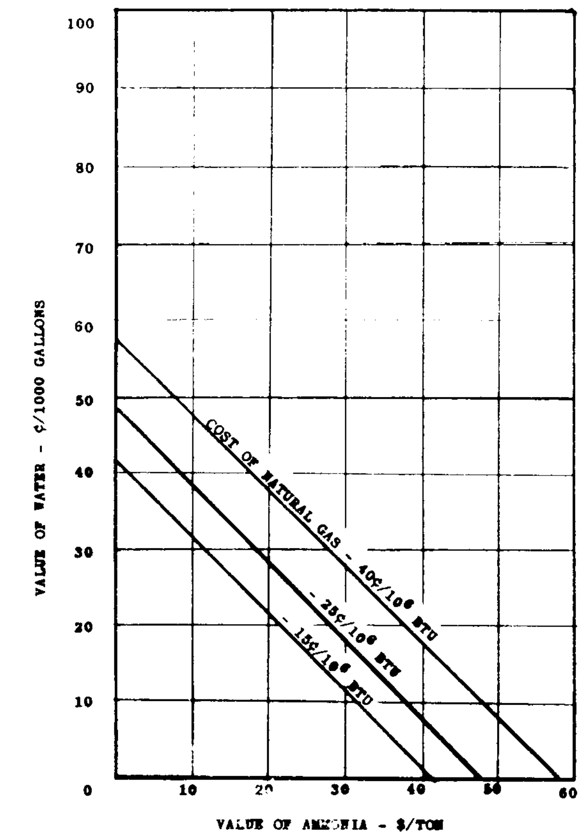


Figure 17c-Dollar Amounts Required to Recover Production Costs, Alternate 5

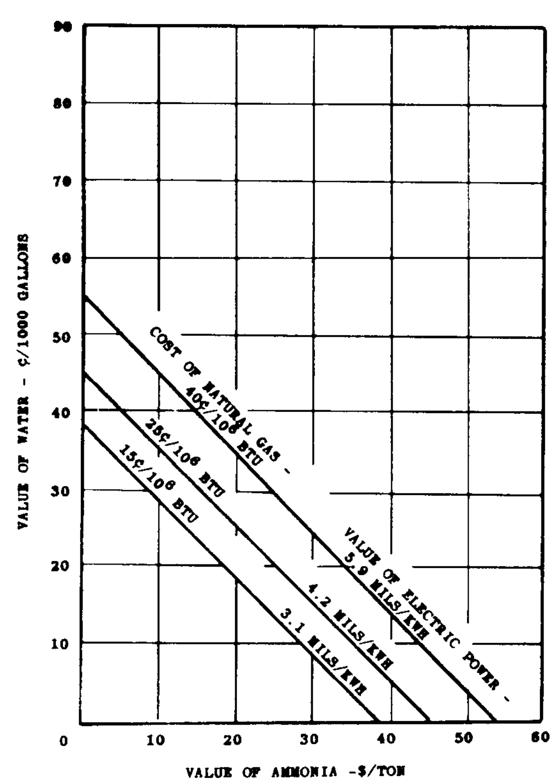


Figure 17d-Dollar Amounts Required to Recover Production Costs, The Value of Generated Power Being Maintained Constant, Alternate 6

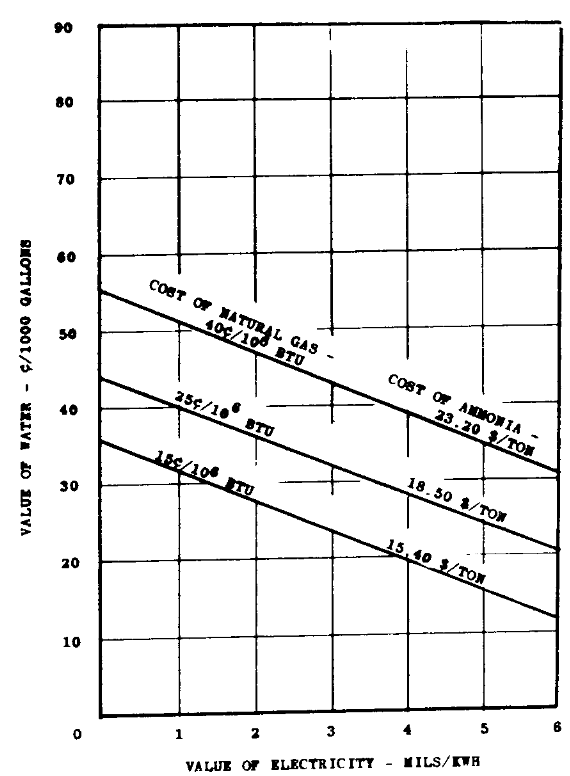


Figure 17e-Dollar Amounts Required to Recover Production Costs, The Value of Ammonia Being Maintained Constant, Alternate 6

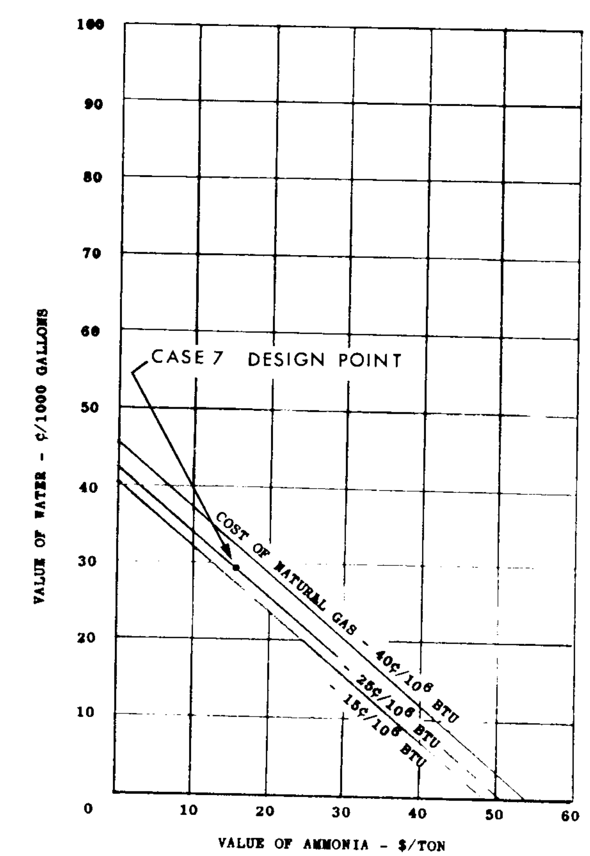


Figure 17f-Dollar Amounts Required to Recover Production Costs, Alternate 7

A portion of the energy balance, illustrating the utilization of the turbine exhaust gases and the recovery of heat in the ammonia plant, is shown in Figure 18. Steam utilization is the same as in Alternate 4A.

2. Cost of Water and Ammonia

Investment and production costs are shown in Tables XV and XVI.

Compared to Alternate 4A, this process requires a small increase in investment. However, the cost of production of water is reduced to 30.50 ¢/Kgal., the same amount as in Alternate 4. Ammonia production costs remain at $ 17.50/ton.

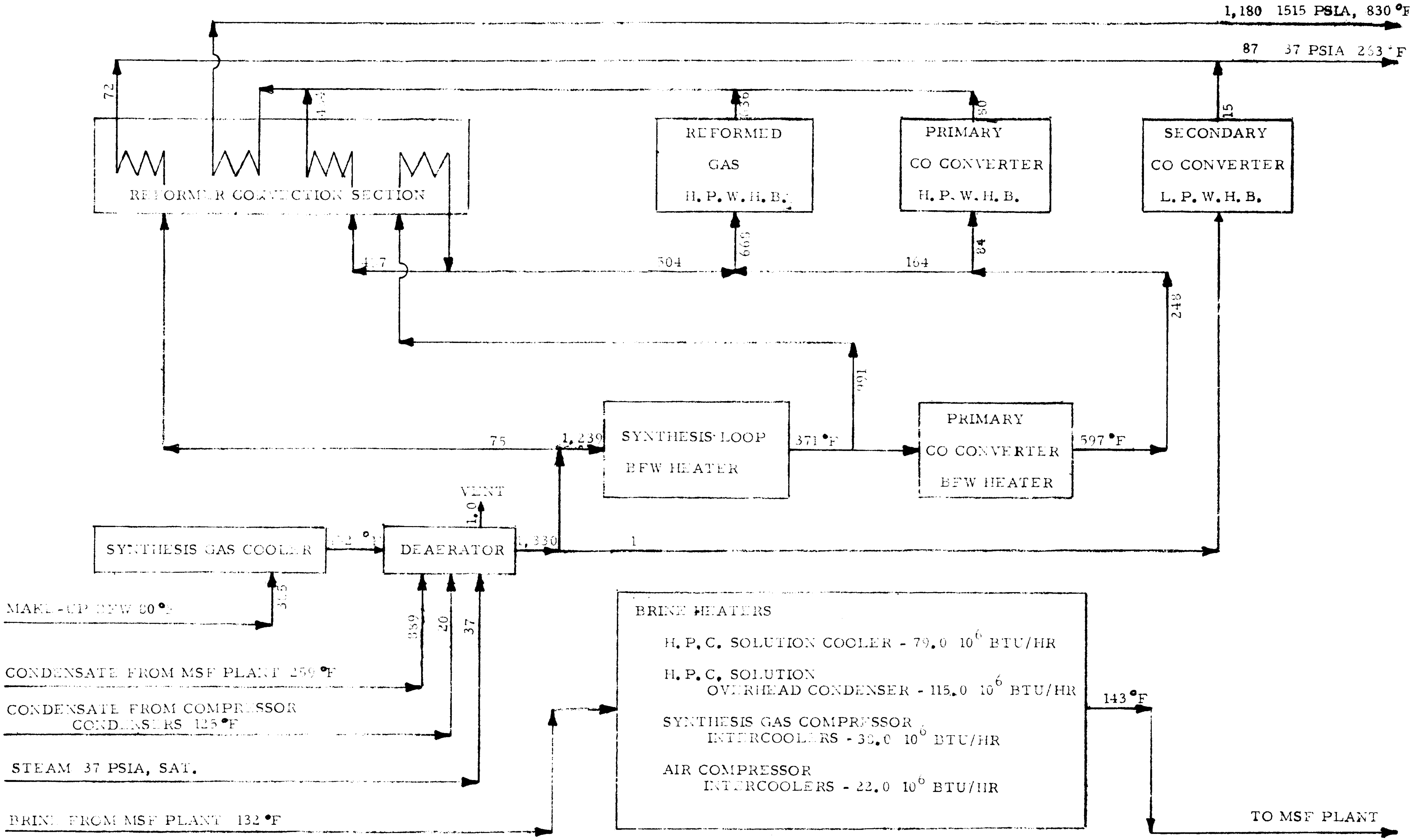
The effect of utilizing a different allocation of production costs, or of a different natural gas price, is demonstrated in Figure 17b.

G. ALTERNATE 5 - 100 MGD OF DESALTED WATER AND 1000 T/D OF AMMONIA

1. Process

a. Common Energy Center

In this Alternate an investigation is undertaken of combining, in a single energy center, facilities for producing high pressure steam to drive vapor compressor and process compressor turbines and also for providing high temperature heat required in the Primary Reforming Furnace and the Convection Section of the ammonia plant.



NOTE: ALL FLOWS IN THOUSANDS OF POUNDS PER HOUR

Figure 18- Recovery of Heat
Alternate 4B

TABLE XV. INVESTMENT AND ANNUAL FIXED CHARGES

ALTERNATE 4B

PRODUCT	WATER	AMMONIA
NOMINAL RATING	100 MGD	2000 T/D
ANNUAL PRODUCTION	33,000 x 10^6 Gal.	660,000 Tons
INVESTMENT ($000)		
Basic plant cost	70,130	31,000
Heat recovery, incremental cost	2,100	
Common off-site facilities	(400)	
Common erection costs	(200)	
Net investment	71,630	31,000
ANNUAL FIXED CHARGES ($000)		
Capital charges	5,010	4,340
Maintenance materials, catalyst and chemicals	1,232	1,150
Maintenance labor	350	220
Operating labor, supervision and administration	507	520
Common maintenance and administration costs	(80)	
Total fixed charges	7,019	6,230

TABLE XVI. UTILITY REQUIREMENTS AND TOTAL COST OF PRODUCTION

ALTERNATE 4B

PLANT	WATER		AMMONIA
NOMINAL RATING	100 MGD		2000 T/D
ANNUAL PRODUCTION	33,000 x 10^6 Gal.		660,000 Tons
NET UTILITY REQUIREMENTS/YR.			
Natural gas - 10^{12} Btu	9.96	(1)	20.56
Power - 10^6 Kwh	85		26.4
Make-up B.F.W. - 10^6 Gal.	40	(1)	336
ANNUAL VARIABLE CHARGES			
Natural gas @ $0.25/$10^6$ Btu	2,490		5,120
Power @ 5.2 mils/Kwh	444		138
Water @ $0.30/$10^3$ Gal.	12		103
Total variable costs	2,946		5,361
COST OF PRODUCTION			
Fixed charges ($000)	7,019		6,230
Variable charges ($000)	2,946		5,361
Total cost of production ($000)	9,965		11,591
Cost per 1000 gallons of water	30.4¢		
Cost per ton of ammonia			$17.50

(1) Fuel (2.64 x 10^{12} Btu) and boiler feedwater required for supplemental firing in the ammonia plant, plus fuel (7.32 x 10^{12} Btu) for the gas turbines

A number of possible configurations of the energy center were reviewed.

The ammonia plant Convection Section is readily combined with that of a large boiler. Heat exchange surfaces for boiler feedwater heating, steam generation, and steam superheating are additive. All ammonia process heat exchanger coils, including those for preheating natural gas, heating the mixed steam and gas, and heating secondary air, are easily incorporated. Suitable automatically operated dampers, and the necessary independent instrumentation, are provided so that it is possible to maintain the ammonia plant process temperatures as needed regardless of variations in throughput of the ammonia plant or the vapor compression section of the water desalination plant. Turn-down ratios of 50% are attainable.

However, a detailed study of specific boiler designs would have to be made before the economic feasibility of installing the Primary Reformer tubes in the radiant section of the energy center boiler can be expressed with any degree of certitude. Factors which would have to be evaluated would be the allowable rate of heat release in a boiler furnace which contains reforming tubes as contrasted to that in a standard boiler, the cost of any change in standard furnace wall construction, and the additional instrumentation and controls necessary to prevent overheating of the reforming tubes.

Therefore, as finally conceived, the energy center consists of a conventional boilér hàving a firing chamber with water cooled walls placed at ground level, and an elevated structure containing all of the heat exchange tubes including those required for the ammonia plant. The Reforming Furnace, with its radiant walls, is placed immediately adjacent to the firing chamber. Hot combustion gases from the reforming furnace are ducted into the boiler furnace and are mixed with the gas being burned there. A uniform gas temperature is obtained, and the mixture of combustion gases passes over the heat exchangers.

This configuration results in a lower total required investment because of the somewhat lower cost of incremental heat exchange surface; the reduction in the number of boiler feedwater pumps and I.D. fans; the reduction in the cost of supporting steel, foundations, instruments and controls; and the elimination of the ammonia plant Convection Section casing.

b. Water Desalination

The desalination process is similar to that utilized in Alternate 4A and 4B, except that the vapor compressors are driven by steam turbines rather than gas turbines. The steam turbine exhaust is used in the brine heaters of the MSF plant.

A total of 95,500 H.P. is generated by steam turbines. Vapor compressors absorb 65,500 H.P., producing 41.8 MGD of water. The remaining power is absorbed by generators which produce all of the electricity needed by the complex. This includes the power to drive all of the pumps in the desalination plants (both VC and MSF), including the supply of 80°F sea water to the ammonia plant coolers, and to drive the H.P.C. pumps.

Intermediate pressure steam, produced in the ammonia plant, is used to drive the boiler feedwater pumps and the boiler fan.

c. <u>Energy Balance</u>

The energy balance at the multi-product complex is shown in Figure 19. (It is understood that process gas, steam and air preheat coils are in the boiler convection section. They are not shown in the drawing, as none of the process streams are included.)

2. <u>Cost of Water and Ammonia</u>

Investment and production costs are shown in Tables XVII and XVIII.

The cost of producing water is 29¢/Kgal., lower than the cost calculated in Alternate 4A or 4B. The cost of producing ammonia is $17.50/ton as in Alternate 2.

The effect of changing the method of allocating production costs, or varying the price of natural gas, is shown in Figure 17c.

NOTE: ALL FLOWS IN THOUSANDS OF POUNDS PER HOUR

Figure 19- Energy Balance
Alternate 5

TABLE XVII. INVESTMENT AND ANNUAL FIXED CHARGES

ALTERNATE 5

PRODUCT	WATER	AMMONIA
NOMINAL RATING	100 MGD	1000 T/D
ANNUAL PRODUCTION	33,000 x 10^6 Btu	330,000 Tons
<u>INVESTMENT ($000)</u>		
Basic plant cost	71,280	16,200
Conversion of vapor compressors and generators to steam drive	(4,350)	
Energy center, allocated costs	4,000	inc.
Heat recovery, incremental cost	1,000	
Common off-site facilities	(400)	
Common erection costs	(200)	
Net investment	71,330	16,200
<u>ANNUAL FIXED CHARGES ($000)</u>		
Capital charges	4,993	2,260
Maintenance materials, catalyst and chemicals	1,147	590
Maintenance labor	356	150
Operating labor, supervision and administration	507	450
Common maintenance and administration costs	(80)	
Total fixed charges	6,923	3,450

TABLE XVIII. UTILITY REQUIREMENTS AND TOTAL COST OF PRODUCTION

ALTERNATE 5

PLANT	WATER		AMMONIA
NOMINAL RATING	100 MGD		1000 T/D
ANNUAL PRODUCTION	33,000 x 10^6 Gal.		330,000
NET UTILITY REQUIREMENTS/YR.			
Natural gas - 10^{12} Btu	10.64		10.28
Power - 10^6 Kwh	-	(1)	13.2
Make-up B.F.W.- 10^6 Gal.	78	(1)	168
ANNUAL VARIABLE CHARGES ($000)			
Natural gas @ $0.25/10 Btu	2,660		2,560
Power @ 5.2 mils/Kwh	-		69
Water @ $0.29/$10^3$ Gal.	22		49
Total variable costs	2,682		2,678
COST OF PRODUCTION			
Fixed charges ($000)	6,923		3,450
Variable charges ($000)	2,682		2,678
Total cost of production ($000)	9,605		6,128
Cost per 1000 gallons of water	29.2¢		
Cost per ton of ammonia			$18.60

(1) Fuel and boiler feedwater required by the Energy Center, net of amounts used in a typical 1000 T/D ammonia plant.

H. ALTERNATE 6 - 100 MGD OF DESALTED WATER, 1000 T/D OF AMMONIA, AND 150 MW(e) OF ELECTRIC POWER

1. Process

The concept of a common energy center is used in the design of a multiple-purpose plant to co-produce desalted water, ammonia, and electricity.

All steam turbines in the ammonia plant and the power plant are non-condensing. All the exhaust steam is used in the brine heaters of the MSF plant, producing 100 MGD of desalted water.

Some of the power produced by the turbogenerator is used to meet the electrical requirements of the complex.

Pumps for the desalting plant require 25,000 Kw. Pumping power to provide sea water for the ammonia plant coolers totals 1000 Kw. One-third of the pumping power required for boiler feedwater is supplied by let-down steam. Energy requirements for the remaining Boiler Feedwater Pumps and the H.P.C. Pumps total 5,000 KW. Therefore, the net power output is 119 MW(e).

The energy balance of the complex is shown in Figure 20.

2. Cost of Water, Ammonia and Electricity

The investment cost of the boiler and its auxiliaries and the sea water inlet station, and all operating costs of the boiler and its auxiliaries, are divided between the power generating facilities and the desalting plant in proportion to the exergy of the steam utilized at each plant (Fruth method). Calculations are based on the extraction

NOTE: ALL FLOWS IN THOUSANDS OF POUNDS PER HOUR

Figure 20- Energy Balance Alternate 6

cycle and turbine efficiency used to prepare the energy balances. The method of calculation is detailed in Appendix B. Approximately 65% of the energy is utilized to produce power and 35% to desalt water. Ammonia plant costs are assumed equal to those which would be achieved in a single-purpose ammonia plant.

Investment and production costs are shown in Tables XIX and XX.

With gas at 25¢ per 10^6 Btu, desalted water is produced for 27.3¢ per 1000 gallons, ammonia for $18.50/ton and electrical power for 4.2 mils/Kwh. This is the lowest cost water of any of the Alternates studied.

The value of desalted water changes if different values are assigned to either ammonia or electricity. This relationship is shown in Figures 17d and 17e. These figures also illustrate the effect of a changed natural gas price.

I. ALTERNATE 7 - 179 MGD OF DESALTED WATER AND 1500 T/D OF AMMONIA (NUCLEAR ENERGY SOURCE)

1. Process

This alternate investigates the feasibility of utilizing nuclear heat, directly, to replace the fossil fuel requirements of an ammonia plant and a desalting plant.

The physical configuration of a high temperature gas-cooled reactor, of the type now being built by

TABLE XIX. INVESTMENT AND ANNUAL FIXED CHARGES

ALTERNATE 6

PRODUCT	WATER	AMMONIA	ELECTRICITY
NOMINAL RATING	101.4 MGD	1000 T/D	150 MW (1)
ANNUAL PRODUCTION	33,200 x 10^6 Gal.	330,000 tons	1,180 x 10^6 Kwh (1)
INVESTMENT ($000)			
Basic plant cost	61,780	16,200	10,800
Energy center, allocated costs (2)	2,300	inc.	4,300
Sea water submarine pipe lines and intake station, allocated costs (2)	(2,850)	inc.	2,850
Heat recovery, incremental costs	1,000		
Common off-site facilities	(550)		(950)
Common erection costs	(250)		(450)
Net investments	61,430	16,200	16,550
ANNUAL FIXED CHARGES ($000)			
Capital charges	4,300	2,260	1,160
Maintenance materials, catalyst and chemicals	857	590	374
Maintenance labor	309	150	210
Operating labor, supervision and administration	362	450	120
Common maintenance and administration costs	(80)		inc.
Energy center, allocated costs of labor, supervision and supplies (2)	150		(150)
Total fixed charges	5,898	3,450	1.714

(1) 31 MW are required for the production of desalted water and ammonia. Therefore, net power rating is 119 MW and net annual production is 935 x 10^6 Kwh.

(2) Costs allocated on the basis of exergy utilization - 65% to the production of electricity and 35% to the production of water.

TABLE XX. UTILITY REQUIREMENTS AND TOTAL COST OF PRODUCTION

ALTERNATE 6

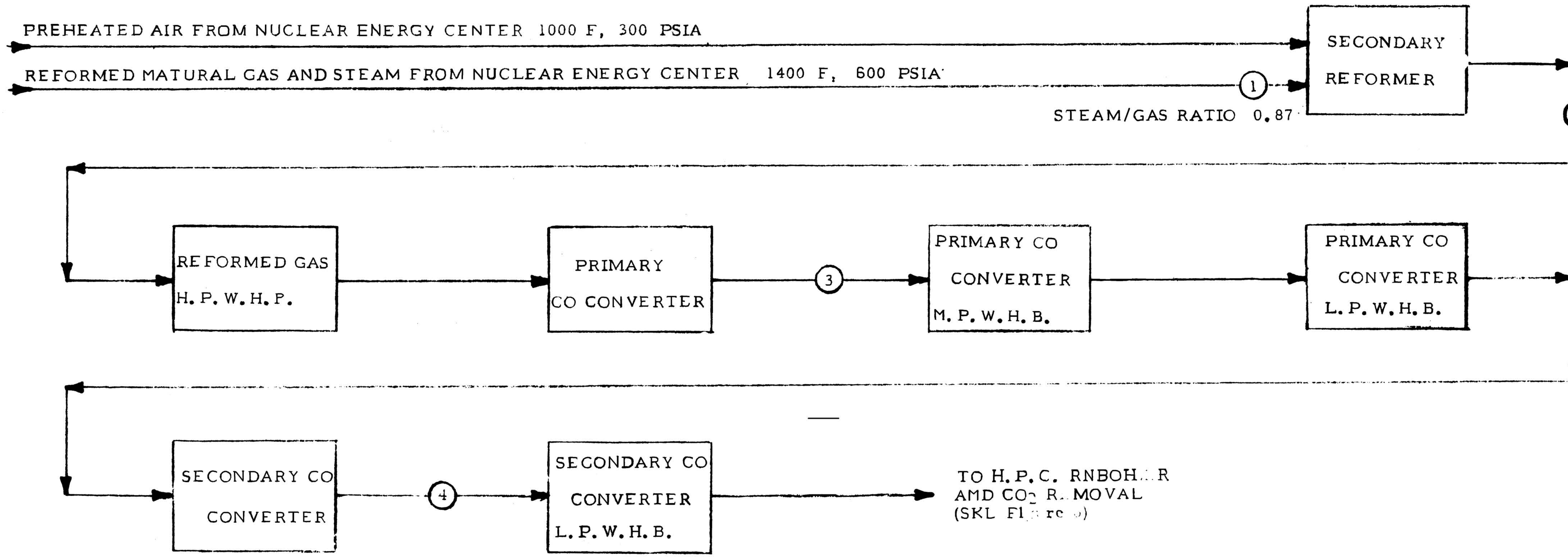
PLANT	WATER	AMMONIA	ELECTRICITY
NOMINAL RATING	101.4	1000 T/D	150 MW (1)
ANNUAL PRODUCTION	33,200 x 10^6 Gal.	330,000 Tons	1,180 x 10^6 KW
NET UTILITY REQUIREMENTS/YR. (2)			
Natural gas - 10^{12} Btu	8.44	10.28	12.68
Power - 10^6 Kwh	210 (3)	13.2	22.1
Make-up B.F.W.-10^6 Gal.	48	168	90
ANNUAL VARIABLE CHARGES ($000)			
Natural gas @ $0.25/$10^6$ Btu	2,100	2,560	3,170
Power @ 4.2 mils/Kwh	880	55	93
Water @ $0.27/$10^3$ Gal.	13	45	24
Total variable costs	2,993	2,660	3,287
COST OF PRODUCTION ($000)			
Fixed charges ($000)	5,898	3,450	1,714
Variable charges ($000)	2,993	2,660	3,287
Total cost of production ($000)	8,891	6,110	5,001
Cost per 1000 gallons of water	26.8¢		
Cost per ton of ammonia		$18.50	
Cost per Kwh of electricity			4.2 mils

(1) Net power output is 119 MW.
(2) Utility requirements of the Energy Center, net of amount used in a typical 1000 T/D ammonia plant, allocated to the production of water and electricity in accordance with exergy consumption.
(3) Sea water, brine and product pumps require 198 x 10^6 Kwh/yr.

Gulf-General Atomic at Fort St. Vrain, Colorado is modified to incorporate the gas reforming tubes and preheat coils of an ammonia plant. The temperature of the circulating helium is increased to 1650°F, a level at which it is possible to obtain satisfactory reforming equilibrium conditions. High pressure steam is generated, for use in the VC desalination plant and in the ammonia plant. Exhaust steam from all compressor and pump drives, and low level heat from the ammonia plant, is utilized in a MSF desalination plant.

The required changes in the design of the nuclear reactor can be made readily.

The ammonia process is similar to that utilized in Alternate 3. However, to compensate for the lower reforming temperature, the reforming pressure is reduced and the steam-gas ratio is increased. The resultant flow diagram for the reforming section of the ammonia plant is shown in Figure 21. (The rate at which heat is absorbed by the reforming tubes is substantially the same whether energy is being transferred by convection from the circulating 700 PSIA helium in the nuclear reactor or by radiation in a conventional reforming furnace.) As less boiler feedwater is required by the ammonia plant, the exchanger following the Primary CO Converter H.P. Boiler is changed from a boiler feedwater heater to a low pressure



GAS ANALYSIS - MOL PER CENT				
COMPONENT	1	2	3	4
H_2	69.4	57.2	60.5	61.4
N_2	0.6	22.1	20.4	19.9
CO	7.[illegible]	11.0	2.6	0.3
CO_2	[illegible]	9.1	15.9	17.8
CH_4	10.4	0.3	0.3	0.3
A	—	0.3	0.3	0.3
TOTAL	100.0	100.[illegible]	100.0	100.0

Figure 21- Gas Reforming Flow Diagram Alternate 7

steam boiler . Purge gas and flash gas are burned in the furnace of a Purge Gas Boiler, designed to accept fuel having a low heating value. The high pressure steam which is generated in the ammonia plant waste heat boilers is superheated in the Purge Gas Boiler.

The desalination process is similar to that utilized in Alternate 5.

The energy balance of the entire complex is shown in Figure 22.

Plant operations are started by burning natural gas in the Purge Gas Boiler. Steam is sent through a by-pass valve into the intermediate pressure steam line to start the helium circulator. (The amount of steam required is much less than the capacity of this boiler.) With the helium compressor in operation, the reactor is activated and the desalting water plant started up. As the power level in the reactor increases, the helium temperature rises. The reforming operation is then started and the ammonia plant placed on stream.

A more complete description of the nuclear reactor and of the entire process may be found in Appendix A. (11).

2. Cost of Water and Electricity

As in the case of the fossil-fueled complexes producing both desalted water and ammonia, it is very difficult to calculate a theoretically preferable method of allocating energy costs.

Figure 22- Energy Balance
Alternate 7

For the purposes of this investigation, 20% of the cost of the nuclear reactor and of the nuclear fuel is assigned to the ammonia plant and 80% to the desalination plant. These percentages are used because 20% of the high temperature nuclear heating is used in the ammonia plant - 14% to support the reforming reaction and 6% to drive ammonia plant compressors.

Capital charge rates of 7% were assumed for the reactor and for the desalination plant, and 14% for the ammonia plant.

Investment costs are shown in Tables XXI and XXII.

Assuming 25¢/10^6 Btu as the price of natural gas and 4.5 mils/Kwh for electric power, water would be produced for 27¢/Kgal and ammonia for $17.50/ton.

A chart indicating the relation of arbitrary assignable dollar values to either ammonia or desalted water as shown in Figure 17f. The same chart indicates the effect of a change in natural gas cost to either 15¢ or 40¢ per 10^6Btu.

As indicated in Appendix A, doubling the size of the nuclear reactor results in a 3¢/Kgal. decrease of the cost of producing water and a $1/ton decrease in the cost of ammonia.

TABLE XXI. INVESTMENT AND ANNUAL FIXED CHARGES

ALTERNATE 7

PRODUCT	WATER	AMMONIA	NUCLEAR ENERGY CENTER
NOMINAL RATING	179 MGD	1500 T/D	840 MW (th)
ANNUAL PRODUCTION	59,000 x 10^6 Gal.	490,000 Tons	
INVESTMENT ($000)			
Basic plant cost	108,640	21,500 (1)	44,000
Conversion of vapor compressors and generators to steam drive	(9,940)		
Special boiler feedwater treatment, pumps and drives			
Common off-site facilities	(400)	(100)	1,400
Common erection costs	(500)	(100)	(200)
Net investment	97,800	21,300	45,200
ANNUAL FIXED CHARGES ($000)			
Capital charges	6,840	2,980	3,160
Maintenance materials, catalyst and chemicals	1,982	830	230
Maintenance labor	542	150	inc. below
Operating labor, supervision and administration	679	450	450
Nuclear liability insurance			300
Nuclear energy center, allocated costs (2)	3,310	830	(4,140)
Total fixed charges	13,353	5,240	

(1) Cost after modification of typical plant; including elimination of convection section, incorporation of heat recovery equipment and purge gas boiler, change of equipment specifications to accomodate lower operating pressure and higher steam-gas ratio,etc.
(2) Costs allocated to the production of water and ammonia in accordance with the utilization of high level energy.

TABLE XXII. UTILITY REQUIREMENTS AND TOTAL COST OF PRODUCTION

ALTERNATE 7

PLANT	WATER	AMMONIA	NUCLEAR ENERGY CENTER
NOMINAL RATING	179 MGD	1500 T/D	840 MW(th)
ANNUAL PRODUCTION	59,000 x 10^6 Gal.	490,000 Tons	
NET UTILITY REQUIREMENTS/YR.			
Natural gas - 10^{12} Btu		9.90	
Power - 10^6 Kwh		19.6	14.0
Make-up B.F.W.- 10^6 Gal.		269	75
Nuclear fuel - 10^{12} Btu			22.6
ANNUAL VARIABLE CHARGES ($000)			
Natural gas @ $0.25/$10^6$ Btu		2,480	
Power @ 4.5 mils/Kwh		90	60
Water @ $0.27/$10^3$ Gal.		72	20
Nuclear fuel @ $0.15/$10^6$ Btu (1)			3,400
Nuclear energy center, allocated costs (2)	2,784	696	(3,480)
Total variable costs	2,784	3,338	
COST OF PRODUCTION			
Fixed charges ($000)	13,353	5,240	
Variable charges ($000)	2,784	3,338	
Total cost of production ($000)	16,137	8,578	
Cost per 1000 gallons of water	27.3¢		
Cost per ton of ammonia		$17.50	

(1) Fuel cycle costs are 0.51 mils/Kw (t)
(2) Costs allocated to the production of water and ammonia in accordance with the amount of high level energy used.

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APPENDIX A

A CONCEPTUAL NUCLEAR ENERGY CENTER FOR THE CO-PRODUCTION OF AGRICULTURAL CHEMICALS AND DESALTED WATER TO SERVE A PILOT FOOD FACTORY *

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INTRODUCTION

The concept of utilizing a nuclear heat energy center for the co-production of power, desalted water and ammonia was first extended by one of the co-authors of this paper in 1963. This original concept, for which a U.S. patent has been issued (1), proposed the direct exchange of nuclear heat to a gaseous mixture of methane and steam to support the endothermic energy requirements for the chemical reforming reaction to carbon monoxide and hydrogen. The simplified schematic flow diagram related to this invention is reproduced herein as Figure 1. Quoting from the background section of the patent narrative: " A broad application of the invention

* Presented at International Atomic Energy Agency Symposium on Nuclear Desalination, Madrid, Spain, November, 1968.

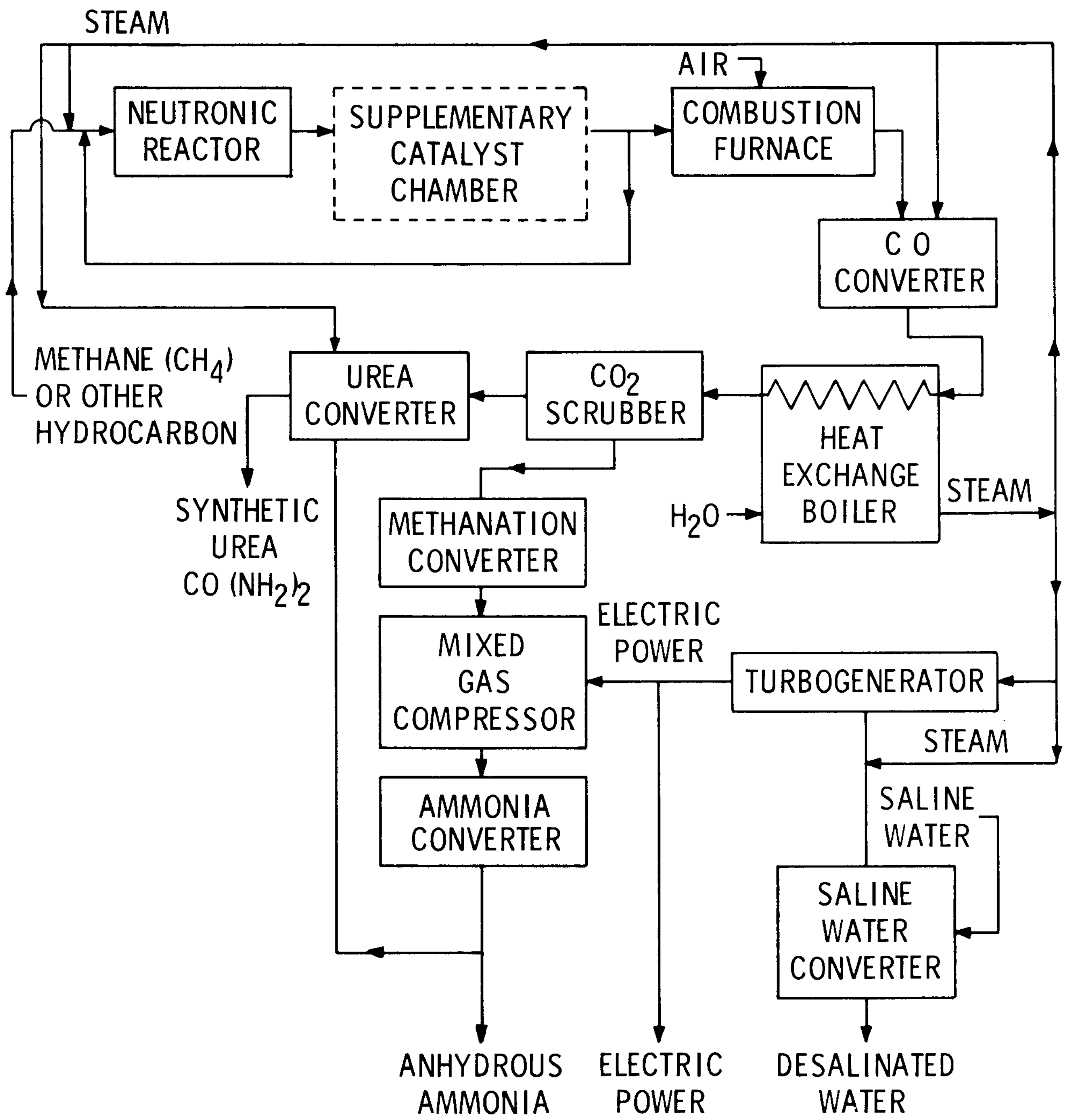


Figure 1—Nuclear Energy Center for the Co-Production of Power, Desalted Water and Ammonia (from U. S. Patent No. 3,394,050)

would utilize the reactor to provide power for transporting saline water from an ocean; convert the saline water to fresh water; produce anhydrous ammonia for addition to the fresh water through an endothermic chemical reaction driven by reactor heat; and transport the ammoniated, desalinated water inland to irrigate arid land. The design of the reactor would be optimized to provide an allocation of energy generated therein to chemical production, electric power production and saline-to-fresh water conversion in proportion to the needs of the geographical area in which the reactor is located."

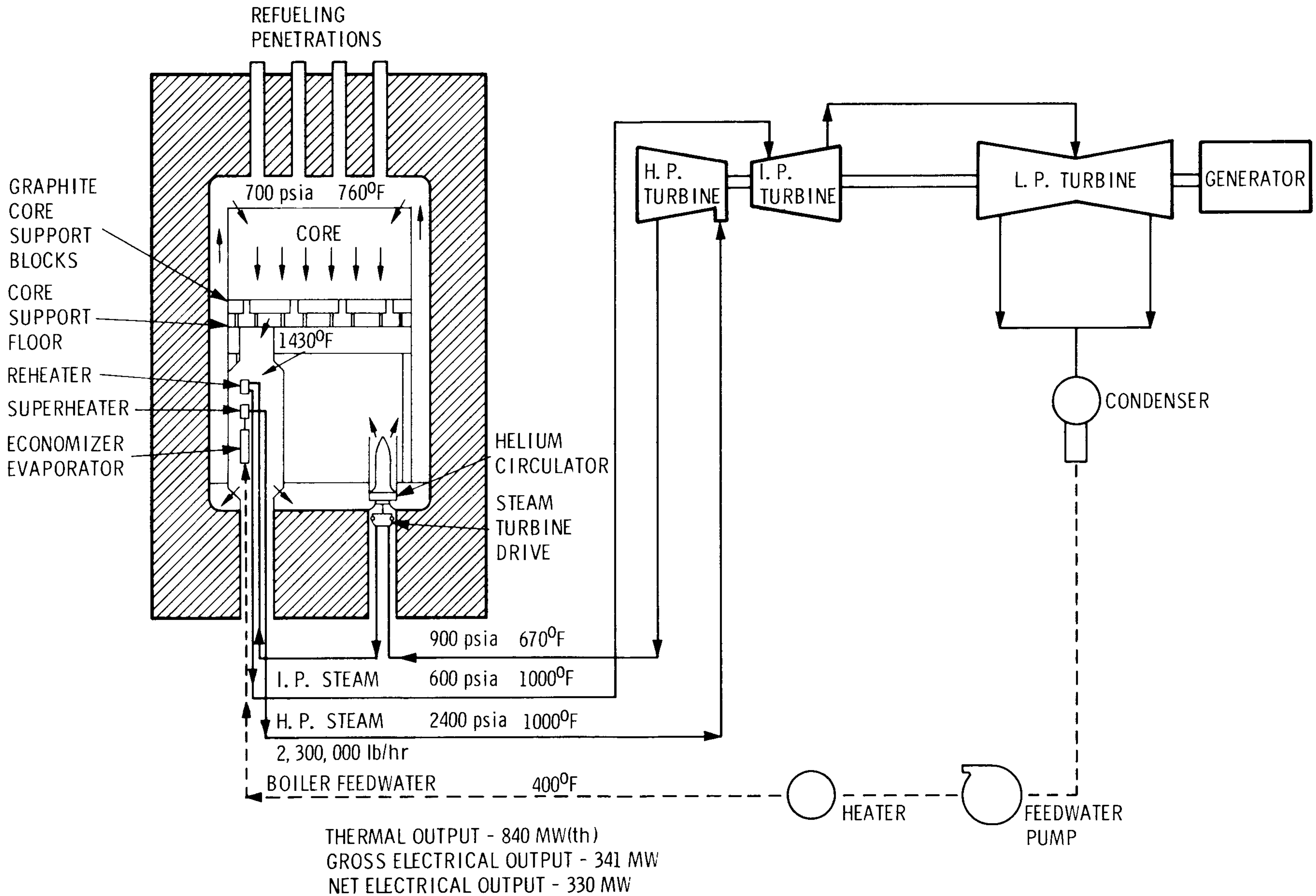
In 1966, as part of a joint study sponsored by the U.S. Office of Saline Water and the Texas Water Development Board, the authors initiated studies of a less advanced application of this concept- one that conceivably could be translated to hardware and compete in the market place on the basis of the application of proven technology. This initial study investigated the feasibility and cost of co-producing 50 Mgal/day (million U.S. gallons per day) of desalted water, 120 MW(e) of electric power and 500 (short) tons/day of ammonia in a single plant complex having a fossil-fired energy center. The specific quantities of desalted water, power and ammonia were selected to correspond to needs in the Lower Rio Grande Valley of Texas for municipal water, incremental electric power and estimated quantities of anhydrous ammonia which could be sold in the region for agricultural fertilizer application. The results of this study (2) provided the following unit costs of products: potable (500 ppm total dissolved solids) water, 26 cents per 1000 gallons; power, 3.7 mills per Kwh; and anhydrous ammonia, $18.10 per ton. It should be mentioned that natural gas, which was assumed as the energy source for this complex, is available in large quantities in this region at a cost of 18 cents per 10^6 Btu.

As a follow-on to this study we have just completed the preliminary examination of the economic effects of scale-up of ammonia plant size on the cost of co-producing desalted seawater with and without electric power production from a common energy center fired by natural gas. In addition we studied one case based on the application of a high-temperature , gas-cooled nuclear reactor of the Gulf-General Atomic Inc. design, modified to provide for the double duty of reforming natural gas and generating high pressure steam adjacent to the reactor core, within the pre-stressed concrete containment shell surrounding the reactor. The results of this dual-purpose nuclear desalting study case are reported in the following paragraphs.

<u>THE DUAL-PURPOSE HTGR</u>

The HTGR reactor design by Gulf-General Atomic Inc. and now under construction at Fort St. Vrain, near Greeley, Colorado, utilizes a circulating stream of helium to transfer heat from the fuel elements to steam generating coils. In this design the helium is heated to 1430°F at a pressure of 700 psia. The helium flows through twelve modules, each of which contain an economizer, evaporator, superheater and reheater coils. Steam is generated at 2400 psia and 1000°F and reheated to 1000°F at 600 psia. The helium, now cooled to 760°F and at a pressure of 685 psia, is recompressed to 700 psia and the cycle is repeated. This reactor is designed to operate at 840 MW(th). Because steam will be generated at high pressure and temperature, the plant will operate with a thermal efficiency of 40%, producing 330 MW(e). A sketch of the Fort St. Vrain reactor is provided in Figure 2.

At the elevated temperature at which the HTGR will operate it becomes practical to utilize nuclear heat to provide the energy necessary to react natural gas (or naphtha) with steam to form hydrogen and carbon oxides. Additional quantities of heat from the reactor can be utilized to generate high pressure steam needed for the reformer reaction and for distillation desalting via the employment of a vapor-compression cycle. With the drives of both the vapor compressors and the ammonia plant compressors exhausting to an MSF distillation plant, rather than to a condenser, a very high thermal efficiency can be obtained.



COURTESY:
GULF-GENERAL ATOMIC INC.
SAN DIEGO, CALIFORNIA, USA

Figure 2—Fort St. Vrain Nuclear Generating Station

This basic concept of utilizing an HTGR reactor in conjunction with a vapor recompression - MSF process for producing ammonia and water (Figure 3) has several advantages :

1. It is flexible. Although the requirement of a high temperature heat transfer imposes a limit to the amount of ammonia which can be produced from a reactor of a given capacity, within this constraint there may be a wide variation in the ratio of ammonia to desalted water or electrical power production.

2. Only that amount of hydrocarbon which is to be directly converted into ammonia is required as a raw material. All hydrocarbon fuel requirements are eliminated. Thus the natural gas requirement is only 20×10^6 Btu/ton, a savings of one-third over that required by conventional ammonia plants. This is a significant conservation of an important non-renewable natural resource.

3. In areas of the world where hydrocarbon costs are relatively high, and where there is a demand for desalted water and/or electrical power, the process appears to be economically attractive when utilizing an 840 MW(th) reactor. Any increase in plant capacity would enhance this economic attractiveness.

4. The required technology and equipment available today and plant design can be started immediately.

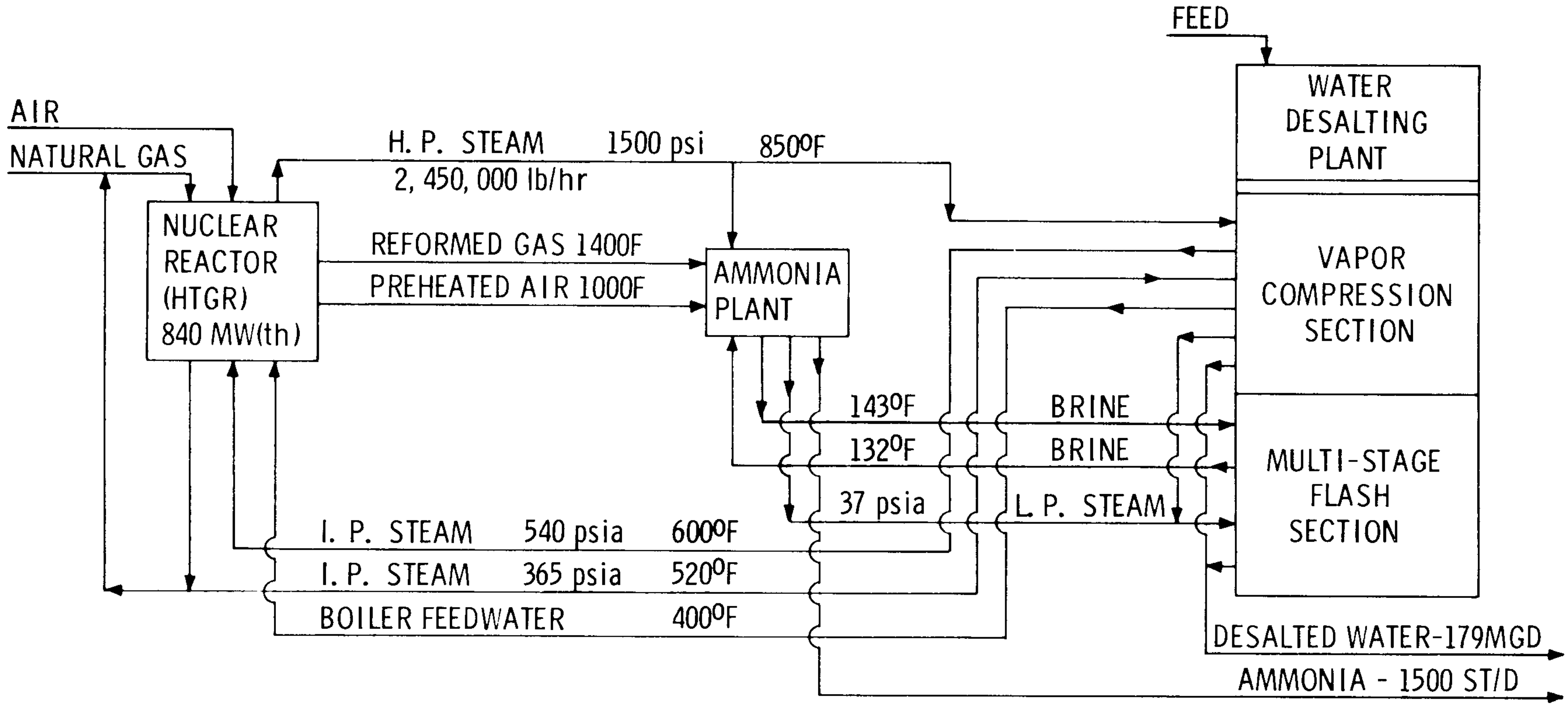
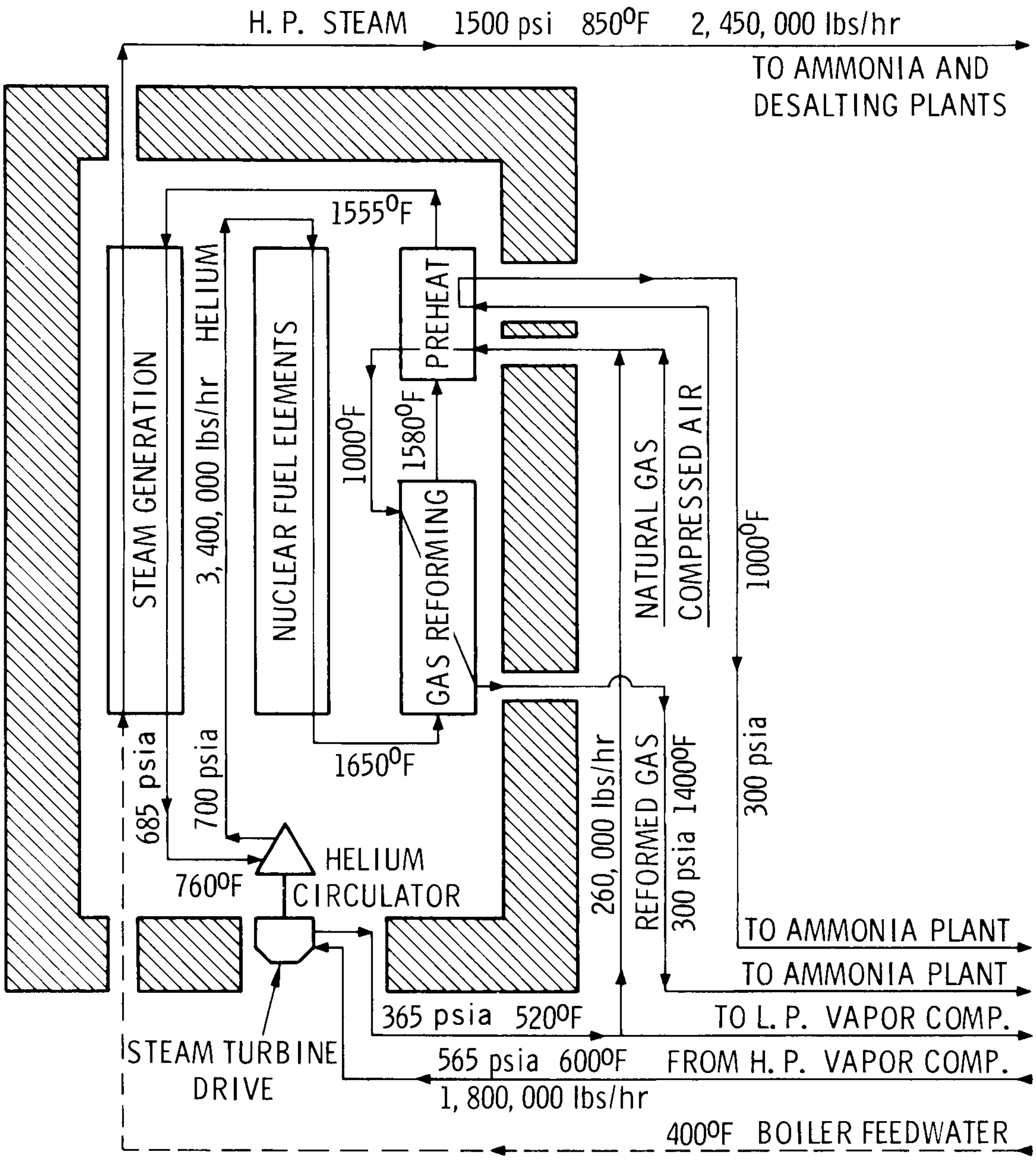


Figure 3—Nuclear Energy Center for the Co-Production of Water and Ammonia

THE CONCEPTUAL NUCLEAR ENERGY CENTER

Some details of the conceptual nuclear energy center for the co-production of ammonia for agricultural chemicals and desalted water to serve a pilot food factory are shown in Figure 4. This is a schematic of the reactor showing the modified process temperature profiles.

It is important to note that in the reforming of natural gas, accomplished by passing a mixture of the hydrocarbon and steam through catalyst-filled tubes, the methane content of gas leaving the reactor is influenced by the temperature, the pressure and the ratio of steam-to-gas. In conventional plants the reforming tubes are hung in gas-fired furnaces in which the walls radiate heat to the tubes to maintain the desired temperature level in the catalyst bed. In modern ammonia plants the gas mixtures leave the furnace at a temperature of about 1500°F and a pressure of about 450 psia. The present designs of the Gulf-General Atomic reactor will permit operation to a helium temperature level of 1650°F. With helium at this temperature and at a pressure of 700 psia one may obtain very high heat transfer rates across the catalyst tubes; thus it is feasible to heat the reacting hydrocarbon and steam to 1400°F. Reducing the pressure of the steam-to-gas mixture to 300 psia and increasing the exit



TOTAL HEAT EVOLVED - 2, 880 x10^6 BTU/HR

HEAT UTILIZATION:

PREHEAT GAS, STEAM AND AIR AND PROVIDE HEAT OF REACTION	14%	20%
GENERATE STEAM FOR AMMONIA PLANT DRIVES	6%	
GENERATE STEAM FOR VAPOR RECOMPRESSION DRIVES		80%

Figure 4—Modified Energy Profile Through Reactor

steam-to-gas ratio to 0.9 enables one to produce a product gas mixture containing only 11 mol percent of unconverted methane (dry basis). This is comparable to the yields obtained in today's conventional primary reformer furnaces and is suitable for processing further into ammonia synthesis gas.

As a result of supplying the heat of reaction for the conversion of methanes into carbon oxides, the helium temperature is reduced to 1580°F. The helium flow is then passed over coils which preheat both the steam-natural gas feed mixture and the required volume of air for the secondary reforming step to 1000°F. The helium leaves the primary reformer section at 1555°F. It is passed over steam generation coils in which additional amounts of heat are transferred from the gas stream produce steam at 1500 psia and 850°F from 400°F boiler feed water. The cooled helium, now at 760°F and 685 psia, is recompressed to 700 psia and recycled to the reactor core.

In the configuration which has been described, about 14% of the reactor heat energy is used to support the reforming reaction. However, the system is designed to permit all of the hot helium flow to pass over these coils to insure maintenance of the reformer tubes at 1500-1600°F at all times during normal operation of the dual-purpose complex. The present Fort St. Vrain design can accomplish this, either by enlarging and utilizing two of the twelve modules for the reformer duty, or by placing both reformer and steam generator coils in each of the modules. Preliminary evaluation indicates that either method is conceptually sound and neither method would cause excessive pressure drop or flow distribution problems.

Another 6% of the reactor heat energy is used to produce high pressure steam for consumption in the ammonia plant. The sum of this steam supply and that produced in the waste heat boilers of the ammonia plant comprise the total steam used to drive the ammonia plant compressors.

As shown in Figure 5, the bulk of the high pressure steam which is generated in the reactor is used to drive the vapor compressor portion of the VC-MSF plant. Six vapor compressors are used. The first three are driven by 36,000 hp non-condensing extraction turbines. Steam, extracted at 540 psia, is used to drive the helium compressors. The remaining steam is exhausted at 37 psia and piped to the MSF portion of the desalting plant. The helium compressor turbines exhaust at 365 psia. A portion of this steam is mixed with natural gas and converted into hydrogen. The remaining steam is used to drive three more 21,000 hp vapor compressors, as well as miscellaneous pumps in the ammonia plant, all exhausting at 37 psia. The basic concepts of the VC-MSF desalting plant have been described in a previous paper. (3)

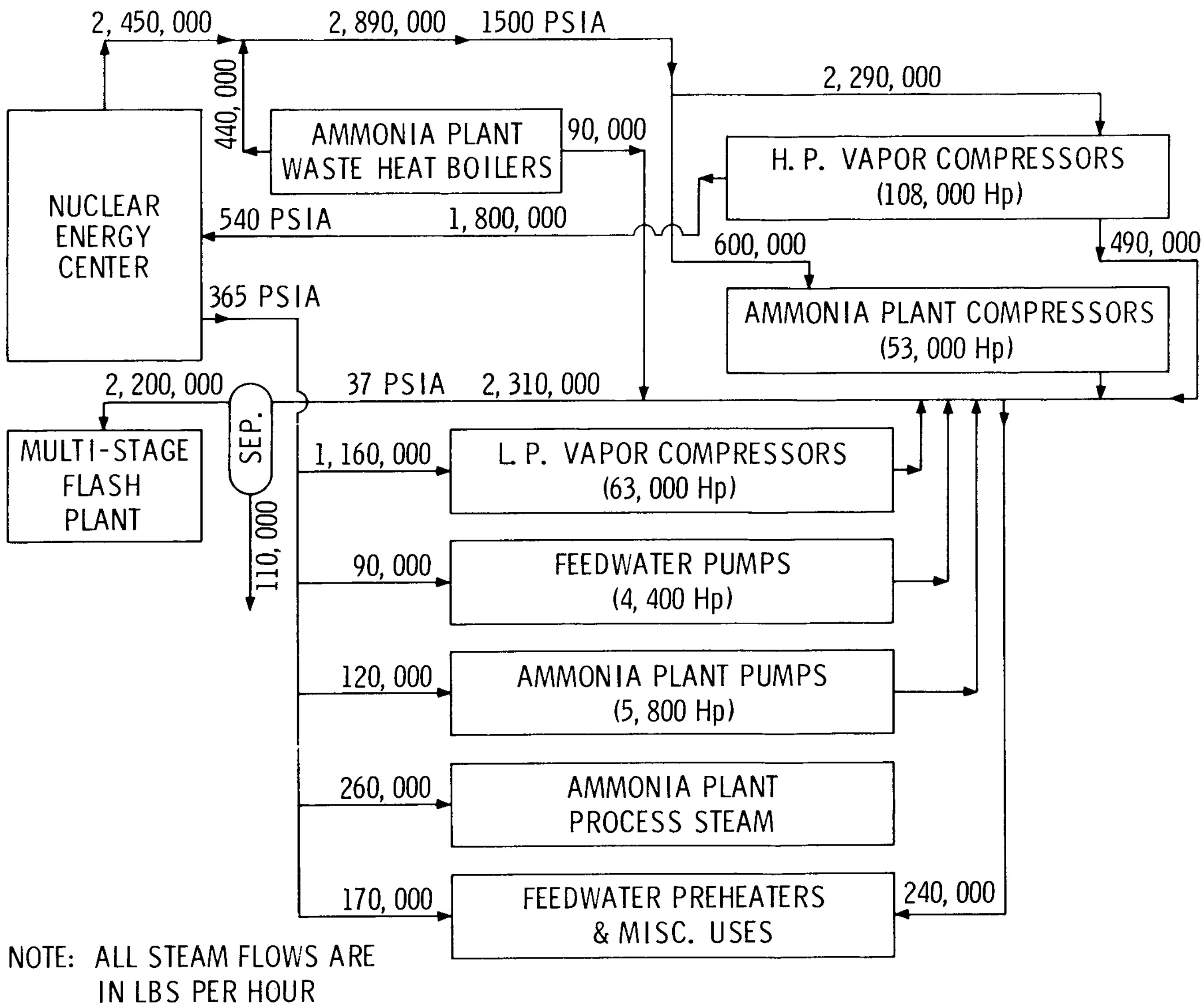


Figure 5—Steam Distribution

The remainder of the ammonia plant is of conventional design, employing a secondary reformer, two stages of carbon monoxide conversion, carbon dioxide removal using hot potassium carbonate, the removal of residual carbon oxides by methanation, and thence compression of the purified mixed gas and high-pressure ammonia synthesis. However, drives for the air, gas and refrigeration compressors in the ammonia plant are performed by non-condensing steam turbines, with the 37 psia exhaust steam being utilized in the brine heaters of the water desalting plant. Waste heat is recovered in both high pressure (1500 psia) and low pressure (37 psia) boilers. Additional amounts of low-level heat are recovered, where economically feasible, by pumping hot brine from the desalting plant to heat exchangers in the ammonia plant. Complete details of this heat recovery scheme will be published by the Office of Saline Water in the near future. (4)

NUCLEAR REACTOR SAFETY

Radioactivity in the circulating helium flow through the reactor is maintained at a very low level (microcuries per millilitre) by a continuous purification system to avoid danger of contamination of the steam, methane or heavier hydrocarbons should leakage occur through the reforming tubes. Alternatively, in no case could a malfunction of the ammonia plant create any difficulties in the operation of the nuclear reactor. Since the reactor operates at a helium temperature of only 1650°F, the reforming tubes would not be damaged even if they approached this environmental temperature. This may be contrasted with conditions in the usual reforming furnace wherein the tubes receive energy from a radiating surface and stoppage of the gas flow - whether locally or throughout the furnace - can result in overheating of the tube wall and cause subsequent tube failures.

ECONOMICS OF THE DUAL-PURPOSE COMPLEX

Table I shows the investment requirements and annual fixed charges for each of the three principal components of the proposed facility - the nuclear heat reactor unit, the water desalting plant and the ammonia plant. Each plant is complete, although storage and distribution facilities are excluded.

Capital charges, which include amortization, interest and taxes, are assumed as 14% for the ammonia plant and 7% for the water desalting plant and for the nuclear heat reactor unit. This is based on the assumption that the ammonia plant will be privately financed whereas the latter plants will be publicly financed. Costs of the nuclear heat reactor unit were obtained from the Gulf-General Atomic Inc. (5) Costs

TABLE I. INVESTMENT REQUIREMENTS AND ANNUAL FIXED CHARGES

Plant	Ammonia	Nuclear Steam Generation	Water Desalting VC-MSF
Nominal Rating	1500 T/D	840 MW(th)	179 Mgal/d
Annual Production (90% Utilization)	490,000 tons		59,000 x 10^6 gallons
Total Required Investment	$21,300,000	$45,200,000	$97,800,000
Annual Fixed Charges			
Capital charges (1)	$2,980,000	$3,160,000	$6,840,000
Operating labor and administration	450,000	450,000	679,000
Maintenance labor	150,000	included	542,000
Maintenance materials, catalysts and chemicals	830,000	230,000	1,982,000
Nuclear liability insurance		300,000	
Transfer costs of nuclear plant	830,000	(4,140,000)	3,310,000
Total fixed charges	$5,240,000	-	$13,353,000

(1) Capital charges include amortization, interest and local taxes. These are assumed to total 14% for the ammonia plant and 7% for the nuclear steam generation and the water desalting plants.

of the VC-MSF water desalting plant were obtained from the Union Carbide Corporation, Nuclear Division.(6)

Total charges of the nuclear unit were allocated on the basis of 20% to the ammonia plant and 80% to the desalting plant. This allocation is used because, as has been indicated, 20% of the heat energy generated by the nuclear reactor is used to reform natural gas and to generate steam to drive the ammonia plant compressors.

Total annual variable charges, and the cost of production of water and ammonia, are shown in Table II. Two tabulations are presented, one with natural gas costing 15 cents per 10^6 Btu and the other with natural gas costing 40 cents per 10^6 Btu. In both cases it has been assumed that electric power would be available at 4.5 mills per Kwh and nuclear fuel at 0.51 mills per Kwh (15.0 cents per 10^6 Btu). The nuclear fuel cycle cost is based on the assumptions of equilibrium operation with U-233 recycle, uranium ore at $8 per pound and cost of separative work at $26 per unit. The results showed 179 Mgal/day desalted water costing 29 cents per 1000 gallons and 1500 tons/day anhydrous ammonia costing $15.50 and $20.50 per ton, respectively, for the cases of 15 cent and 40 cent cost of natural gas.

TABLE II. ANNUAL VARIABLE CHARGES AND COST OF PRODUCTION OF WATER AND AMMONIA

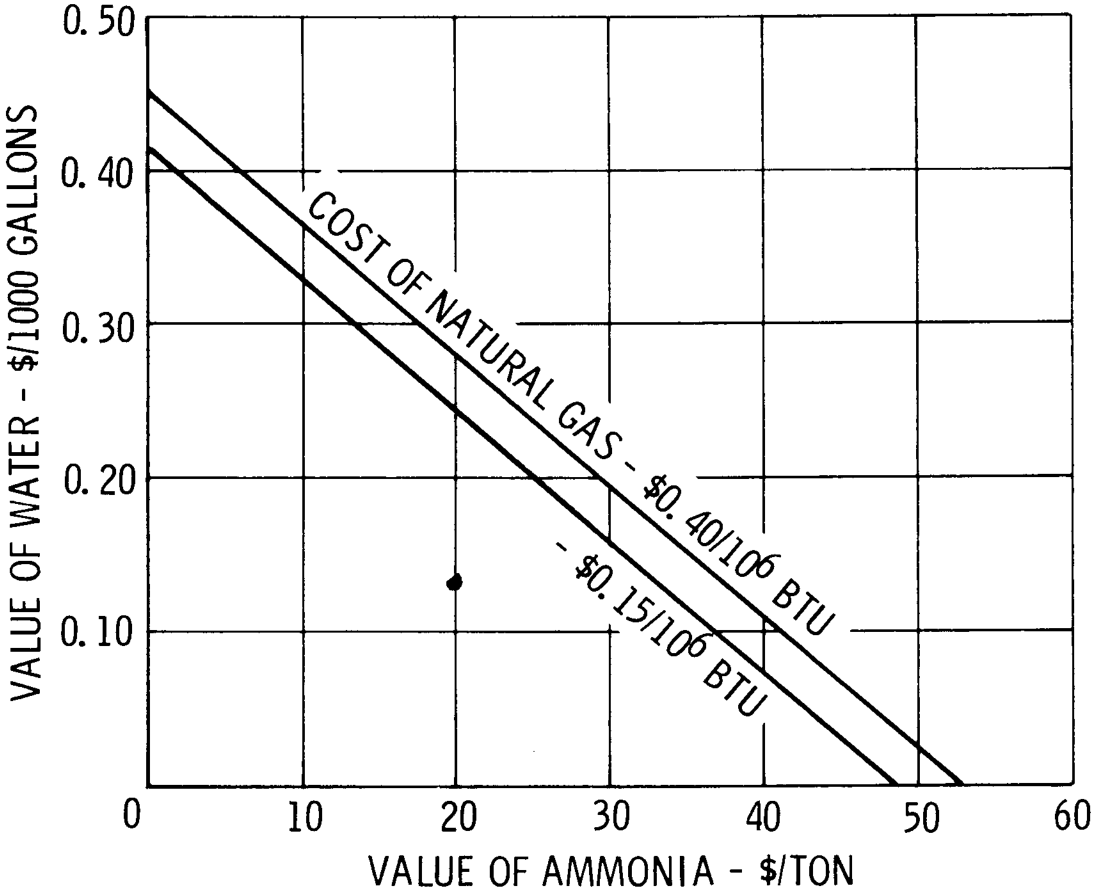
Plant	Ammonia	Nuclear Steam Generation	Water Desalting VC-MSF
Nominal Rating	1500 T/D	840 MW(th)	179 Mgal/d
Annual Raw Material and Energy Requirements			
Natural gas -10^{12} Btu	9.9		
Nuclear fuel -10^{12} Btu		22.6	
Electric power -10^{6} Kwh	19.6	14.0	
Boiler feedwater-10^{6} gal.	269	75	
Prod. Cost-Low Gas Price			
Natural gas @ \$0.15/10^{6}Btu	\$1,480,000		
Nuclear fuel @\$0.15/10^{6}Btu(1)		\$3,400,000	
Electric power @ \$0.0045/Kwh (2)	90,000	60,000	
Boiler feedwater @ \$0.29/1000 gal.	80,000	20,000	
Transfer costs of nuclear plant	700,000	(3,480,000)	2,780,000
Total variable costs	\$2,350,000	-	\$2,780,000
Total annual charges (3)	\$7,590,000	-	\$16,133,000
Cost of production			
-\$/ton ammonia	15.5		
-\$/1000 gallons			0.27
Prod. Cost-High Gas Price			
Natural gas @ \$0.40/10^{6}Btu	\$3,960,000		
Other variable costs, as detailed above	870,000		\$2,780,000
Total annual charges (3)	\$10,070,000		\$16,133,000
Cost of Production			
-\$/ton ammonia	20.5		
-\$/1000 gallons			0.27

(1) Based on equilibrium operation with U-233 recycle, uranium ore at \$8 per pound; separative work at \$26 per unit; and a 7% working capital charge. This cost is equivalent to 0.51 mills per KW (th).
(2) Assuming electrical power generation from nuclear energy
(3) Total fixed charges as noted in Table I.

Since the energy center serves to produce two distinct products, it should be possible to assign an arbitrary selling price to either one of the two products, with the second absorbing the difference in cost. A chart indicating the relation of dollar values is shown in Figure 6.

For comparative purposes, it may be noted that the ammonia produced in the Gulf Coast area of the United States, where natural gas costs 18 to 20 cents per 10^6 Btu, is reported to be sold generally for about $30 per ton (excluding storage charges), although much lower prices have been noted. The total cost of production, before general expenses and profit, is about $20-25 per ton. The effect of natural gas cost at 40 cents per 10^6 Btu would be to add $6.50 to the cost of ammonia production.

These costs represent an upper limit of the cost of producing desalted water and ammonia by the process outlined here. Today a Fort St. Vrain reactor design in considered small. As reactors increase in size, costs decrease significantly. For example, doubling the size of the nuclear heat reactor would probably reduce its basic price from $52 to $39 per KW(th). (7) While fuel costs and liability insurance would be expected to increase in proportion to power output, most other costs would change very little with an increase in plant size. A 1670 MW(th) reactor could produce about 360 Mgal/day of desalted water and 3000 tons/day of ammonia.

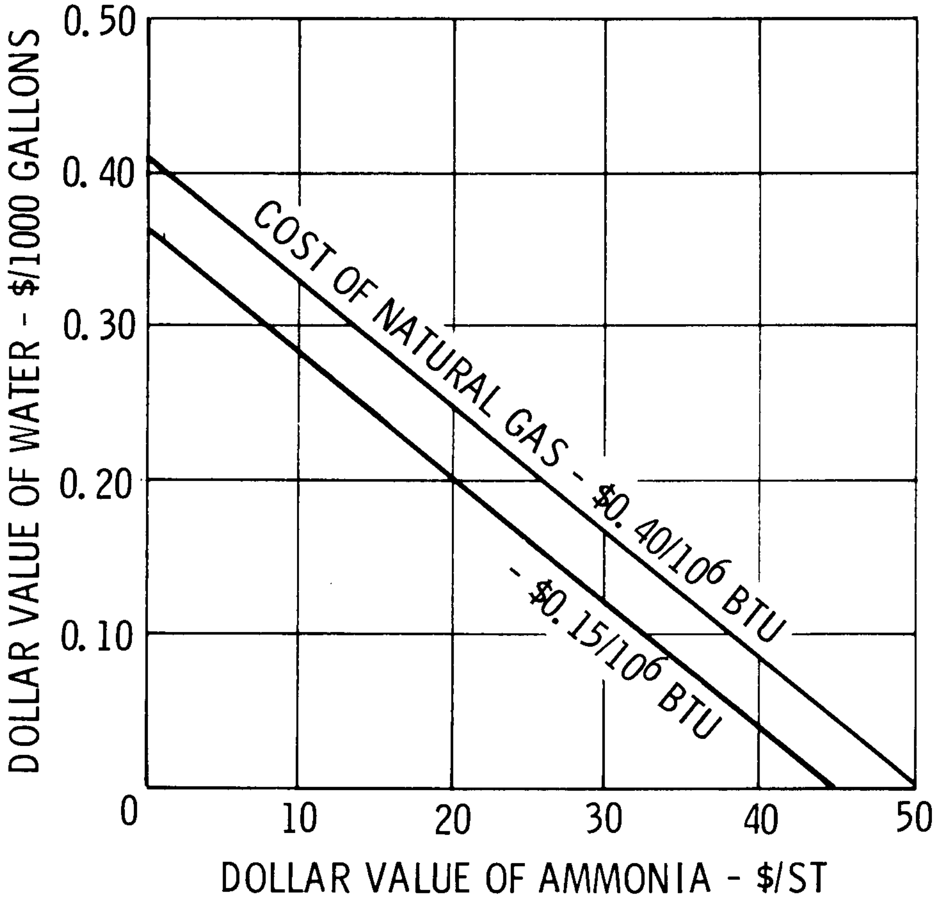


DOES NOT INCLUDE COSTS OF STORAGE OR DISTRIBUTION

Figure 6—Dollar Amounts Required to Recover Production Costs When Utilizing a 840 MW (th) HTG Reactor

On this basis, the cost of water would decrease to 24 cents per 1000 gallons and the cost of ammonia would decrease to \$14.50-\$19.50 per ton, for 15-40 cent natural gas cost. Figure 7 indicates the relation of dollar values of ammonia and desalted water. If the dual-purpose production facility were to be further expanded to include proportionately large quantities of electric power production, for elemental phosphorous production for example, the effect of scale-up would be to reduce the costs of water and ammonia production still further. Elemental phosphorous can be combined with ammonia to produce diammonium phosphate, a highly concentrated and widely useful fertilizer. Basically, the energy center described herein generates nuclear heat to produce carbon oxides, hydrogen and high-pressure steam. The steam can be used in many kinds of chemical processing operations or to generate electric power in a conventional power plant. The carbon oxides and hydrogen can be used to produce ammonia for agricultural chemicals or methanol for motor fuel or oxo chemicals or all of these products. As the technology of fuel cells develops, it may be found desirable to route the hot hydrogen and carbon monoxide gas mixtures and air to fuel cells for the very efficient production of electric power.

Figure 8 shows how this basic concept can be expanded to produce both the desalted water and the fertilizer required for a complete agricultural complex. All of the process and all of the required engineering technology is available today.



DOES NOT INCLUDE COSTS OF
STORAGE OR DISTRIBUTION

Figure 7—Dollar Amounts Required to Recover Production Costs When Utilizing a 1680 MW (th) HTG Reactor

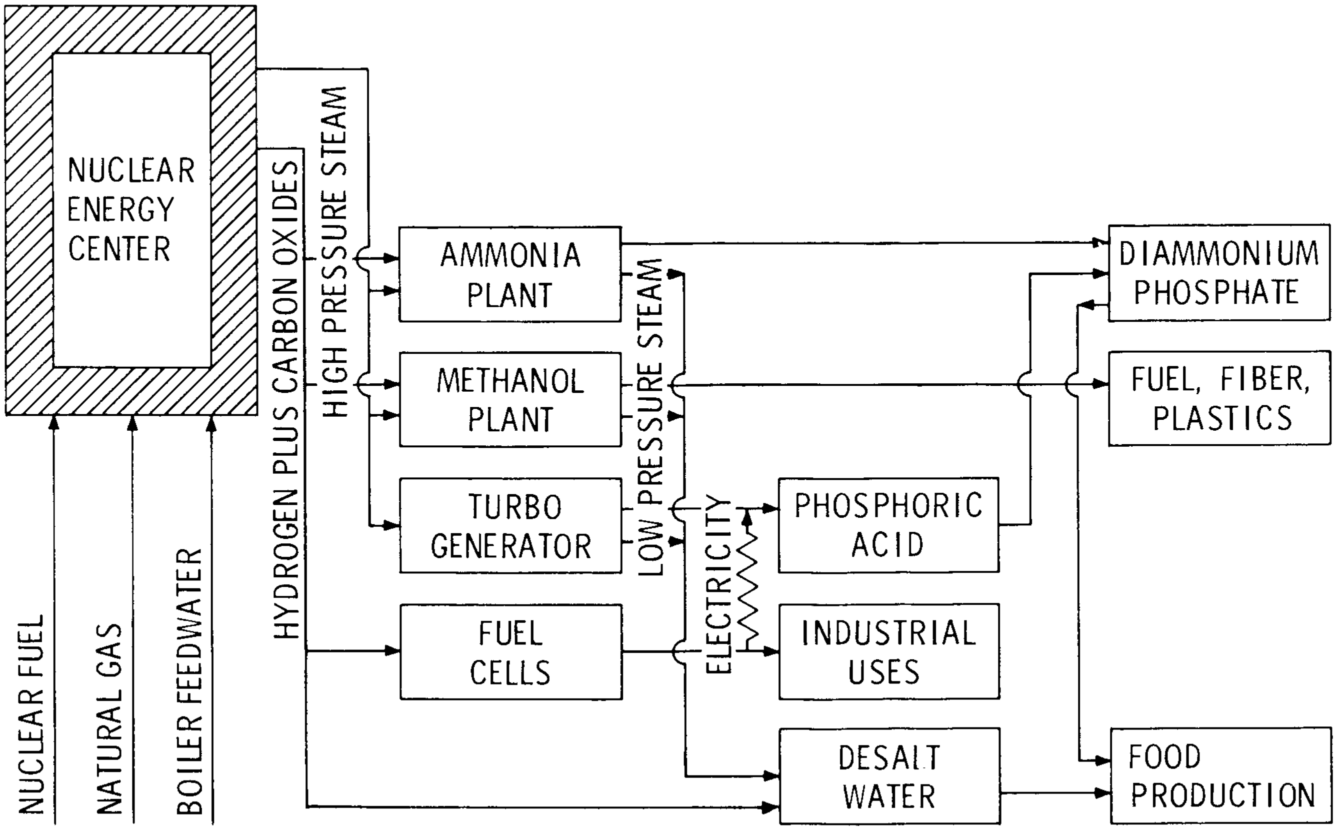


Figure 8—Nuclear Energy Center for the Production of Water and Fertilizers for an Agricultural Complex

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(5) Gulf-General Atomic Inc., Private communication (February 6, 1968).

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APPENDIX B

PRORATION OF ENERGY COST ON THE BASIS OF AVAILABLE ENERGY (EXERGY)

ALTERNATE 6

Condition of steam at turbine throttle-	1465 PSIA, 825°F
Pressure of steam at turbine exhaust when operating with condenser -	2 1/2" Hg
Turbine efficiency-	80%
No. of extraction stages -	Two
Pressure to 1st stage B.F.W. heater-	105 PSIA
Pressure to 2nd stage B.F.W. heater-	315 PSIA
Pressure of steam to desalination plant-	37 PSIA

Flow of steam to throttle-	2,110,000 Lbs/hr
Flow of steam to 1st stage B.F.W. heater-	157,000 Lbs/hr
Flow of steam to 2nd stage B.F.W. heater-	167,000 Lbs/hr
Flow of steam to desalination plant(total, including 26,000 lbs/hr to stream jets)-	1,760,000 Lbs/hr

h_A= Enthalpy of steam to throttle=	1380 Btu/lb
h_{C_2}= Enthalpy of steam to condenser=	958 Btu/lb
h_{Y_2}= Enthalpy of steam to 1st stage B.F.W. heater =	1177 Btu/lb
h_{Y_3} =Enthalpy of steam to 2nd stage B.F.W. heater=	1250 Btu/hr
h_{B_2} =Enthalpy of steam to desalination plant=	1177 Btu/hr

Y_1 = Fraction of steam to condenser= 0.0

Y_2 = Fraction of steam to 1st stage B.F.W. heater= 0.074

Y_3 = Fraction of steam to 2nd stage B.F.W. heater= 0.079

Y_b = Fraction of steam to desalination plant= 0.847

e'_A = corrected specific available energy at turbine inlet

$$= (h_A - h_{C_2}) - \sum_{j=1}^{n} Y_j (h_j - h_{c_2})$$

$h_A - h_{C_2} = 1380-958 =$	422
$Y_1 (h_1 - h_{C_2}) =$	0
$Y_2 (h_2 - h_{C_2}) = 0.074 (1177-958) =$	16.2
$Y_3 (h_3 - h_{C_2}) = 0.079 (1250-958) =$	23.8

e'_A = 422 - (16.2 + 23.8) = 382 Btu/lb

e'_B = corrected specific available energy of steam to desalination plant

$$= Y_b (h_{B_2} - h_{C_2})$$

$$= 0.847 (1177-958) = \quad 135 \text{ Btu/lb}$$

Therefore:

x_w = fraction of cost of common items attributable to water produced = e'_B / e'_A 135/382 = 0.35

x_e = fraction of cost of common items attributable to electricity produced $= 1 - e'_B / e'_A = 1.0 - .35 =$ 0.65

ANNEX A

SEAWATER DESALTING BY DISTILLATION PROCESSES, INCLUDING VAPOR COMPRESSION, AS APPLIED TO INTEGRAL CHEMICAL COMPLEXES

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INTRODUCTION

The combination of a seawater distillation plant with an ammonia production facility is similar to a dual-purpose power and water production plant. In the power case a back-pressure steam turbine is used for power production, and the exhaust steam is used for the brine heater of the distillation plant. Similarly, in the ammonia production plant various back-pressure steam turbine drives are used in the process, and the exhaust steam is used for the brine heater. Additional amounts of heat are obtained from energy usually dissipated to cooling water or the atmosphere.

When the water requirements exceed the amount of available heat to produce it, the water production is enhanced by the production of additional high-pressure steam which may then be best utilized by driving steam turbines which in turn drive vapor compressors for the vapor re-compression distillation process.

A conservative approach was taken in specifying the conventional multistage flash evaporator (MSF) for the production of potable water when sufficient exhaust steam is available to meet the assumed demand. The conceptual design performed by the Foster Wheeler Corporation for the Office of Saline Water (1) was used as a basis on which to build.

When additional water production was required, the vapor compression process under development at Oak Ridge (2) was applied, updated to reflect costs based on the latest estimates by Oak Ridge and others.

DISCUSSION OF THE MSF PROCESS USING EXHAUST STEAM AND WASTE HEAT

The MSF process used in this study has a performance ratio of 16.25 lb of product water per 1000 Btu of steam input to the brine heater. Some prefer to specify this ratio as 15 lb product per lb of steam. The MSF plant is a conventional single effect, multistage type using four heat reject stages and sixty-eight* recovery stages.

Sulfuric acid is added to the makeup feed which proceeds to the heat reject section after the CO_2 degassing tank. The stream is then deaerated and joins the recycle stream; these combined streams flow successively through the heat recovery tubes and on to the brine heater. The heated brine then passes directly to the flashing side of the first stage chamber and in succession through the sixty-eight stages.

Scale prevention is attained by the injection of sulfuric acid into the seawater makeup stream in quantities about equal to the total alkalinity of the seawater. The lowered pH results in the conversion of bicarbonates and carbonate ions to CO_2 which is stripped in the degassifier and the deaerator.

* The use of more than 50 stages total has not been proven out in practice nor is it now recommended by the Office of Saline Water.

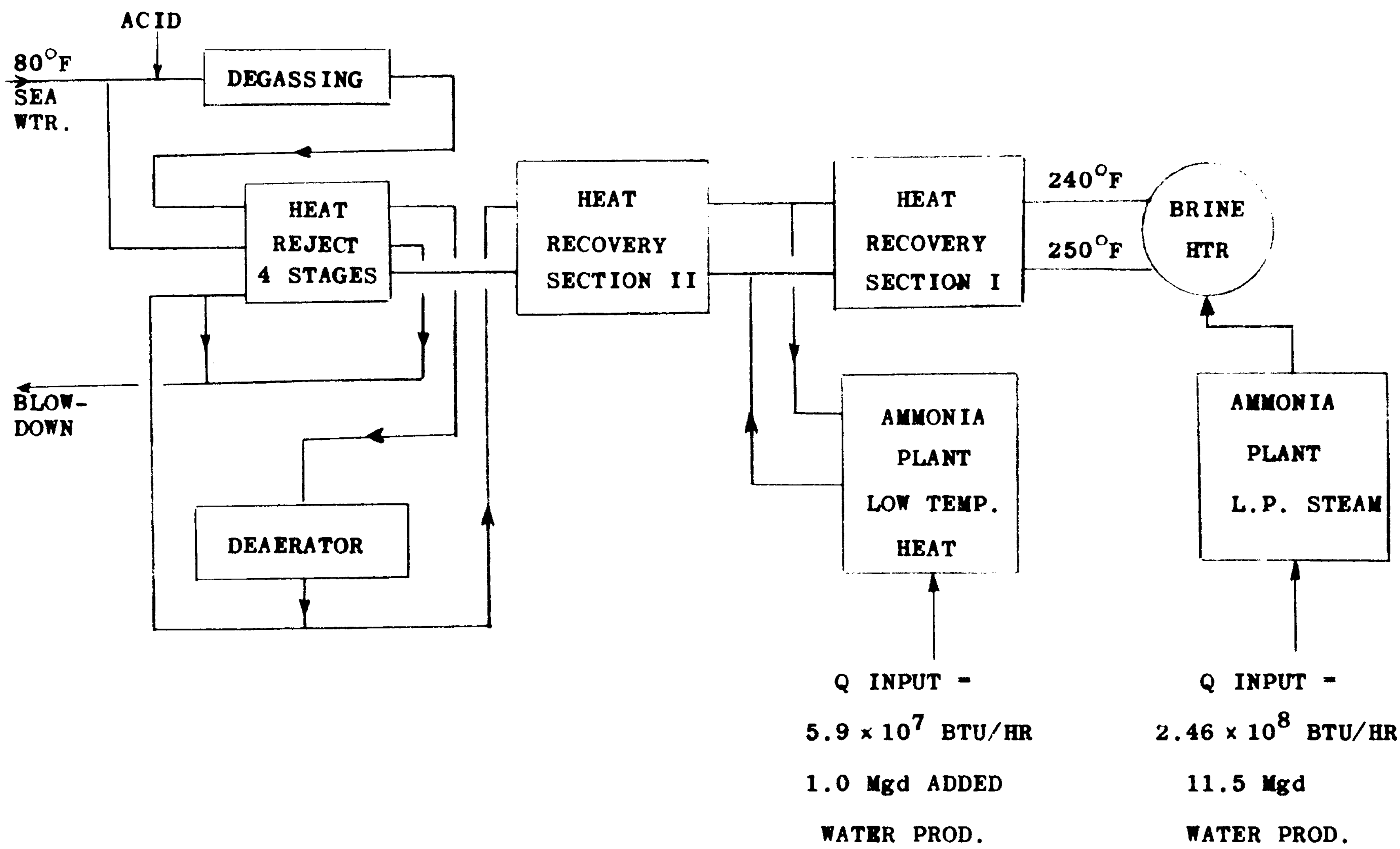
A basic MSF flow sheet integrated with a 500 ton/day ammonia production plant is shown in Figure 1. The MSF plant provides a nominal 11.5 Mgd capacity from 260°F steam sources which requires 2.46×10^8 Btu/hr. Low temperature sources of 5.9×10^7 Btu/hr are used to heat seawater from 132°F to 143°F; this produces an additional 1.0 Mgd of potable water.

An economic evaluation was made to determine the cutoff point of the recovery of waste heat from the various sources of the ammonia process. The following procedure was used: first, the value of steam to the brine heater was assigned; and secondly, the value of heat at the intermediate points of the MSF evaporator was determined from the value of the steam at the brine heater. This value of intermediate steam was prorated on the flashing range that the steam would see in the evaporator. As an example, if 15¢/MBtu steam is assumed at the brine heater at 260°F, the value of steam at 143°F with blowdown temperature of 95°F would be 31 percent of the value at 260°F. This would result in a value of steam of 4.7¢/MBtu. Next, the capital cost of the required equipment changes, including piping and heat exchangers, for injection of this heat into the MSF plant was determined. Assuming a fixed charge rate of 7 percent, the annual cost of recovering the heat was determined and the cutoff point for the recovery of waste heat was thus compared with the value of the steam

Figure 1

FLOW DIAGRAM OF INTEGRATED WATER AND AMMONIA PLANT

500 TONS/DAY AMMONIA, 12.5 Mgd WATER



at that point in the flash plant. This procedure was utilized in all the cases studied.

DISCUSSION OF THE VC-VTE PROCESS

In the basic vapor recompression cycle, the flow sheet employs a gas turbine drive for a vapor compressor which operates across a two-vertical-tube effect evaporator and a multistage flash plant which is primarily used for feed heating. A fired waste heat boiler is used on exhaust gases of the gas turbine which produces steam at approximately 650°F and 260°F. The high pressure steam is used to drive a turbogenerator, producing all of the electricity needed by the process. The steam turbine exhausts into the low pressure header, the total quantity being supplied as additional heat for the vertical-tube evaporator. The vapor compressor receives the low-pressure steam from the second effect and increases its saturation temperature to 256°F. The multistage flash plant is used for feed heating incoming seawater, employing a once-through system without recirculation. Sulfuric acid is injected in the feed stream for scale control before the feed is decarbonated and deaerated.

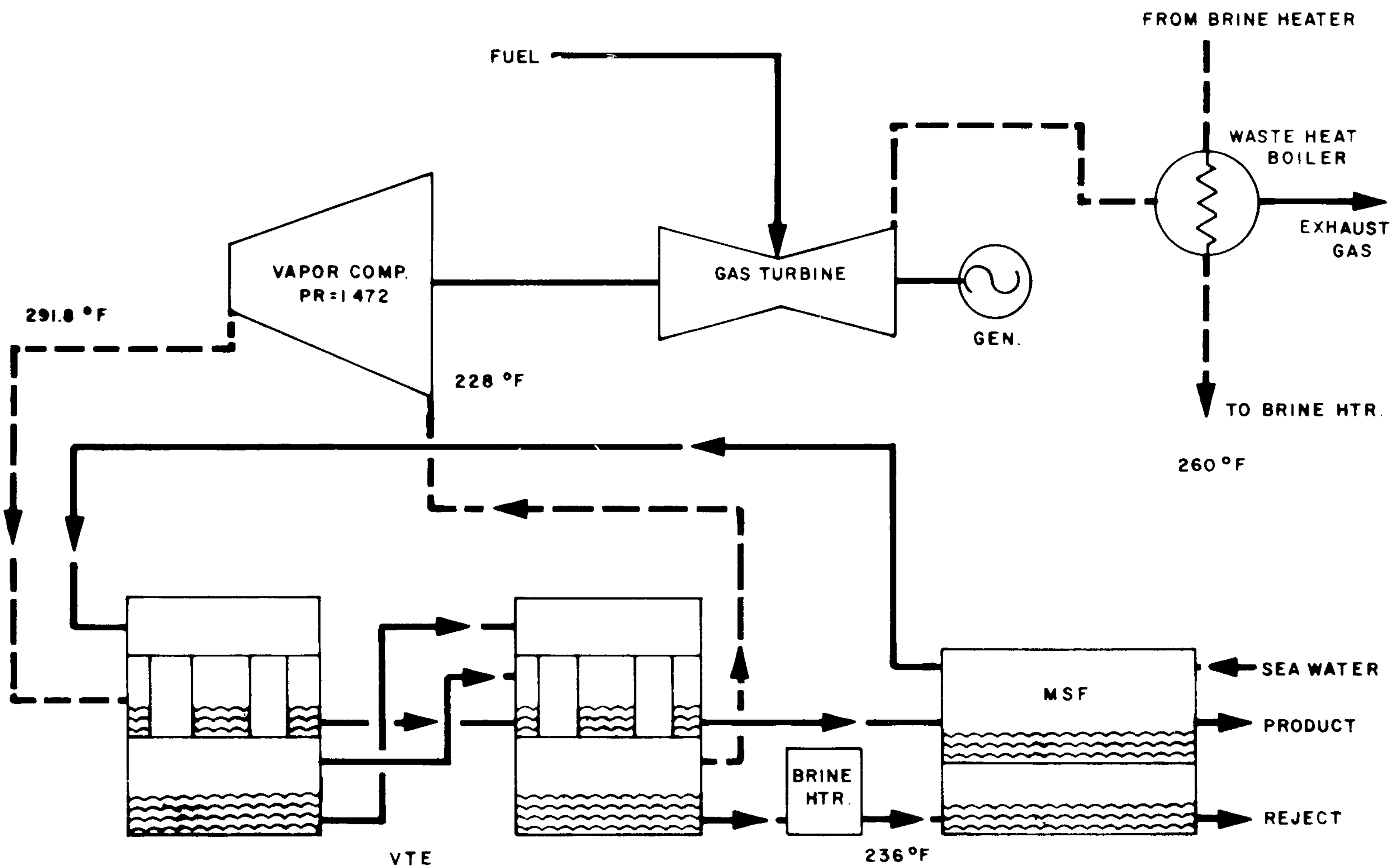
The vertical-tube evaporator consists of 3-inch O.D., double-fluted, 90/10 copper-nickel tubes with an effective length of 12 feet 6 inches. The brine feed to the first effect is spread through distribution nozzles to the inside of the fluted tubes. The steam from the low-pressure boiler and from the vapor compressor outlet is used to heat this incoming brine to the level at which evaporization occurs. Essentially, all of the heat of vaporization from the steam is transferred to the incoming brine, producing an equivalent amount of product from the brine. The reject brine from the first effect is used as feed for the second effect, and the steam generated from the first effect is used for heat in the second effect.

The MSF train is a conventional, commercially available system similar to the Foster Wheeler concept discussed in the MSF portion of this write-up. The waste process heat recovered and used in the MSF portion of the plant was evaluated by the same method as discussed above.

Figure 2 is a schematic diagram of a typical vapor compression system similar to that employed in these studies.

SUMMARY OF RESULTS

The attached tables summarize the results of the studies performed for the combination ammonia and/or power and seawater distillation plants. Table I is a listing of the various Alternates that are referred to in the preceding section of this report and lists the available energy at 260°F and 143°F and the quantities of water that could be



TYPICAL TURBINE POWERED VC SYSTEM

TABLE I. MSF AND VC-VTE PRODUCTION RATES

Alternate	260°F Steam lb/hr	143°F Heat Btu/hr	Water Production: From 260°F Steam Mgd (1)	From 143°F Heat Mgd (2)	VC Mgd	Total Mgd
1	263,000	5.9×10^7	11.5	1.0		12.5
2	432,000	1.27×10^8	19.6	2.0		21.6
3	612,000	1.88×10^8	27.9	2.9		30.8
4	864,000	2.54×10^8	39.3	3.9		43.2
4A	864,000	2.54×10^8	38.7	3.9	57.4	100.0
4B	877,000	2.54×10^8	39.2	3.9	56.9	100.0
5	1,276,000	1.27×10^8	56.3	1.9	41.8	100.0
6	2,262,000	1.27×10^8	99.5	1.9		101.4
7	2,197,000	1.92×10^8	96.8	2.9	79.0	178.7

(1) Performance ratio of 16.25 lbs/1000 Btu

(2) Performance ratio at 5.03 lbs/1000 Btu

produced utilizing the MSF plant with a performance ratio of 16.25 lb per 1000 Btu and a combination of MSF and vapor compression system. Table II is a listing of the annual operating costs of the MSF plants only. These costs do not include the cost of utilities. The fixed charge rate assumed for capital is 7 percent. Table III is a listing of annual costs of the various cases for the vapor compression system without the cost of fuel included. Table IV is a listing of capital costs of both the MSF and the vapor compression plants for each case.

REFERENCES

(1) "Foster Wheeler Corporation Conceptual Design of a 50MGD Desalination Plant," Office of Saline Water, September, 1965.

(2) "Preliminary Design of a Diesel-Powered Vapor-Compression Plant for Evaporation of Seawater," Office of Saline Water Research and Development Progress Report No. 276, August, 1967.

TABLE II. ANNUAL FIXED COSTS OF MSF PLANTS ONLY

(Thousand Dollars)

Alternate No.	1	2	3	4	4A	4B	5	6	7
Capital Charge 7%	802.2	1230.6	1750.0	2258.9	2258.9	2286.2	2847.6	4324.6	4244.8
Operating Labor	126.9	165.9	198.1	234.6	234.6	236.1	274.4	362.2	359.1
Maintenance Labor	56.3	87.9	125.0	161.3	161.3	163.3	203.4	308.9	303.2
Maintenance Materials and Chemicals	123.4	202.9	289.0	391.4	391.4	395.9	517.9	856.7	841.8

Note: Plants require 6 Kwh per 10^6 gallons

TABLE III. ANNUAL VC PLANT COSTS WITHOUT FUEL COSTS

GAS TURBINE DRIVES

Alternate No.	4A	4B	5	7
MGD	57.4	56.9	41.7	79.0
Capital Charges at 7%	$2,646,000	$2,622,000	$2,142,000	$3,360,000
Operating Labor	272,500	271,200	232,400	319,700
Maintenance Labor	189,000	187,300	153,000	240,000
Maintenance materials and chemicals	843,000	835,700	628,700	1,139,000
Fuel Requirement Btu/yr HHV				
Gas Turbine	8.12×10^{12}	8.05×10^{12}	5.91×10^{12}	11.17×10^{12}
Boiler trimming	8.44×10^{12}	8.37×10^{10}	6.14×10^{10}	11.62×10^{10}

Note: Plants generate own electrical power requirements

TABLE IV. CAPITAL COST BREAKDOWN

Case	$\$ \times 10^6$ MSF Capital Cost From 260° F Steam	$\$ \times 10^6$ MSF Capital Cost From 143° Heat	$\$ \times 10^6$ VC Capital Cost	$\$ \times 10^6$ Total
I	10.8	0.45	-	11.25
II	16.0	1.58	-	17.58
III	22.8	2.20	-	25.0
IV	29.5	2.77	-	32.27
IVa	29.5	2.77	37.8	70.07
IVb	29.89	2.77	37.47	70.13
V	39.1	1.58	30.6	71.28
VI	60.2	1.58	-	61.78
VII	58.44	2.20	48.0	108.64

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