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Final Report

Ammonia Production Systems Integration

(22 June 1965—22 February 1966)

AD 484255

22 February 1966

Contract DA-44-009-AMC-1766(X)

For the U. S. Army Engineer Reactors Group,
Army Nuclear Power Program
Fort Belvoir, Virginia

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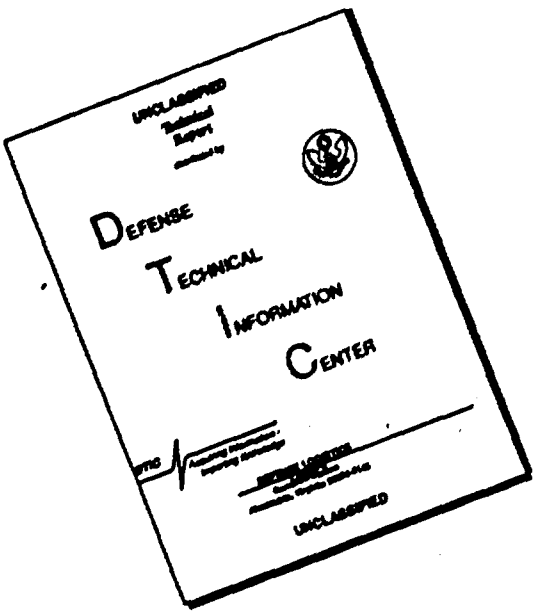
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Final Report

Ammonia Production Systems Integration

(22 June 1965—22 February 1966)

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22 February 1966

Contract DA-44-009-AMC-1166(X)

For the U. S. Army Engineer Reactors Group,
Army Nuclear Power Program
Fort Belvoir, Virginia

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EDR 4496

PREFACE

This document is the final report of the Ammonia Production Systems Integration study contract conducted under U. S. Army Engineer Research and Development Laboratories Contract DA-44-009-AMC-1166(X). The reporting period covered extends from 22 June 1965 to 22 February 1966. The following subcontractors contributed to the study.

Dr. R. L. Harvin Mr. J. P. Bocard Girdler Corporation Louisville, Kentucky	Survey of ammonia synthesis processes and design of the selected ammonia synthesis subsystem
Dr. G. Lund Gas Equipment Engineering Corporation Milford, Connecticut	Survey of air fractionation processes and design of the selected nitrogen generation subsystem.

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I. SUMMARY

Conceptual designs of 11 mobile ammonia fuel production systems were prepared by Allison Division of General Motors under U.S. Army Engineer Research and Development Laboratories Contract No. DA-44-009-AMC-1166(X). These systems represent various combinations of total electric power input to the system, mobility class, electrical power frequency, and cooling medium.

This report presents a description of each of the 11 systems including weight, cost, and performance characteristics. Summary data for each of the 11 systems are presented in Table 1-I. These systems were designed for an ammonia production capability optimized primarily on the basis of system power input and cost. Each subsystem was studied individually with several alternate processes considered. Efficiency was sacrificed in the higher power range to achieve minimum system cost. The significant results of each subsystem study are as follows.

- Ammonia Synthesis—The process selected used an ejector driven ammonia refrigeration unit with minimum power requirements. The optimum ammonia loop pressure was 6000 psia.
- Nitrogen Generation—The process selected requires a minimum amount of power by utilizing the expansion of the electrolyzer oxygen as the source of low temperature refrigeration. The final product nitrogen purity was 99.83%.
- Hydrogen Production—Use of the basic module developed under Contract DA-44-009-AMC-1021(X) resulted in subsystem designs with operating power requirements from 19.9 to 23.2 kwh/lb H_2. Most of the other required components can be procured with a minimum of development effort.
- Water Purification—The vapor compression process was selected because it is well developed and flexible enough to handle the entire range of feedwater purity without scaling.
- Power Conditioning—Main transformers with relatively high efficiencies were selected. Silicon control rectifiers were selected for rectification, on-off power control, and d-c voltage regulation.
- Control—The pneumatically operated system which was selected will provide the plant coordination and control necessary for automatic operation. The control scheme will also maximize plant production rates for the total power available and operating parameter restrictions.

The system integration and packaging studies combined the several subsystems into a minimum number of separate packages (3 to 5) consistent with optimization studies. All packages are within the specified size and weight restraints. The performance of the integrated systems was analyzed and the detailed results are presented.

Table 1-I.
Ammonia production system summary data.

System	Mobility class	Power input (Mwe)	Ammonia output (lb NH_3/hr)	Conversion efficiency (%)	System weight (lb)	No. of packages	Production cost* ($)	Development cost ($)
1	I	3	729	53.5	98,300	3	1,005,000	3,600,000
2**	I	3	730	53.5	95,400	3	1,010,000	3,600,000
3	II	3	758	55.5	118,400	3	1,270,000	4,000,000
4	I	6	1380	50.5	155,000	5	1,505,000	4,500,000
5	II	6	1410	51.5	166,000	3	1,535,000	4,600,000
6**	II	6	1420	52	160,000	3	1,515,000	4,600,000
7***	II	6	1410	51.5	146,000	3	1,500,000	4,800,000
8	III	6	1410	51.5	167,000	3	1,535,000	4,600,000
9	II	10	2290	50.5	264,000	5	2,065,000	5,400,000
10	III	10	2290	50.5	259,000	4	2,060,000	5,400,000
11	III	15	3300	48.5	360,000	5	2,575,000	6,700,000

*Cost on basis of 100 units/yr production rate.

**Systems 2 and 6 are water-cooled—other systems are air-cooled.

***System 7 power input frequency is 400 cps—other systems are 60 cps.

Allison

II. INTRODUCTION

The Allison Division of General Motors has prepared conceptual designs of mobile ammonia fuel production systems optimized to meet mobility input power and other requirements specified by the U.S. Army Engineer Reseat and Development Laboratories under Contract No. DA-44-009-AMC-1166(X). These syste consist of the following subsystems:

- Ammonia synthesis
- Nitrogen generation
- Hydrogen production
- Water purification
- Power conditioning
- Control

A simplified ammonia fuel production system schematic is shown in Figure 2-1.

The system designs maximize the yield of ammonia fuel for a given electrical power input consistent with mobility restraints, and other considerations such as reliability, maintainability, response time, and cost. These designs are based on the following criteria except where specific perturbations were requested.

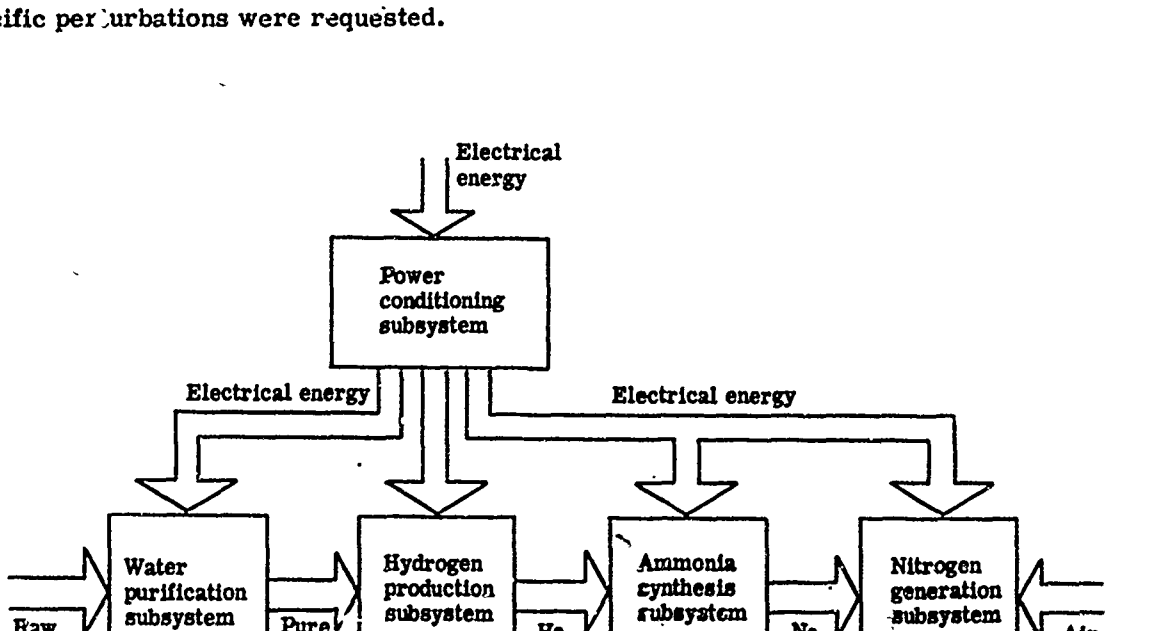


Figure 2-1. Ammonia fuel production schematic.

- The electrical power input is 4160 v, 3 phase, 60 cps.
- Source materials are air and water.
- The systems are capable of operating in all climates from sea level to 10,000 ft. The design point and off-design conditions are indicated in Table 2-I.
- The systems consist of a number of separable, skid-mounted packages capable of being transported by road, sea, or air. Package size and weight limitations for three different mobility levels are indicated in Table 2-II.
- The systems are designed to minimize response time, i.e., system shutdown, disassembly of packages for transport, assembly of separate packages after relocation, and start-up of system. Plant service conditions are defined in Table 2-III.
- With the exception of the electrolyzer modules in the hydrogen production subsystem, the system designs are based on state-of-the-art technology.

Table 2-I.
Feed and product stream conditions.

Design point	
Air	77°F, 14.7 psia, and 50% relative humidity
Water*	77°F
Product	liquid ammonia at 77°F and 146 psia
Off-design conditions	
Air	125°F max, -65°F min, 14.7 psia max, and 10 psia min
Water	105°F max and 32°F min

*Available for cooling purposes in Systems 2 and 6 (Figure 2-1).

Table 2-II.
Mobility class definition package size and weight limitations.

Class	Length (ft)	Width (ft)	Height (ft)	Weight (lb)
I	25	8	8	35,000
II	30	10	8	65,000
III	35	12	8	100,000

Table 2-III.
Plant service pattern.

Mobility class	Min operating time between moves (hr)	Average operating time between moves (hr)	Average transit time (hr)
I	24	48	8
II	72	144	24
III	144	288	48

Eleven fuel production system designs are presented herein. These designs represent various combinations of electrical power input, mobility class, electrical power frequency, and cooling medium. The combination for each system is indicated in Figure 2-2. Ammonia fuel production capacity; system cost, weight, size, and packaging arrangement; response time; and control characteristics have been determined for each system and are presented in Section III.

To provide a basis for these integrated system designs, each subsystem was studied individually. Several alternate processes were considered for each subsystem. Preliminary subsystem designs were prepared for each subsystem based on the product rates presented in Table 2-IV.

Table 2-IV.
Preliminary subsystem design-product rate.

Subsystem	
Ammonia synthesis	600, 1200, 2000, and 3000 lb NH_3/hr
Nitrogen generation	500, 1000, 1670, and 2500 lb N_2/hr
Water purification	125 to 700 gals H_2O/hr

The results of the initial studies and the preliminary subsystem designs are presented herein. Integration of the individual subsystems into eleven optimized system designs required modification of the product rates and preliminary designs. The cost and weight optimization studies, which include the nuclear powerplant cost and weight, are presented in the classified Supplement to this report.

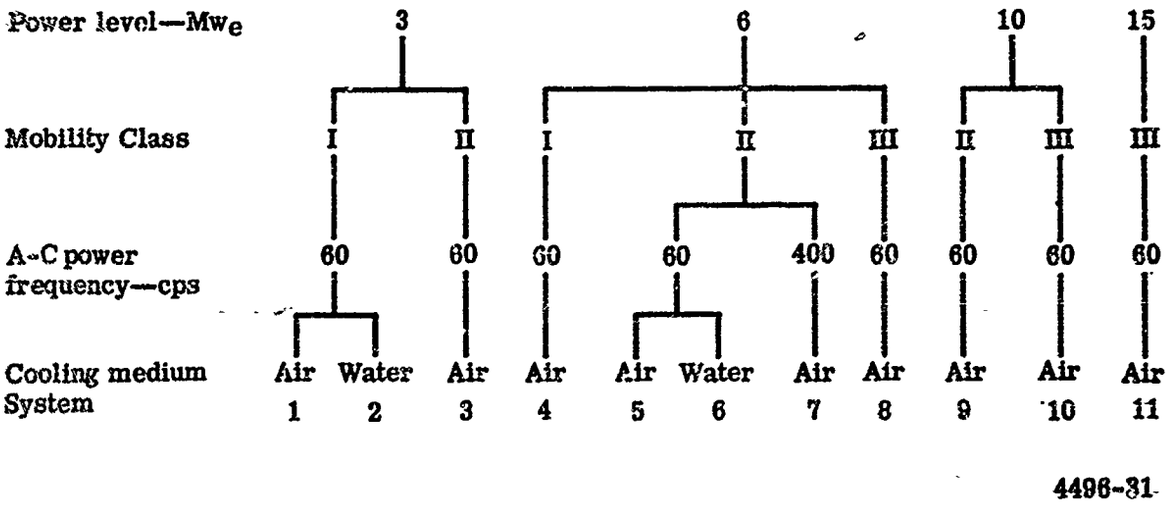


Figure 2-2. Ammonia production systems definition.

III. AREAS OF INVESTIGATION

Initially, in the Ammonia Production Systems Integration study, conceptual designs of each of the subsystems were prepared at specified ratings or outputs. Subsequently, these subsystem designs were integrated and optimized to meet the several requirements of eleven different ammonia fuel production system conceptual designs. Information on the subsystem and system designs is presented herein. Cited references are given at the end of each subsection.

Ammonia Synthesis Subsystem

DESCRIPTION

Ammonia is produced by the chemical combination of hydrogen and nitrogen. The reaction is generally carried out at elevated pressure (1500 to 15,000 psia) and temperature (700 to 1100°F) over a catalyst.

Figure 3.1-1 shows a typical ammonia synthesis process. The hydrogen-nitrogen mixture is compressed, mixed with unreacted recycle gas, and introduced into the synthesis loop. The recycle compressor used to bring the gas back up to system pressure may be either a separate compressor or an additional cylinder on the frame of the main synthesis gas compressor. The gas mixture is then passed through a filter for removal of compressor oil and other contaminants. The gas stream is then cooled in a heat exchanger and an ammonia-refrigerated condenser—the gaseous ammonia contained in the recycle gas is condensed and additional impurities not caught in the filter are washed out of the gas by the condensing ammonia. A standard ammonia refrigeration system is used to cool the refrigerated condenser. The compressor for this system may be either a separate unit or added to the frame of the main synthesis gas compressor.

The gas separated from the liquid ammonia in the separator at the base of the heat exchanger flows back through the heat exchanger to precool the gas entering the refrigerated condenser. The gas leaving the heat exchanger then enters the ammonia converter.

Liquid ammonia from the separator flows to the let-down drum where its pressure is reduced and dissolved gases are permitted to flash off. These gases may be recovered and in some cases added to the feed gas. Liquid ammonia from the let-down drum flows to storage.

The reaction of nitrogen and hydrogen takes place in the ammonia converter. Since the reaction is not complete in one pass, the unreacted synthesis gas must be recycled through the process. Before passing over the catalyst, the gas entering the converter is heated to reaction temperature by heat exchange with the reacted gas. The partially cooled and reacted gases from the converter flow to the condenser where additional liquid ammonia is condensed. The condensed ammonia is separated from the recycle gas and combined with the ammonia from the other separator.

A provision is made to prevent the inert gas in the feed stream from building up in the loop. Some of the inert gas is soluble in the product and is eliminated from the loop in this manner. However, if sufficient inert gas is not dissolved in the liquid ammonia to maintain the proper inert gas concentration in the system, a small gas purge is employed. The remaining gas is recycled through the process loop.

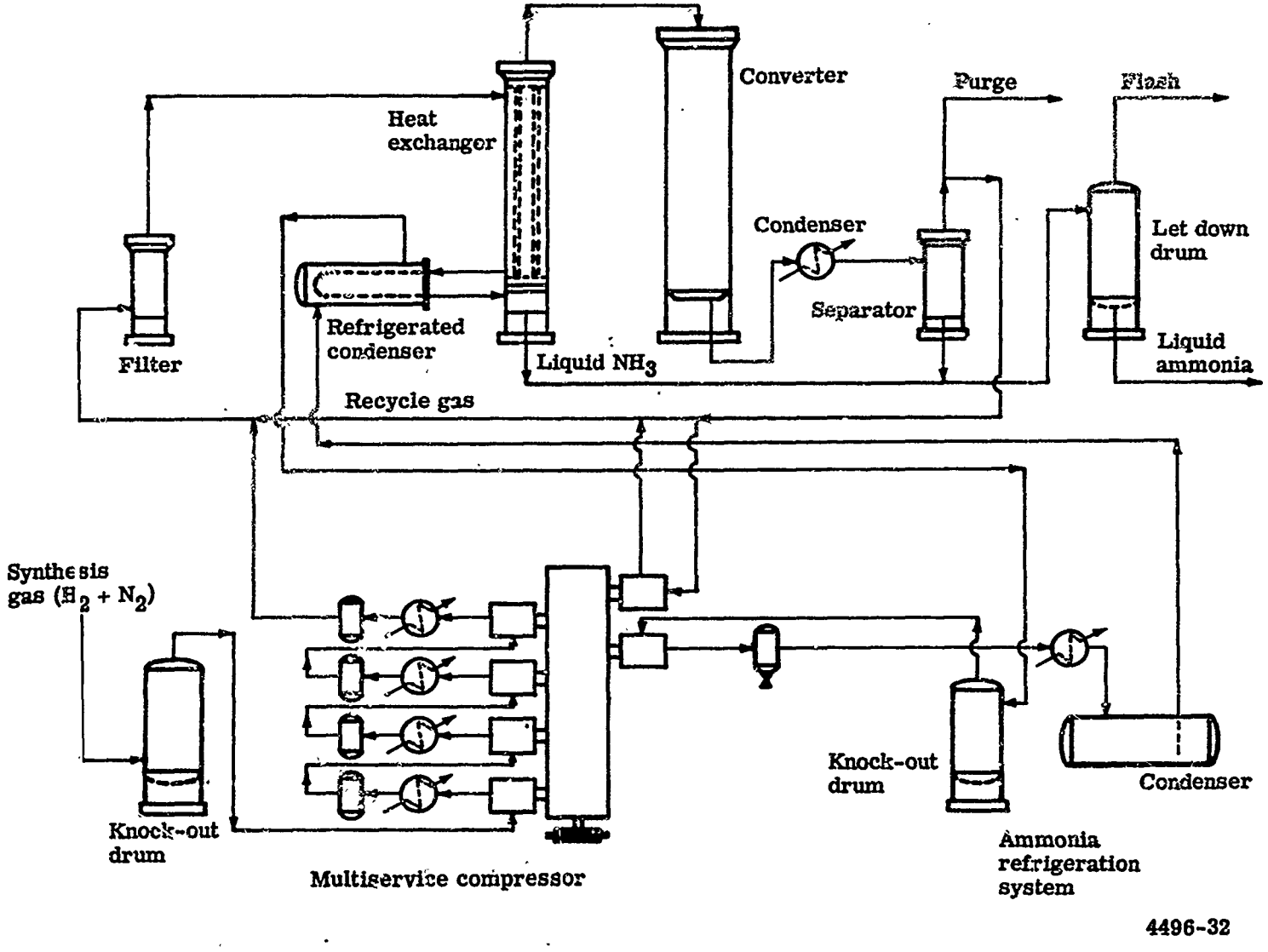


Figure 3.1-1. Typical ammonia synthesis process.

INVESTIGATION

A number of processes were evaluated for the synthesis of ammonia. The processes differed primarily in the operating pressure and temperature levels. During the process evaluation, major emphasis was placed on the power requirements and the applicability of the process for Energy Depot application. The catalyst performance was also considered with variations in loop pressure, catalyst temperature, space velocity, recycle stream hydrogen-nitrogen ratios, and inert concentrations in the loop. The effects of feed gas purity on the subsystem operation and equipment were considered. The investigation included the selection of equipment for heat recovery and the compression of synthesis gas.

Process Comparison

The processes investigated range in operating pressures from 1500 to 15,000 psi. A brief comparison of the various commercial processes is contained in Table 3.1-I, and discussions of the important features of each process follow.

Table 3.1-I.
Ammonia synthesis processes.

Process	Pressure (atm)	Temperature (°F)	Conversion (%)
Casale	500 - 900	932	15 - 18
Claude	900 - 1000	932 - 1200	40 - 85
Fauser	200	932	12 - 23
Haber-Bosch	200 - 350	1022	8
Modified Haber-Bosch	200 - 300	932	20 - 22
Mont Cenis	100	750 - 800	9 - 20

The notable feature of the Casale process is the ability to liquefy the ammonia by water cooling alone. In the higher pressure cases, this can be accomplished by air cooling. Thus, no refrigeration is needed. The refrigeration power saving, however, is offset by higher synthesis gas compression requirements. Excessive catalyst heating is controlled with 2 or 3% ammonia in the inlet gas to the converter and by proper design of the converter.

The Claude process is characterized by the very high loop pressure of 900 to 1000 atm. It is a one pass process with the unconverted gas, approximately 20% of the feed, being wasted as exhaust to the atmosphere. A special converter design is required to dissipate the heat. The process results in greater compactness, simplicity, and ease of construction; elimination of expensive heat exchangers; and removal of ammonia by water or air cooling. The high pressure operation, however, results in decreased equipment life and increased apparatus maintenance.

The Fauser process is primarily a standard synthesis loop process employing moderate pressures, refrigeration, and recirculation. Its chief differences from the standard process are the use of hydrogen from electrolytic cells and nitrogen from air liquefaction followed by deoxidation over a copper catalyst.

The Haber-Bosch process and the similar Modified Haber-Bosch process are the standard processes used in modern ammonia synthesis technology. The fresh feed is added to the loop prior to the secondary or refrigerated condenser. The condensed ammonia is separated from the vapor which is heated by heat exchange before passing through the converter. The exit gas after cooling is recirculated by a compressor which supplies the pressure necessary for gas flow through the loop. The pressure of operation is moderate which increases the component reliability, reduces maintenance, and increases the equipment life. Refrigeration is necessary to increase the conversion.

The Mont Cenis process makes use of the lowest loop pressures of commercial importance. A more active catalyst, iron cyanide, is used in this process; promoted iron catalysts are used in the other processes. To obtain a moderate degree of conversion at the low loop pressures, it is necessary to refrigerate the synthesis gas stream to a very cold temperature to condense the ammonia and reduce the ammonia concentration. The temperature of reaction is 100 to 200°F lower than the other processes. At these lower pressures, the tendency for run-away temperatures, which would deactivate the catalyst, is reduced. At higher loop pressures, the tendency is pronounced and, because of the higher catalyst activity, temperature control becomes a major problem. To control temperature at higher operating pressures, sophisticated methods of heat exchange are required which greatly complicate the converter internals. For this reason, commercial use of iron cyanide catalyst in processes which operate at higher loop pressures than the Mont Cenis process has not been realized.

In conclusion, the modified Haber-Bosch ammonia synthesis process appears most suitable for the Energy Depot application. The process utilizes moderate pressures and temperatures and minimizes hydrogen loss. The final selection of an ammonia synthesis process will depend on further consideration of loop operating pressure and process suitability.

Operating Pressure Evaluation

Commercial experience has shown that catalyst life is dependent on the operation pressure. The thermodynamic tendency toward greater reaction at the higher pressures causes hot spots in the catalyst bed which result in the destruction of catalyst activity. The catalyst life may be shortened by a factor of 10 by increasing the pressure from 5000 to 15,000 psi. At the lower pressures where the temperature can be more easily controlled, additional catalyst can be placed in the converter to lengthen the time interval between catalyst changes.

Selection of an operating pressure level becomes quite complex when the basic cases previously described and the many modifications possible to these processes are considered. For a given set of conditions, a process can be tailor-made to give the desired characteristics. In this study, major emphasis was placed on minimizing power consumption. For this reason, the power consumption for ammonia synthesis was calculated for various loop pressures ranging from about 1900 to 15,000 psi. The calculations were based on general data concerning the various processes which are available in the literature. The synthesis gas was assumed to be available at 300 psi. The power requirements for synthesis gas compression to loop pressure, recirculation of the loop gases, and the refrigeration requirements necessary to reduce the ammonia concentration to the converter inlet conditions were calculated. The results are presented in Figure 3.1-2 where the theoretical horsepower per ton per day of ammonia produced is plotted against loop operating pressure.

As can be seen in Figure 3.1-2, the power for synthesis gas compression increases as loop pressure is increased from 300 psi. The recycle power requirement and the refrigeration power rise quite rapidly at the lower loop pressures. The increased recycle power requirement is the result of greater synthesis gas recycle required because of the lower ammonia conversion and equilibrium concentrations at the lower loop pressures. The increased refrigeration requirement results from the need for lower temperatures to reduce the ammonia

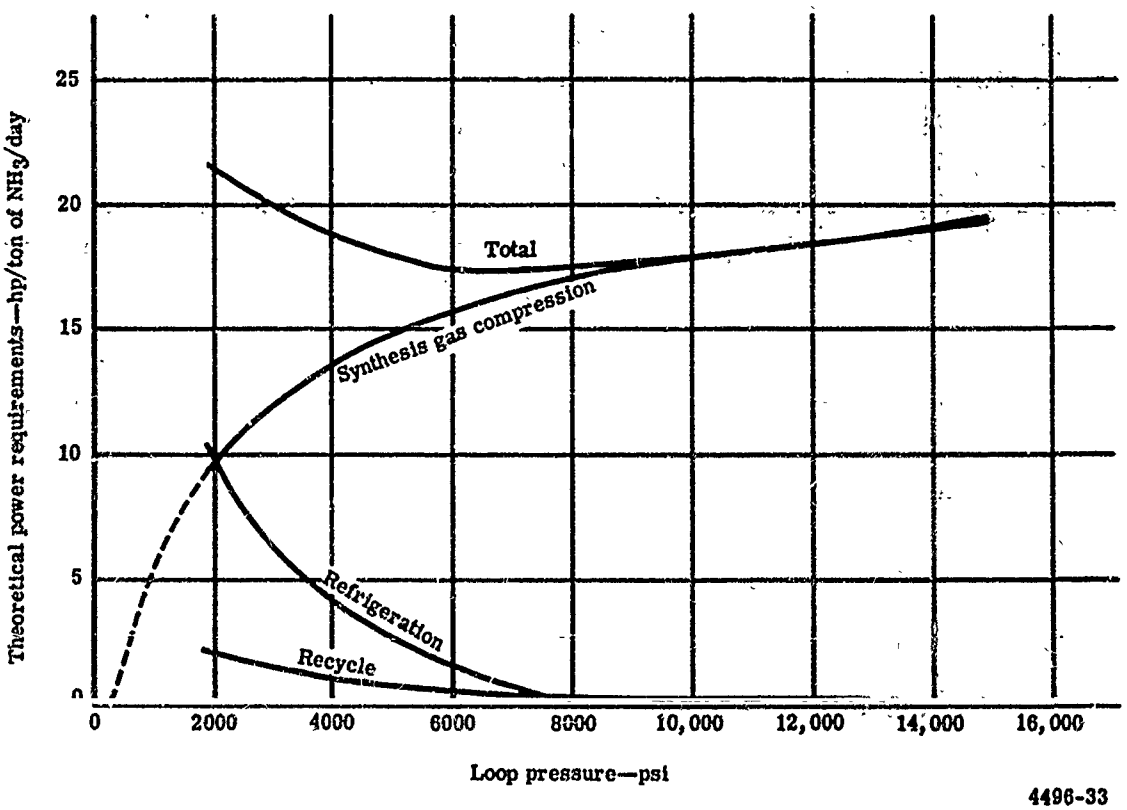


Figure 3.1-2. Ammonia synthesis subsystem power requirements.

concentration in the recycle gas and from the larger recycle stream which requires refrigeration at the lower loop pressures. The total power requirement is at a minimum and fairly constant for loop pressures between 5000 and 10,000 psi. Thus, the selection of the operating pressure for the ammonia synthesis subsystem will be governed by factors other than power requirements.

Process Selection

In the selection of the synthesis process, it was decided that a process with mechanical refrigeration would not be desirable. If mechanical refrigeration were included, it would complicate operation, lengthen start-up time, reduce reliability, and possibly add to the overall weight. Referring to Figure 3.1-2, it can be seen that at 8000 psi no refrigeration is needed; the ammonia concentration can be sufficiently reduced by water or air-cooling.

The use of an ejector to aspirate a refrigerated secondary condenser allows loop operation at a lower pressure level without mechanical refrigeration. By taking the product liquid ammonia at high pressure, evaporating it to provide cooling for the ammonia synthesis converter, and expanding it as the motive fluid in an ejector to a lower pressure, a significant amount of refrigeration can be produced. Estimates indicated that refrigeration equivalent to approximately 1.5 thp/ton/day could be achieved by using this scheme. A review of Figure 3.1-2 showed that by reducing the loop pressure to 6000 psi the refrigeration requirements could be met by the ejector-refrigerant scheme. This scheme would result in the addition of an ejector which has no moving parts and a high degree of reliability. The operation of the unit would be extremely simple. Start-up and shutdown time would not be affected. There would be no increase in hydrogen loss and there would be no particular problems in operating at off-design conditions. Thus, it was decided to proceed with a design loop pressure of 6000 psi to take advantage of the power savings afforded by the ejector scheme.

An ammonia synthesis process similar to the Modified Haber-Bosch process was, therefore, selected for the Energy Depot application. The selected process differs from the Modified Haber-Bosch process in two respects: first, the loop operating pressure is slightly higher and second, an ejector is used to drive the refrigeration system.

Effect of Oxygen Impurity in Feed Gas

The performance of the ammonia synthesis loop under a given set of conditions can be altered quite significantly by the presence of various impurities in the feed gas. Oxygen, when present in the loop even in small quantities, can cause deterioration of the catalyst. Since the hydrogen is produced by electrolysis and the nitrogen is available from air liquefaction, oxygen can be expected in the feed gases. For a suitable catalyst life, this oxygen must be removed before the gas stream enters the loop.

A study was made to determine the effect of oxygen content on the size of the deoxidation equipment. The proposed deoxo unit operating conditions were a temperature of 250°F or higher, a pressure of 700 to 1000 psi, and 0.75 mole % oxygen in the feed gas stream. Information available from a catalyst vendor indicates that space velocities to 10,000 ft^3 of gas/hr-ft^3 of catalyst can be safely used under the proposed conditions.

The deoxidation reaction consists of combining the oxygen with hydrogen. Since both feed streams could conceivably contain oxygen, treatment was proposed after stream combination. To attain the elevated temperature without using heat transfer equipment, it was necessary to pass through one stage of feed gas compression. At these design conditions, it was estimated that the deoxo catalyst could reduce an oxygen content of 0.75 mole % (equivalent to 3% oxygen content in the nitrogen feed gas) to about 1 ppm. No correlation of space velocity and inlet oxygen content was available below 0.75 mole % oxygen. Because of this, no weight savings can be assumed for feed gases containing less than 0.75 mole % oxygen unless no oxygen at all is contained in the feed gas. At oxygen concentrations greater than 0.75 mole %, it can be assumed that space velocities less than 10,000 ft^3 of gas/hr-ft^3 of catalyst must be used resulting in heavier deoxo units. An estimate of the variation in deoxo unit weight for higher feed gas oxygen concentrations is shown in Figure 3.1-3. Since the deoxo unit represents only about 1% of the ammonia subsystem weight, an increased concentration of oxygen in the feed gas will have little effect on the subsystem weight.

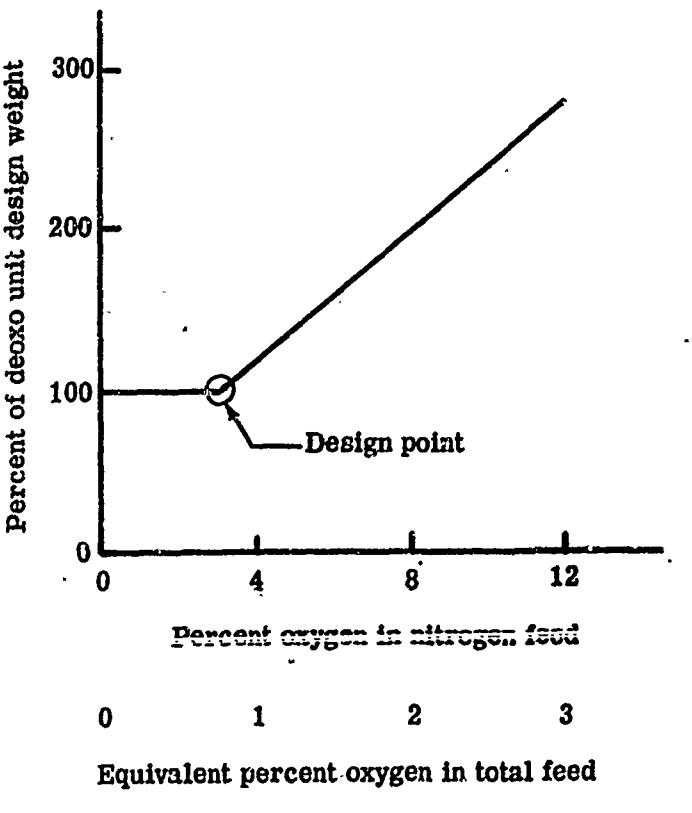


Figure 3.1-3. Ammonia synthesis subsystem deoxo unit weight variation.

Study of Loop Inert Concentrations

Another feed gas impurity that can be expected is the inert gas argon, since the nitrogen is obtained from air liquefaction. This gas does not have a detrimental effect on the catalyst, but rather reduces the overall reaction through a dilution effect. If the concentration of inerts in the feed gas is high, the loop inert concentration will build up and reduce the performance.

Increased inert concentrations necessarily cause greater amounts of gases to be recycled if production is to be maintained. In commercial applications the upper level of inerts tolerated is approximately 20 mole %. Since the inerts do not enter into the reaction and their presence is detrimental, they must be removed from the loop.

Inert removal is accomplished in two ways: by purging some of the loop gas and through solubility in the condensed ammonia. In commercial plants where the feed gas can contain inert concentrations up to several percent, solubility in the liquid ammonia product is insufficient for proper control. Thus, high pressure purging of the recycle stream is required. Hydrogen, nitrogen, and ammonia (as well as the inerts) are lost from the loop. Where the power cost for hydrogen is high, as it is in this application, purging must be held to a minimum. Since the feed gas in this application is relatively pure, the inert level in the loop can be quite easily controlled by solubility alone. However, hydrogen and nitrogen are also soluble in the liquid ammonia and are, therefore, lost with the inert. It was found that the ratio of hydrogen-to-inert soluble in the liquid ammonia was dependent on the loop inert level. A study was made to determine this effect for inert levels in the loop to 22 mole % in the recycle stream.. The hydrogen-to-inert ratio varied from 50 at 1% inerts to 2 at 20.5%.

Figure 3.1-4 contains the results of the study in graphical form. Calculations showed that the hydrogen inert ratios in purge streams are up to 20% higher than the ratios in the liquid, indicating the desirability of reducing inert levels by solubility. An inert level of 4.55 mole % was selected as the process design point.

It should be pointed out that although the ratio of hydrogen-to-inert is high in the loop at low inert levels in the recycle stream the actual amount of inert that must leave the loop should be no greater than the amount entering in the fresh feed. By proper recycling of the flash gas from the various pressure let-down points, the loop inert level can be regulated and the hydrogen loss minimized.

Caustic Removal

Since the hydrogen production subsystem (Section 3.3) employs a caustic electrolyte, it is possible to have some caustic carry-over in the hydrogen feed stream to the ammonia subsystem. Commercial experience has indicated that caustic carry-over can cause a problem by

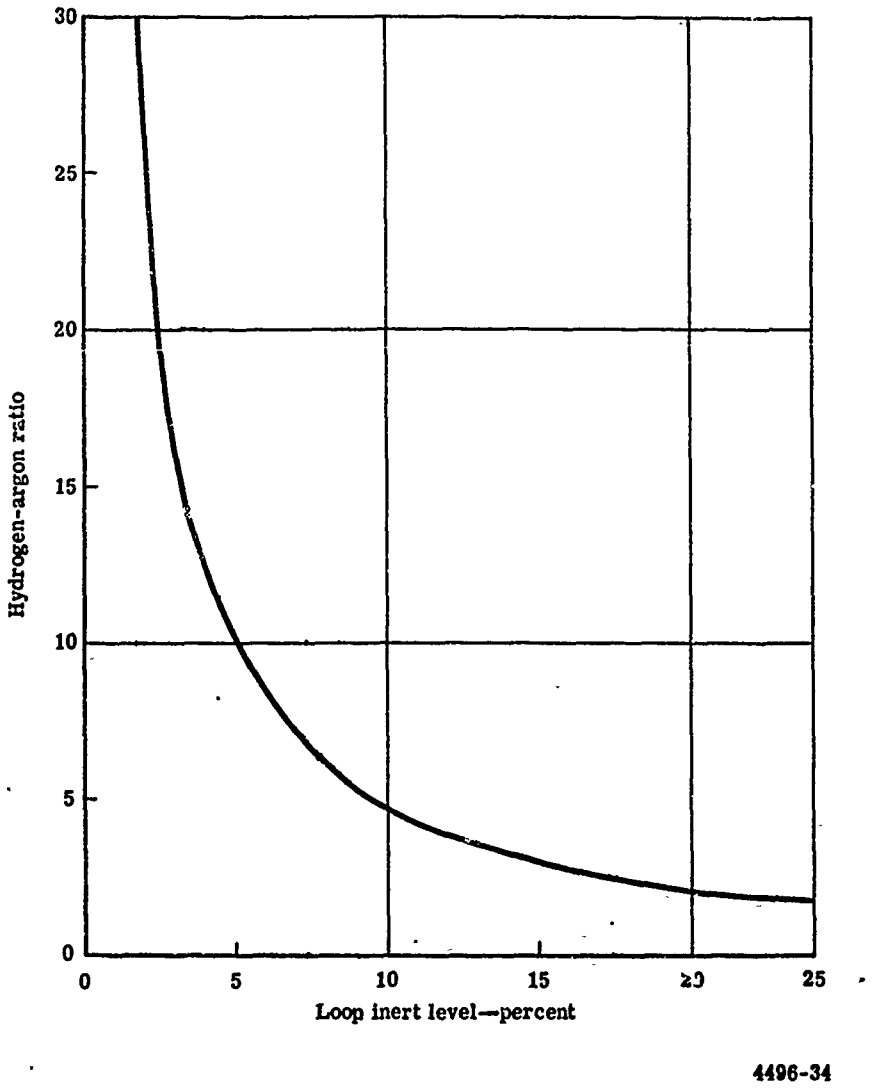


Figure 3.1-4. Solubility ratio of hydrogen-to-argon in liquid ammonia.

depositing caustic in the synthesis gas compressor if it is not removed before compression. This is not expected to present a problem in the present design for two reasons. First, the hydrogen subsystem separator-reservoirs are provided with demisters to minimize entrainment in the product streams. Second, the hydrogen stream is cooled before being mixed with the nitrogen stream for the synthesis gas compression. Experience[1] indicates that cooling of the water-saturated hydrogen stream partially condenses the contained water and scrubs out the caustic. In the proposed ammonia subsystems, it is expected that the hydrogen precooler and associated knock-out drum will remove any caustic carry-over which passes through the demister. However, should operating experience indicate that the proposed scheme is not adequate, a scrubber may be added in which the feedwater-to-hydrogen subsystem would be used to scrub the hydrogen stream.

Compressor Study

An important piece of equipment for the ammonia synthesis subsystem is the multiservice compressor. This compressor must be capable of compressing the fresh feed gas from its supply pressure to that of the synthesis loop. It must also provide the necessary amount of recirculation. This type of service is fairly common for water-cooled compressors of commercial capacities. It is more difficult to find air-cooled equipment for this duty particularly when the units must be rugged, lightweight, and easily transportable.

A number of compressor manufacturers were contacted. An air-cooled multiservice compressor offered by Joy Manufacturing Company was selected as the basic unit. This compressor was selected because it had the highest capacity of any of the air-cooled compressors offered for the required 6000 psi discharge pressure.

The maximum capability of a single Joy air-cooled compressor is 250 hp. By connecting two compressors to a double-ended motor, the total power capability can be increased up to 500 hp for a single unit. Multiple units are necessary when the power requirement exceeds 500 hp.

Commercially available water-cooled compressors with a 300 bhp capability are too large in dimension and too heavy for the applicable mobility classifications. Joy Manufacturing Company representatives said that for water-cooled systems the lighter weight air-cooled designs could be modified by the addition of water jackets in place of the cooling fins on the cylinders. This would result in a water-cooled unit with essentially no increase in weight over the air-cooled compressor plus coolers. In addition, as much as a 3% reduction in power consumption could accrue because of better heat rejection with water-cooling.

In the separate subsystem investigations, the nitrogen compression to 315 psia was included in the multiservice compressor in the ammonia synthesis subsystem and also in a separate compressor in the nitrogen generation subsystem (Section 3.2). During the Integration and Packaging phase of the contract, the best location was selected for the nitrogen compression service in each of the eleven system of interest.

The variation in brake horsepower for compressing N_2 with feed gas pressures from 15 to 315 psia (the synthesis gas compressor suction pressure) is shown in Figure 3.1-5. On the basis of this information and the studies reported in the Nitrogen Subsystem section, an optimum nitrogen subsystem process was selected which had a nitrogen stream pressure of 17 psia. This pressure was used as the basis for subsequent design and selection of compressors for the various ammonia subsystems.

Either a recycle compressor or an ejector is used in commercial installations to recycle the synthesis gas stream. The recycle compressor mode was selected for the Energy Depot

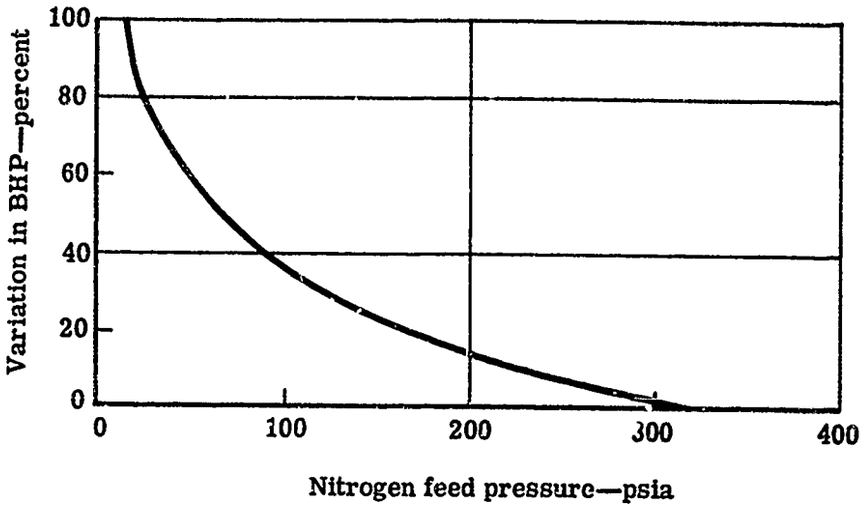


Figure 3.1-5. Compressor power vs nitrogen suction pressure.

systems because the pressure drop around the loop is fairly high (200 psi); the recycle stream is much larger than the feed stream which would be the motive stream for the ejector; and the recycle compressor power requirement is fairly small (Figure 3.1-2).

Catalyst Performance Studies

Ammonia synthesis catalyst performance data were readily available in published information for loop pressures up to about 5000 psi but were lacking for higher pressures. However, a Girdler computer program incorporating the basic thermodynamic and kinetic data was available by which various correlations of the parameters could be generated. The computer program was run to determine the change in expected ammonia concentration with changes of temperature, pressure, space velocity, inert concentration, inlet ammonia concentration, and hydrogen-nitrogen ratio. The data were used to prepare graphs showing the individual effects of the parameters on the catalyst performance. Table 3.1-II contains the values assigned to the various parameters while making the studies.

Table 3.1-II.
Catalyst performance study parametric values.

Subsystem loop operating pressure, psi (atm)	6,000 (400)
Effective catalyst temperature, °F	932
Space velocity, ft^3 of gas/hr-ft^3 of catalyst	80,500
Inlet NH_3 concentration, mole %	5.5
Inlet inert concentration, mole %	4.55
Inlet gas H_2 - N_2 ratio	3.0

The results of the study are shown in Figures 3.1-6 through 3.1-11.

It should be noted that in the case of the 6000 psi loop the effect of space velocity was determined at five different temperatures. From these data, it is possible to determine the effects of additional catalyst, off-design operation, etc.

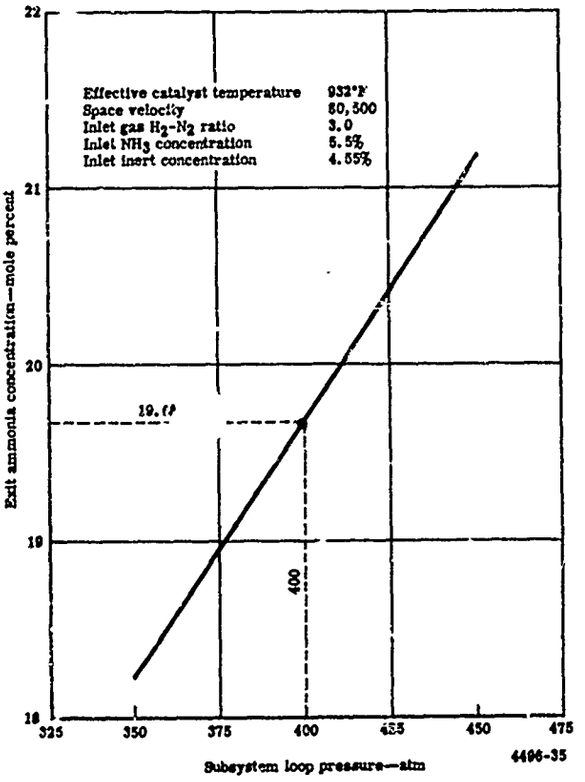


Figure 3.1-6. Catalyst performance with pressure.

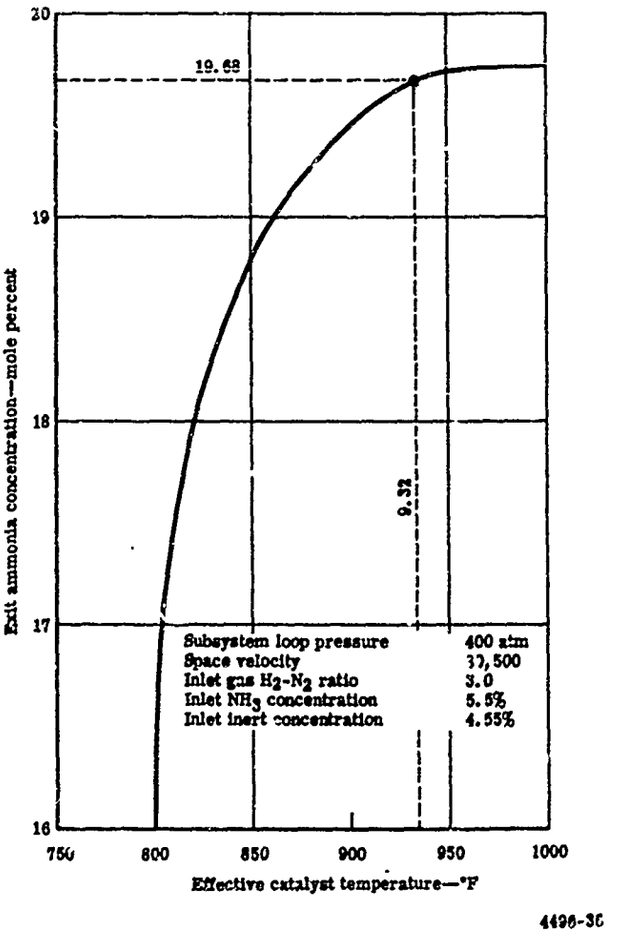


Figure 3.1-7. Catalyst performance with temperature.

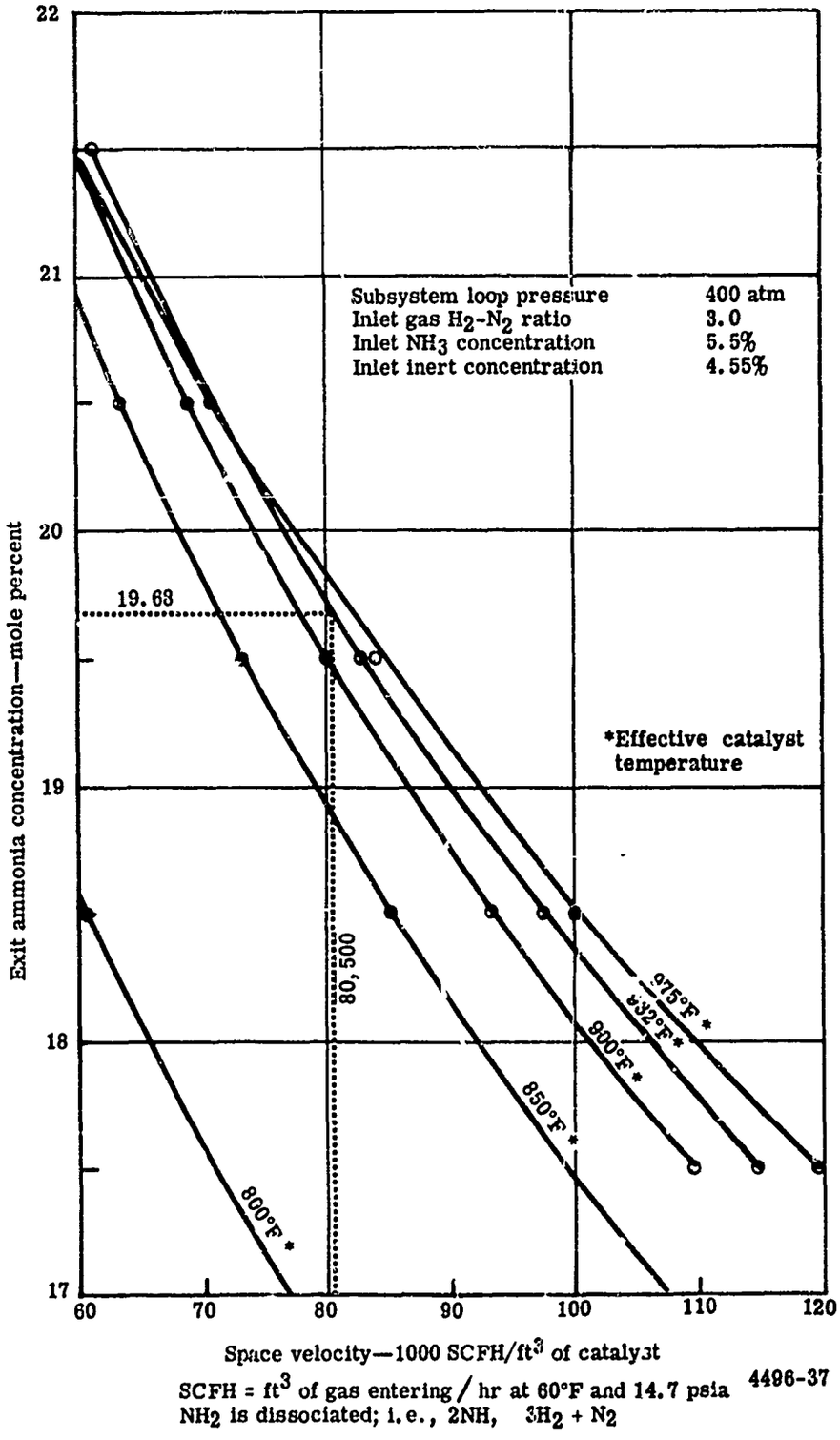


Figure 3.1-8. Catalyst performance with space velocity.

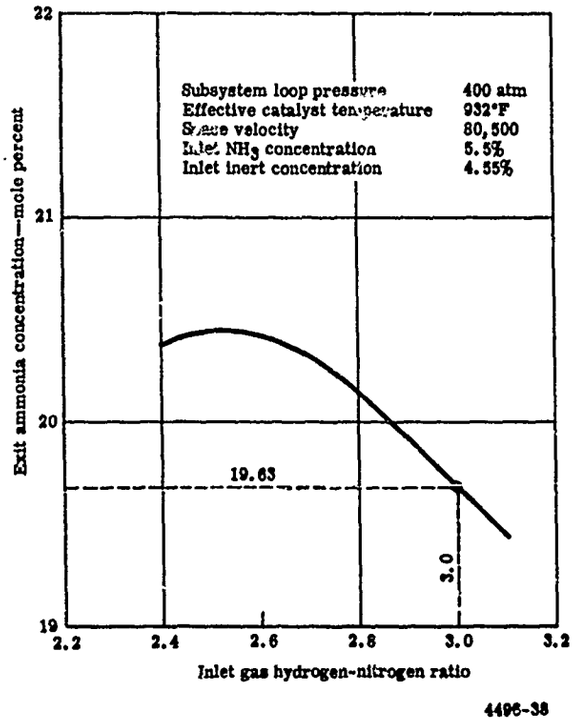


Figure 3.1-9. Catalyst performance with hydrogen-nitrogen ratio.

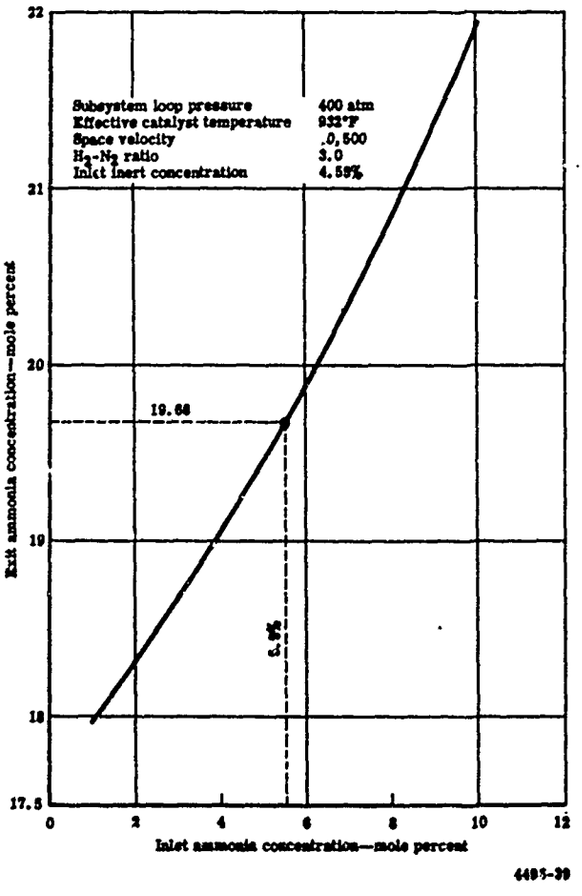


Figure 3.1-10. Catalyst performance with inlet ammonia concentration.

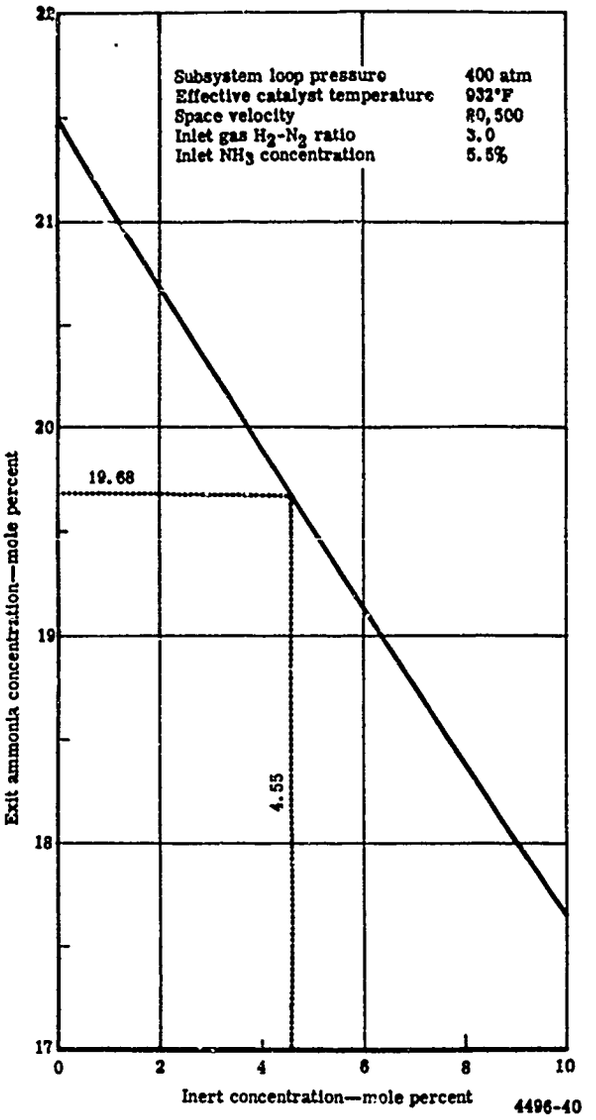


Figure 3.1-11. Catalyst performance with inert concentration.

The effect of the hydrogen-nitrogen molar ratio in the ammonia synthesis subsystem synthesis gas recycle loop is shown in Figure 3.1-9. The maximum conversion occurs in the converter when the hydrogen-nitrogen ratio in the loop is approximately 2.5:1. In design of the subsystem, a loop hydrogen-nitrogen ratio of 3:1 (the feed gas ratio) was used and no advantage was taken of the slight increase in conversion at the 2.5:1 ratio. This results, therefore, in some conservatism in the design since the control subsystem (Section 3.6) will optimize the hydrogen-nitrogen ratio in the loop to the maximum synthesis rate in Figure 3.1-9. Note that while the hydrogen-nitrogen ratio in the loop may be 2.5:1, the feed gas ratio must be at the 3:1 stoichiometric ratio for the production of ammonia.

The catalyst studies were based on Haldor-Topsoe type KM-I catalyst. This is a triple promoted catalyst based on the promoter set K_2O-CaO-Al_2O_3. The catalyst has a particle size of 3 to 6 mm and a bulk density of approximately 150 lb/ft^3. A required catalyst life of greater than three years was used so that catalyst changes in the field would not be necessary.

Converter Temperature Profile

This study was initiated after the material and heat balances had been calculated for the entire loop. The temperature profile was determined with the assumption that heat loss through the converter walls amounted to 5% of the heat released by the synthesis reaction and that the catalyst was in two equal volume beds. The results are presented in Figure 3.1-12.

Heat Recovery Schemes

In the synthesis of ammonia, the reaction generates more heat than is required for thermal regeneration in the system. One restraint was placed on the means of recovering this excess heat; the ammonia synthesis subsystem must use the recovered energy. For example, no credit could be taken for exported steam. Various means were considered but only two of these, steam generation and ammonia refrigeration, were considered feasible for serious consideration.

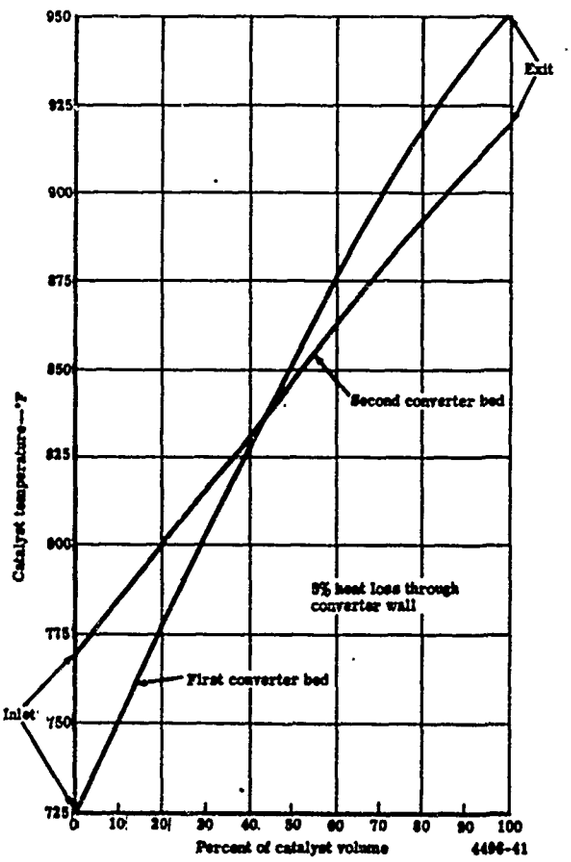


Figure 3.1-12. Converter catalyst temperature profile.

Steam Generation

The most widely used means of reclaiming the heat of the ammonia synthesis reaction is through steam generation. One process, Fauser-Montecatini, claims as much as 0.85 lb of moderate pressure steam per 1.0 lb of ammonia produced. To recover this amount of heat (approximately 68% of the heat of reaction), the process requires the use of elaborate heat exchange equipment and precise control of the catalyst bed temperatures. In the small converters envisioned for this application, burying heat transfer coils in the catalyst bed becomes a difficult design problem. The most practical means of handling heat removal from small-scale converters appears to be by employing a heat exchanger separate from the catalyst bed either in the converter vessel or in a connecting vessel.

Under this condition, the amount of steam generation was based on recovering 50% of the heat of reaction. This resulted in 0.62 lb of steam at 250 psig per 1.0 lb of ammonia produced. Limiting recovery to 50% of the heat of reaction allows the use of reasonable temperature driving forces and avoids the use of large heat transfer areas with a resultant increase in both weight and volume.

To generate and make use of this steam, it would require a turbine drive together with a condenser, condensate pump, and controls. It was envisioned that the turbine could drive a compressor (requiring a speed reducer), cooling fan, or electrical generator. This additional equipment, however, would complicate the plant operation and reduce the reliability. Failure of the system would greatly curtail ammonia production since the converter cooling would be reduced. Based on a turbine steam use rate of 25 lb/hr per hp, a maximum of 14.8 bhp could be developed for the smallest capacity plant (600 lb/hr). This small gain in power (maximum of 7% of the ammonia subsystem power requirement) is not considered practical when weighed against the disadvantages previously listed and against the advantages of the ammonia refrigeration system described below.

Ammonia Refrigeration

A heat recovery scheme that was considered during the process selection studies consisted of using the high pressure liquid ammonia condensed in the loop to compress vapors from a refrigeration system. To accomplish this, the liquid is vaporized at high pressure, taking heat from the synthesis converter, and then is used as the motive fluid in an ejector. The amount of refrigeration available is directly related to the amount of heat recovered and the characteristics of the ejector. By calculating the maximum refrigeration available, it was possible using Figure 3.1-2 to determine that this refrigeration scheme would allow the loop operating pressure to be reduced from 8000 to about 6000 psi. Since no power is required for the ejector system to provide the necessary refrigeration for the ammonia subsystem to operate at 6000 psi, a power savings is realized which is equivalent to the refrigeration horsepower previously calculated for mechanical refrigeration at 6000 psi.

The proposed scheme would use the liquid ammonia product from the primary and secondary separators as a cooling medium for the converter. The high pressure liquid ammonia would be vaporized and superheated and, on expansion through the ejector, would supply the motive force for the refrigerant stream.

Again the heat recovery was assumed to be 50% as in the steam generation case. The amount of refrigeration horsepower developed on expanding the high pressure ammonia to 300 psi amounts to a maximum reduction in power to the ammonia subsystem of 8.5%. Use of this system does not add to the complexity of plant operation and does not decrease the reliability. As long as ammonia is being produced (requiring heat removal), ammonia will be available for heat removal. There are no moving parts and the size and weight of the ejectors are small. There are no mechanical efficiencies to be applied as there are for turbogenerator systems. Thus, it is believed that from a practical viewpoint this system offers the most efficient use of the waste heat generated in the ammonia subsystem.

DESIGN PARAMETERS

In addition to the general overall design parameters presented in the Introduction (Tables 2-I, 2-II, and 2-III), the ammonia subsystem designs were developed to meet the requirements presented in Figure 3.1-13. These designs were later modified in the Integration and Packaging studies to develop the necessary ammonia subsystems for the eleven fuel production systems of interest.

Since the feed gases for the synthesis of ammonia are to be hydrogen (produced from the electrolysis of water) and nitrogen (produced by fractionation of air), the only impurity expected in the hydrogen is oxygen. The chief impurities in the nitrogen stream are argon and oxygen.

After determining the effect on power requirements for nitrogen supply pressures of 0 to 300 psig, it was established that the nitrogen would be available from the nitrogen generation subsystem at 17 psia (see Section 3.2). The hydrogen pressure was fixed at 300 psig and was not considered a variable.

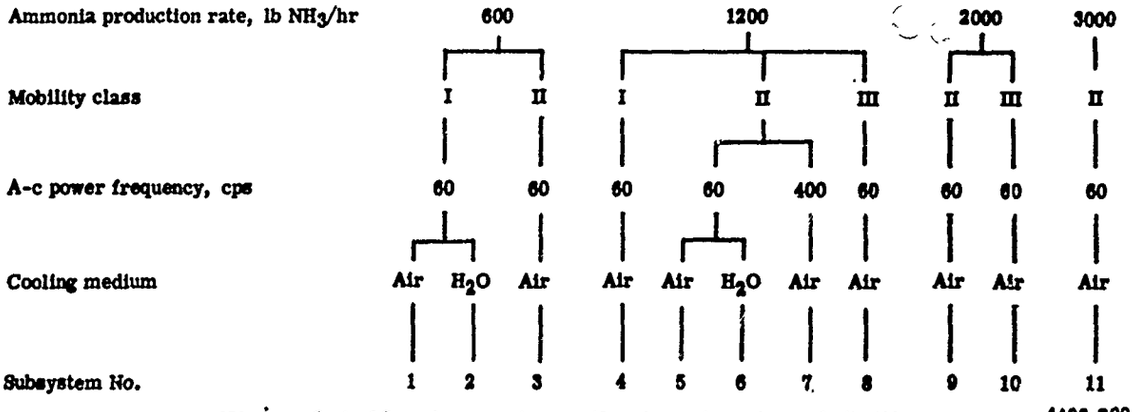


Figure 3.1-13. Ammonia synthesis subsystem definition. 4496-209

Because of the high power requirement for producing hydrogen, losses were kept to a minimum. In most processes studied, the hydrogen loss from the loop ran less than 1% of the feed. Process calculations were based on 0.56% argon in the nitrogen feed.

DESIGN

The proposed design may best be described as a modified Haber-Bosch process even though the operating pressure at 6000 psi is slightly higher than normal. It is similar to processes normally used commercially since the converter gas is recycled and advantage is taken of heat exchange and recovery. The converter design differs quite extensively from that used in the normal Haber-Bosch process, with the chief difference being the method of temperature control.

Process Flow Scheme

The basic process flow scheme is the same for all eleven cases and is outlined in Figure 3.1-14. The nitrogen feed gas at 17 psia and 80°F is controlled for stoichiometric flow with the feed hydrogen. It flows to the knock-out drum, D-5, for separation of any entrained liquids or solids before being compressed to 300 psi in the compressor, C-1. The hydrogen feed is cooled from 167 to 120°F in the hydrogen precooler, E-12, and joins the nitrogen feed coming from the compressor in the knock-out drum, D-4. From this vessel, the combined stream is compressed in three stages to loop pressure. Intercoolers follow each stage of compression to cool the stream and to condense and remove water vapor. After the first stage of compression, the gas flows through a gas filter, D-3, and a deoxo unit, D-1, where any oxygen present is reacted with hydrogen to form water. The temperature rise during compression is sufficient for the deoxidation reaction to occur. The water formed is condensed and removed in the interstage cooler, E-9, and its accompanying separator. The feed gas stream joins the recycle gas stream in the loop at the recycle compressor suction.

The recycle compressor furnishes the means for circulating the synthesis gas around the loop. From the compressor, the gas flows through a filter, D-6, to remove any entrained compressor oil and is then cooled to 75°F in the refrigerated secondary condenser, E-1. The ammonia that is condensed is separated from the gas stream in the secondary separator, S-2. The gas is heated to 725°F in the converter heat exchanger, E-2. A bypass around E-2 is provided for temperature control at the converter bed inlets.

In the first converter, the ammonia concentration changes from 5.5 to 13.8%. The gas is then cooled in the ammonia vaporizer, E-3, to 770°F. In the second converter, the ammonia concentration increases to 19.7% and the temperature rises to 919°F. The gas is cooled to 240°F in E-2 and to 110°F in the primary condenser, E-4. The condensed ammonia is separated in the primary separator, S-1, and the gas returns to the recycle compressor.

The liquid ammonia from the two separators, S-1 and S-2, is reduced in pressure and is then vaporized in E-3 by the exit gas from the first converter. The ammonia then passes through the ejector, EJ-1, where the pressure is reduced to 300 psi and a secondary stream of ammonia refrigerant from E-1 is compressed from 96 to 315 psia. The ammonia is cooled and condensed in the product condenser, E-5, and the uncondensed gas is separated in the product cooler, S-3. A portion of the liquid ammonia is used as refrigerant while the remainder is rejoined with the vapor and sent to storage.

Figure 3.1-14 includes the quantities of material flowing at the various points in the process. The heat balance was calculated assuming that heat loss was 5% of the heat of reaction in the converter and 2.5% in the rest of the loop.

Equipment Description

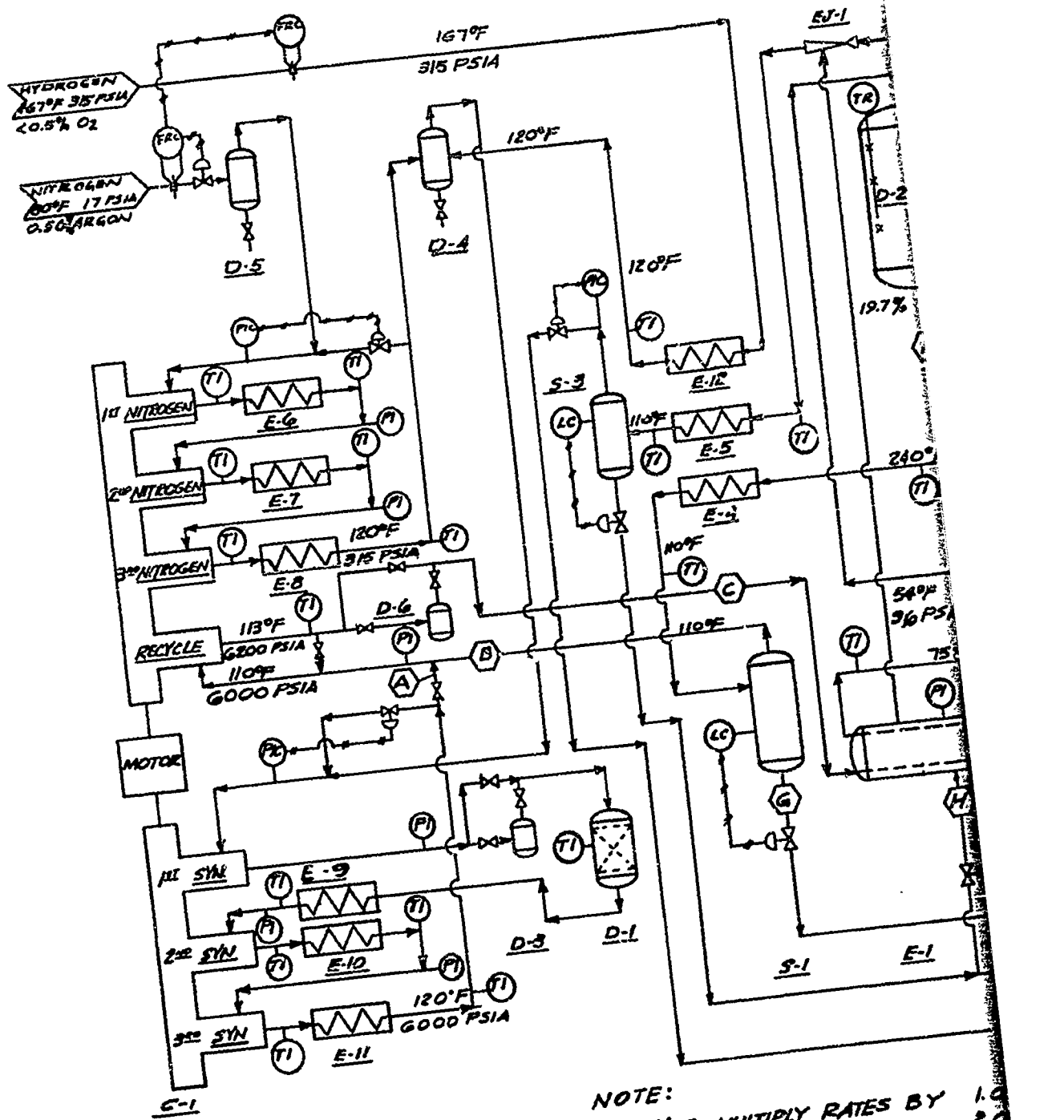
The equipment for the ammonia synthesis subsystem was designed from information developed for commercial installations. The reliability of these designs has been proved by numerous operating hours. Although considerable engineering and testing would be required for proving the Energy Depot subsystem design, development work would be at a minimum.

Compressor

The basic compressor used in these designs is the Joy Manufacturing Company unit of the WL-80 class. This unit is air-cooled and can be made with either single or double throw cranks. The maximum horsepower rating for the compressor with double throw crankcases is 150 bhp with single-acting cylinders and 250 bhp with double-acting cylinders. The single throw crankcase has one-half the horsepower rating of the double throw unit. Pressurized lubrication is provided by a pump.

The basic multiservice compressor, including intercoolers and aftercoolers, was sized for 720 lb NH_3/hr, 20% above the subsystem design point value of 600 lb NH_3/hr. This was done to prevent step changes in compressor weight from being required in the integrated systems, if the overall system production rates were slightly larger than the subsystem design points. When multiple compressor units were used for the larger capacity plants, the 20% over-rating of the basic compressor unit resulted in a capability of 1440 and 3600 lb NH_3/hr for the 1200 and 3000 lb/hr subsystems. The 2000 lb/hr subsystem was designed for 2160 lb NH_3/hr, an 8% over-capacity capability. However, these design values do not represent the maximum capability of the basic compressor unit. If necessary, the capability of each unit can be increased to 500 bhp.

Compression of the synthesis gas for the 600 lb/hr capacity case from 300 to 6000 psi requires six cylinders arranged in two throw-cranks. The horsepower requirement loads one unit to the point where it is necessary to put the other services on a second compressor. The second unit

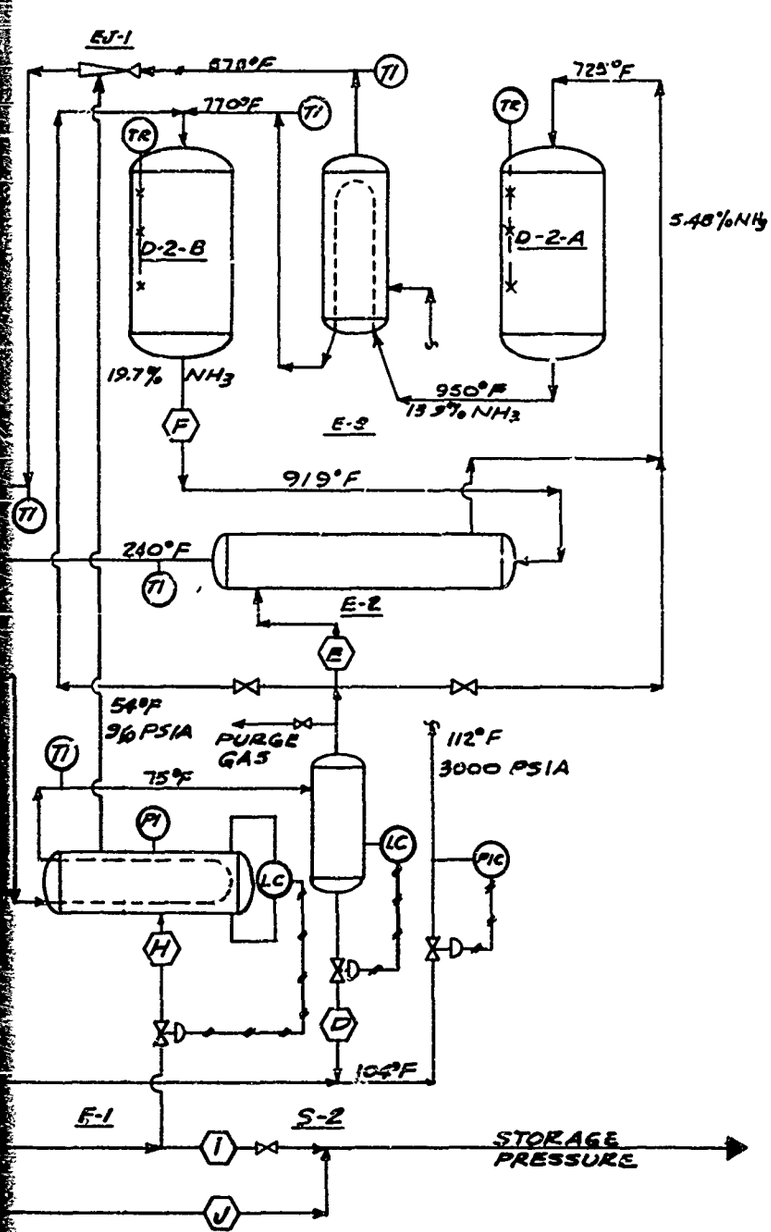


NOTE:
FOR 600#/HR MULTIPLY RATES BY
1200#/HR 1.0
2000#/HR 2.0
3000#/HR 3.5
5.0

FLOW RATES LBMOLS/HR FOR 600 LBNH3/H

STREAM	A	B	C	D	E	F
HYDROGEN	53.673	146.798	200.471	0.136	200.335	147.894
NITROGEN	17.916	48.914	66.830	0.052	66.778	49.165
ARGON	0.100	13.387	13.487	0.013	13.474	13.474
AMMONIA	-	22.197	22.197	5.933	16.264	51.492
TOTAL	71.689	231.296	302.985	6.134	296.851	261.625

Figu



Equipment List

C-1	Multiservice compression
D-1	Deoxo unit
D-2A&B	Converter
D-3	Gas filter
D-4	Knockout drum
D-5	Knockout drum
D-6	Gas filter
E-1	Secondary condenser
E-2	Converter heat exchanger
E-3	Ammonia vaporizer
E-4	Primary condenser
E-5	Product condenser
E-6 through E-11	Compressor interstage & aftercoolers
E-12	Hydrogen precooler
EJ-1	Ejector
S-1	Primary separator
S-2	Secondary separator
S-3	Product separator

Instrument legend

FRC	Flow recording controller
LC	Level controller
PI	Pressure indicator
TI	Temperature indicator
TR	Temperature recorder
PIC	Pressure indicating controller

ES BY 1.0
2.0
3.33
5.0

600 LB NH3/HR					
E	F	G	H	I	J
0.335	147.894	0.696	0.004	0.009	0.823
2.778	49.165	0.251	0.002	0.004	0.299
474	13.474	0.087	0.001	0.001	0.099
264	51.492	29.295	15.081	30.778	4.450
6851	761.625	30.929	15.088	30.792	5.671

Figure 3.1-14 Ammonia synthesis subsystem process flow diagram.

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is of single throw-crank design. It has three double-acting cylinders for nitrogen and cylinder blow-by compression to 300 psi and one single-acting cylinder for recycle compression. The two basic compressor units are driven by a common double-ended electric motor. This configuration is suitable for all cases of 600 lb/hr capacity. For cases of higher capacities, multiples of this basic unit are used.

For Subsystems 1 through 3, only one compressor unit is required. Two units are needed for Subsystems 4 through 8. For Subsystems 9 and 10, there are two units for the nitrogen, blow-by, and synthesis gas services. Each of these units is similar in design, but of higher capacity (larger cylinders) than the basic unit. Recirculation is accomplished by a third unit consisting of only recycle cylinders. For Subsystem 11, there are three compressors for synthesis gas service. The nitrogen, blow-by, and recycle service is on a fourth double unit of larger capacity than the basic unit. All of the horsepower ratings for the altered units are within the manufacturer's recommended limits.

Each section of a double unit is equipped with a fan for moving cooling air over the cylinders. In some cases, the compressors also drive the fans for the fin-fan air-cooled units used for the various cooling services throughout the loop.

In the water-cooled cases (Subsystems 2 and 6), the compressor is the same basic unit. However, the cylinder fins have been replaced by water jackets.

All of the compressors are to run at 1160 rpm and are driven by induction motors which are direct-coupled to the compressors. With 400-cps input electrical power (Subsystem 7), two alternatives existed. A small high speed motor with a gearbox transmission for speed reduction, or a many-poled, large, special application motor directly coupled may be used.

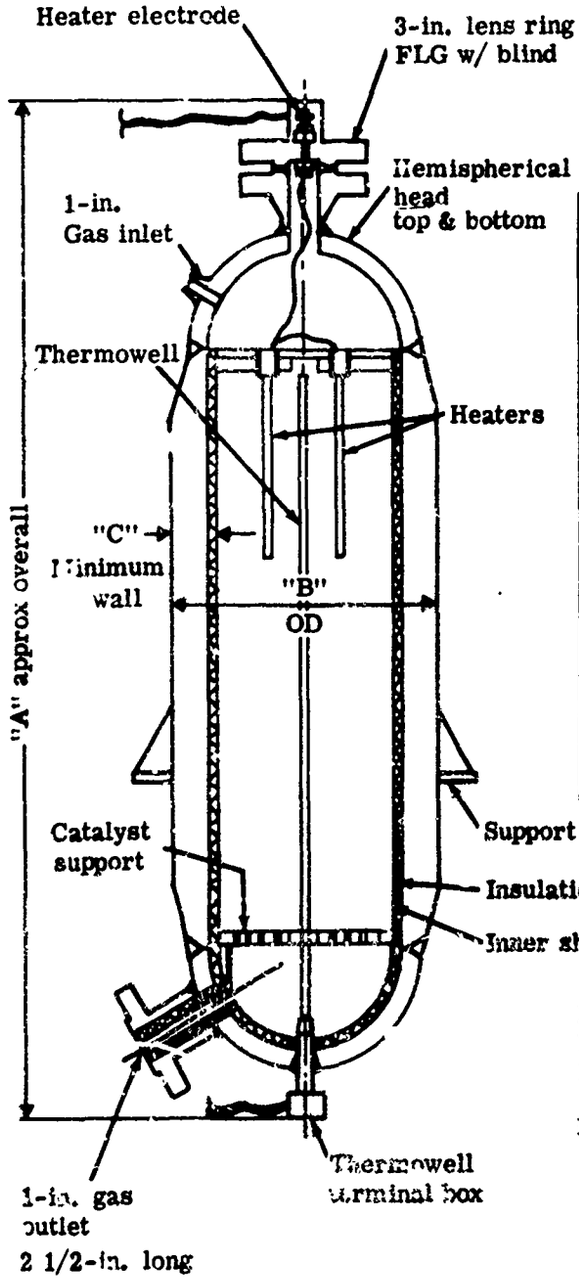
An investigation made by Joy indicated that the weight and bulk of a 400-hp speed reduction system would be approximately equivalent to a motor operating directly at 1200 rpm. Therefore, the direct drive 1200-rpm, 400-cps motor was selected because it appeared to offer a simpler system.

Converter

The converter design departs slightly from the usual commercial configuration because of the small capacities and low height specifications. In the proposed design, there are no heat exchange or cooling coils buried in a single catalyst bed as is customary in large plants. Instead, two separate converters in series are used. After partial conversion, the gas is cooled in a separate heat exchange vessel. The conversion is then completed in a second converter.

A preliminary converter design is shown in Figure 3.1-15. Each converter is internally insulated to reduce the heat loss to 5% of the heat of reaction. Start-up heaters are imbedded

in the catalyst bed to reduce the start-up time for heating the catalyst. The converters are straight cylindrical vessels with a minimum of openings. The internals are simply designed. The gas flow is from top to bottom with no baffles or reverse flows. Because of this design and the internal insulation which reduces the rate of wall heating, no undue stressing is anticipated during rapid start-up.



	Production rate—lb/hr			
	600	1200	2000	3000
"A" ft and in.	6-2	6-6	6-9	7-1
"B" in.	13 1/2	17 3/4	22	26 1/4
"C" in.	1 7/8	2 3/8	3	3 1/2
Catalyst cu ft	1.4	2.8	4.6	7.0
Approximate gross wt including catalyst—lb	1450	2400	3650	5250
Start-up heater kw	50	75	100	125

Design pressure 6600 psi
Design temperature-out shell 650°F
Design temperature-internals 1000°F

Material
Outer shell SA336 Gr. F22(2 1/4 Cr 1 Mo)
Internals 304SS

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Figure 3.1-15. Ammonia converter schematic design.

Heat Exchange Vessels

With the exception of the fin-fan air-cooled units, all of the heat exchangers and coolers are of the shell and tube types. Depending on the application, they may be either of fixed head with U-tubes or of floating head design. The high pressure service was placed in the tubes where the smaller diameters would result in thinner walls.

Fin-Fan Air-Cooled Exchanger

Elimination of the waste heat from the loop is accomplished by fin-fan air-cooled exchangers (except for water-cooled subsystems 2 and 6). The proposed units vary in cross sectional or face area and the number of rows of tubes, depending on the duty required and the allowable space. Each unit may incorporate several cooler functions, depending on the subsystem considered. The tubes are of the correct wall thickness for the specific operating temperature and pressure level. Thus, adjacent tubes may vary. The multiservice would be accomplished by proper manifolding. Louvers would be provided to regulate the quantity of air flow for changes in ambient temperatures.

Ejector

The ejector used in the proposed design is for the purpose of compressing the vaporized ammonia used as the refrigerant in the secondary condenser. It is of special design (machined from bar stock) because of the motive pressure—3000 psi. It is necessary to reduce the ammonia pressure from 6000 to 3000 psi for this application because at the high pressure the small quantity of flow in the smallest system (600 lb NH_3/hr) requires an orifice smaller than it is practical to machine. The same motive pressure was used in the larger systems to allow for loop operation at less than 6000 psig when the catalyst is new and more active or for half capacity operation.

The ejector is made of 316 SS Three sizes were recommended by the manufacturer for the four capacities considered. The largest size can serve two capacities by changing the orifice size.

Miscellaneous Vessels

Included in this category are the gas filters, knock-out drums, deoxo unit, and separators. They are all cylindrical carbon steel vessels of standard design. The most critical part of the design of these items from a process point of view is supplying sufficient residence time in the vessel for the specified function to occur.

Instruments

Where practical, the instruments envisioned for this design are of the standard small-case variety. The compressor comes complete with instruments. Table 3.1-III shows those instruments required in addition to those required for compressor operation. The number of instruments is the same for all ammonia subsystems.

Table 3.1-III.
Required instruments.

Instrument	Required No.
Flow recording controller	2
Pressure indicating controller	4
Level controller	4
Temperature recorder	2
Temperature indicator	20
Pressure indicator	7

Air operated valves are anticipated. Automatic analysis of the hydrogen-nitrogen ratio in the converter gas will control the hydrogen-nitrogen ratio of the feed gases. Steady-state operation of the plant will be automatic.

Piping and Structure

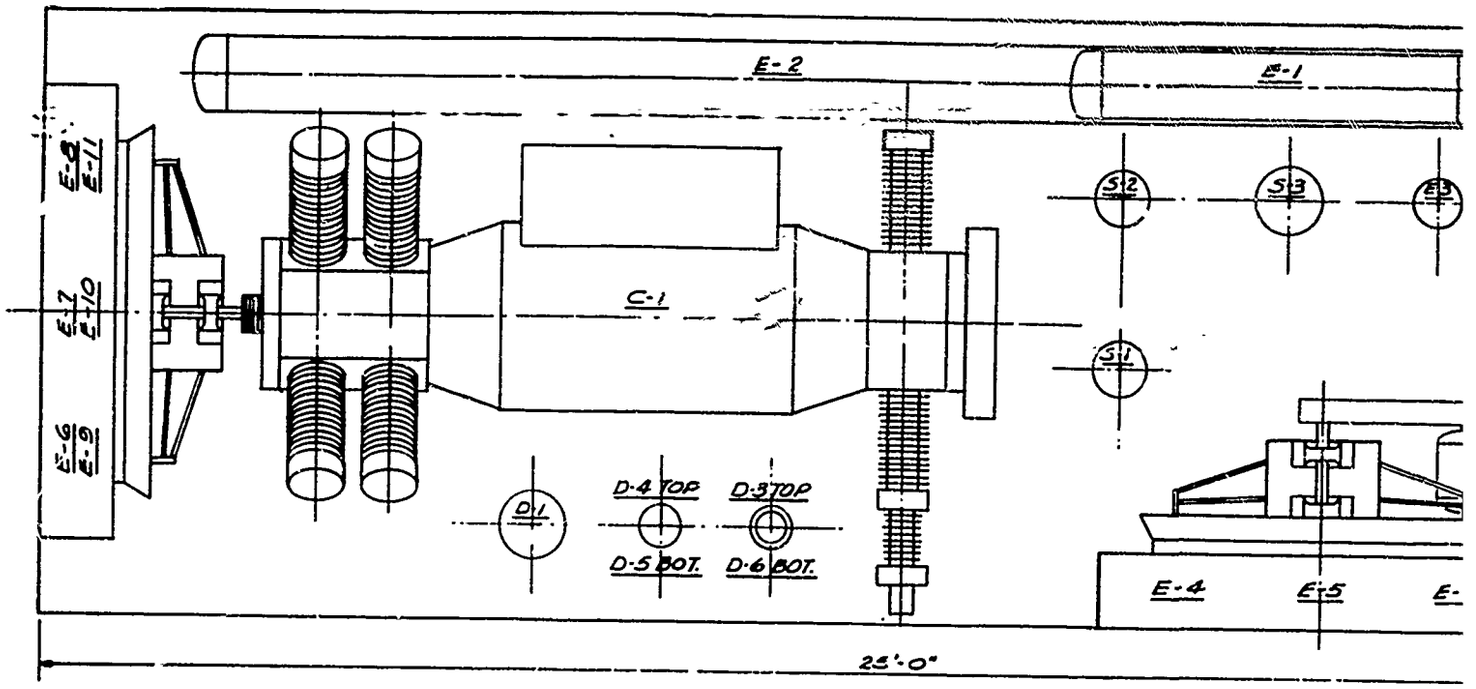
The piping used on this design will have a minimum nominal diameter of 1 in. to maintain ruggedness. The maximum line sizes are anticipated to be 2.5 in. in diameter. To maintain ruggedness, Schedule 40 will be the minimum wall thickness used on piping. The weight of piping used for ammonia synthesis subsystems using two skids will be slightly higher than for single skid cases. This is caused by the longer runs between the various vessels.

The skids are designed to have aluminum beams at least 8 in. thick. Stiffness is imparted by ample use of cross braces and by building a box type frame around the entire package.

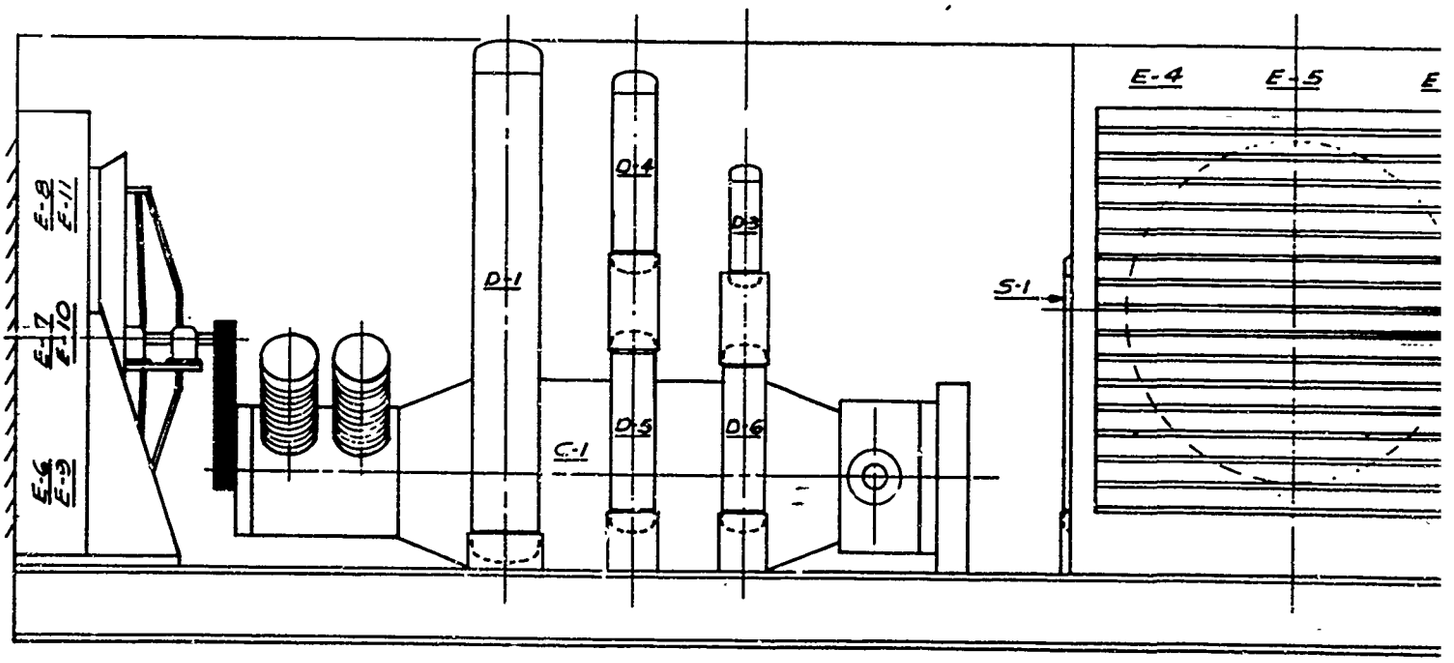
Equipment Layout

The plant layout for the eleven subsystems is presented in Figures 3.1-16 through 3.1-25. Nomenclature for the vessels and symbols is included.

The spacing of the equipment was planned so as to have sufficient room for interconnecting piping. The required room was based on fabrication experience. For clarity, piping was not shown. Equipment was placed to minimize the piping runs and the number of lines connecting the skids. High pressure lines between skids were eliminated where practical.

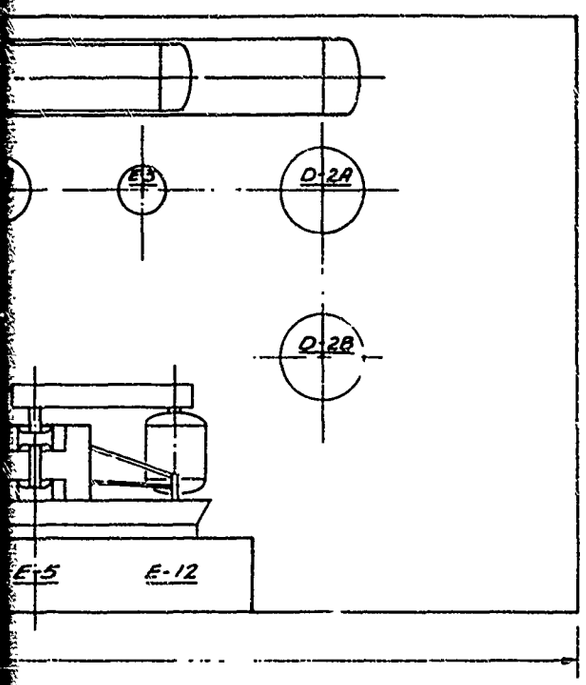


PLAN



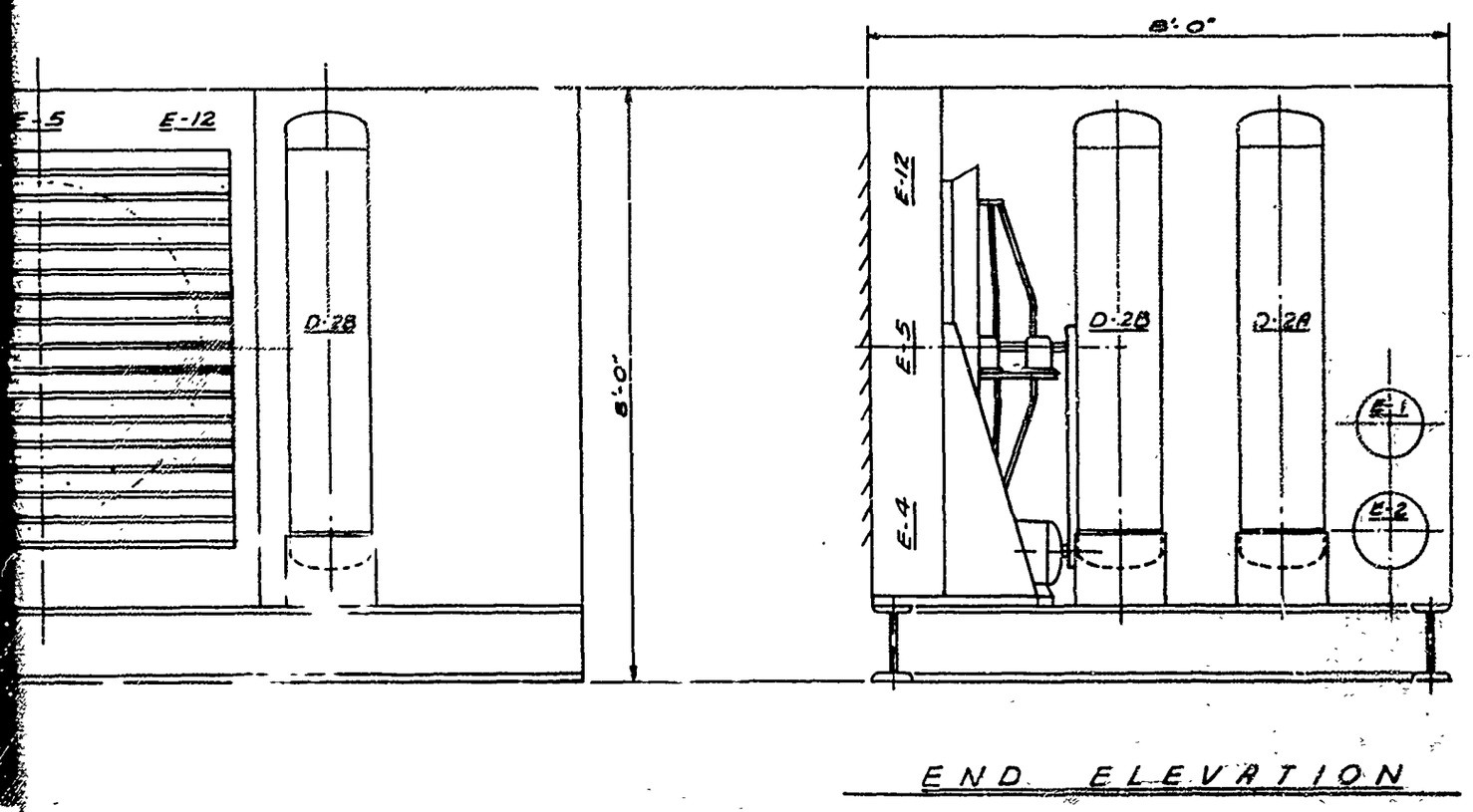
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SIDE ELEVATION



Equipment legend

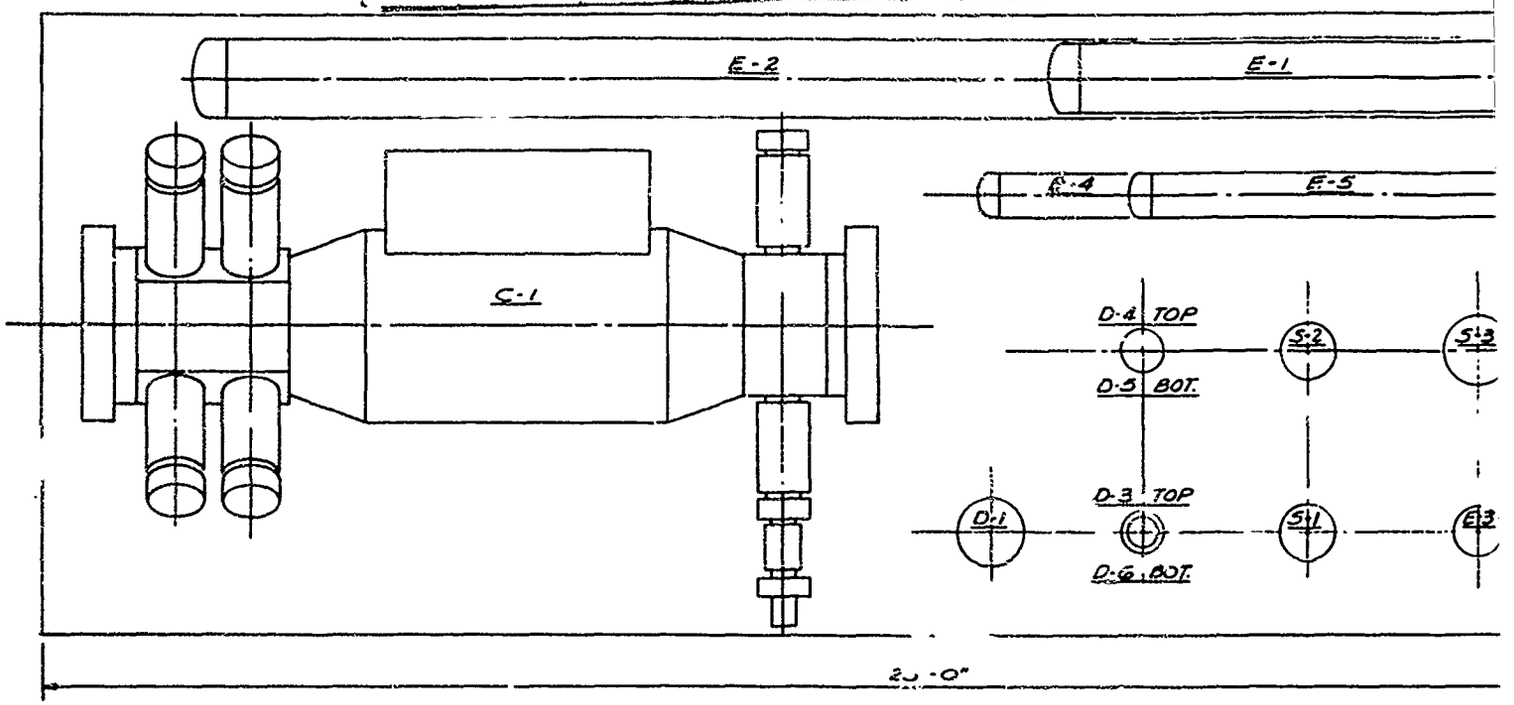
C-1	Multiservice compressor
D-1	Deoxo unit
D-2A&B	Ammonia converter
D-3	Gas filter
D-4	Knock-out drum
D-5	Knock-out drum
D-6	Gas filter
E-1	Secondary condenser
E-2	Converter heat exchanger
E-3	Ammonia vaporizer
E-4	Primary condenser
E-5	Product condenser
E-6 through E-11	Compressor intercoolers & aftercoolers
E-12	Hydrogen precooler
S-1	Primary separator
S-2	Secondary separator
S-3	Product separator



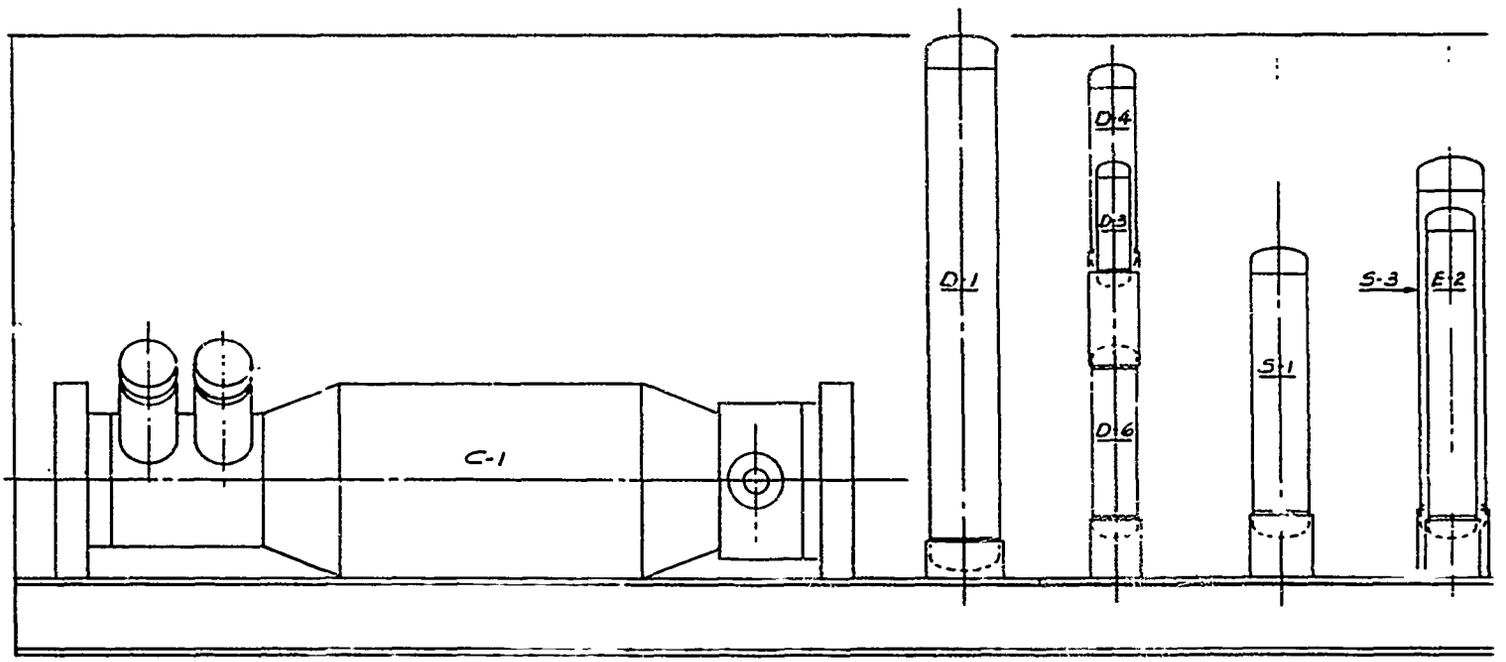
END ELEVATION

Figure 3.1-16. Equipment layout for ammonia subsystems 1 and 3.

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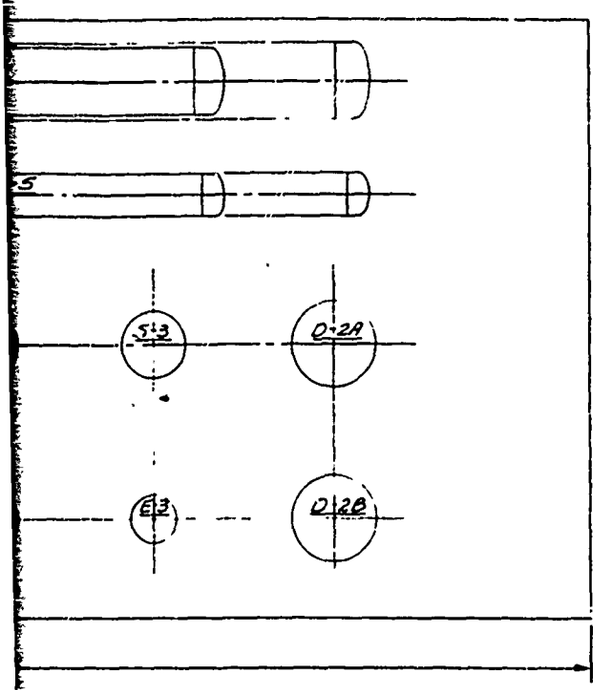


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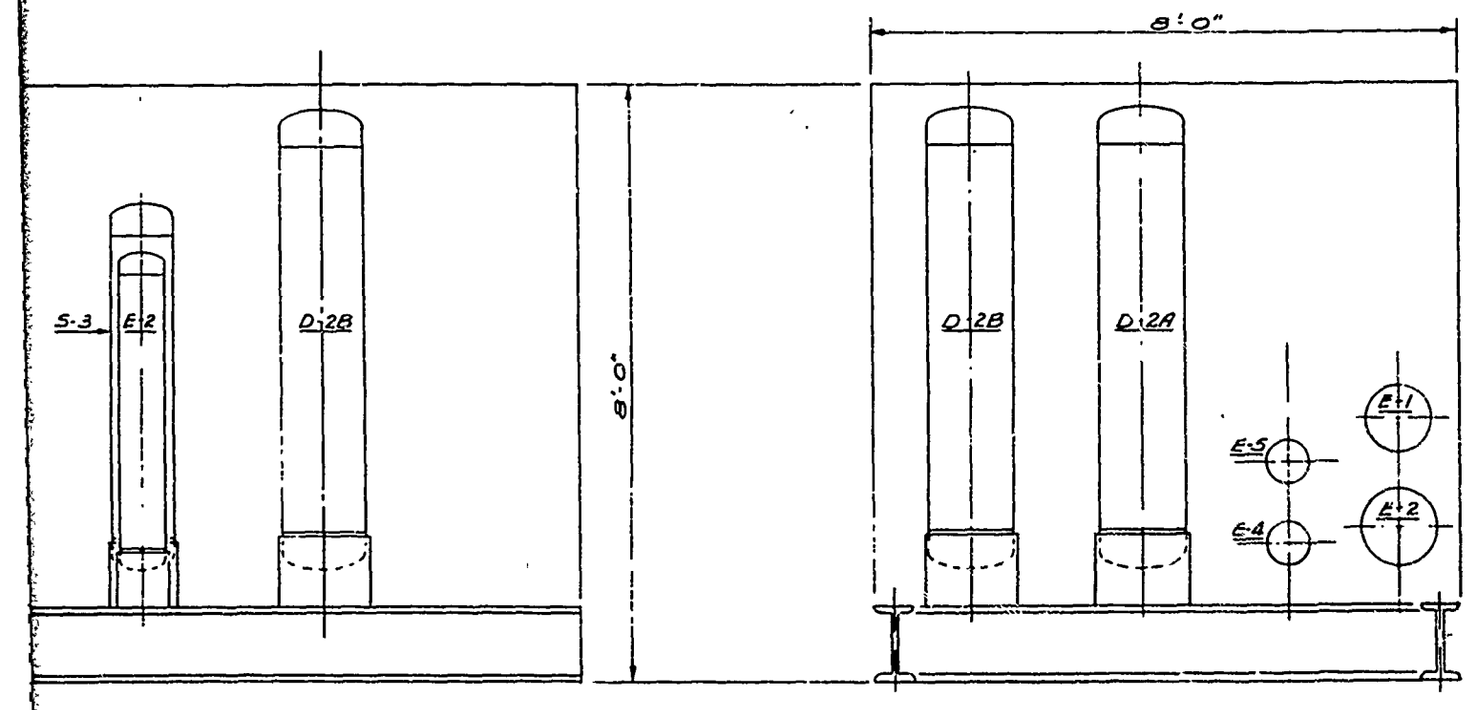
4496-45

SIDE ELEVATION



Equipment legend

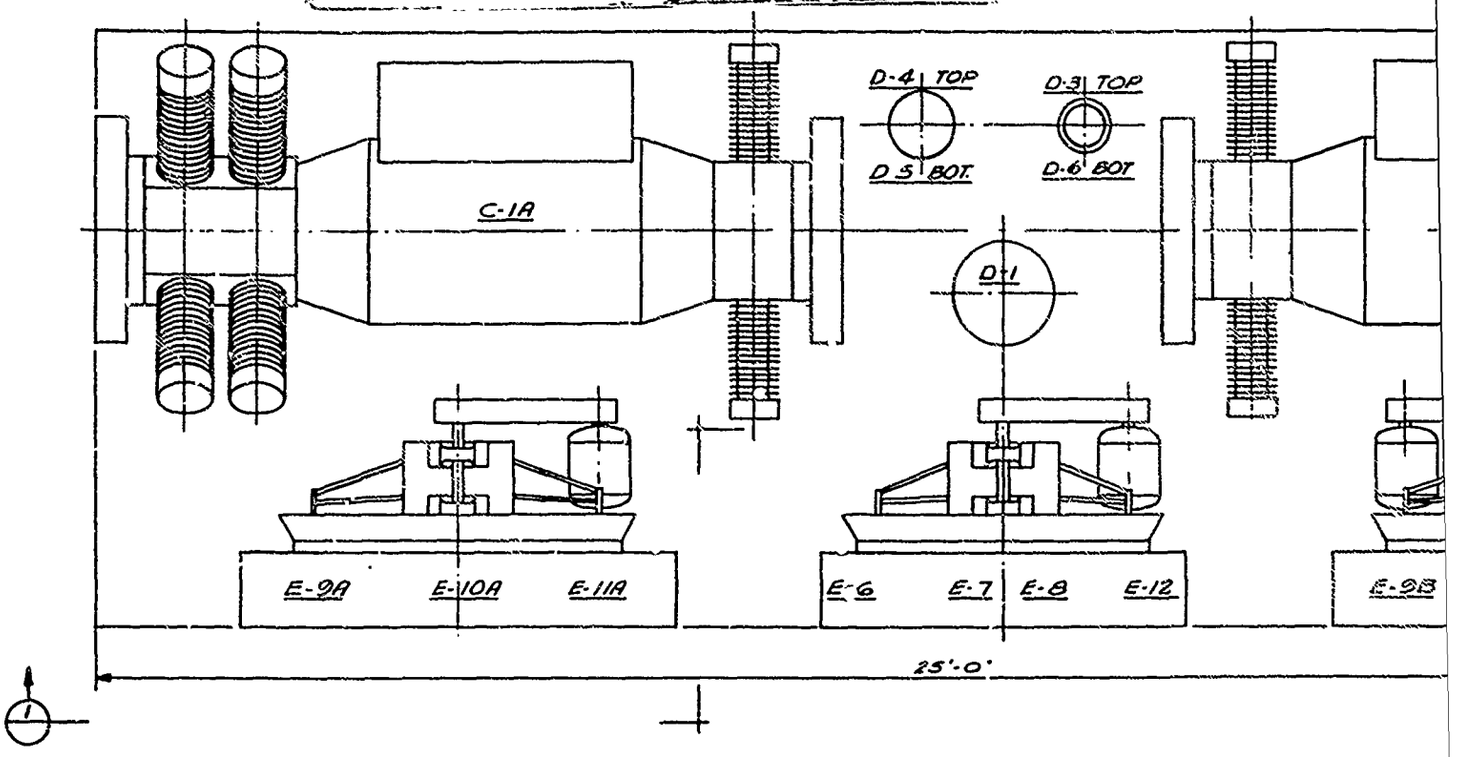
C-1	Multiservice compressor
D-1	Deoxo unit
D-2A&B	Ammonia converter
D-3	Gas filter
D-4	Knock-out drum
D-5	Knock-out drum
D-6	Gas filter
E-1	Secondary condenser
E-2	Converter heat exchanger
E-3	Ammonia vaporizer
E-4	Primary condenser
E-5	Product condenser & hydrogen precooler
S-1	Primary separator
S-2	Secondary separator
S-3	Product separator



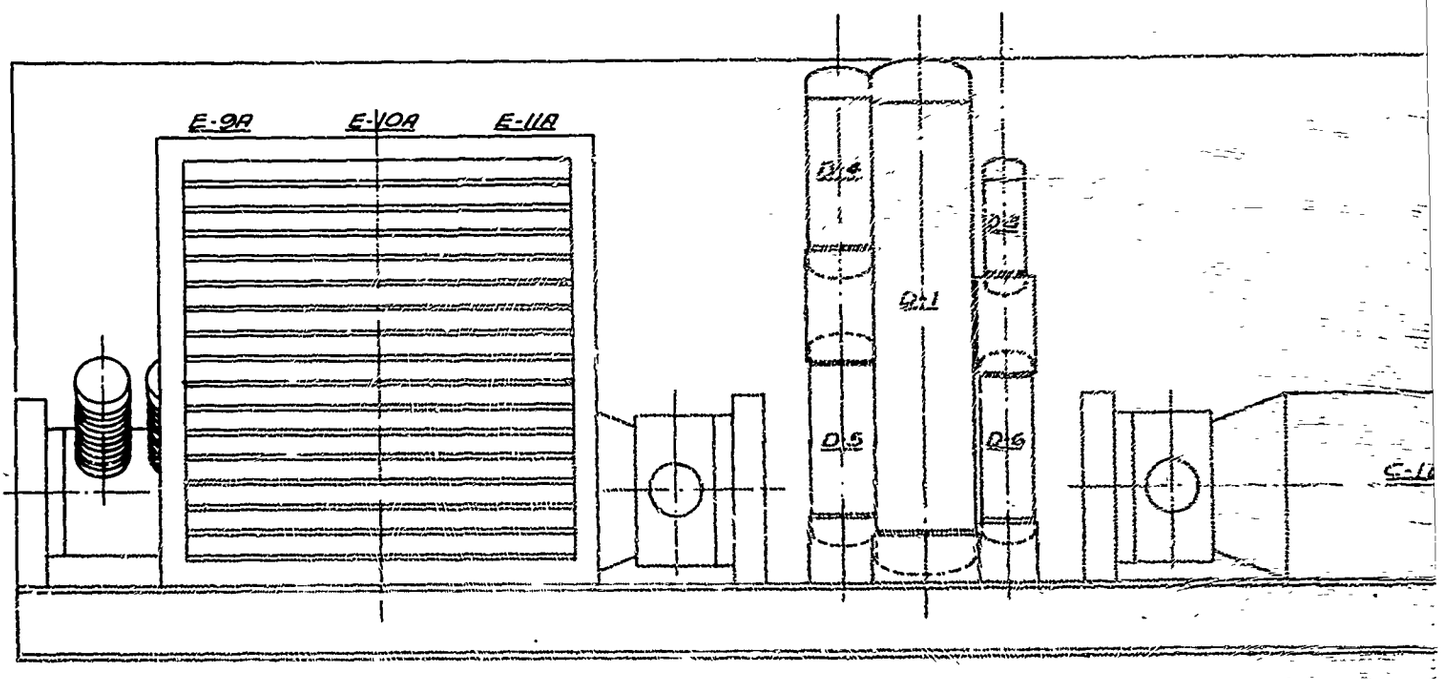
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Figure 3.1-17. Equipment layout for ammonia subsystem 2.

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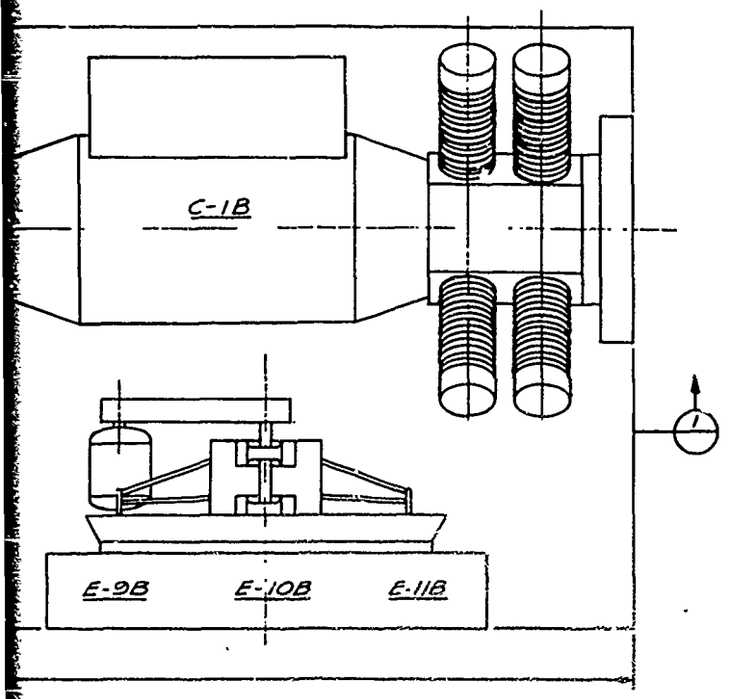


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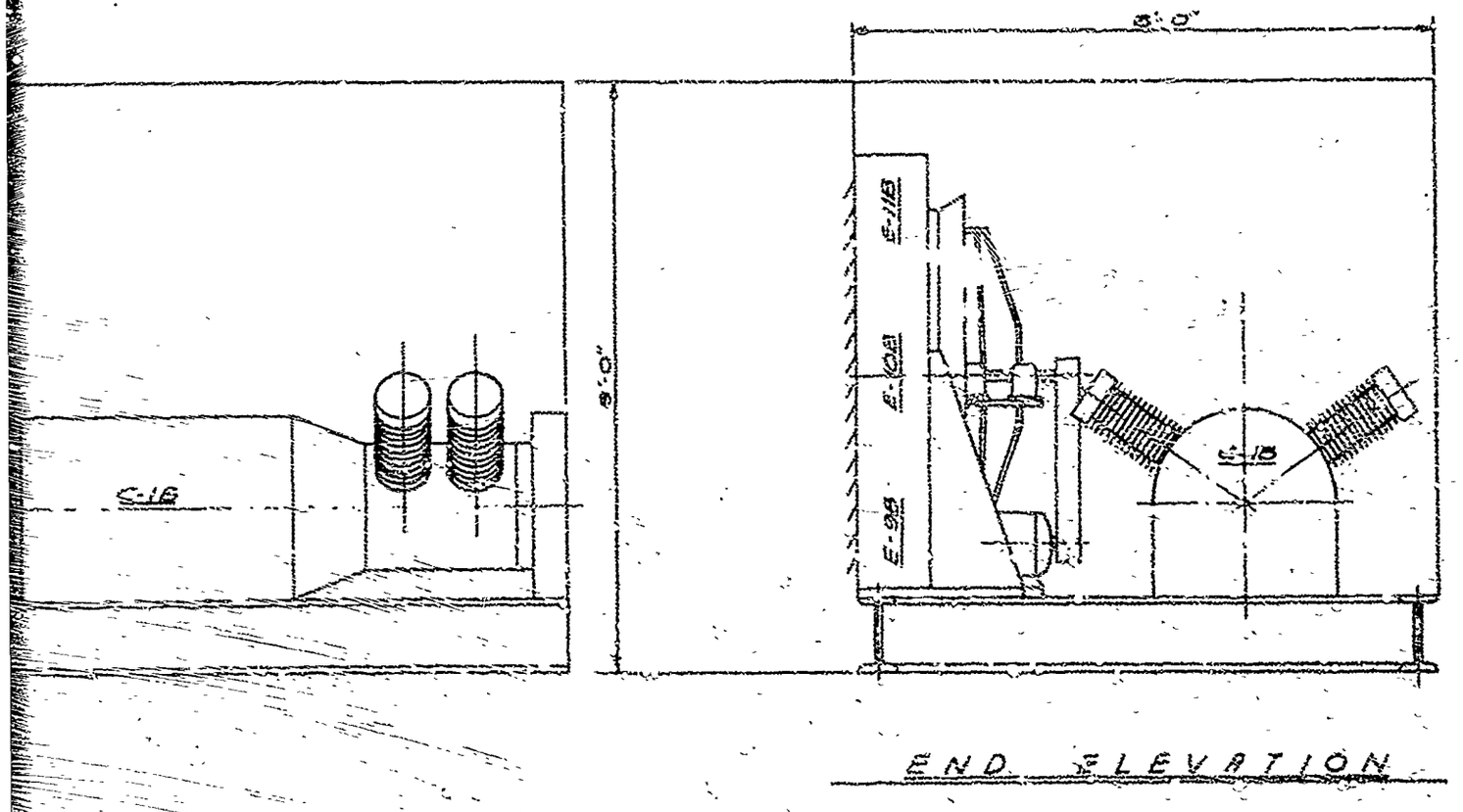
1498-46

SECTION "1-1"



Equipment legend

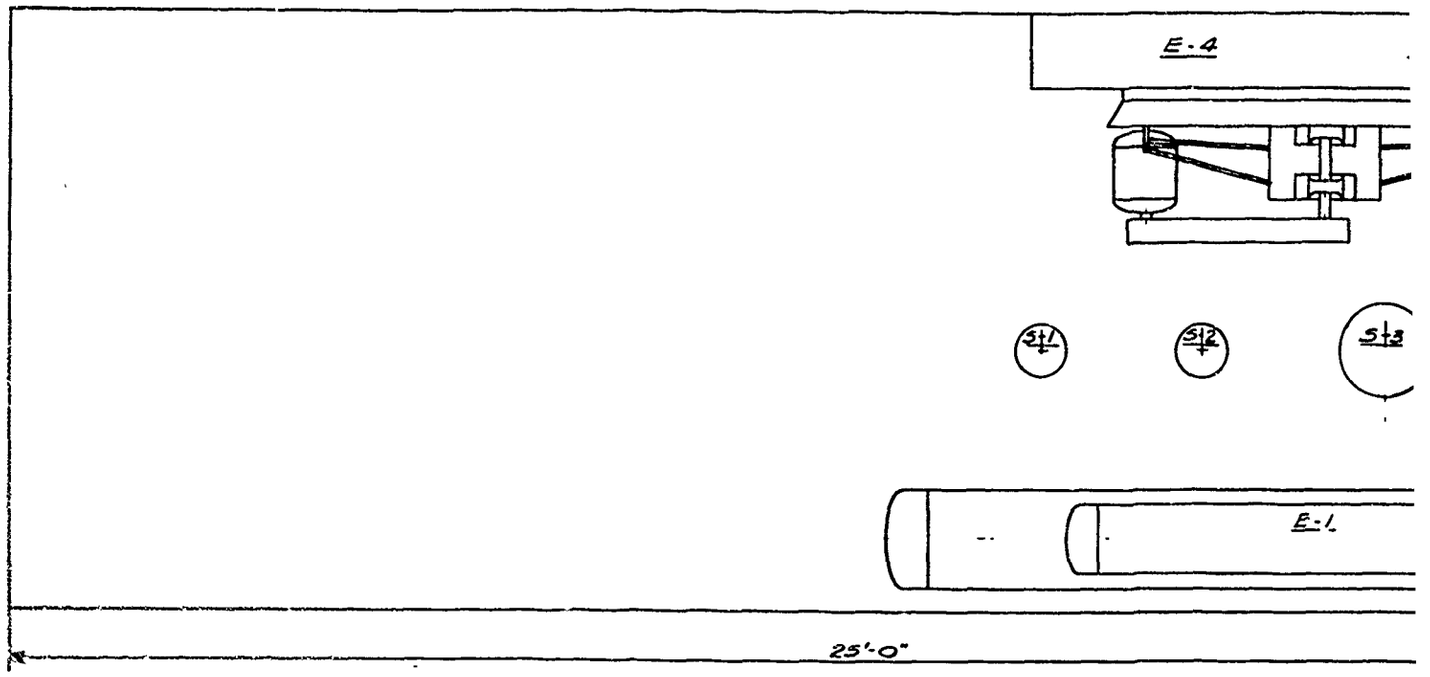
C-1A&B	Multiservice compressor
D-1	Deoxo unit
D-2A&B	Ammonia converter
D-3	Gas filter
D-4	Knock-out drum
D-5	Knock-out drum
D-6	Gas filter
E-1	Secondary condenser
E-2A&B	Converter heat exchanger
E-3	Ammonia vaporizer
E-4	Primary condenser
E-5	Product condenser
E-6 through E-11	Compressor interstage & aftercoolers
E-12	Hydrogen precooler
S-1	Primary separator
S-2	Secondary separator
S-3	Product separator



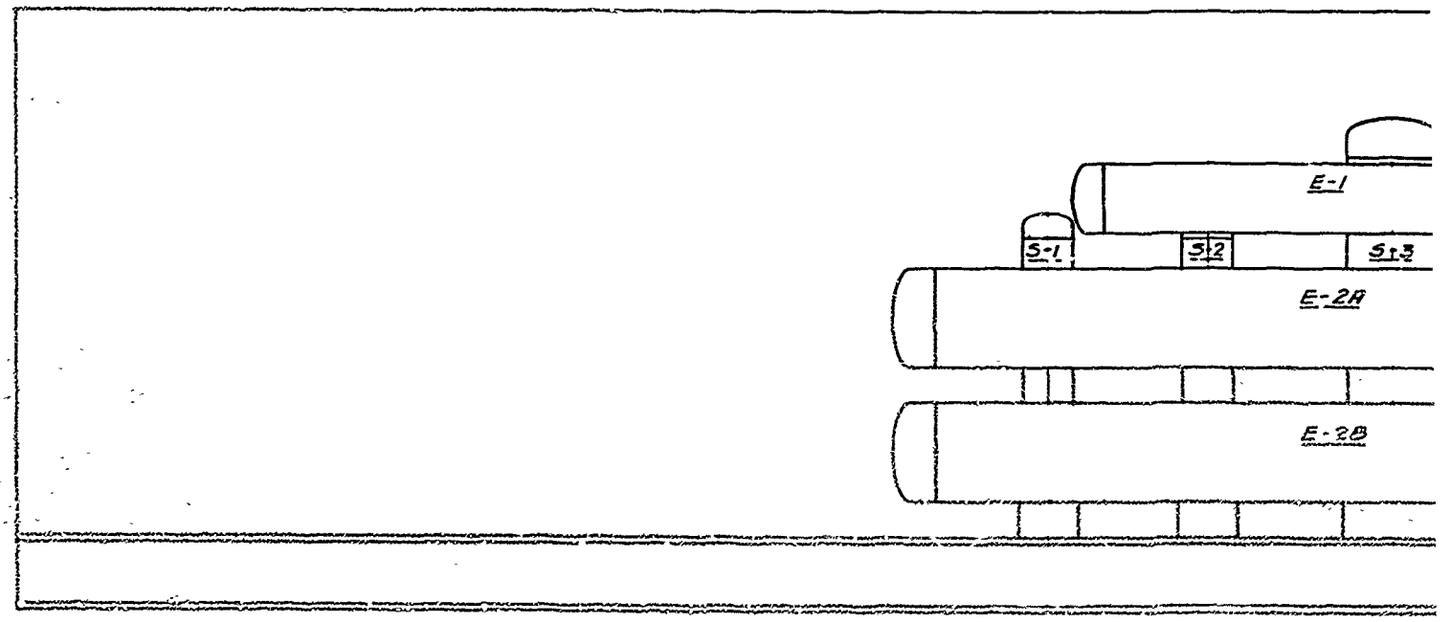
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Figure 3.1-16. Equipment layout for ammonia subsystem 4—Skid 1.

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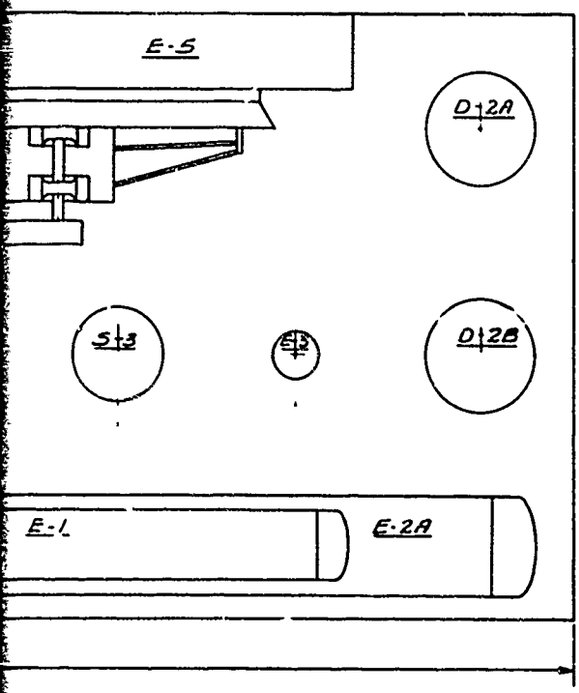


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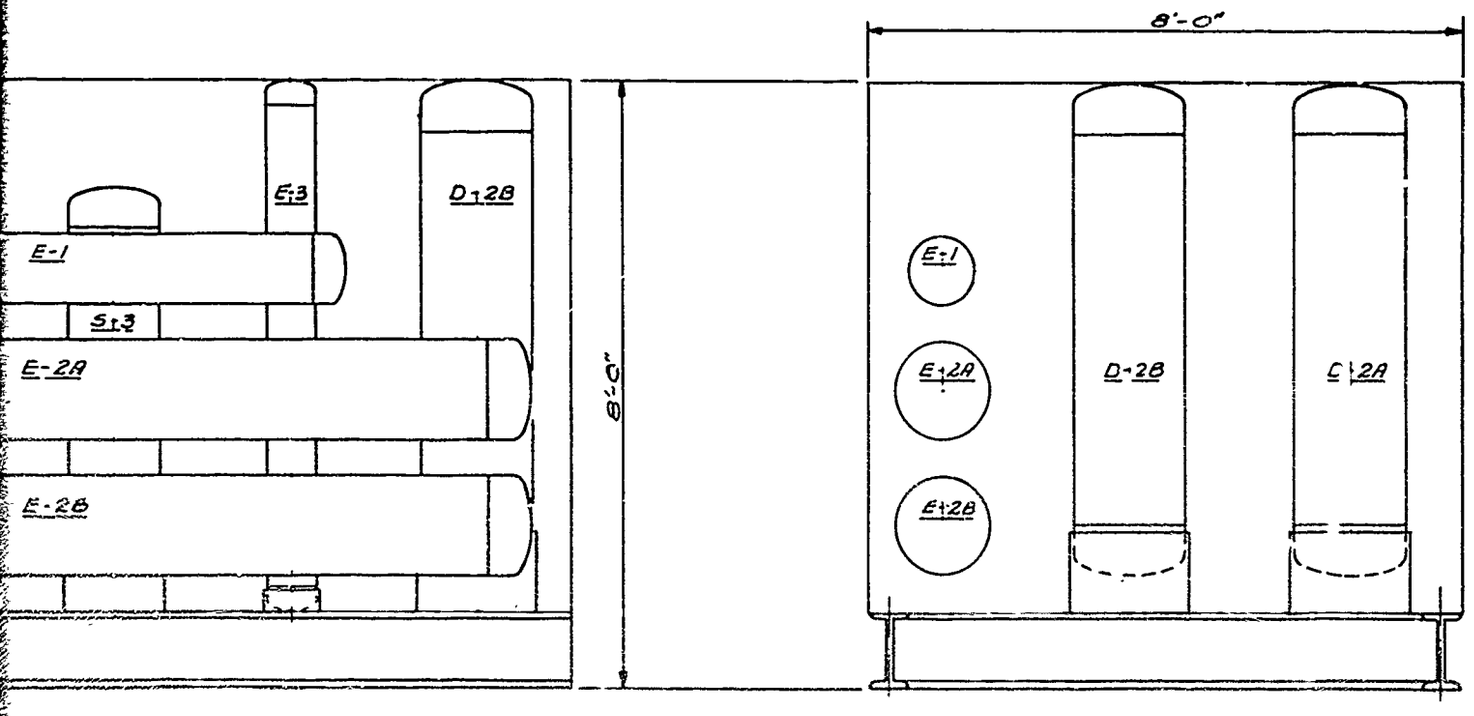
4496-17

SIDE ELEVATION



Equipment legend

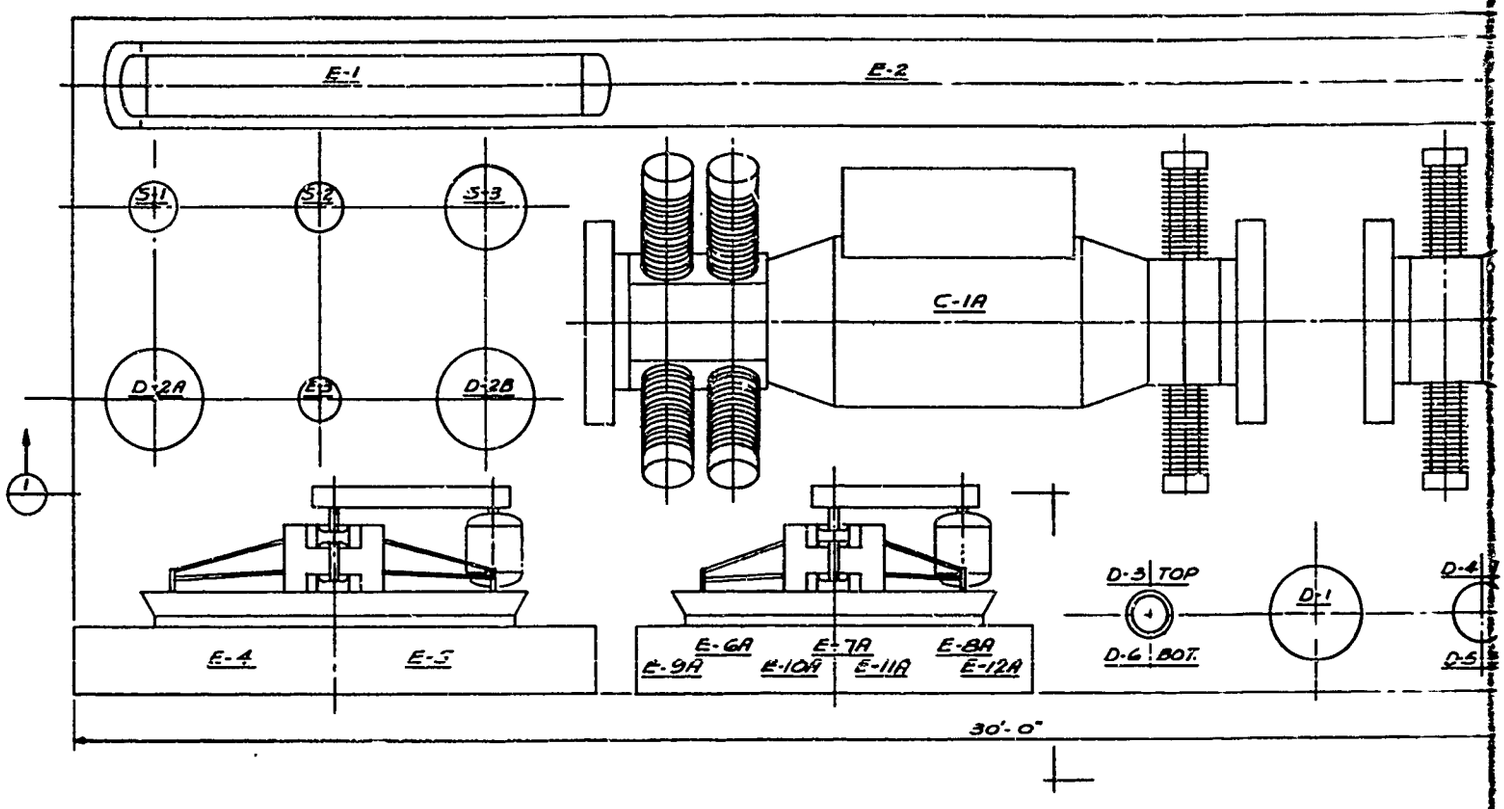
C-1A&B	Multiservice compressor
D-1	Deoxo unit
D-2A&B	Ammonia converter
D-3	Gas filter
D-4	Knock-out drum
D-5	Knock-out drum
D-6	Gas filter
E-1	Secondary condenser
E-2A&B	Converter heat exchanger
E-3	Ammonia vaporizer
E-4	Primary condenser
E-5	Product condenser
E-6 through E-11	Compressor interstage & aftercoolers
E-12	Hydrogen precooler
S-1	Primary separator
S-2	Secondary separator
S-3	Product separator



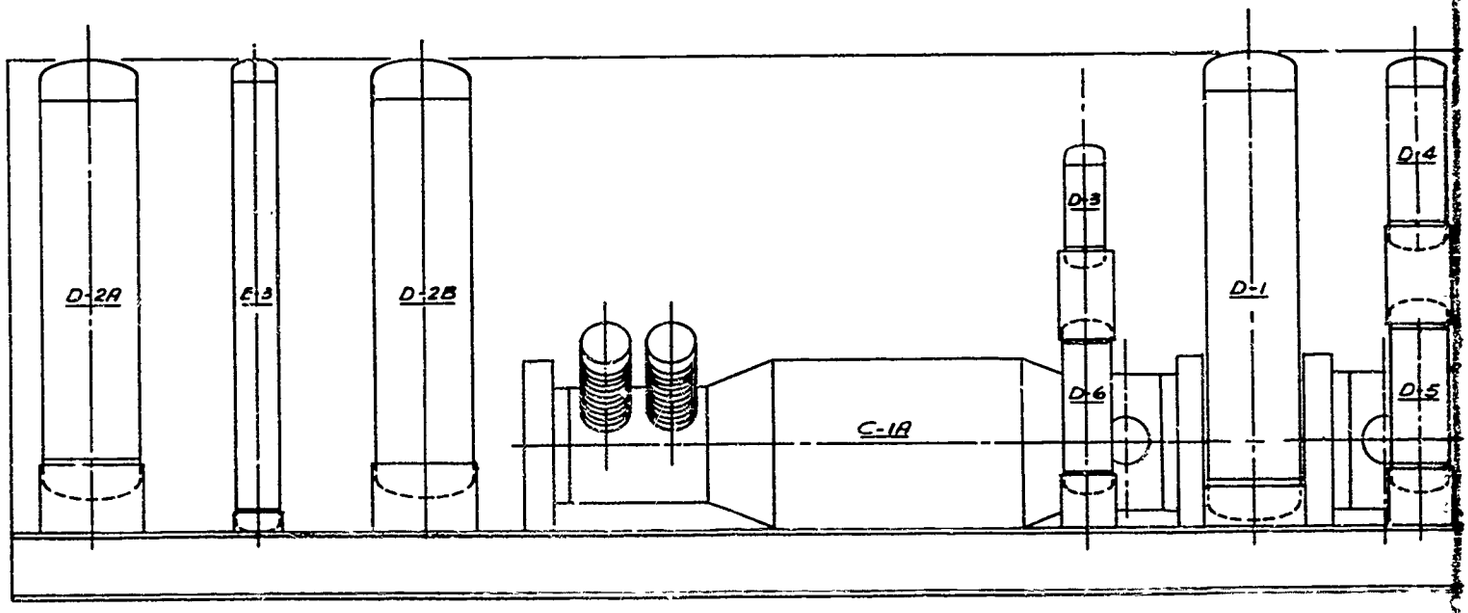
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Figure 3.1-18. Equipment layout for ammonia subsystem 1- Skid 2.

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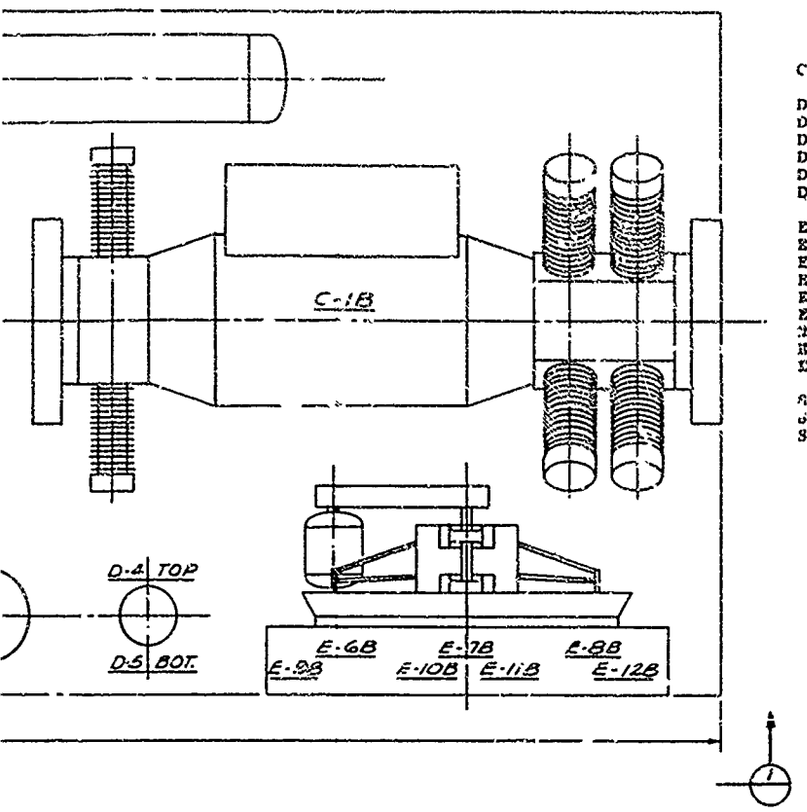


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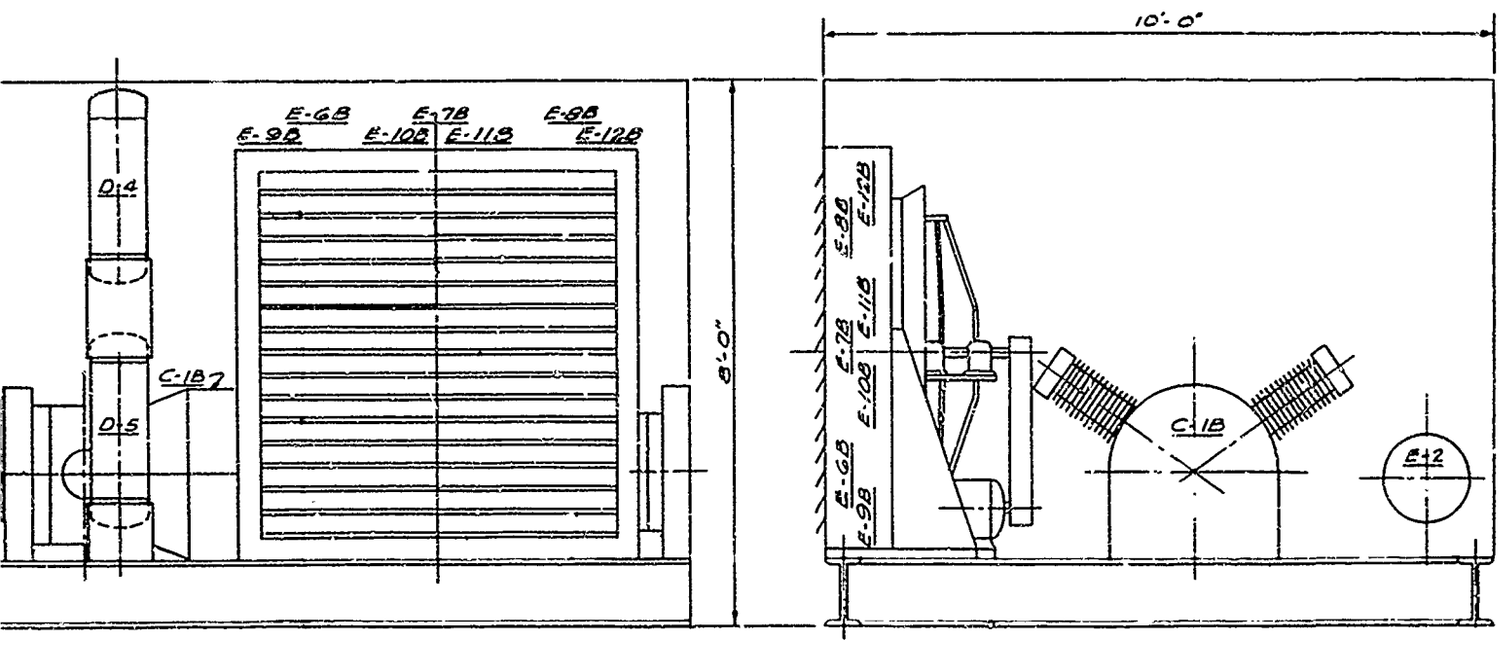
SECTION "1-1"

4852-48



Equipment legend

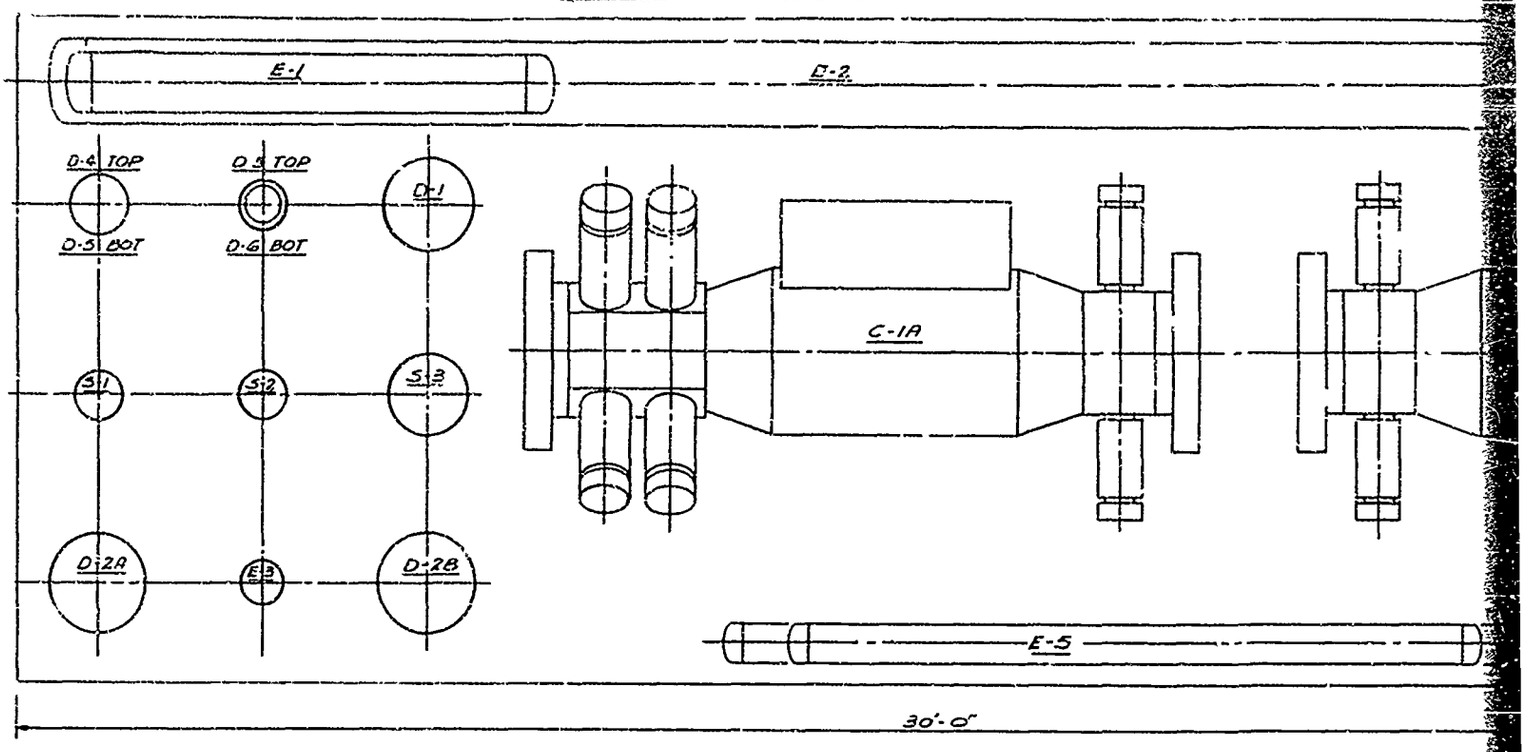
C-1A&B	Multiservice compressor
D-1	Deore unit
D-3A&B	Ammonia converter
D-4	Gas filter
D-5	Knock-out drum
D-6	Knock-out drum
	Gas filter
E-1	Secondary condenser
E-2	Converter heat exchanger
E-3	Ammonia vaporizer
E-4	Primary condenser
E-5	Product condenser
E-6A&B through E-12A&B	Compressor intercoolers & aftercoolers
E-13A&B	Hydrogen precooler
S-1	Primary separator
S-2	Secondary separator
S-3	Product separator



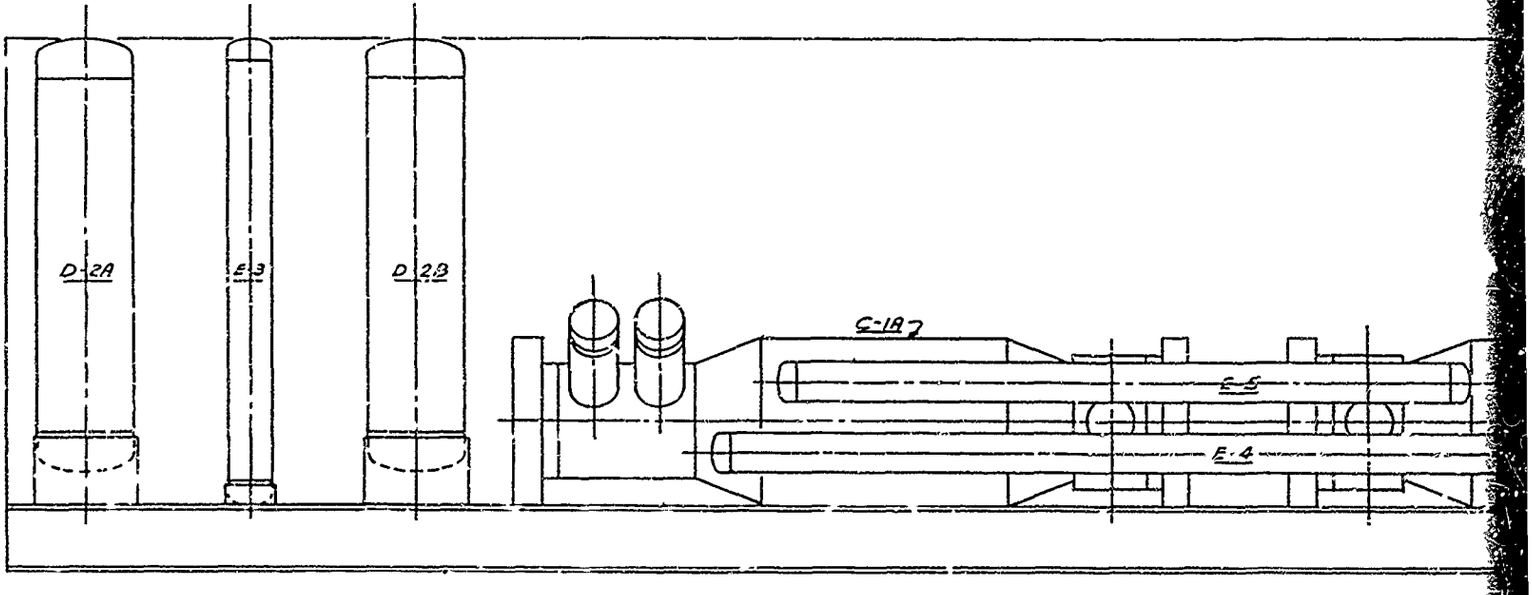
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Figure 3.1-20. Equipment layout for ammonia subsystems 5, 7, and 8.

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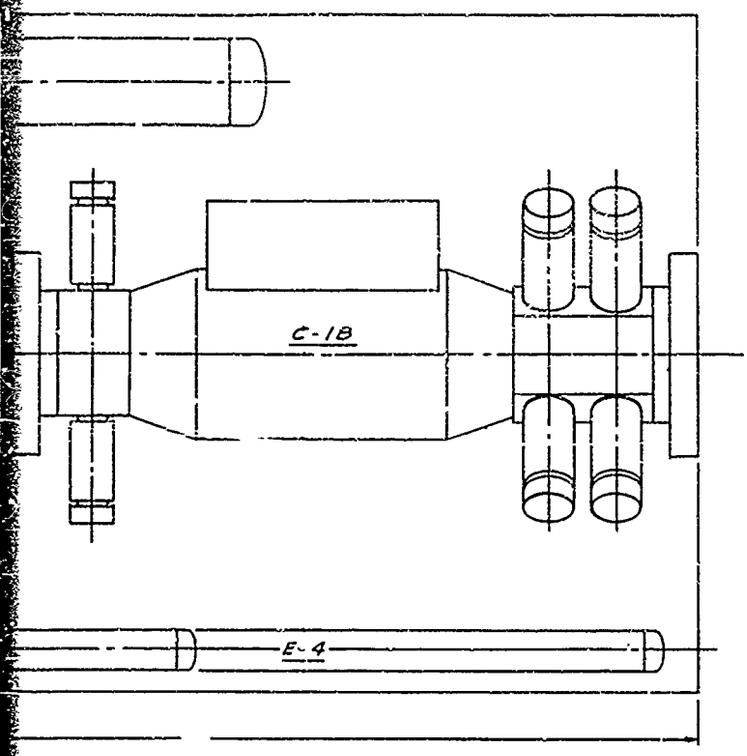


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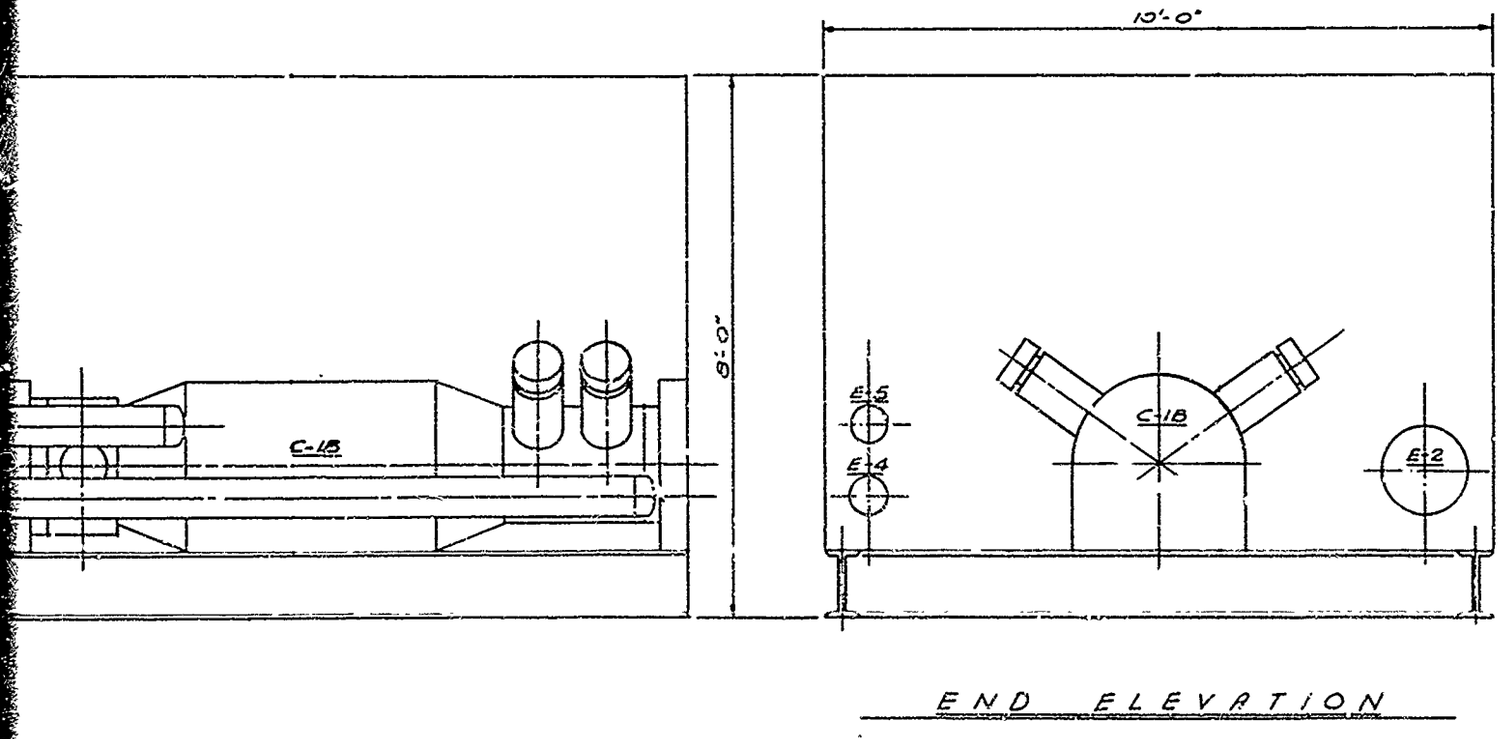
4496-49

SIDE ELEVATION



Equipment legend

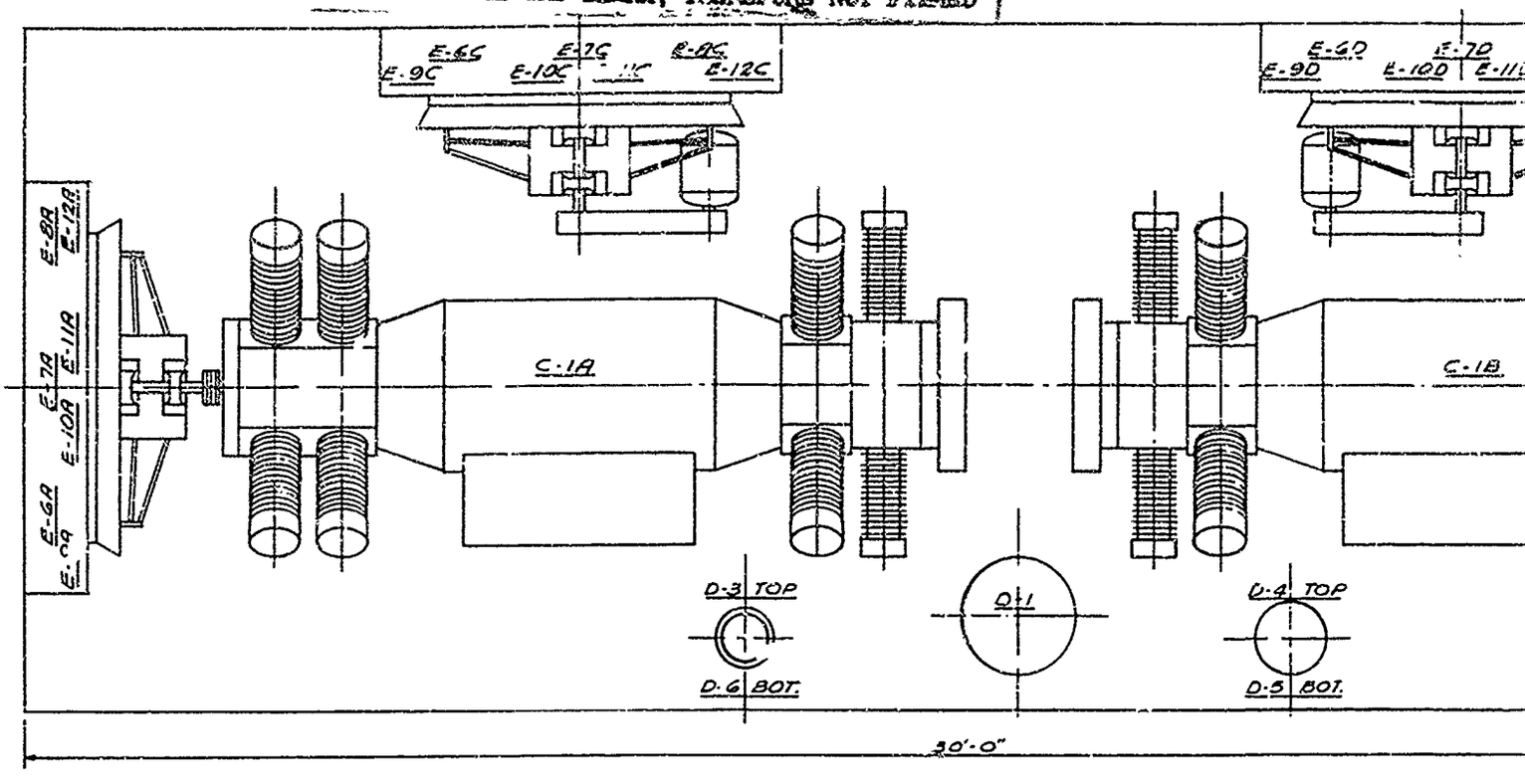
C-1A&B	Multistage compressor
D-1	Deoxo unit
D-2A&B	Ammonia converter
D-3	Gas filter
D-4	Knock-out drum
D-5	Knock-out drum
D-6	Gas filter
E-1	Secondary condenser
E-2	Converter heat exchanger
E-3	Ammonia vaporizer
E-4	Primary condenser
E-5	Product condenser & hydrogen precooler
S-1	Primary separator
S-2	Secondary separator
S-3	Product separator



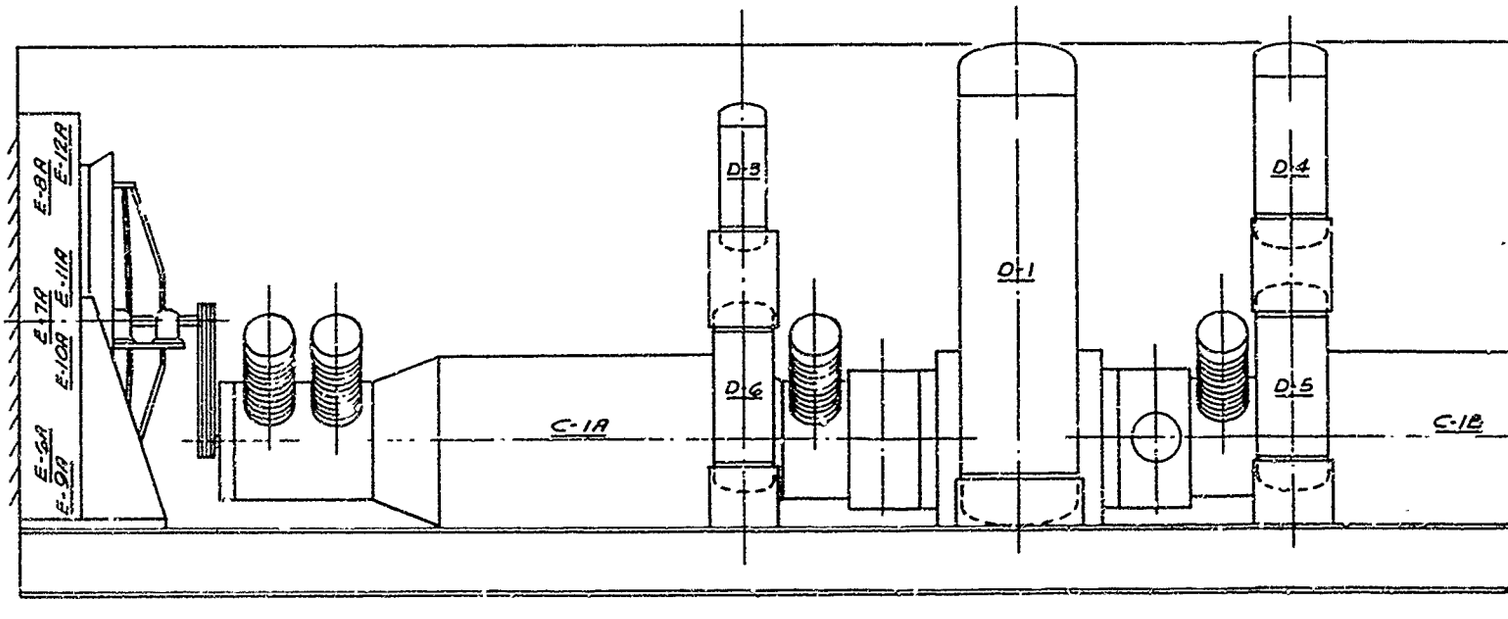
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Figure 3.1-21. Equipment layout for ammonia subsystem 6.

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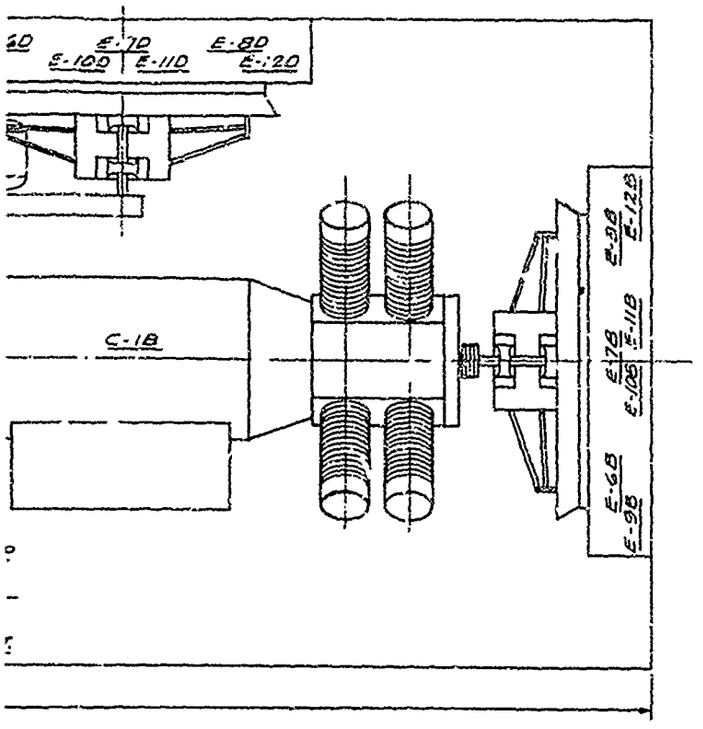


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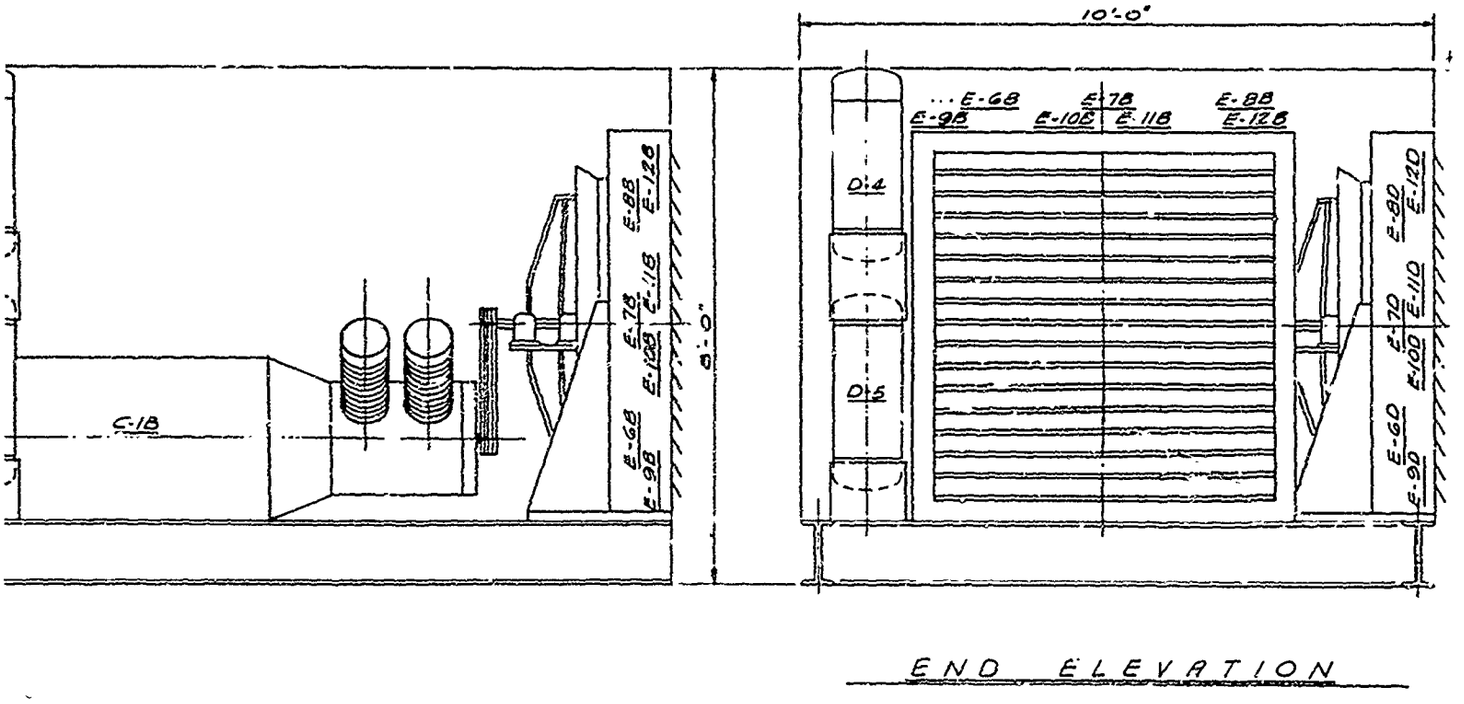
4496-50

SIDE ELEVATION



Equipment legend

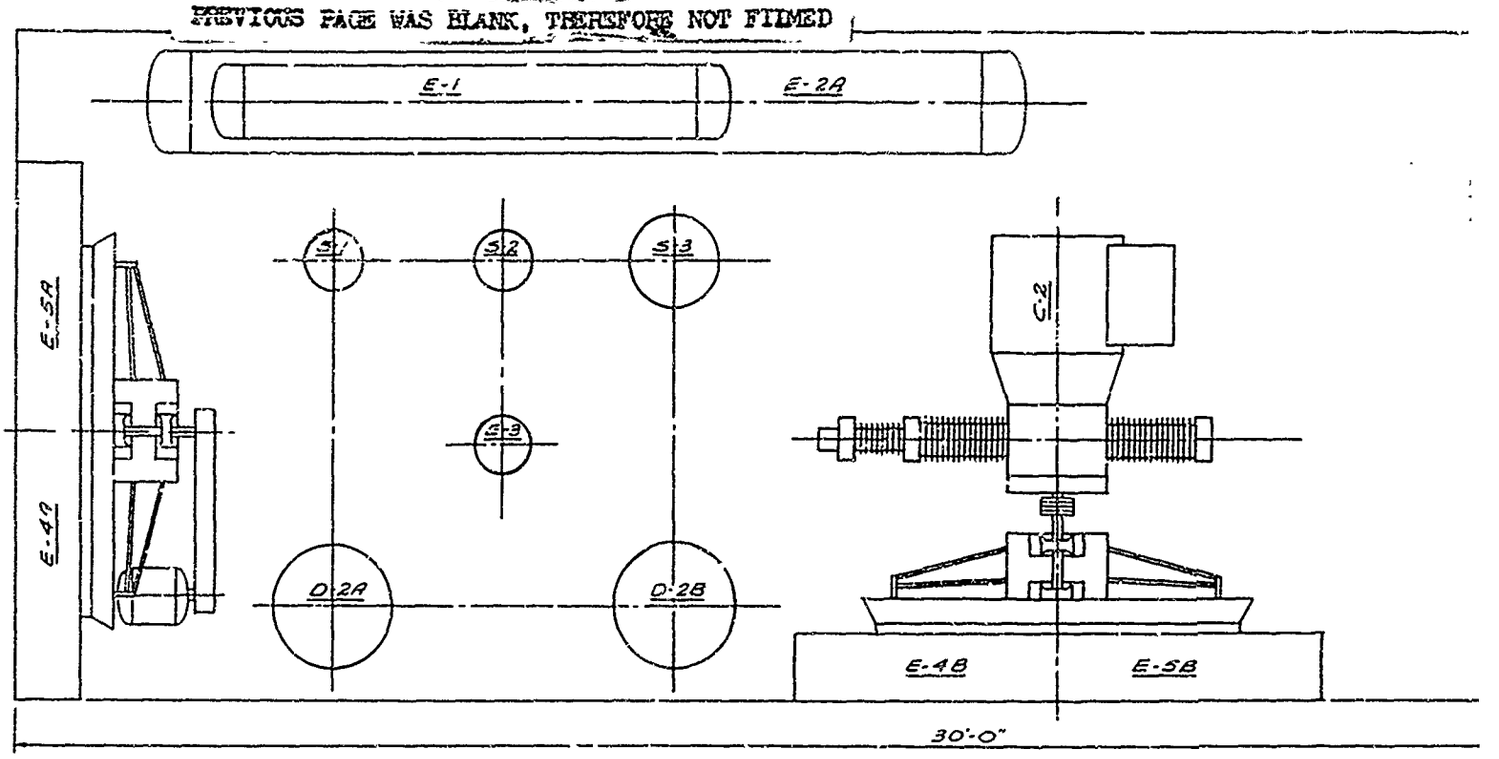
C-1A&B	Multiservice compressor
C-2	Recirculation compressor
D-1	Deoxo unit
D-2A&B	Ammonia converter
D-3	Gas filter
D-4	Knock-out drum
D-5	Knock-out drum
D-6	Gas filter
E-1	Secondary condenser
E-2A&B	Converter heat exchanger
E-3	Ammonia vaporizer
E-4A&B	Primary condenser
E-5A&B	Product condenser
E-6A, B, C, D through E-11A, B, C, D	Compressor intercooler & aftercoolers
E-12A, B, C, D	Hydrogen precooler
S-1	Primary separator
S-2	Secondary separator
S-3	Product separator



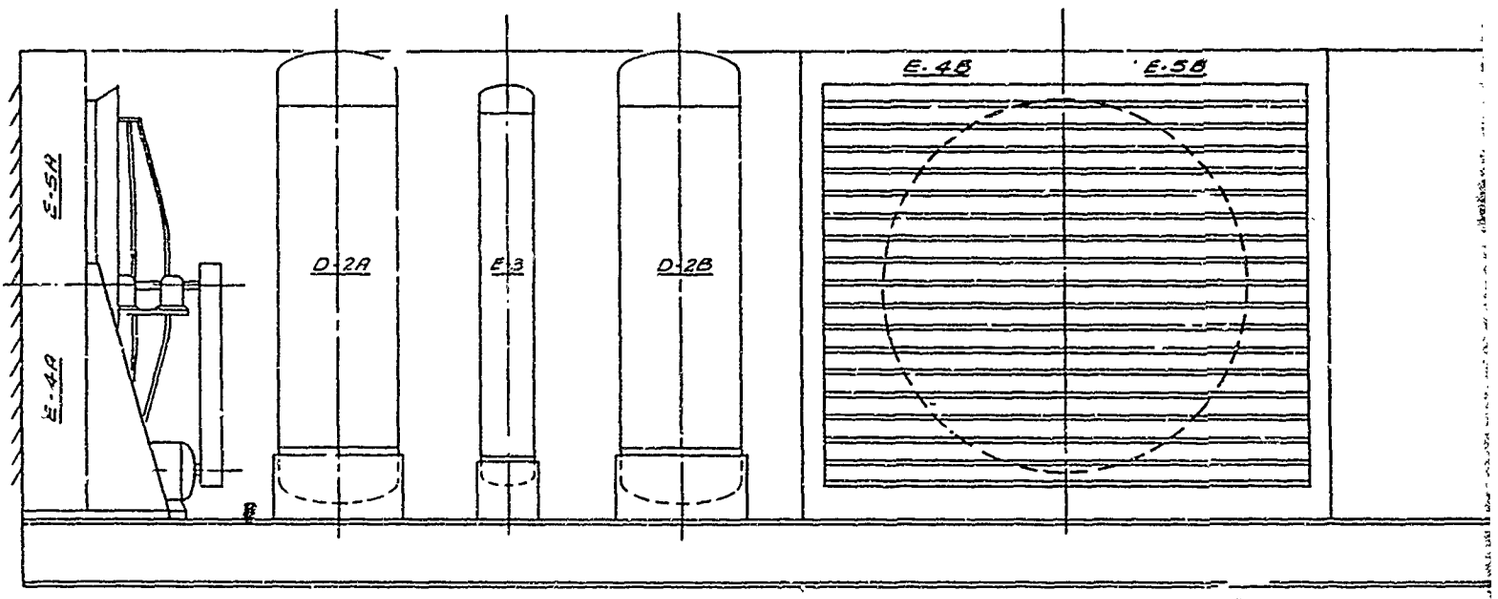
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Figure 3.1-22. Equipment layout for ammonia subsystems 5 and 10—Skid 1.

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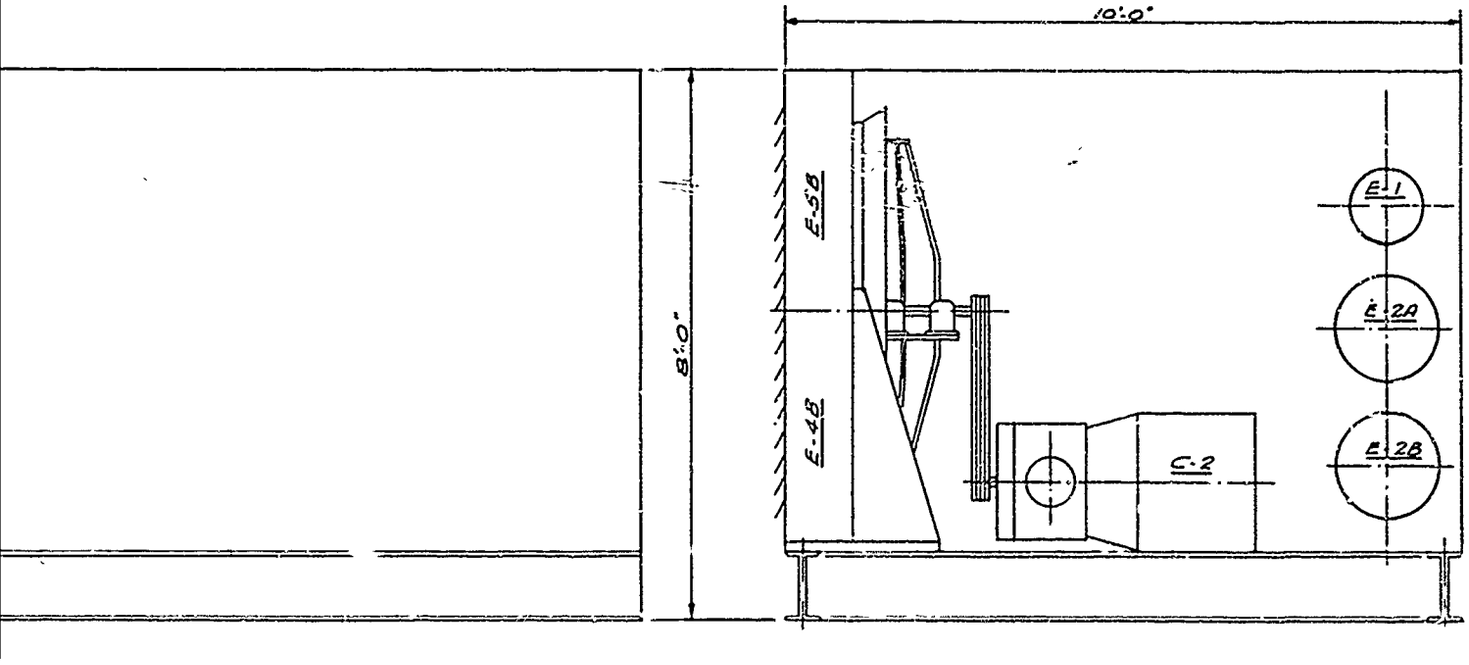


SIDE ELEVATION

4496-51

Equipment legend

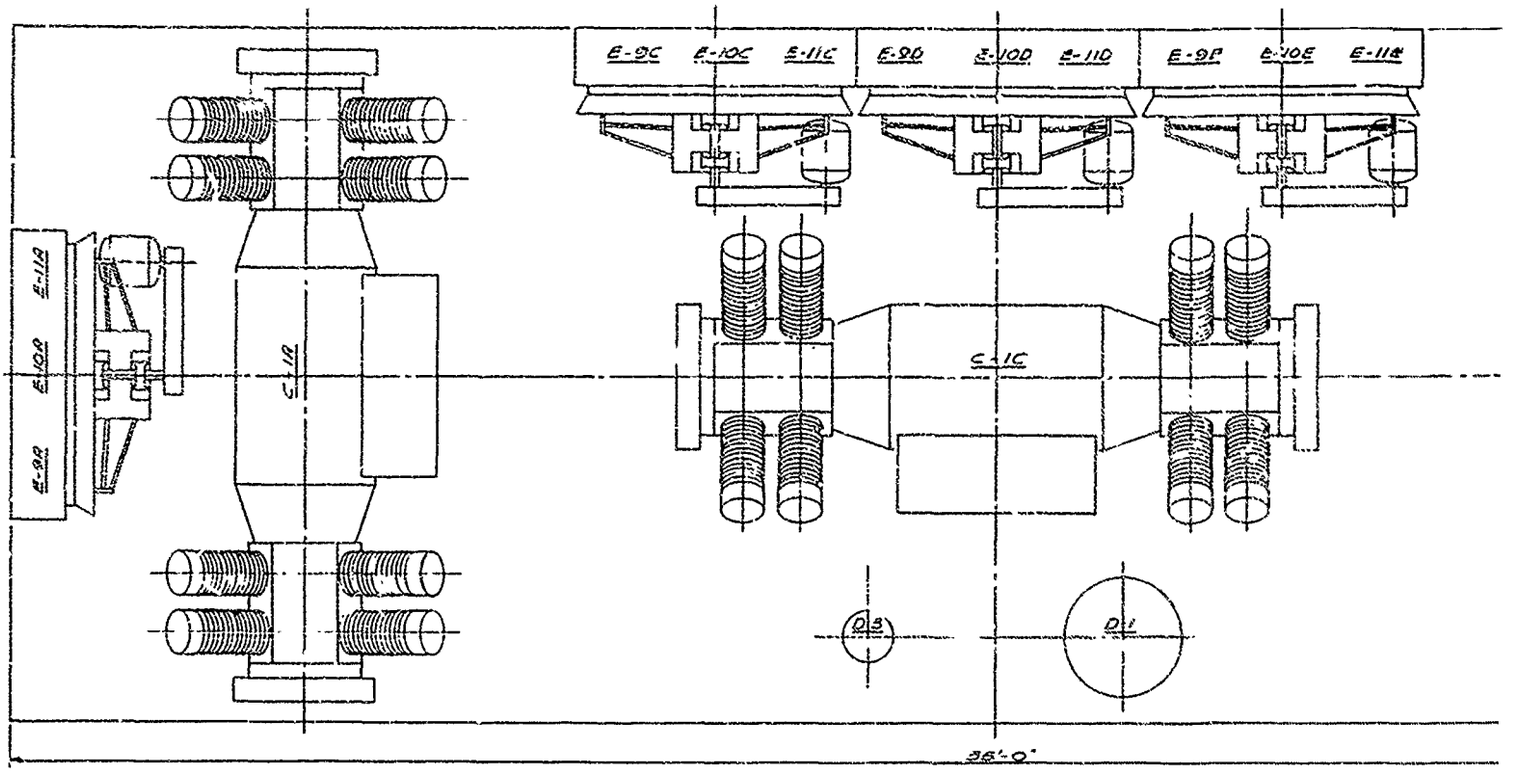
C-1A&B	Multiservice compressor
C-2	Recirculation compressor
D-1	Deoxo unit
D-2A&B	Ammonia converter
D-3	Gas filter
D-4	Knock-out drum
D-5	Knock-out drum
D-6	Gas filter
E-1	Secondary condenser
E-2A&B	Converter heat exchanger
E-3	Ammonia vaporizer
E-4A&B	Primary condenser
E-5A&B	Product condenser
E-6A, B, C, D through E-11A, B, C, D	Compressor intercooler & aftercoolers
E-12A, B, C, D	Hydrogen precooler
S-1	Primary separator
S-2	Secondary separator
S-3	Product separator



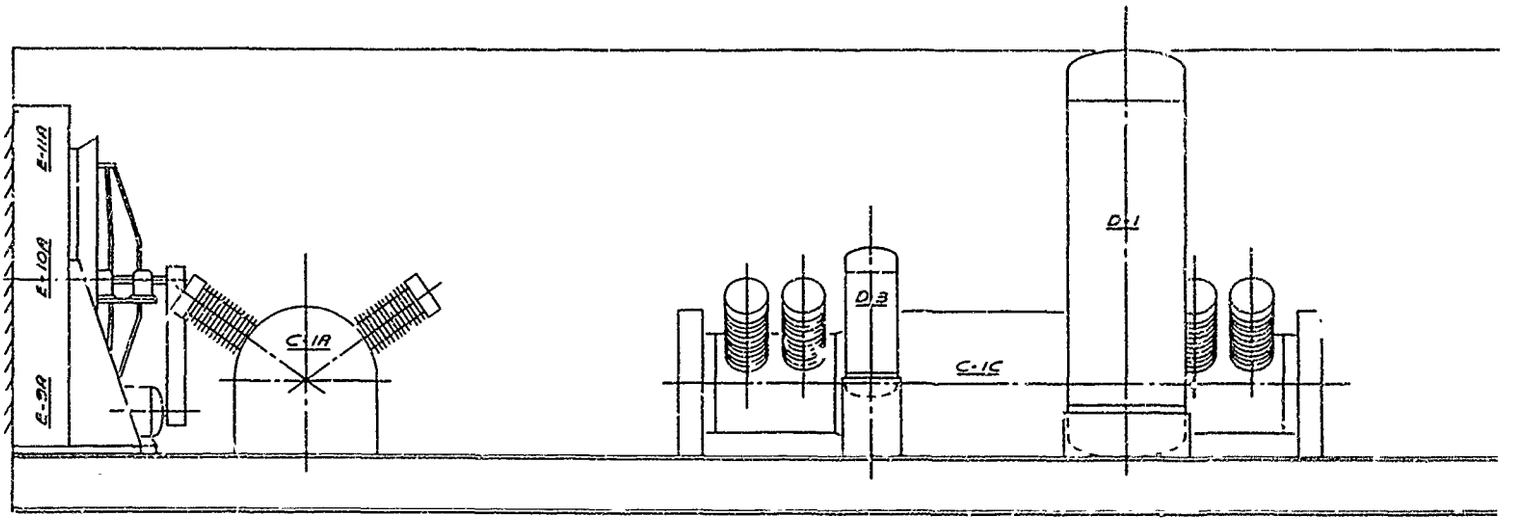
END ELEVATION

Figure 3.1-23. Equipment layout for ammonia subsystems 9 and 10 - Sht 2.

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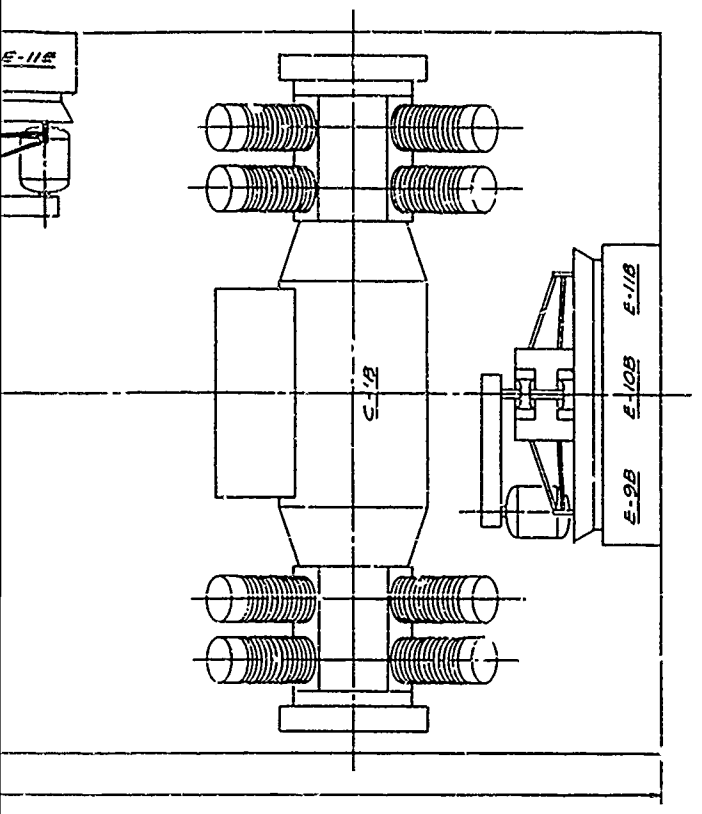


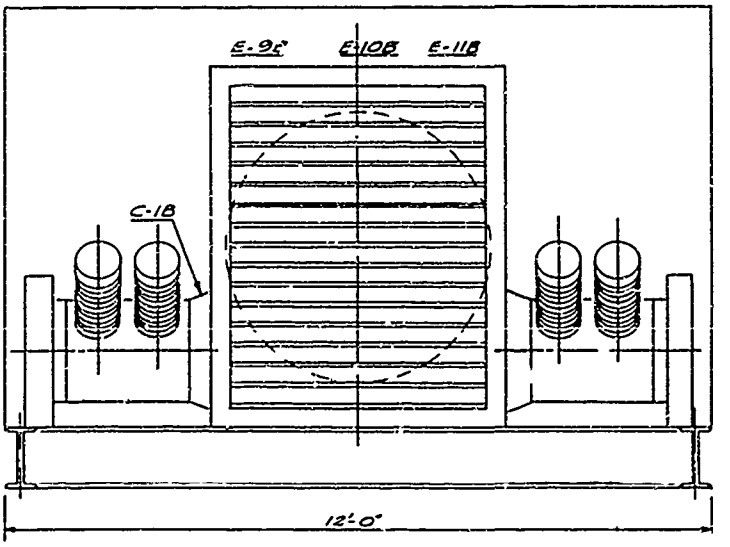
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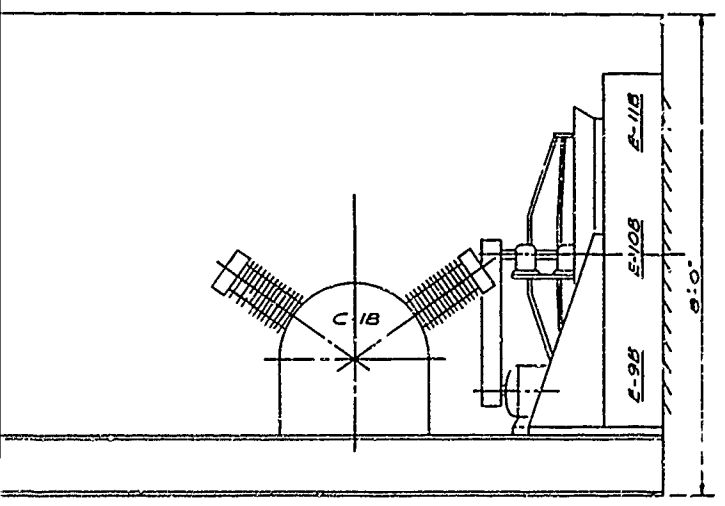
4496-53

SIDE ELEVATION





END ELEVATION

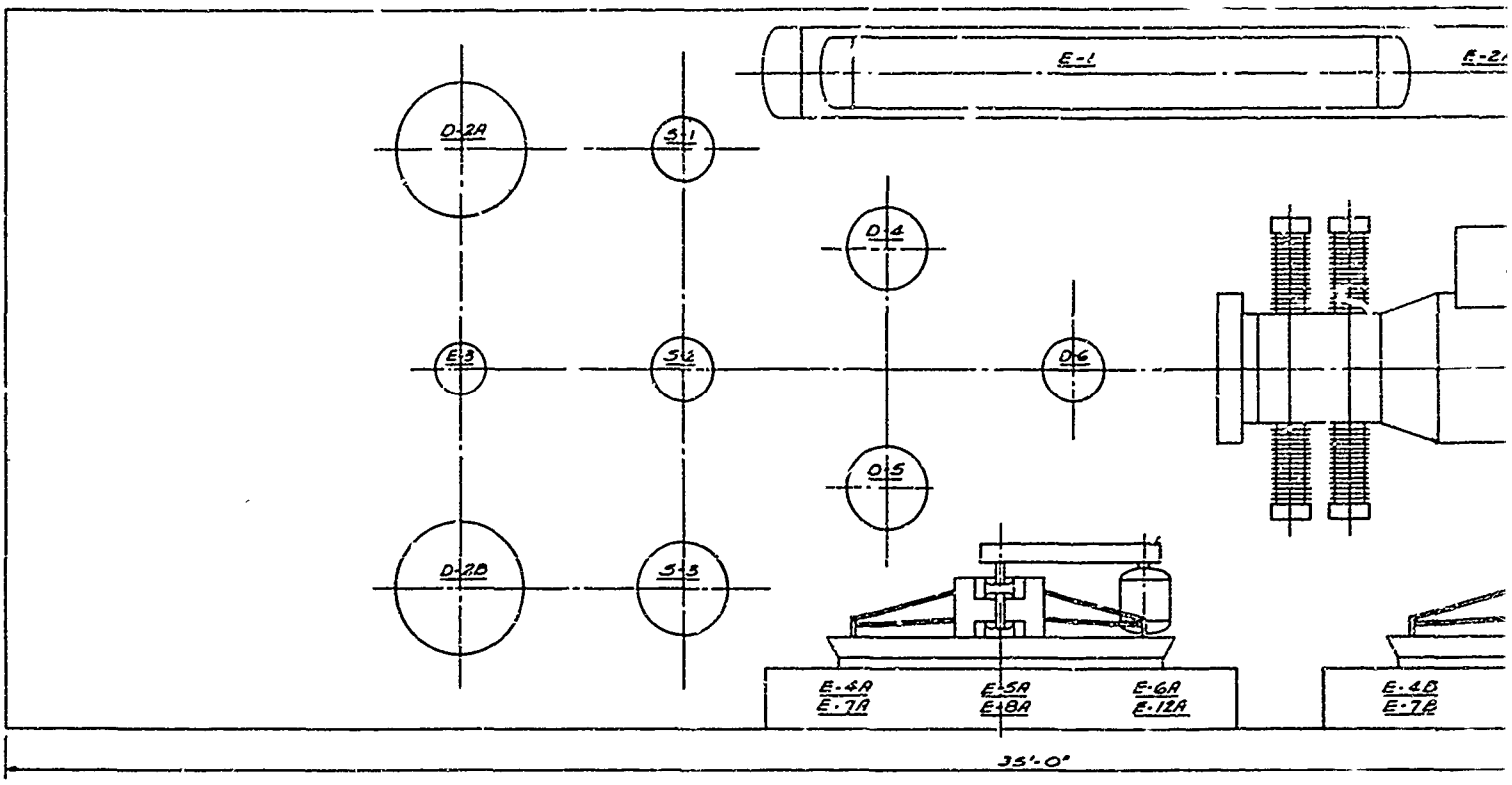


Equipment legend

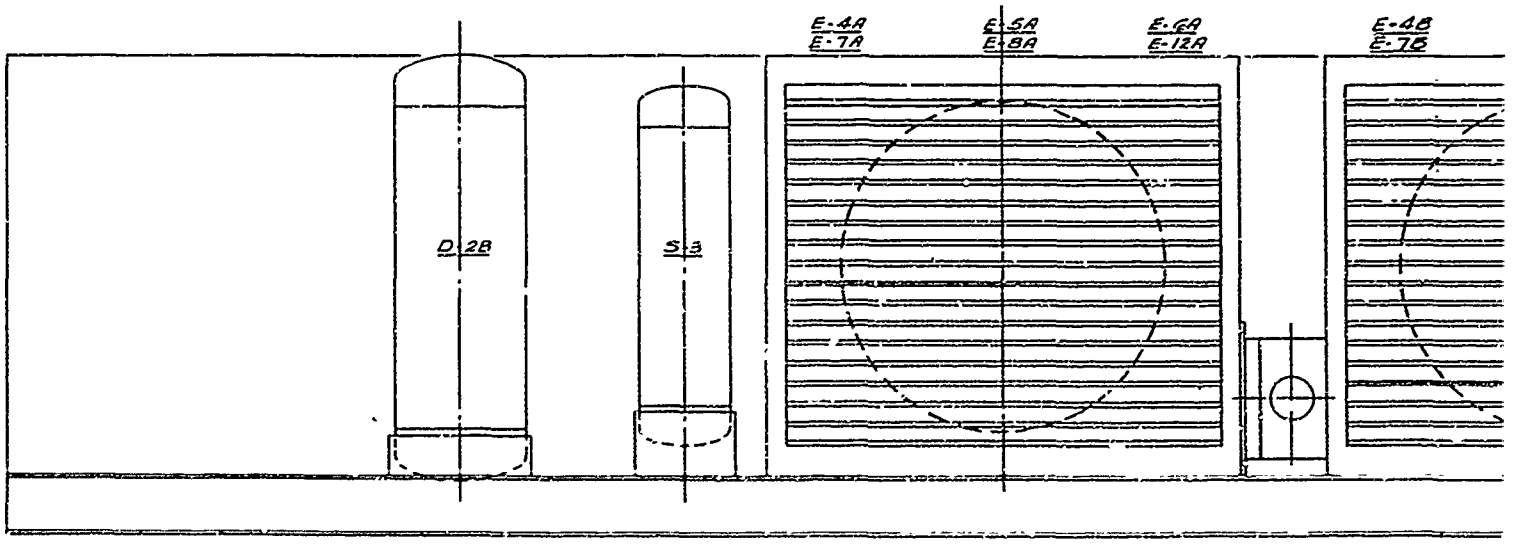
C-1A, B, C	Synthesis gas compressor
C-2	Nitrogen compressor
D-1	Deoxo unit
E-2A&B	Ammonia converter
D-3	Gas filter
D-4	Knock-out drum
D-5	Knock-out drum
D-6	Gas filter
E-1	Secondary condenser
E-2A&B	Converter heat exchanger
E-3	Ammonia vaporizer
E-4A, B, C	Primary condenser
E-5A, B, C	Product condenser
E-6A, B, C through E-8A, B, C	Nitrogen compressor intercooler & aftercooler
E-9A, B, C, D, E through E-11 A, B, C, D, E	Synthesis g s compressor Intercooler & aftercooler
E-12A, B, C	Hydrogen precooler
S-1	Primary separator
S-2	Secondary separator
S-3	Product separator

Figure 3.1-24. Equipment layout for ammonia subsystem 11—Skid 1.

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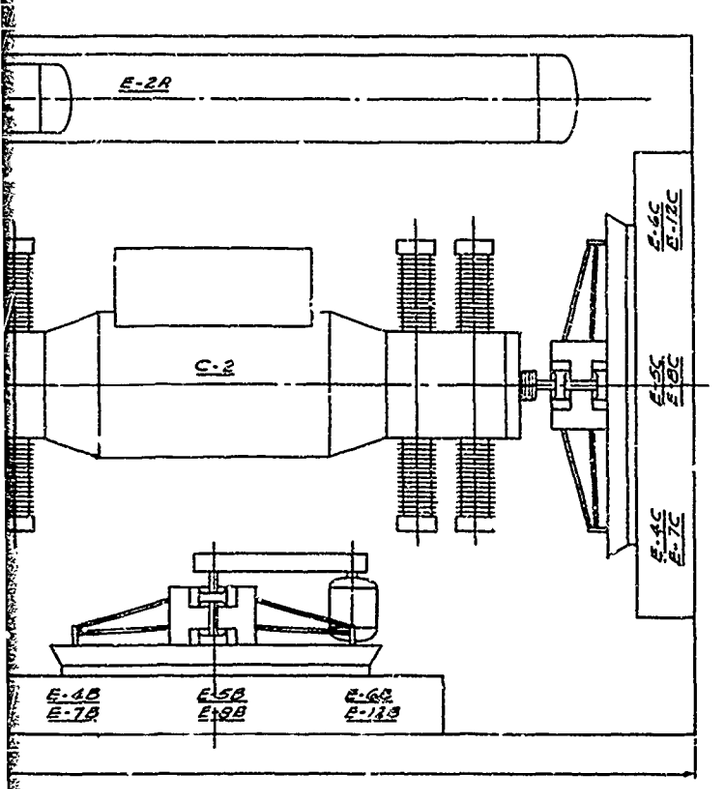


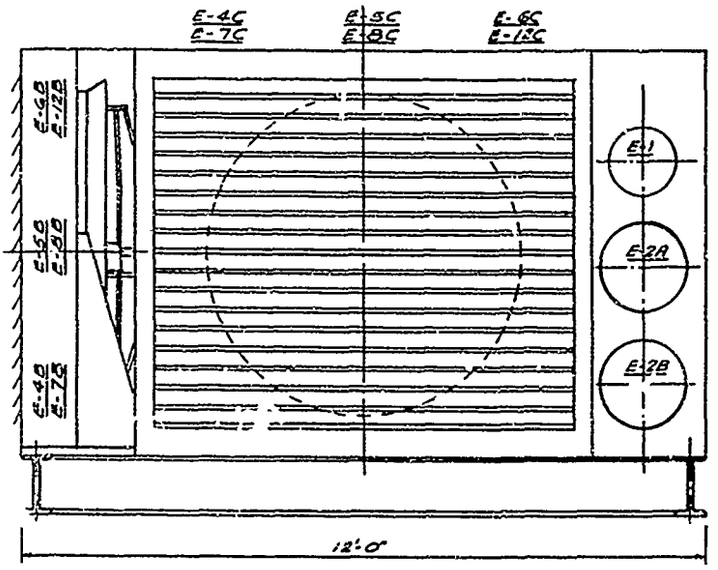
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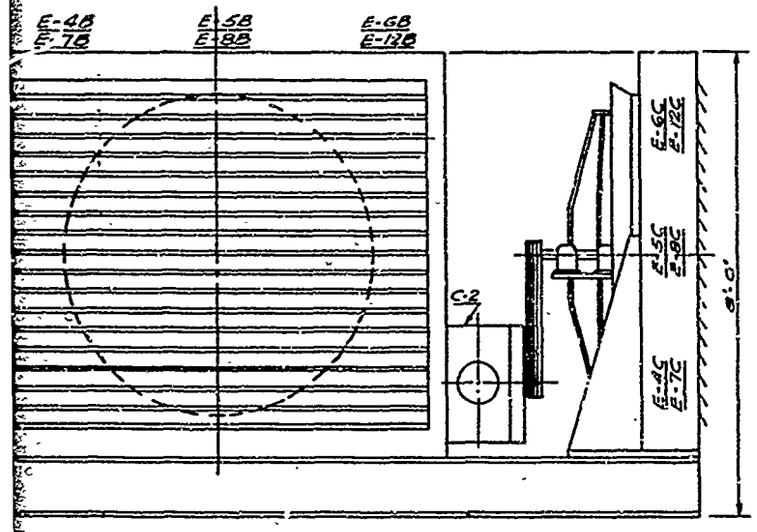
4426-53

SIDE ELEVATION





END ELEVATION



Equipment legend

C-1A, B, C	Synthesis gas compressor
C-2	Nitrogen compressor
D-1	Deoxo unit
D-2A, B	Ammonia converter
D-3	Gas filter
D-4	Knock-out drum
D-5	Knock-out drum
D-6	Gas filter
E-1	Secondary condenser
E-2A&B	Converter heat exchanger
E-3	Ammonia vaporizer
E-4A, B, C	Primary condenser
E-5A, B, C	Product condenser
E-6A, B, C through E-8A, B, C	Nitrogen compressor intercooler & aftercooler
E-9A, B, C, D, E through E-11A, B, C, D, E	Synthesis gas compressor intercooler & aftercooler
E-12A, B, C	Hydrogen precooler
S-1	Primary separator
S-2	Secondary separator
S-3	Product separator

Figure 3.1-25. Equipment layout for ammonia subsystem 11—Skid 2.

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Equipment Weight

Table 3.1-IV presents the weight of the various pieces of equipment of the eleven cases. Included are the weights of the instruments, piping, and structural steel. For the cases where two skids are used, the weights have been separated accordingly.

PREDICTED PERFORMANCE

The utilities required for the ammonia synthesis subsystem consist of pure hydrogen and nitrogen, electricity, and instrument air. Water for cooling is also required for Subsystems 2 and 6. The water must be available at the plant boundary at 50 psia to ensure proper flow through the cooling equipment. The utilities are summarized in Table 3.1-V.

In subsystems 2 and 6, which are water-cooled, the power requirement is slightly lower than for air-cooling because of a slightly higher compressor efficiency with water-cooling.

The heat rejection rates of the various coolers are presented in Table 3.1-VI. Also shown are the cooling water requirements for the water-cooled subsystems. The compressor inter and aftercooler heat load is slightly lower for the water-cooled subsystems because of the slightly higher compressor efficiency.

DISCUSSION

The catalyst for the ammonia synthesis reaction can be one of several different kinds. The one considered in the design calculations is type KM-I manufactured by Haldor-Topsoe Corporation. This is not to infer that only this catalyst would perform satisfactorily. Experience has shown that this catalyst has been active, had long life, and exhibited a high degree of ruggedness. However, it has not been used in service where the cycling has been as frequent as anticipated for the mobile Energy Depot. Upon discussion of this problem with a manufacturer, the opinion was offered that only by actual testing could the effect of rapid cycling be determined. The effects are expected to take the form of physical damage or break up the catalyst's particles rather than a reduction of catalyst activity. If physical destruction occurs, there should be a measurable increase in the pressure drop across the converter. When this pressure drop becomes too great, it would be necessary to change the catalyst.

There are many factors which govern the quantity of catalyst needed for a given operation—pressure and temperature of operation, inert level, inlet ammonia concentration, etc. In arriving at a suitable catalyst volume, it is necessary to take into account all of these factors and then add a sufficient excess of catalyst (approximately 15%) to account for uncertainties and assure a reasonable life. The active life of the catalyst is expected to be in excess of three years.

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DISCUSSION

The catalyst for the ammonia synthesis reaction can be one of several different kinds. The one considered in the design calculations is type KM-I manufactured by Haldor-Topsoe Corporation. This is not to infer that only this catalyst would perform satisfactorily. Experience has shown that this catalyst has been active, had long life, and exhibited a high degree of ruggedness. However, it has not been used in service where the cycling has been as frequent as anticipated for the mobile Energy Depot. Upon discussion of this problem with a manufacturer, the opinion was offered that only by actual testing could the effect of rapid cycling be determined. The effects are expected to take the form of physical damage or break up the catalyst's particles rather than a reduction of catalyst activity. If physical destruction occurs, there should be a measurable increase in the pressure drop across the converter. When this pressure drop becomes too great, it would be necessary to change the catalyst.

There are many factors which govern the quantity of catalyst needed for a given operation--pressure and temperature of operation, inert level, inlet ammonia concentration, etc. In arriving at a suitable catalyst volume, it is necessary to take into account all of these factors and then add a sufficient excess of catalyst (approximately 15%) to account for uncertainties and assure a reasonable life. The active life of the catalyst is expected to be in excess of three years.

Table 3.1-IV.

Ammonia synthesis subsystem equipment weight.

Subsystem	Component No.*	Subsystem No.									
		1 & 3	2	4		5, 7, 8	6	9 & 10		11	
Skid		1	1	1	2	1	1	1	2	1	2
Compressor (air-cooled)	C-1	7,900	—	15,300	—	15,300	—	20,600	—	28,950	—
Compressor (including water jacket)	C-1	—	9,450	—	—	—	18,900	—	—	—	—
Compressor (air-cooled)	Recirculation	—	—	—	—	—	—	—	2,740	—	—
Compressor (air-cooled)	Recirculation, nitrogen, & blow-by	—	—	—	—	—	—	—	—	—	9,800
Deoxo	D-1	400	400	660	—	680	680	1,095	—	1,630	—
Converter (including catalyst)	D-2 A & B	2,900	2,900	—	4,800	4,800	4,800	—	7,300	—	10,500
Filter	D-3	25	25	50	—	50	50	85	—	150	—
Knock-out drum	D-4	60	60	120	—	120	120	165	—	—	245
Knock-out drum	D-5	60	60	120	—	120	120	165	—	—	245
Filter	D-6	200	200	355	—	355	355	—	610	—	1,090
Secondary condenser	E-1	475	475	—	770	770	770	—	1,175	—	1,710
Converter heat exchanger	E-2	4,900	4,900	—	8,800	8,380	8,380	—	13,500	—	18,600
Ammonia vaporizer	E-3	300	300	—	465	465	465	—	690	—	970
Fin-fan air-cooled unit	E-6-11	1,600	—	—	—	—	—	—	—	—	—
Fin-fan air-cooled unit	E-4, 5, & 12	2,400	—	—	—	—	—	—	—	—	—
Fin-fan air-cooled unit	E-6-11	—	—	2,400	—	—	—	—	—	6,000	—
Fin-fan air-cooled unit	E-6-5 & 12	—	—	900	—	—	—	—	—	—	—
Fin-fan air-cooled unit	E-4 & 5	—	—	—	4,800	4,800	—	—	9,600	—	—
Fin-fan air-cooled unit	E-6-12	—	—	—	—	3,200	—	4,800	—	—	—
Fin-fan air-cooled unit	E-4-8 & 12	—	—	—	—	—	—	—	—	—	14,400
Primary condenser (water-cooled)	E-4	—	185	—	—	—	360	—	—	—	—
Product cooler (water-cooled)	E-5 & 12	—	110	—	—	—	210	—	—	—	—
Ejector	EJ-1	20	20	—	35	35	35	—	55	—	55
Separators	S-1 & 2	700	700	—	1,140	1,140	1,140	—	1,920	—	3,200
Separator	S-3	225	225	—	330	330	330	—	470	—	585
Fan & motor drive		250	—	750	250	750	—	500	250	1,250	500
Instruments		1,500	1,500	1,000	500	1,500	1,500	900	600	500	1,000
Piping		2,800	2,800	3,100	3,200	5,200	5,000	4,200	5,690	5,700	9,300
Structure		2,500	2,500	2,500	2,500	3,500	3,500	3,000	3,500	4,200	6,000
Total		29,215	26,810	27,275	27,590	51,495	46,715	37,130	47,400	48,380	79,200

*See Figure 3.1-14.

Table 3.1-V.
Ammonia synthesis subsystem summary of utilities.

Subsystem	Power (kw)	Hydrogen (scfh)	Nitrogen (scfh)
1	230	20,400	6,850
2	220	20,400	6,850
3	230	20,400	6,850
4	461	40,800	13,700
5	459	40,800	13,700
6	440	40,800	13,700
7	459	40,800	13,700
8	459	40,800	13,700
9	681	68,000	22,800
10	681	68,000	22,800
11	1145	102,000	34,250

Table 3.1-VI.
Ammonia synthesis subsystem heat rejection (BTU/hr).

Subsystem	E-4*	E-5*	E-6 thru 11*	E-12*	Total	Cooling water (gph)
1	478,000	525,000	464,000	17,500	1,484,500	—
2	478,000	525,000	450,000	17,500	1,470,500	5,350
3	478,000	525,000	464,000	17,500	1,484,500	—
4	956,000	1,050,000	928,000	35,000	2,969,000	—
5	956,000	1,050,000	928,000	35,000	2,969,000	—
6	956,000	1,050,000	900,000	35,000	2,941,000	10,700
7	956,000	1,050,000	928,000	35,000	2,969,000	—
8	956,000	1,050,000	928,000	35,000	2,969,000	—
9	1,600,000	1,750,000	1,392,000	58,300	4,800,300	—
10	1,600,000	1,750,000	1,392,000	58,300	4,800,300	—
11	2,390,000	2,625,000	2,320,000	87,500	7,422,500	—

*See Figure 3.1-14.

The loop, while designed to operate at pressures of about 6000 psig, will normally approach this level only after the conversion catalyst has aged. When using active catalyst and operating at rated capacity, the pressure in the loop may be expected to be in the range of 5200 to 5500

psig. At lower production rates, a corresponding decrease in the loop pressure may be expected; for example, at half rate, a loop pressure of 4500 to 4800 psig may be anticipated.

The various packages are of a totally enclosed design. The side members would act as trusses. The side and end panels would be removable to permit free access and ample flow of air. The entire package could be mounted on wheels for easy moving. Leveling jacks, either hydraulically or manually operated, would assist in the alignment of the various skids.

Block valves and quick disconnects on both ends of interpackage piping connections would prevent air from entering the loop. Flexible, reinforced hoses are used to allow for some variation in the position of packages. Quick couplings would also be used on the electrical connections.

For short shutdown periods when no work is to be performed on the loop, it is not necessary to reduce the loop pressure. Under these conditions the loop pressure may be locked in at 1000 to 2000 psi to reduce the start-up time.

Power consumption for the ammonia synthesis subsystem remains relatively constant with variations in feed rate and ambient conditions. The synthesis gas compressor, which is the major load, operates at constant inlet pressure and speed; a bypass from compressor outlet-to-inlet is controlled and maintains net flow into the synthesis loop consistent with synthesis gas feed rate. Other ways to vary the capacity of an electrically driven compressor would be by using clearance pockets, valve lifters, or a variable speed drive. These alternatives would add considerable complexity to the already compact compressor unit and increase the weight. The additional complexity is not felt to warrant the savings in power.

It should be reemphasized that this ammonia loop is designed to eliminate voluntary purging of inert gases from the high pressure section of the loop. The inert level is controlled by solubility of the inert gases in the liquid ammonia product. Since the liquid and gaseous products leaving the product separators are joined together and sent to storage, any purging of inerts from the system will occur at the storage area.

This study has indicated that portable ammonia synthesis loops can be built in compliance with the specified mobility restraints. Each of the eleven subsystems that were studied is feasible, although much engineering remains to be performed on the equipment prior to actual construction. No development work will be needed since commercial processes have been used in this design. However, testing of a prototype unit will be required.

The air-cooled cases, though requiring slightly more power, appear preferable to the water-cooled cases for several reasons. Fouling of the coolers will become a factor when below standard water is used for extended periods of time. There will be no water treatment, other than filtering for the cooling water. However, the fouling of the heat transfer surface of the

coolers has been minimized by limiting the maximum cooling water temperature to 110°F. The availability of cooling water at temperatures below freezing also appears to present a problem.

It can also be concluded from this study that some means of refrigeration must be supplied to the ammonia storage system. If the ammonia temperature rises to 125°F, the ammonia vapor pressure will be 308 psia. At this pressure, it will be necessary for the pressure in the product cooler and separator to rise to maintain flow from the separator to the product storage.

For the larger plants, multiples of the basic compressor unit have been used. Even though this is less desirable from a weight and cost standpoint than a single compressor unit for each size plant, it is probably better from a reliability standpoint. A 20% weight reduction and a 2% increase in efficiency might be possible by developing a single unit for the larger plants rather than using multiple smaller units. Also, compressors which are generally available for use on a mobile Energy Depot system are not designed with lightweight requirements in mind. Thus, an additional 20% weight reduction might accrue with a compressor designed specifically for the Energy Depot application. A reduction in the number of skids from two to one for the larger ammonia systems may also be possible with a large single compressor.

A trade-off analysis indicates that the savings which can be made by using a large single compressor, rather than multiple smaller units, will more than offset the cost of a compressor development program. This analysis is reported in Section 3.8. Therefore, it appears that additional studies should be made. These studies should include the preliminary design of a large, lightweight compressor to fit specific dimensional mobility restraints and a trade-off analysis which considers the effect of this unit on overall system weight, mobility, cost, and reliability.

CALCULATIONS

The calculations for designing an ammonia loop are quite numerous and involved, especially when multiple cases are to be considered. The following calculations are typical of those used for the various cases in this design work.

Ammonia Loop Material Balance

The following material balance is for 600 lb/hr of ammonia. Simple factors applied to the results will give balances for other capacities.

- Given:
  - Feed 3 mols of H_2 to 1 mol of N_2
    N_2 purity = 99.44% with 0.56% Ar
  - Loop Pressure = 6000 psia
  - Product = 600 lb/hr NH_3

- Method: Girdler Computer Program CP-2
- Data input:

Mol fraction Ar in feed	0.0014
Synthesis equilibrium temperature	932°F
Fraction of theoretical conversion	0.65
Inert mol fraction	0.0515
Secondary condenser temperature	75°F
Primary condenser temperature	110°F
Compressor discharge pressure	6130 psia
Equipment and line pressure drop	200 psi

- Data output:

Process stream	Process stream conditions (lb mols/hr) H_2	N_2	Ar	NH_3	Total
Recycle	146.8	48.9	13.4	22.2	231.3
Make-up	53.7	17.9	0.1	—	71.7
Secondary condenser feed	200.5	66.8	13.5	22.2	303.0
Secondary separator—liquid product	0.14	0.05	0.01	5.93	6.13
Secondary separator—vapor product	200.3	66.8	13.5	16.3	296.9
Converter No. 2 product	147.4	49.2	13.5	51.5	261.6
Primary separator—liquid product	0.6	0.3	0.1	29.3	30.3
Primary separator—vapor product	146.8	48.9	13.4	22.2	231.3

The computer program subroutines take into account the following:

- Effect of temperature on the equilibrium constant
- Effect of temperature and pressure on the fugacity of H_2, N_2, and NH_3
- Effect of inerts and pressure on the equilibrium ammonia content
- Solubility of H_2, N_2, and Ar in liquid NH_3
- Ammonia vapor in H_2 and N_2 mixtures over liquid NH_3

Ammonia Loop Enthalpy Balance

The enthalpy of the gas or liquid streams can be calculated at any place in the loop by designating the temperature, pressure, and composition. The following example indicates the method relating to the inlet and exit conditions of the primary condenser.

- Method: Girdler Computer Program CP-2
- Data input: Same as for ammonia loop material balance plus other temperatures (110 and 240°F) at which enthalpy data is desired.

- Data output:

Computer run	Temperature F)	State	Process stream conditions (lb moles/hr) at 6000 psia H_2	N_2	Ar	NH_3	Total	Enthalpy (BTU/hr)
1	240	Vapor	147.4	49.2	13.5	51.5	261.6	29,000,000
2	110	Vapor	146.8	48.9	13.4	22.2	231.3	23,450,000
		Liquid	0.6	0.3	0.1	29.3	30.3	5,072,000
								28,522,000

Heat Exchanger Duty

The heat exchanger and cooler duties were determined by subtracting the exit enthalpy from the inlet enthalpy as indicated in the following example for the primary condenser:

Duty = Q = Enthalpy in - Enthalpy out
= 29,000,000 - 28,522,000
= 478,000 BTU/hr

Sizing Air-Cooled Heat Exchanger

The air-cooled heat exchangers (i.e., the primary condenser) were sized according to the method and sample calculation presented in Reference 2.

Sizing Water-Cooled Heat Exchanger

The shell and tube heat exchangers were sized by using the Girdler Computer Program CP-5.

- Data input:

	Tube	Shell
Temperature in, °F	900	75
Temperature out, °F	240	725
H_2, lb moles/hr	147.4	200.3
N_2, lb moles/hr	49.2	66.8
Ar, lb moles/hr	13.5	13.5
NH_3, lb moles/hr	51.5	16.3

Diameter of shell	= 8.0 in.	No. of tubes	= 51
No. of shells	= 1	Baffle pitch	= 3.0 in.
No. of passes	= 1.0	Tube size	= 7/8 in.
Length of tubes	= 17 ft	Fouling factor	= 0.0013

Baffle cut = 0.30

- Data output:

  Tubes required = 49.98
  Coefficient = 63.5
  Duty = 1,475,000 BTU/hr
  Area = 136.5 sq ft

Program takes into account gas properties, tube, and shell pressure drop, etc. If given exchanger does not supply sufficient area, the calculations are repeated for a larger size.

Catalyst Requirement

Performance of ammonia synthesis catalyst was determined from a computer program based on information supplied by a catalyst vendor. The program takes into account the reaction kinetics and equilibrium constant, the catalyst activity, the effect of inert concentration, pressure and temperature, etc.

- Data input:

  % hydrogen % nitrogen % ammonia
  Temperature Pressure Gas flow

- Data output:

  Ammonia concentration, temperature, and catalyst volume are pointed out for increments of ammonia concentration.

Filter, Knock-out Drum, and Separator Sizing

The flow area of the filters, knock-out drums, and separators was sized for the allowable vapor velocity. The vapor velocity was calculated from the basic equation

$$\text{Vapor velocity} = K\sqrt{\frac{\rho_L - \rho_G}{\rho_G}}$$

The calculations for gas filter D-6 are

$$\text{Maximum velocity} = 0.35\sqrt{\frac{35.3 - 10.4}{10.4}} = 0.55 \text{ f}$$

where

Gas flow = 0.087 ft^3/sec
Area = 0.087/0.55 = 0.158 ft^2

The same basic equations were used for sizing filters, knock-out drums, and separators.

Ejector Sizing

The ejector was sized by the ejector vendor for the designated temperatures, pressures, and flows.

Vessel Wall Thickness

Vessel wall thicknesses (t) were calculated from the A. S. M. E. Code, Section III. An example for the converter, D-2, is as follows:

Design pressure, p = 6600 psi
Design temperature = 650°F
Allowable membrane stress, Sm = 16,600 psi
Radius, r = 5-1/2 in.

$$t = \frac{pr}{Sm + 0.5\ p}$$

$$t = \frac{6600\ (5.5)}{16,600 + 0.5\ (6600)}$$

t = 1.83 in. (Use 1-7/8-in. wall thickness.)

Compressor Horsepower

Compressor horsepower results were calculated from the standard equation as exemplified by the calculations for the synthesis gas.

Suction Pressure = 314.7 psia
Discharge Pressure = 6000 psia
Gas Flow = 28.5 ft^3/min
No. of Stages = 3
Cp/Cv = 1.40
Efficiency = 66.5%

$$THP = \frac{0.004364\ (3)\ (1.40)\ (28.5)\ (314.7)}{1.4^{-1}} \left[\left(\frac{6000}{314.7} \right)^{\frac{1.4-1}{3\ (1.4)}} - 1 \right]$$

THP = 133
BHP = 133/0.665 = 200

REFERENCES

1. Tyler, R. D., et al. Final Technical Report Ammonia Production Feasibility Study. Allison Division, General Motors. EDR 4200. November 1965.

2. Cook, E. M. "Rating Methods for Selection of Air-Cooled Heat Exchangers." Chemical Engineering. Vol 71 (3 August 1964), p. 97.

Nitrogen Generation Subsystem

DESCRIPTION

The nitrogen required for ammonia synthesis is generated in the nitrogen generation subsystem. This is done by extracting nitrogen from air by low temperature air fractionation. Air fractionation plants of this type involve the following basic operations:

- Compression of ambient air to the pressure required by the process to effect the required purification, refrigeration, and separation
- Purification to remove oil, water, and carbon dioxide from the air before or while cooling it to liquefaction temperature
- Heat exchange to conserve and recover the low level refrigeration introduced for cooling and liquefaction of air
- Refrigeration to cool the air to liquefaction temperature
- Fractionation of liquid air to separate the nitrogen from the oxygen
- Cooling by air or water for removal of heat generated in the process

A simplified schematic of the air fractionation process selected for this application is shown in Figure 3.2-1. The process, referred to as Process N in the following discussion, operates at low pressure and obtains its low temperature refrigeration through the expansion of the high pressure oxygen produced in the electrolyzer in the hydrogen production subsystem. The air is fractionated using a double column technique; the product nitrogen is pressurized using a gas compressor. Inlet ambient feed air is compressed, purified, and then cooled by the cold product streams until it partially liquefies. Additional refrigeration is supplied by the expansion of the electrolyzer oxygen. The partially liquefied air is distilled in a double-column system to produce purified nitrogen and an oxygen-rich waste gas stream. These cold product streams are used to cool the inlet air—the oxygen-rich stream is then vented and the purified nitrogen stream is delivered to the nitrogen gas compressor.

INVESTIGATION

Processes for the fractionation of air to produce nitrogen differ in their operating pressure, the method used to achieve low temperature refrigeration, the type of column used to fractionate the air, and the method used to bring the nitrogen to the pressure required by the ammonia synthesis subsystem.

Nitrogen Generation Process Evaluation

Operating Pressure

Depending on the type of refrigeration source employed, the operating pressure may range from 75 to 2000 psia. The operating pressure influences the design of the compression unit, the purification unit, and the heat exchangers.

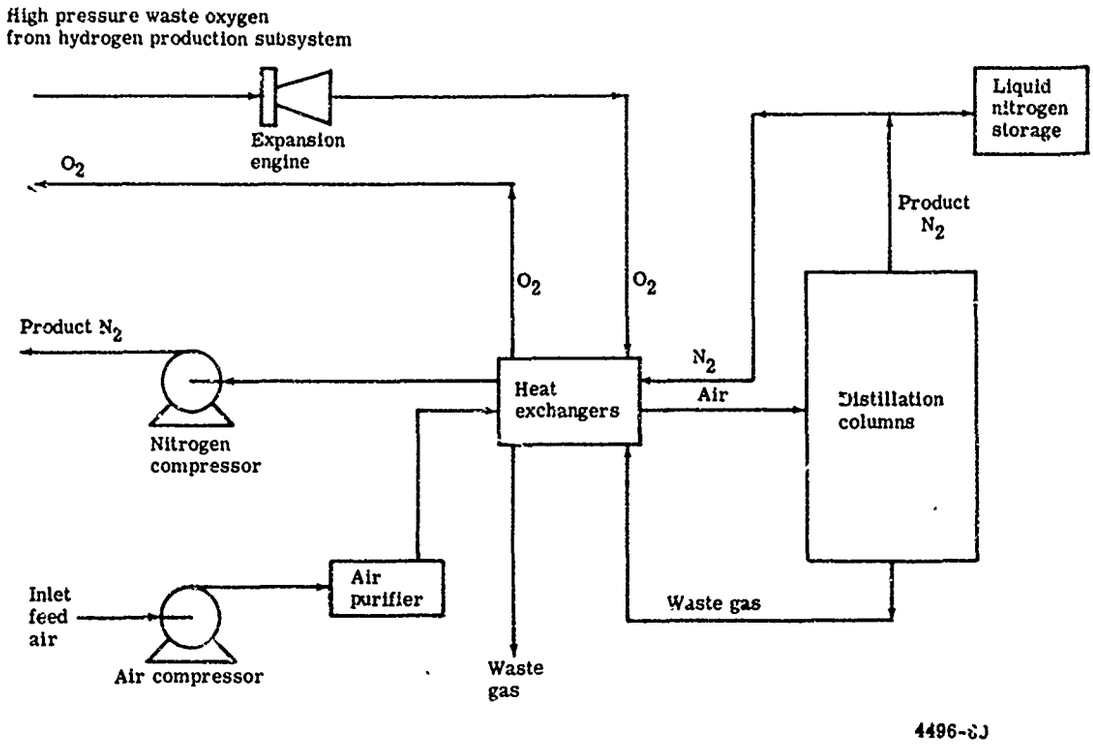


Figure 3.2-1. Simplified schematic of nitrogen generation process selected (Process N).

Refrigeration Techniques

Low temperature refrigeration for liquefying air may be achieved in a variety of ways. The four basic sources are as follows.

- Joule-Thomson Expansion—The isenthalpic expansion of air through a throttle valve or similar device leads to a decrease in air temperature. This scheme, with heat exchangers and suitable insulation to reduce heat leak, can be used to cool air to the liquefaction temperature. A relatively high operating pressure is required to obtain a useful amount of cooling with this technique.
- Expansion Engine—An isentropic expansion of air can be accomplished in an expansion engine of the reciprocating or turbine type. Such an engine is considerably more complicated than a throttle valve, but also results in considerably more cooling of the process air. Reciprocating type expansion engines are particularly suitable for high pressures and low airflows. For high gas flows at relatively low pressures, a turbine type expander is more desirable.
- Stirling Cycle Refrigeration Unit—This type of refrigeration unit combines compression, expansion, and heat exchanger components into a single unit to produce cooling. A closed loop, noncondensing gas refrigeration cycle is used. Generally, helium or hydrogen is employed as the working fluid. This type of unit can produce refrigeration at very low temperatures.

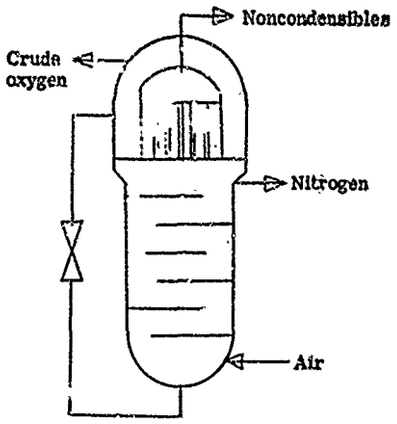
- Evaporative Type Refrigeration Unit—The evaporation of a low boiling temperature compound can be used as a source of refrigeration. Equipment is required to compress and recondense the vapors. The recondensed refrigerant fluid is returned to the evaporator to close the fluid loop. The lower temperature limit for a single fluid system is approximately -100°F. By means of a cascade system employing two or more refrigerants with successively lower boiling points, refrigeration can be supplied at temperatures below -200°F.

Air Fractionation Columns

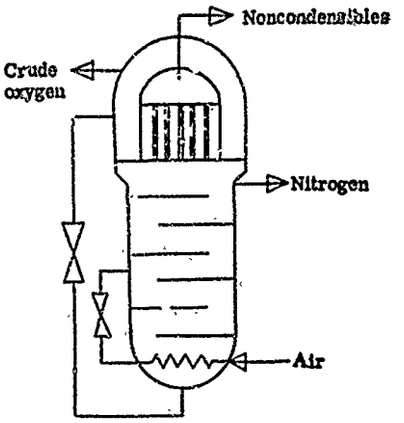
For the fractionation of liquid air, either a single or a double column may be employed. In a single column system, the nitrogen recovered from the process air is considerably less than when a double column system is used. Typical single columns for air fractionation are shown in Figure 3.2-2a and b. The feed air enters at the bottom of the column and rises through the trays against a stream of liquid reflux. As the gas ascends it becomes richer in nitrogen, the more volatile component. At the top of the column, the gas, which is essentially pure nitrogen, is partially liquefied in a condenser to provide reflux for the column. The nitrogen product may be withdrawn from the top of the column as gas or may be liquefied and withdrawn as liquid. The crude oxygen at the bottom of the column is throttled through a valve to a lower pressure and is then fed to the condenser at the top of the column to provide the refrigeration needed to condense the nitrogen. Crude oxygen vapor is withdrawn from the condenser. In the single column shown in Figure 3.2-2a, the feed air is fed directly into the bottom of the column. For this type of column, the crude oxygen bottoms contains a minimum of about 62% nitrogen and the maximum recovery of nitrogen is about 57%.

Figure 3.2-2b shows a single column with a reboiler. With this type of column, the feed air is liquefied in the reboiler and fed to an intermediate point in the column. This design requires that the feed air be at a pressure greater than the column pressure. The nitrogen in the crude oxygen may be as low as 45% and the nitrogen recovered can be as high as 78%.

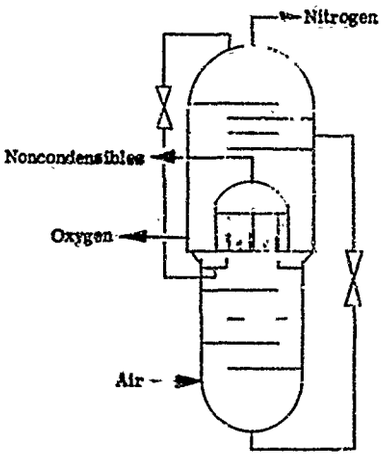
A higher nitrogen recovery can be achieved by means of a double column. A typical double column is illustrated in Figure 3.2-2c. The feed air enters at the bottom of the lower (high pressure) column and becomes enriched in nitrogen as it rises to the condenser at the top. The nitrogen is then condensed—part is used as reflux for the lower column and the remainder is fed to the upper column as reflux. The crude oxygen from the bottom of the lower column is fed to the upper column at an intermediate point. Pure nitrogen product gas is withdrawn from the top of the upper column and relatively pure oxygen gas or liquid is withdrawn from the bottom of the upper column. Typical operating pressures would be 65 psig for the lower (high pressure) column and 5 psig for the upper (low pressure) column. With a double column, the nitrogen in the oxygen product can be reduced to less than 4% and the nitrogen recovered can be in excess of 98%.



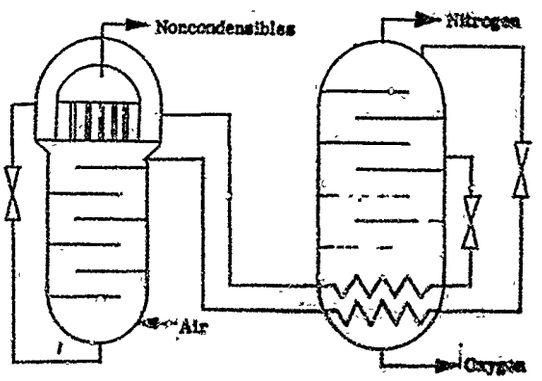
a. Single column



b. Single column with reboiler



c. Double column



d. Split double column

Figure 3.2-2. Fractionation columns used in air separation.

If the two sections of a double column are assembled vertically, as shown in Figure 3.2-2c, the column can become quite high. In cases where there are rigid height limitations, the column can be split as shown in Figure 3.2-2d. In this arrangement, the crude oxygen from the bottom of the high pressure column is fed to the condenser at the top of the column. The crude oxygen vapor from the condenser is liquefied in a condenser at the bottom of the low pressure column and is then fed as liquid to an intermediate point on the low pressure column. Nitrogen vapor from the top of the high pressure column is also liquefied in a condenser at the bottom of the low pressure column and is then fed to the top of the low pressure column as reflux. The nitrogen recovery for the split double column is essentially the same as for the standard double column, but the required pressure for the feed air is slightly higher.

Product Nitrogen Pressurization

The advantages of any particular air fractionation process depend on whether the product is to be delivered as low pressure gas, high pressure gas, or liquid. In the Energy Depot application, the nitrogen will eventually be required at elevated pressure for the ammonia synthesis. There are two ways to obtain high pressure nitrogen. The gas may be withdrawn from the air fractionation system at low pressure and then compressed or it may be pressurized as a liquid in the cold section of the air fractionation system by means of a liquid pump, warmed and vaporized in a heat exchanger, and withdrawn from the system as gas at elevated pressure.

Nitrogen Generation Processes

Several air fractionation schemes can be devised using various combinations of the previously discussed processes. Seventeen schemes were examined to isolate the one scheme exhibiting the greatest advantages. Care was taken to make certain that the effects of operating pressure, type of refrigeration, type of column, and method of pressurizing the product nitrogen were analyzed thoroughly for each of the seventeen processes studied. Code letters were used to identify these processes, as given in Table 3.2-I. Five code letters were combined as follows to identify each process.

- First letter identifies operating pressure range.
- Second and third letters identify refrigeration source.
- Fourth letter identifies type of air fractionation column.
- Fifth letter identifies nitrogen compression method.

For example, Code LEWSC specifies a low pressure process employing an expander on the waste gas stream, a single column, and a compressor to boost the nitrogen product to synthesis pressure.

Table 3.2-I.
Definition of nitrogen generation process code letters.

Characteristic	Code letter	Identification
Process air pressure	L	0-150 psig (low)
	I	150-600 psig (intermediate)
	H	600-2000 psig (high)
Refrigeration source	JO	Joule-Thomson expansion only
	JR	Joule-Thomson expansion with auxiliary refrigeration
	EA	Expansion engine operating on process air
	EW	Expansion engine operating on waste gas
	EN	Expansion engine operating on nitrogen stream
	EO	Expansion engine operating on waste oxygen from hydrogen production subsystem
	SC	Stirling cycle refrigeration unit
	CR	Cascade refrigeration
Fractionation column	S	Single column
	T	Double column
Nitrogen compression	C	Delivery of relatively low pressure gas from plant with subsequent compression
	P	Pressurizing nitrogen by means of liquid pump in air separation subsystem

Schematic diagrams of four of the seventeen processes studied are shown in Figures 3.2-3 through 3.2-6. Nitrogen generation Process A, shown in Figure 3.2-3, is a low pressure process wherein nitrogen is separated from air in a single column with the waste gas expanded to provide refrigeration. Air is compressed and purified to remove contaminants, moisture, and carbon dioxide. The air enters the exchangers where it is cooled by the separated gas streams in counterflow. Low level cooling of the air is done in Exchanger C against the low temperature, low pressure waste gas from the expander. After leaving the exchangers, the air goes to the single column where the nitrogen gas product is withdrawn overhead together with the stream of liquid nitrogen which flows to the accumulator from which it is withdrawn for plant start-up. The nitrogen gas product stream flows at essentially column pressure through the exchangers and is then pressurized by a nitrogen compressor. The waste stream flows as liquid from the bottom of the column through a valve into the condenser where it is vaporized at lower pressure and condenses the nitrogen reflux for the column. The waste gas leaving Exchanger C cools the incoming air in Exchangers B and A. The waste gas is then vented to the atmosphere at essentially ambient conditions.

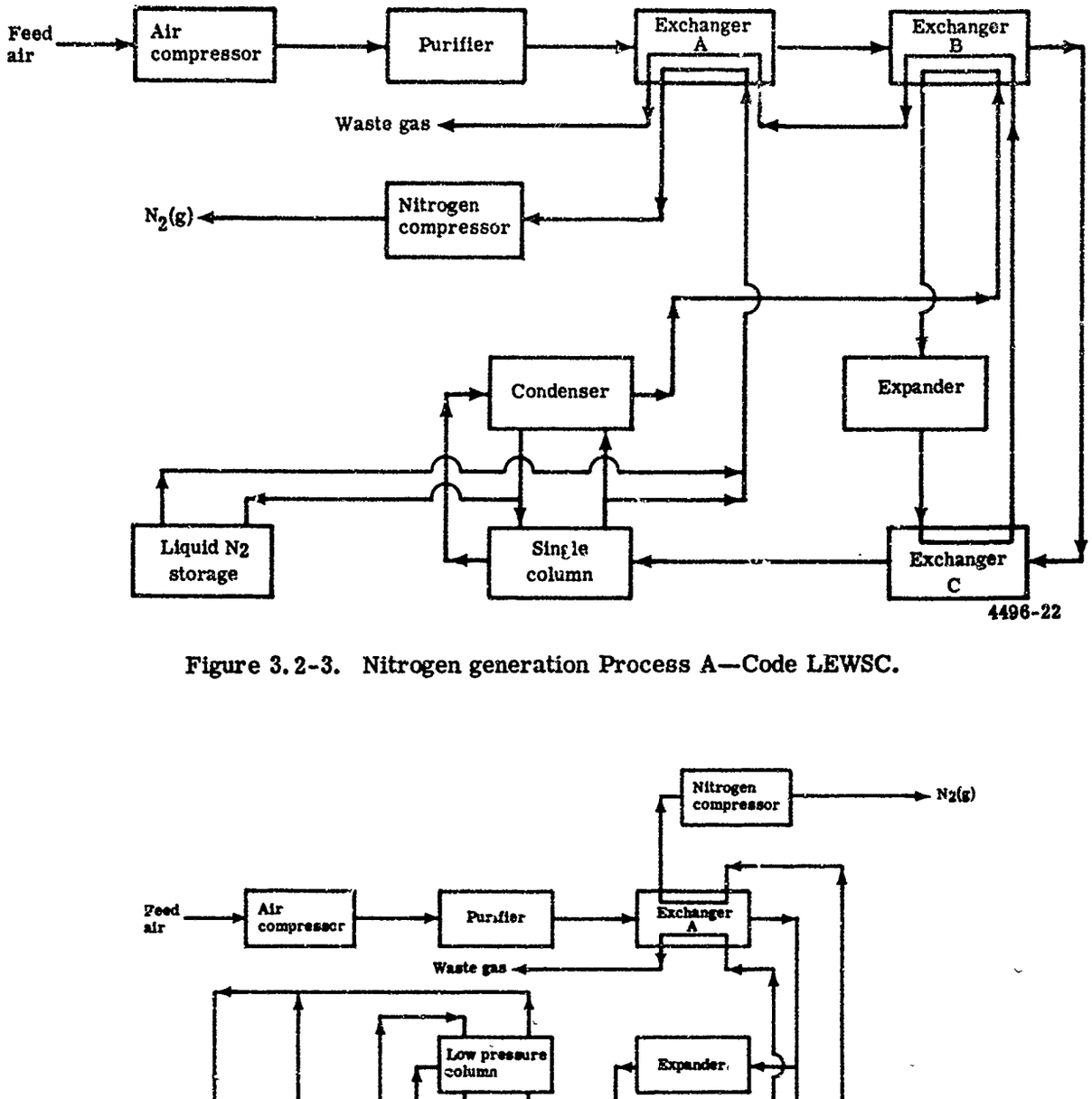


Figure 3.2-3. Nitrogen generation Process A—Code LEWSC.

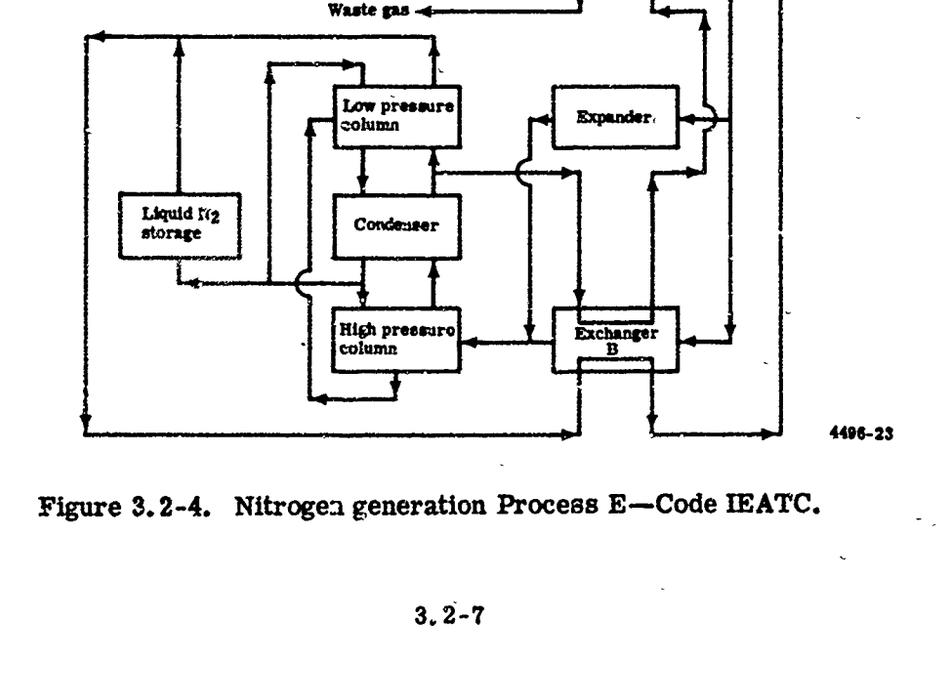


Figure 3.2-4. Nitrogen generation Process E—Code IEATC.

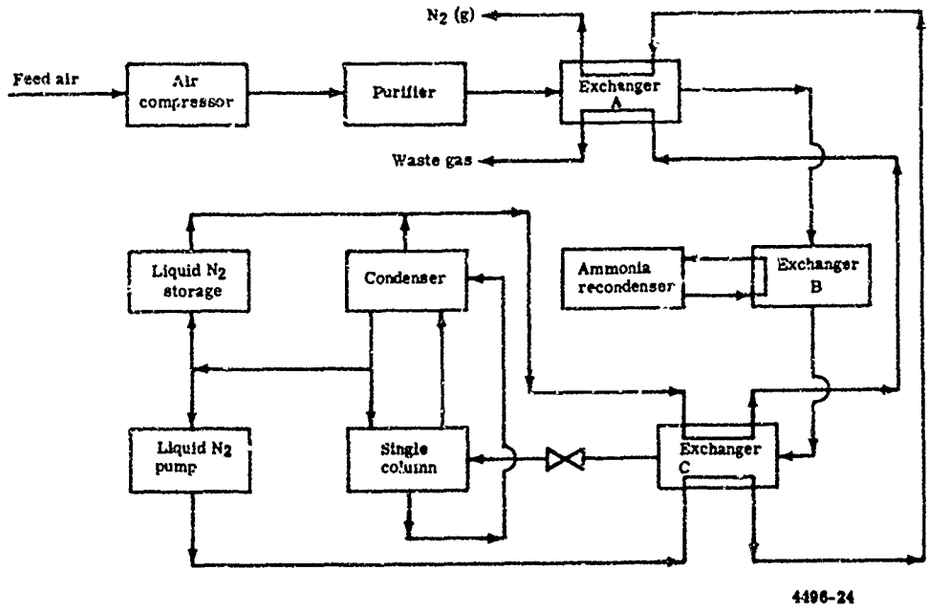


Figure 3.2-5. Nitrogen generation Process K—Code HJRSP.

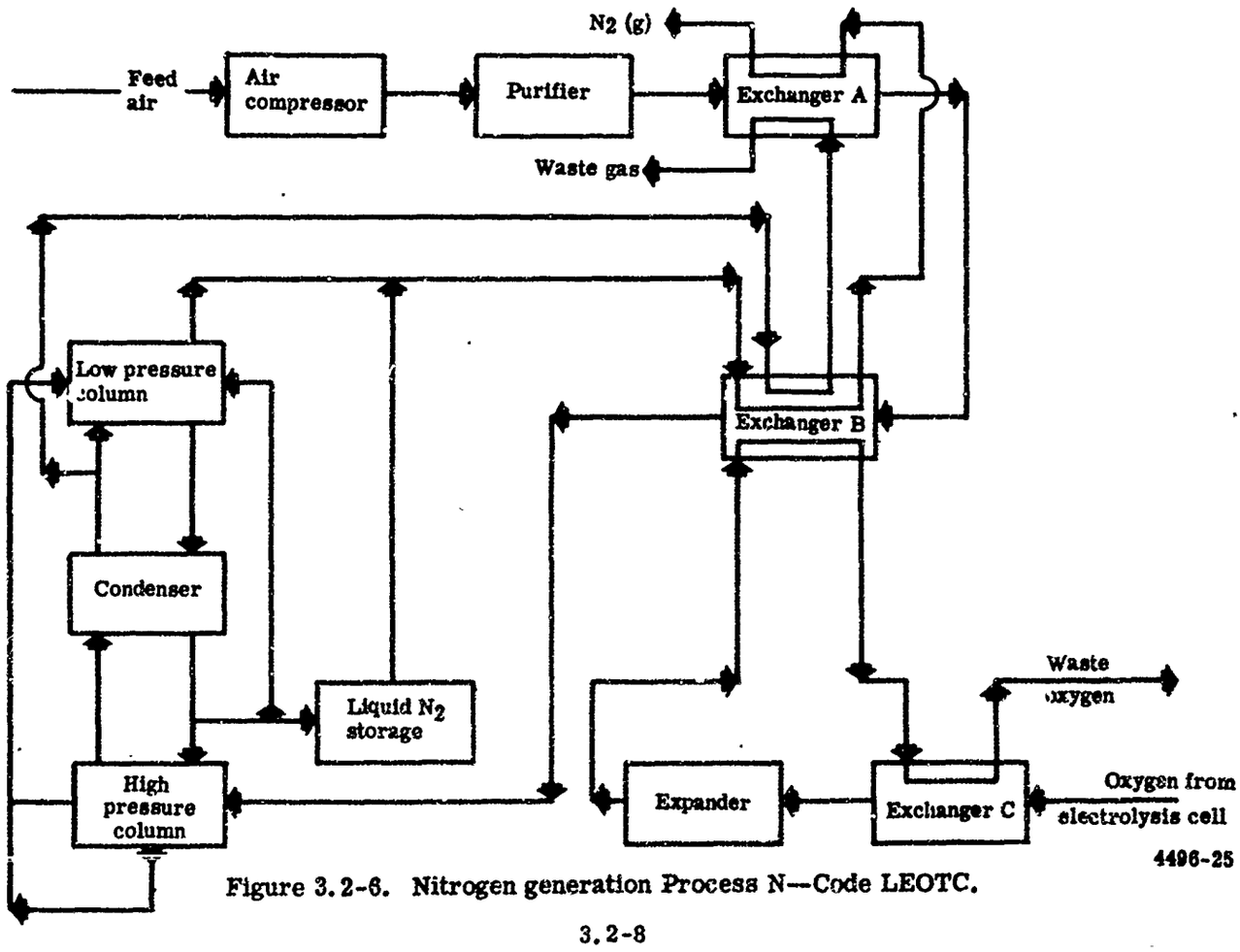


Figure 3.2-6. Nitrogen generation Process N—Code LEOTC.

Nitrogen generation Process E shown in Figure 3.2-4 is the basic Claude cycle. After compression to an operating pressure in the intermediate range, the air is passed through a purifier unit for removal of oil, water, and carbon dioxide. The purified air flows through Exchanger A for precooling by means of the returning product streams; a portion of the air is then directed to an expansion engine. The remainder of the air flows through Exchanger B for further cooling by means of the product streams from the column system. The air from Exchanger B is throttled to column pressure through a valve and joins the air discharge from the expander. The combined stream enters the column and is separated into nitrogen and oxygen product streams which cool the incoming process streams in Exchangers A and B. The product streams are discharged from the system at essentially atmospheric temperature and pressure. The gaseous nitrogen is compressed to synthesis pressure by means of a compressor.

Nitrogen generation Process K shown in Figure 3.2-5 is a high pressure process in which Joule-Thomson refrigeration is supplemented by ammonia refrigeration. A single column is used for fractionation and a liquid pump is used to pressurize the nitrogen product. Atmospheric air is compressed by means of a compressor and is then passed through a purifier unit in which oil, water, and carbon dioxide are removed. The purified air then passes through three heat exchangers in which it is cooled by returning product streams (Exchangers A and C) and ammonia refrigeration (Exchanger B). The cold, high pressure air is expanded through an expansion valve and is fed to the air fractionation column where a nitrogen fraction and an oxygen-rich fraction are separated. The oxygen-rich gas from the column condenser returns through Exchangers C and A and leaves the system at essentially atmospheric temperature and pressure. The nitrogen product is withdrawn from the column by means of a liquid nitrogen pump. The high pressure nitrogen passes through Exchangers C and A and leaves the system at high pressure and a temperature close to ambient.

Nitrogen generation Process N shown in Figure 3.2-6 is a low pressure process with a double column using oxygen available from the hydrogen production subsystem as the source of refrigeration. Atmospheric air is compressed and purified and enters Exchanger A where it is cooled by returning product streams. The air then passes through refrigerated Exchanger B for final cooling and liquefaction. The cooled air is fed to the column system and the gaseous nitrogen and oxygen-rich product streams are withdrawn and returned through the exchangers. Refrigeration for the air separation process is obtained from the electrolyzer oxygen stream by cooling in Exchanger C, expanding through an expander, passing the expanded oxygen through a passage in refrigerated Exchanger B, and then returning through oxygen Exchanger C.

The four processes discussed are typical and illustrate the general equipment arrangements commonly employed. In all cases, provisions were made to accumulate liquid nitrogen for use in cooling the plant quickly when restarting after shutdown. Also, nitrogen product compression was accommodated in the nitrogen plant since the decision of whether to compress in

the multipurpose ammonia plant compressor or a separate nitrogen compressor could not be made until the subsystem data were complete.

Process Selection

Simplified process calculations were made for the seventeen air fractionation processes to determine power requirements. In performing process calculations, the following procedure was used.

- Known values of pressure, temperature, and enthalpy were obtained from the assumptions and technical data.
- By means of an overall nitrogen balance, the amounts of process air, waste gas, and liquid nitrogen draw off required for one mole of nitrogen product were calculated.
- By means of an overall heat balance, the amount of refrigeration required for the process was calculated.
- The total heat leak into the system was distributed between the basic components using engineering judgement.
- By means of a heat balance on the column system, the enthalpy of the column feed stream was calculated.
- By means of heat balances on other components of the system, the enthalpy and temperature at other points in the system were established, and temperature differences were checked to make sure they were reasonable and positive, i.e., no violation of the second law of thermodynamics.
- Work requirements for the process air compressor, nitrogen compressor or liquid pump, and refrigeration unit were calculated and summed.

In addition to the power requirements outlined above, there is a requirement for cooling, air purification, and control. The various processes considered require cooling for the air and nitrogen compressors, Stirling cycle refrigeration units (when used), and refrigerant condensers. A basic power requirement for an air-cooled exchanger was developed and applied to the total cooling requirements for each process. The calculation method for Process N is described at the end of this subsection under Calculations.

All of the preceding power values were summed to determine the total power required by each of the seventeen processes. The results of the process calculations are given in Table 3.2-II. These data show that Process N (Code LEOTC) requires the least amount of power to produce one pound-mole of nitrogen—this is not surprising since high pressure waste oxygen was available to produce refrigeration. In addition, the results of the process calculations indicate the Processes G, O, and Q are not feasible due to a lack of refrigeration.

Table 3.2-II.

Power requirements for seventeen air fractionation processes.

($BTU/lb\text{-}mole\ N_2$)

Process	Code	Air compressor	N_2 compressor or pump	Process refrigeration	Air cooling	Auxiliary refrigeration	Air purification	Controls	Total power	Relative power
A	LEWSC	12340	5897	0	220	506	2370	284	21617	1.30
B	LENSC	12301	9950	0	265	506	2370	284	25676	1.54
C	LSCTC	4910	9950	2150	212	274	1050	284	18780	1.13
D	LEATC	10806	9950	0	246	494	2290	284	24070	1.45
E	IEATC	7104	9950	0	203	223	1035	284	18799	1.13
F	HJOTC	11351	9950	0	258	224	1050	284	23117	1.39
G	LEWSP	X	X	X	X	X	X	X	X	X
H	LSCSP	7402	434	21400	445	506	2370	284	32841	1.97
I	LEASP	42908	524	0	550	1280	5950	284	51496	3.08
J	IEASP	22540	435	0	272	506	2370	284	26607	1.59
K	HJRSP	26810	435	600	344	506	2370	284	31350	1.87
L	LCRSP	15072	435	15803	286	506	2370	284	34756	2.09
M	LCRTC	6650	9950	1016	221	224	1050	284	19985	1.20
N	LEOTC	4960	9950	0	177	224	1050	284	16645	1.00
O	LEOSP	X	X	X	X	X	X	X	X	X
P	IEASC	11740	5540	0	248	506	2370	284	20688	1.24
Q	LEOSC	X	X	X	X	X	X	X	X	X

X These processes are not feasible.

Since Process N was found to require the least amount of power, the relative merits of this process were checked for the following variations in conditions:

- Nitrogen product discharge pressure of 8000 psi instead of 3000 psi
- Recovery of expander work
- Recovery of work from electrolyzer oxygen

For all conditions, Process N remained the preferred process. This process compared favorably with all the other processes with respect to reliability of components, simplicity of operation, and weight. A consideration of the effect of plant size on power requirements for the various processes made it clear that size would have little effect on the relative power requirements and Process N would remain the best for all plant sizes investigated. Consequently, Process N was selected.

Investigation of Selected Nitrogen Generation Process

Air Fractionation Column

For air fractionation columns, various internal contact devices can be used to bring about intermixing of the liquid and vapor. The three major types of contact devices are bubble cap trays, sieve trays, and packing materials.

Weedman and Dodge[1] made an extensive investigation of various types of packing materials for use in air fractionation columns. Their experimental column was 2 in. in diameter and, for most of the packings studied, the height of a theoretical plate (HETP) was greater than 2 in. The exception was some Stedman type packings for which HETP values of 1 to 2 in. were obtained. In larger towers with Stedman packing, however, it proved very difficult to maintain steady operating conditions. Lobo and Williams[2] give a comprehensive review of work done on packing materials and, in connection with the Stedman packings, state that "All in all, the performance of the Stedman tower was very disappointing." Since stability of operation with packed columns in air fractionation service has proved very poor and HETP values are higher than can be achieved with a tray design, further consideration of a packed tower was abandoned and attention was directed to bubble cap and sieve trays.

For the particular application of concern to this study, the bubble cap tray had two important advantages over the sieve type tray. First, the liquid holdup on the bubble cap trays is less than on the sieve tray and the amount of liquid required to fill the column is considerably less. Also, the lower holdup makes a bubble cap column quick to respond to changes in operating conditions. A second advantage of the bubble cap tray is that liquid will be retained on a tray during interruptions in flow—a sieve tray will lose its liquid very quickly. These factors led to the choice of bubble cap trays for the air fractionation column, even though their cost is somewhat higher.

For the design of the air fractionation columns, a series of computer analyses were made to determine the effect of the number of stages and operating conditions on the amounts and purities of the products. This was done for the purpose of fixing the optimum number of stages and the optimum operating conditions. In making these calculations, air was treated as a ternary mixture consisting of 78% nitrogen, 21% oxygen, and 1.0% argon. An example is provided in the Calculations subsection which illustrates the data used and results obtained in the computer analysis of column performance.

The pertinent results of the computer analysis are presented in Table 3.2-III. The following important column performance characteristics were noted from the results.

- For a given set of feed and product conditions, the number of trays in the high pressure column has a great effect on the overhead product purity from the high pressure column and very high purities are easily attained. This is illustrated by the data for Problem 9 which are plotted in Figure 3.2-7.
- The number of trays in the low pressure column has a relatively small effect on the impurity concentration in the overhead nitrogen product from the low pressure column. The use of a large number of trays does not improve this product purity appreciably. This is illustrated by the data from Problems 2 and 8 which are plotted in Figure 3.2-8.

Table 3.2-III.

Results of computer analysis of air fractionation columns.

Problem	Case	Column*	No. of theoretical trays	Amount overhead**	Amount bottoms**	Feed No. 1†	Feed No. 2†	Mole fraction in overhead			Mole fraction in bottoms		
								Nitrogen	Oxygen	Argon	Nitrogen	Oxygen	Argon
1	1	HP	10	0.34L	0.66L	Air		0.99508	0.39412×10^{-2}	0.97714×10^{-3}	0.66903	0.31632	0.014657
1	2	HP	30	0.34L	0.66L	Air		0.99999	0.63865×10^{-6}	0.64509×10^{-5}	0.66650	0.31834	0.015157
1	3	HP	60	0.34L	0.66L	Air		1.00000	0.26092×10^{-12}	0.18748×10^{-8}	0.68647	0.31837	0.015162
2	1	LP	30	0.77V	0.23V	1-1-B	1-1-0	0.99423	0.32125×10^{-2}	0.25590×10^{-2}	0.061872	0.90318	0.034948
3	1	LP	50	0.77V	0.23V	‡	‡	0.99169	0.67602×10^{-2}	0.15492×10^{-2}	0.070185	0.89179	0.038017
4	1	LP	90	0.77V	0.23V	1-3-B	1-2-0	0.97685	0.21472×10^{-1}	0.16785×10^{-2}	0.11222	0.85158	0.036197
5	1	LP	30	0.73V	0.27V	1-1-B	1-1-0	0.99719	$0.13336 \times 10^{-}$	0.14788×10^{-2}	0.19219	0.77475	0.033067
6	1	LP	60	0.73V	0.27V	‡	‡	0.99669	0.21843×10^{-4}	0.92252×10^{-3}	0.18681	0.77684	0.034347
7	1	LP	90	0.73V	0.27V	1-2-B	1-2-0	0.99789	0.11042×10^{-4}	0.21020×10^{-2}	0.18587	0.78251	0.031623
8	1	LP	25	0.77V	0.23V	1-1-B	1-1-0	0.99373	0.37854×10^{-2}	0.24750×10^{-2}	0.063789	0.90104	0.035322
8	2	LP	20	0.77V	0.23V	1-1-B	1-1-0	0.99235	0.52849×10^{-2}	0.23681×10^{-2}	0.068408	0.89601	0.035579
8	3	LP	15	0.77V	0.23V	1-1-B	1-1-0	0.98852	0.90458×10^{-2}	0.23312×10^{-2}	0.080826	0.88347	0.035705
9	1	HP	10	0.34L	0.66L	Air		0.99527	0.37667×10^{-2}	0.95533×10^{-3}	0.66638	0.31645	0.014871
9	2	HP	30	0.34L	0.66L	Air		0.99983	0.69449×10^{-4}	0.10255×10^{-3}	0.66665	0.31825	0.015104
9	3	HP	60	0.34L	0.66L	Air		0.99999	0.11838×10^{-5}	0.96015×10^{-5}	0.66652	0.31832	0.015155
10	1	HP	13	0.34V	0.66L	Air		0.99937	0.41265×10^{-3}	0.21950×10^{-3}	0.66686	0.31802	0.015046
10	2	HP	13	0.40V	0.60L	Air		0.99734	0.17988×10^{-2}	0.86273×10^{-3}	0.63479	0.34910	0.016107
10	3	HP	13	0.45V	0.55L	Air		0.98800	0.91234×10^{-2}	0.28762×10^{-2}	0.61021	0.37396	0.015813
11	1	LP	24	0.76V	0.23V	10-1-B	10-1-0	0.98946	0.79368×10^{-2}	0.26040×10^{-2}	0.10080	0.86511	0.03398
12	1	LP	24	0.764V	0.236V	10-2-B	10-2-0	0.99830	0.68876×10^{-3}	0.10107×10^{-2}	0.071696	0.88932	0.038183
13	1	LP	24	0.764V	0.236V	10-3-B	10-3-0	0.99546	0.31136×10^{-2}	0.14280×10^{-2}	0.062657	0.87960	0.037747
14	1	LP	24	0.688V	0.312V	10-2-B	10-2-0	0.99693	0.61093×10^{-3}	0.45786×10^{-3}	0.29620	0.67270	0.031105
14	2	LP	24	0.840V	0.160V	10-2-B	10-2-0	0.92740	0.64686×10^{-1}	0.79129×10^{-2}	0.0038025	0.97516	0.021035

*HP = high pressure column, LP = low pressure column.

**0.34L = 0.34 lb-moles of liquid, 0.77V = 0.77 lb-moles of vapor.

†Feed 1-1-B means the feed is the bottoms (B) product obtained from Problem 1, Case 1.
Feed 1-2-0 means the feed is the overhead (O) product obtained from Problem 1, Case 2.
Feed No. 1 enters low pressure column in lower half of column. Feed No. 2 enters at top tray and becomes reflux for low pressure column.

‡Feed conditions obtained from interpolation of data for Problem 1, Cases 1, 2, and 3 to obtain values for 20 theoretical trays.

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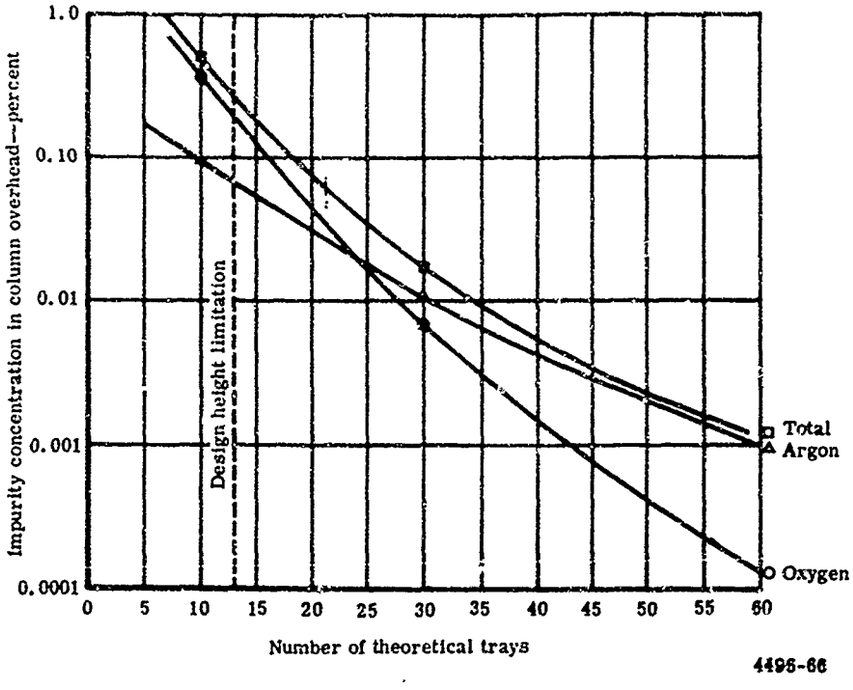


Figure 3.2-7. Impurities in high pressure nitrogen column overhead product as a function of number of trays.

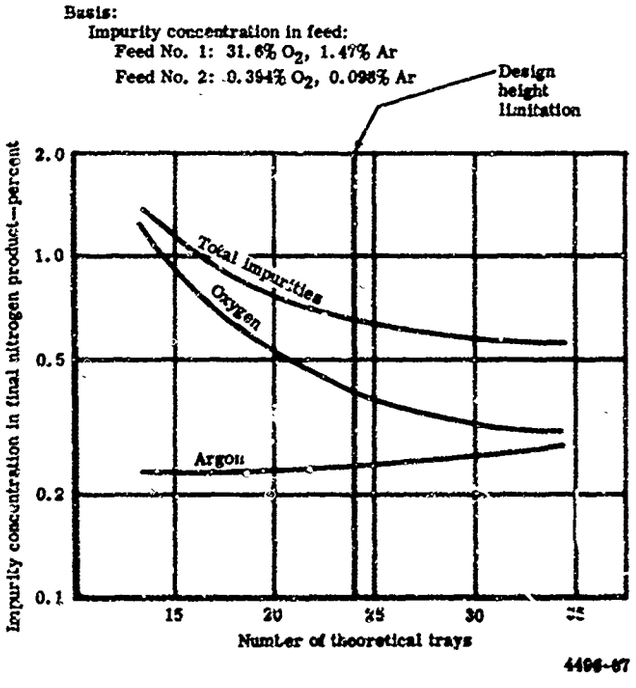


Figure 3.2-8. Impurities in final nitrogen product as a function of number of trays in low pressure column.

A standard double column would exceed the plant height limitation of 8 ft. Consideration was given to the possibility of using a column which could be put upright for operation and laid down during shipment. However, this scheme presented almost insurmountable mechanical problems, complicated operation, and made inefficient use of available space. These conclusions, together with the observations on effectiveness of number of trays in improving purity, led to the decision to use a split column and put the maximum possible number of trays in each of the two column units. Design calculations indicated that with a maximum plant height of 8 ft, 17 actual trays could be put into the high pressure column and 30 actual trays could be put into the low pressure column. With a reasonable tray efficiency of 80%, the equivalent number of theoretical trays is 13 and 24 for the high pressure and low pressure columns, respectively.

The influence of the overhead draw-off rate on the overhead product purity for the high pressure column is shown in Problem 10 of Table 3.2-III and Figure 3.2-9. The draw-off rate has a strong effect on product purity. Since the products from the high pressure column are the feeds for the low pressure column, the performance of the low pressure column for conditions obtained from Problem 10 was investigated to determine the optimum operating conditions. The results of this investigation are shown in Problems 11, 12 and 13 of Table 3.2-III and Figure 3.2-9. The concentrations of nitrogen, oxygen, and argon in the overhead final nitrogen product are plotted as a function of the ratio of high pressure column overhead product draw-off rate (feed No. 2 to the low pressure column) to total feed rate. The plot indicates that final nitrogen product purity is at a maximum of 99.83% for a ratio of approximately 0.40. The corresponding oxygen concentration is 0.069% and the corresponding argon concentration is 0.101%.

The effect on the system of increasing the final nitrogen product purity by relaxing the mobility height constraints was assessed by considering the amount of hydrogen needed to remove the impurity oxygen. At the operating conditions previously described, there are 690 parts of oxygen per million parts of product (0.069%). At a nitrogen product rate of 500 lb/hr, the corresponding amount of oxygen to be removed is about 0.35 lb/hr. Approximately 0.04 lb/hr of hydrogen are required to remove this oxygen. Generating this amount of hydrogen requires about 0.9 kw of electrical power on the basis of an average power consumption of 22 kw-hr/lb H_2. Therefore, if the height constraints were removed so that ultrapure nitrogen was produced, the maximum power saving would be 0.9 kw, a value that appears too low to justify removing the constraints.

Since it is very probable that the final nitrogen product draw-off rate from the low pressure column will vary depending on ammonia synthesis requirements, computer studies were made of the effect of the draw-off rate on nitrogen product purity. Data from computer problems 12 and 14 give an indication of this effect. In Figure 3.2-10, nitrogen product purity is plotted as a function of nitrogen draw-off rate from the low pressure column for a given set of feed conditions (high pressure column product conditions). The data indicate that the nitrogen product purity falls off quite rapidly at draw-off rates above the design point, but remains relatively stable at draw-off rates below the design point.

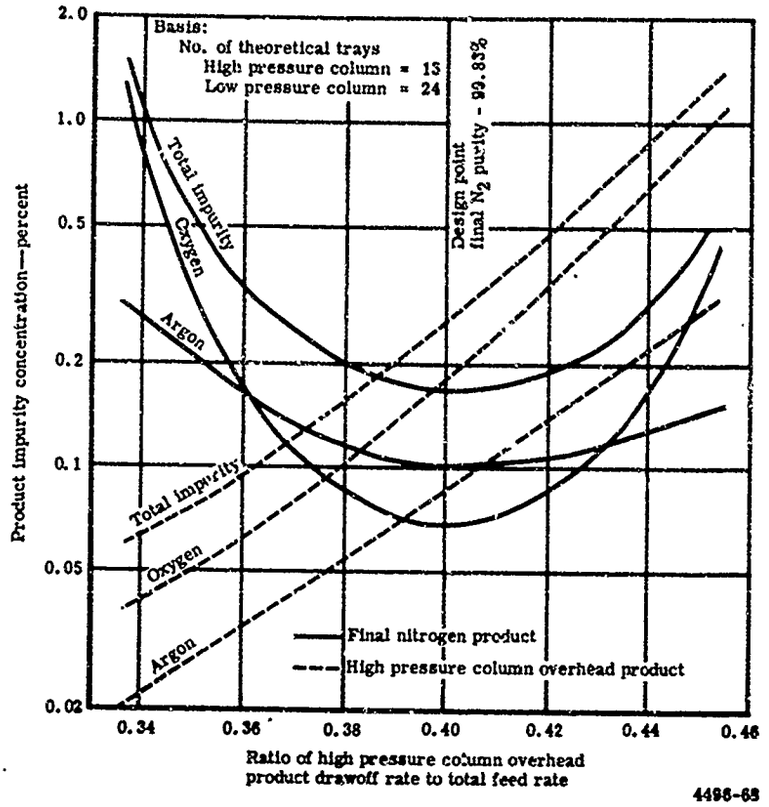


Figure 3.2-9. Product purity as a function of high pressure column overhead product draw-off rate.

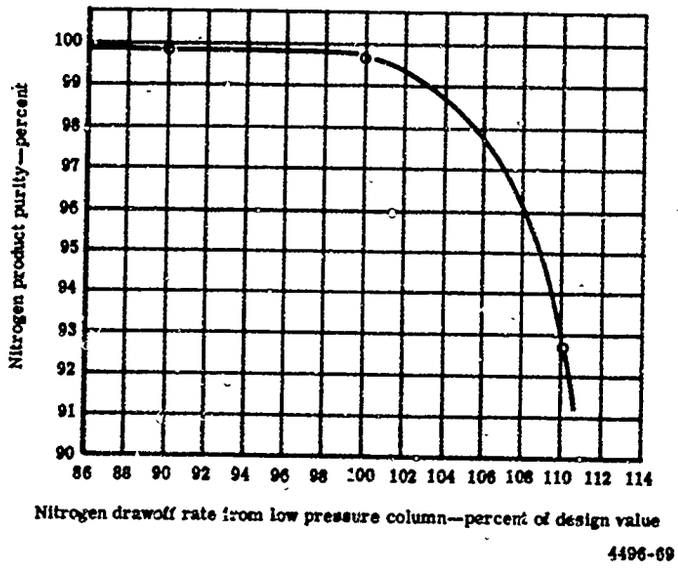


Figure 3.2-10. Nitrogen product purity as a function of low pressure column draw-off rate for design feed conditions.

In the analysis of column performance, the concentration gradient in the columns is of interest. In Figure 3.2-11, the concentration of nitrogen, oxygen, and argon is plotted as a function of tray number to show the nature of the gradients in the column. The data are for the low pressure column at normal operating conditions. The apparent unusual behavior of the argon concentration curve is normal and frequently used to separate argon from air.[3]

Process Air Purification

It is necessary to remove water and carbon dioxide from the process air to avoid obstructions due to freeze-out at the low temperature levels involved. Various methods have historically been used for the removal of these impurities.

The following are the more common methods for carbon dioxide removal:

- Scrubbing with caustic solution
- Solid type absorbents such as soda lime
- Freeze-out exchangers in alternate service
- Reversing exchangers
- Adsorbent beds with reactivation of adsorbent

The first two methods require repeated replenishment of chemicals and would be undesirable for the plants under study because of logistic problems.

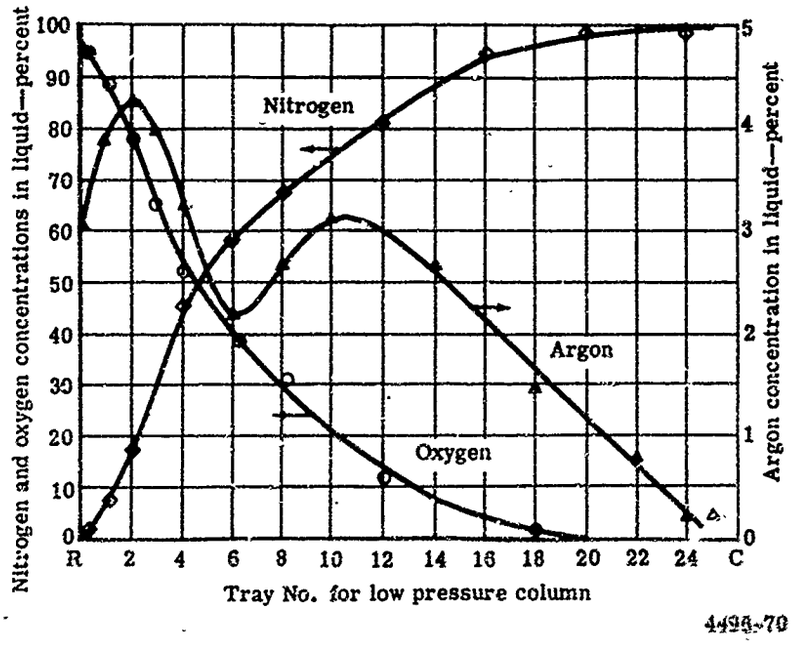


Figure 3.2-11. Nitrogen column concentration gradients as a function of tray number for low pressure column.

Freeze-out exchangers would work, but have some undesirable features. Two exchangers for alternate service are required and this increases size and weight requirements. Also, from an operational viewpoint, switching of the exchangers is difficult to accomplish and tends to shock the air fractionation section thermally. These considerations eliminated this method of carbon dioxide removal.

Reversing exchangers are often used for carbon dioxide removal when the product stream is relatively small and large amounts of waste gas are available for picking up deposited carbon dioxide and water in the exchanger. The use of a product stream for pickup of impurities in the reversing exchanger introduces these impurities into the product and leads to contamination with air unless a purging operation is used. Since nitrogen product purity is of major importance and the waste gas stream is small, reversing exchangers were not considered to be suitable.

The adsorbent method for carbon dioxide removal was, therefore, selected. The adsorbent used is molecular sieve which can be operated at ambient or lower temperatures. From an operational viewpoint, ambient temperature adsorption is easier to control, is less likely to shock the system thermally, and permits greater accessibility and ease of maintenance.

There are two basic approaches in general use for the removal of water from the process air:

- Solid dessicant in a dual bed unit with one bed being reactivated while the other is in service
- Refrigeration using reversing exchangers

The undesirable features of the reversing exchanger scheme mentioned previously and the fact that the adsorbent used for water removal can be combined with the adsorbent for carbon dioxide removal in a single bed form the basis for selection of the adsorption method.

Several possible adsorbent materials can be used for water removal. Molecular sieve and activated alumina have proved to be the best materials. Molecular sieve has greater capacity for water than does activated alumina, but requires considerably higher temperatures for reactivation. In the present application, reactivation time and energy considerations favor the use of activated alumina for water removal. Therefore, a dual unit with a two-layer bed in each unit was selected. The process air first passes through the activated alumina for removal of water and then through molecular sieve for removal of carbon dioxide. Both impurities are reduced to a level of less than 5 ppm (0.0005%). Two adsorber vessels are required for the purifier—one is in service while the other is reactivated.

Power Recovery From Waste Oxygen

Waste oxygen from the hydrogen production subsystem is delivered at approximately 300 psig. Since oxygen for refrigeration in the nitrogen generation subsystem is required to be at a

pressure of 75 psig or less, there is the potential for recovery of work from the oxygen by expanding it isentropically from 300 psig to 75 psig. This possibility was investigated, but it was concluded that it was not practical.

Since the work recovered by the expansion of a gas is proportional to absolute temperature, the maximum work output would result from operation at the maximum available temperature. However, water must be removed from the oxygen before expansion to prevent the formation of ice which would subsequently plug and damage the components. Simultaneously maintaining high temperature and removing the water introduces complications. Either an adsorbent type purifier would be needed which, during regeneration, would consume a large fraction of the power generated by the expansion, or a combination switching and regenerative heat exchanger system could be devised. The latter would increase the complexity, cost, and space requirements considerably. In addition, the expander vendors do not include provisions for recovering this work unless the amount of work recovered is much larger than in the present designs.

The possibility of using expanded oxygen to cool the product nitrogen and thus reduce the power required for compression to synthesis pressure was also studied, but again the net reduction in power consumption did not justify the additional equipment required and the increased complexity of operation.

DESIGN PARAMETERS

In addition to the general overall design parameters presented in Section II, the following specifications were followed in the evaluation of processes.

- Nitrogen subsystem designs will be developed for the following production rates and mobility classes:

Nitrogen production rate—lb/hr	500	1000	1670	2500
Mobility Classes	I II	I II III	II III	III

- Process air enters the cold box at 90°F.
- Gaseous product streams leave the cold box at 80°F.
- Nitrogen product purity is 99.5% (final purity 99.83%).
- The waste gas stream contains 68% nitrogen when a single column is used and 6% nitrogen when a double column is used.
- Feed air is a binary mixture containing 79% nitrogen and 21% oxygen except for the column calculations where argon was included.
- For the enthalpy, h, of a mixture of nitrogen and oxygen at pressure P and temperature T, plot $(h)_{P,T}$ versus x using data for nitrogen, oxygen, and air; interpolate for the enthalpy of the mixture.
- Heat leak (Q_L) into the cold box is 5 BTU/lb process air for a process using a double column and 4 BTU/lb of process air for a system using a single column. (For the largest capacity plant design, a heat leak of 2.5 BTU/lb process air was used.)

- Based on the service pattern given in Section II (Table 2-III) and using minimum operating time as a basis for calculations, the ratio of operating time (θ) to down time (θ') is constant at 3.0. For these conditions, the liquid nitrogen draw-off rate is given by the equation (derived later):

$$W_N = \frac{W_A\, q_L}{518.4}$$

where

W_N = liquid nitrogen draw-off rate, lb/hr

W_A = feed airflow rate, lb/hr

q_L = heat leak, BTU/hr air

- The temperature difference across column condensers is 3°F.
- Pressure drops across the cold box exchanger system are as follows:

  ΔP for process air = 3 psi

  ΔP for low pressure nitrogen product = 5 psi

  ΔP for nitrogen product at high pressure column pressure = 4 psi

  ΔP for waste gas = 6.5 psi
- Efficiencies of machines are as follows:

  Expansion turbines, 75%

  Compressors, 70%

  Liquid pumps, 75%
- Calculations based on one pound-mole of nitrogen product.
- Process air compressor discharge pressure is 5 psi greater than the inlet pressure to the cold box.
- The ultimate nitrogen product pressure is 3000 psi.
- For gas compressors, the maximum pressure ratio per stage as a function of number of stages is as follows:

No. stages	Maximum pressure ratio per stage
1	6.8
2	5.3
3	4.3
4	3.6
5	3.1
6	2.7

Technical data used in the calculations were obtained from standard thermodynamic curves. The properties of air, oxygen, and nitrogen were obtained from the Elliott curves for pressures below 180 psia and from the Kellogg curves for pressures above 180 psia. Equilibrium constants for the ternary system, nitrogen-oxygen-argon, were obtained from a report prepared by Wilson, Silverberg, and Zellner.[4]

DESIGN

The essential features of Process N are illustrated by the flow diagram of Figure 3.2-12. The pieces of equipment are defined in Table 3.2-IV.

Table 3.2-IV.
Nitrogen generation subsystem equipment list.

B-1	Motor for air compressor	E-2	Refrigerated heat exchanger
B-2	Motor for nitrogen compressor	E-3	Oxygen reversing heat exchanger
B-3	Motor for refrigeration compressor	E-4	Crude oxygen subcooler
B-4	Motor for reactivation blower	E-5	Nitrogen subcooler
B-5	Motor for air cooling fan	E-6	Air compressor aftercooler
C-1	Air compressor	E-7	Refrigerant condenser
C-2	Nitrogen compressor	E-8	Reactivation gas cooler
C-3	Refrigerant compressor	E-9	Secondary reactivation gas cooler
C-4	Reactivation blower	E-10	Oxygen cooler
C-5	Cooling air fan	E-11	Air precooler
CD-1	Continuous drainer (D-1)	E-12	Nitrogen compressor aftercooler
CD-2	Continuous drainer (D-3)	F-1	Dust filter
CD-3	Continuous drainer (E-9)	F-2	Air intake filter
CD-4	Continuous drainer (D-6)	F-3	Reactivation air filter
D-1	Separator (air compressor)	H-1	Reactivation heater
D-2	Refrigerant receiver	R-1	Air purifiers
D-3	Separator (precooler)	ST	Motor starters (not shown in Figure 3.2-12)
D-4	Accumulator (expander)		
D-5	Liquid nitrogen storage tank	T-1	High pressure column
D-6	Separator (oxygen)	T-2	Low pressure column
D-7	Silencer	X-1	Expansion turbine
E-1	Main heat exchanger		

Detailed process calculations were performed for Process N to establish process conditions throughout the system for nitrogen production rates of 500, 1000, 1670, and 2500 lb N_2/hr. The conditions established are tabulated in Table 3.2-V through 3.2-VIII. The process points to which the conditions apply are given in Figure 3.2-12. Equipment selection and design were based on these process conditions.

Ambient air is compressed by means of a compressor (C-1) to a pressure of 82 psig and then cooled in an air-cooled aftercooler (E-6). From the aftercooler, the process air passes through a separator (D-1) which removes entrained water and oil. Further cooling of the process air is accomplished in a refrigerated precooler (E-11) by means of R-12 (dichlorodifluoromethane, CCl_2F_2) refrigeration. A refrigerant recondensing unit, which includes a

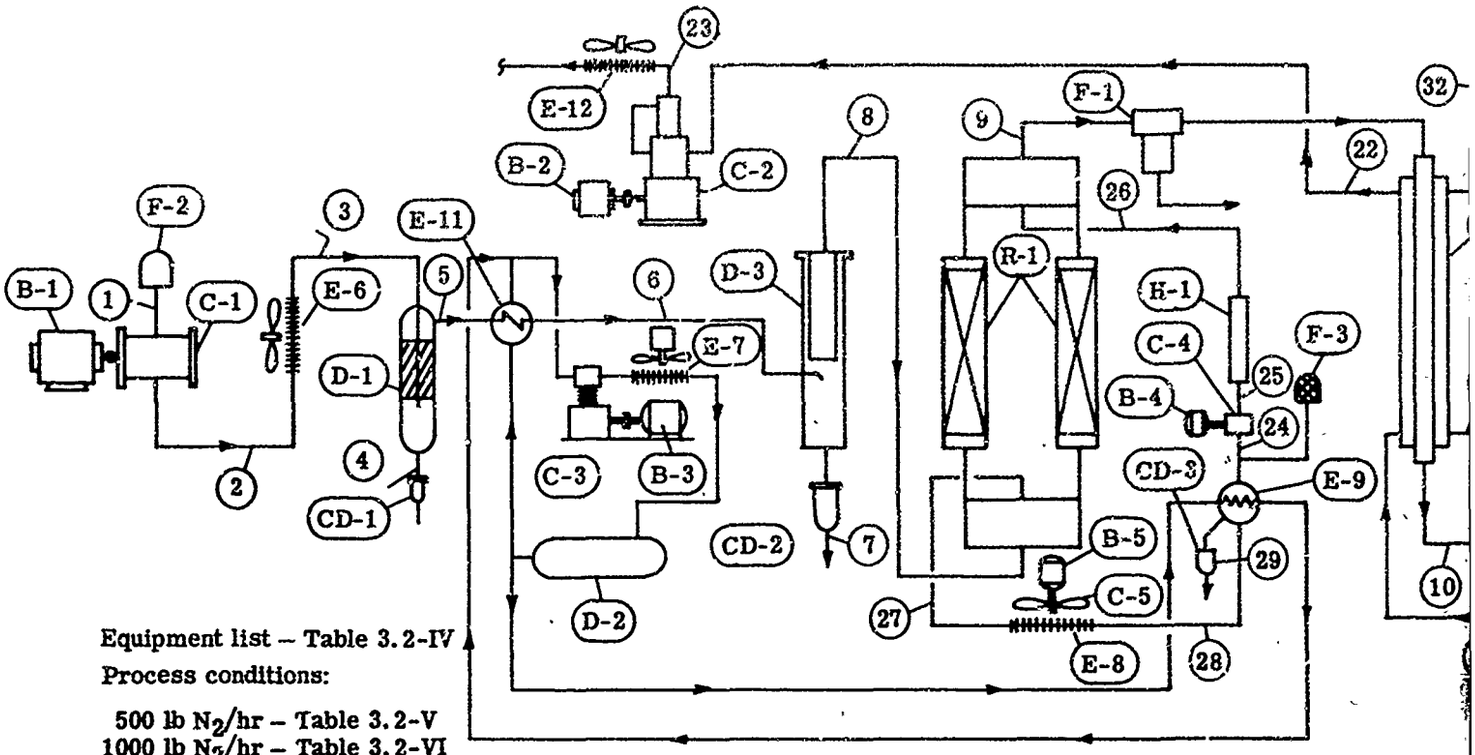
compressor (C-3), an air-cooled condenser (E-7), and the liquid receiver (D-2) is provided to supply the refrigeration needed for precooling of the process air.

Table 3.2-V.

Nitrogen subsystem process conditions, 500-lb N_2/hr capacity.

Point*	Fluid	Pressure (psia)	Temperature (°F)	Enthalpy (BTU/lb)	Flow (lb/hr)	Water rate (lb/hr)	Remarks
1	Air	14.7	77	128.5	676.8	6.8	1. Enthalpy is on dry basis.
2	Air	97.0	242	167.8	676.8	6.3	2. WA = water.
3	Air	96.2	97	132.8	676.8	6.8	3. OL = oxygen-rich liquid.
4	WA	96.2	97			2.9	4. NL = nitrogen liquid.
5	Air	96.0	97	132.8	676.8	3.9	5. NG = gaseous nitrogen.
6	Air	95.5	75	127.5	676.8	2.0	6. WG = waste gas.
7	WA	95.5	75			1.9	7. RO = electrolyzer O_2.
8	Air	95.0	75	127.5	676.8	2.0	
9	Air	92.0	90	131.1	676.8	0	
10	Air	90.0	-236.5	50.27	676.8	0	
11	Air	89.0	-279	34.23	676.8	0	
12	OL	89.0	-280	-39.55	414.8	0	
13	OL	45.0	-293	-44.5	414.8	0	
14	NL	88.0	-286	-36.6	262	0	
15	NL	22.0	-314	-36.6	10.69	0	
16	NL	88.0	-286	-36.6	251.31	0	
17	NG	22.0	-314	34.0	0.21	0	
18	NL	87.9	-290	36.0	251.31	0	
19	NG	22.0	-314	34.0	502.9	0	
20	NG	21.0	-291	40.5	502.9	0	
21	NG	20.0	-261	48.5	502.9	0	
22	NG	17.0	80	133.8	502.9	0	
23	NG	325.0	396	211.6	502.9	0	
24	Air	17.0	75	132.5	210.0	3.4	
25	Air	21.0	109	136.1	210.0	3.4	
26	Air	20.5	109/450		210.0	3.4	
27	Air	18.0	90/300		210.0	3.4	
28	Air	17.5	97	133.2	210.0	7.3	
29	WA	17.0	75			3.9	
30	WG	23.5	-289.5	35.0	163.4	0	
31	WG	21.0	-261	42.4	163.4	0	
32	WG	17.0	80	118.0	163.4	0	
33	OL	46.0	-289	43.1	414.8	0	
34	RO	290.0	150	128.5	857.2	6.26	
35	RO	289.0	97	116.7	857.2	6.26	
36	WA	289.0	97			4.82	
37	RO	71.0	97	120.7	857.2	1.44	
38	RO	68.0	-238	45.5	857.2	0	
39	RO	20.0	-292	35.0	857.2	0	
40	RO	18.0	-261	42.2	857.2	0	
41	RO	15.0	87	118.9	857.2	1.44	

*See Figure 3.2-12.



Equipment list – Table 3.2-IV

Process conditions:

500 lb N_2/hr – Table 3.2-V
1000 lb N_2/hr – Table 3.2-VI
1670 lb N_2/hr – Table 3.2-VII
2500 lb N_2/hr – Table 3.2-VIII

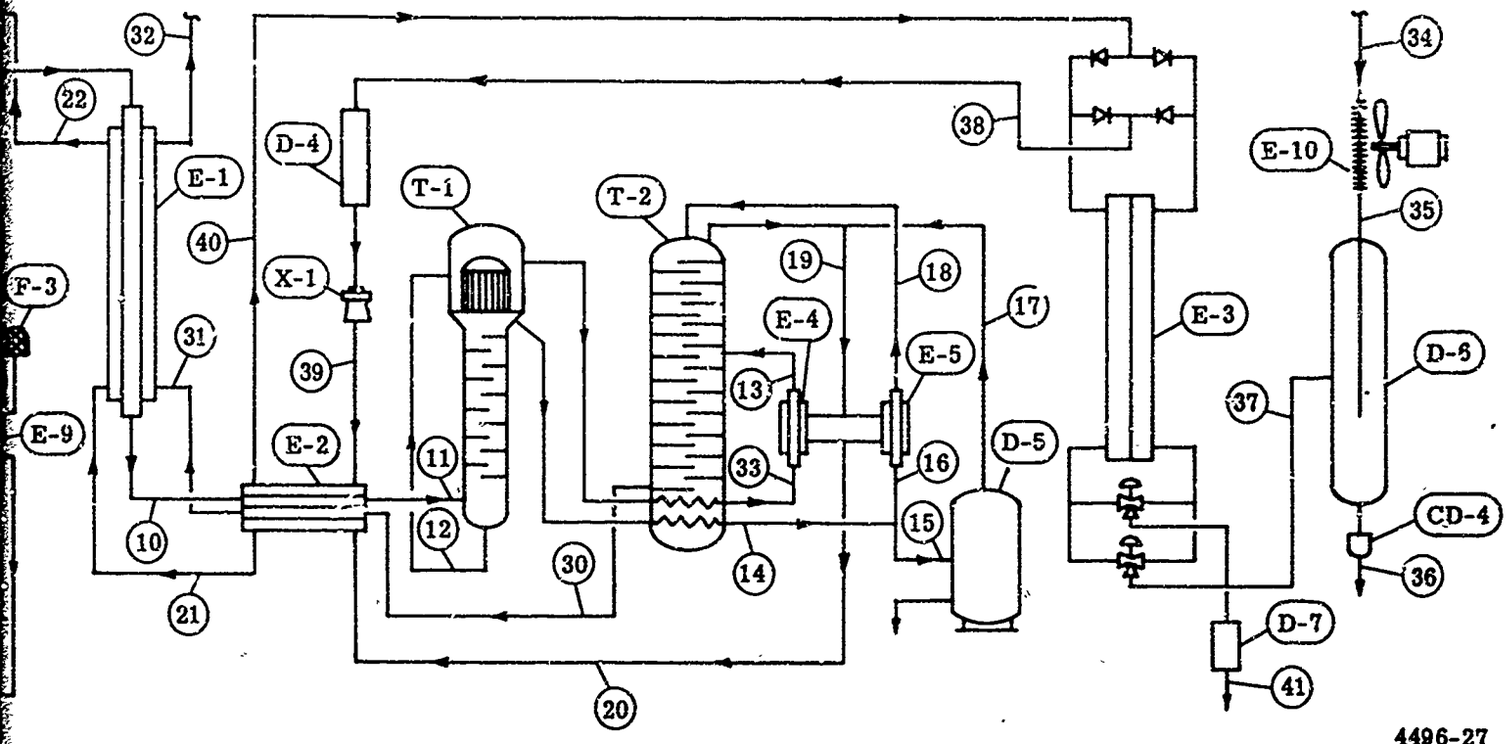


Figure 3.2-12. Nitrogen generation subsystem flow sheet.

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Table 3.2-VI.

Nitrogen subsystem process conditions, 1000-lb N_2/hr capacity.

Point*	Fluid	Pressure (psia)	Temperature (°F)	Enthalpy (BTU/lb)	Flow (lb/hr)	Water rate (lb/hr)
1	Air	14.7	77	128.5	1347.0	13.6
2	Air	97.0	242	167.8	1347.0	13.6
3	Air	96.2	97	132.8	1347.0	7.7
4	WA	96.2	97			5.9
5	Air	96.0	97	132.8	1347.0	7.7
6	Air	95.5	75	127.5	1347.0	3.9
7	WA	95.5	75			3.8
8	Air	95.0	75	127.5	1347.0	3.9
9	Air	92.0	90	131.1	1347.0	0
10	Air	90.0	-245.5	47.8	1347.0	0
11	Air	89.0	-279.0	35.37	1347.0	0
12	OL	89.0	-280.0	-39.55	827.0	0
13	OL	45.0	-293	-44.5	827.0	0
14	NL	86.0	-285	-36.6	520.0	0
15	NL	22.0	-314	-36.6	10.36	0
16	NL	88.0	-286	-36.6	503.64	0
17	NG	22.0	-314	34.0	0.32	0
18	NL	37.0	-290	-38.0	503.64	0
19	NG	22.0	-314	34.0	1005.7	0
20	NG	21.0	-291	40.5	1005.7	0
21	NG	20.0	-268.5	46.6	1005.7	0
22	NG	17.0	80	133.8	1005.7	0.
23	NG	325.0	396	211.6	1005.7	0
24	Air	17.0	75	132.5	420	6.8
25	Air	21.0	109	136.1	420	6.8
26	Air	20.5	109/450		420	6.8
27	Air	18.0	90/300		420	6.8
28	Air	17.5	97	133.2	420	14.6
29	WA	17.0	75			7.8
30	WG	23.5	-289.5	35.0	325	0
31	WG	21.0	-268.5	40.6	325	0
32	WG	17.0	80	118.0	325	0
33	OL	46.0	-289	-43.1	827.0	0
34	RO	290.0	150	128.5	1714.3	12.4
35	RC	289.0	97	116.7	1714.3	12.4
36	WA	289.0	97			9.6
37	RO	80.0	97	120.7	1714.3	2.88
38	RC	57.0	-247	44.1	1714.3	0
39	RO	20.0	-292	35.0	1714.3	0
40	RO	18.0	-288.5	40.6	1714.3	0
41	RO	15.0	87	118.9	1714.3	2.88

*See Figure 3.2-12.

Remarks

1. Enthalpy is on dry basis.
2. WA water.
3. OL = oxygen-rich liquid.
4. NL = nitrogen liquid.
5. NG = gaseous nitrogen.
6. WG = waste gas.
7. RO = electrolyzer O_2.

Table 3.2-VII.

Nitrogen subsystem process conditions, 1670-lb N_2/hr capacity.

Point*	Fluid	Pressure (psia)	Temperature (°F)	Enthalpy (BTU/lb)	Flow (lb/hr)	Water rate (lb/hr)	Remarks
1	Air	14.7	77	128.5	2243.7	22.6	1. Enthalpy is on dry basis.
2	Air	97.0	242	167.8	2243.7	22.6	2. WA = water.
3	Air	96.2	97	132.8	2243.7	12.9	3. OL = oxygen-rich liquid.
4	WA	96.2	97			9.7	4. NL = nitrogen liquid.
5	Air	95.0	97	132.8	2243.7	12.9	5. NG = gaseous nitrogen.
6	Air	95.5	75	127.5	2243.7	6.4	6. WG = waste gas.
7	WA	95.5	75			6.5	7. RO = electrolyzer O_2.
8	Air	95.0	75	127.5	2243.7	6.4	
9	Air	92.0	90	131.1	2243.7	0	
10	Air	90.0	-250	16.6	2243.7	0	
11	Air	89.0	-279	36.04	2243.7	0	
12	OL	89.0	-280.6	-39.55	1373.7	0	
13	OL	45.0	-293	-44.5	1373.7	0	
14	NL	88.0	-286	-36.6	870.0	0	
15	NL	22.0	-314	-36.6	22.72	0	
16	NL	88.0	-286	36.6	847.3	0	
17	NG	22.0	-314	34.0	0.44	0	
18	NL	87.0	-290	-38.0	847.3	0	
19	NG	22.0	-314	34.0	1678.6	0	
20	NG	21.0	-291	40.5	1679.6	0	
21	NG	20.0	-272	45.5	1679.6	0	
22	NG	17.0	80	133.8	1379.6	0	
23	NG	325.0	396	211.6	1679.6	0	
24	Air	17.0	75	132.5	700	11.3	
25	Air	21.0	109	136.11	700	11.3	
26	Air	20.5	109/450		700	11.3	
27	Air	18.0	90/300		700	11.3	
28	Air	17.5	97	133.2	700	24.4	
29	WA	17.0	75			13.0	
30	WG	23.5	-289.5	35.0	541.8	0	
31	WG	21.0	-272	39.9	541.8	0	
32	WG	17.0	80	118.0	541.8	0	
33	OL	46.0	-289	-43.1	1373.7	0	
34	RO	290.0	150	128.5	2862.9	20.8	
35	RO	289.0	97	116.7	2862.9	20.8	
36	WA	289.0	97			16.1	
37	RO	53.0	97	120.7	2862.9	4.82	
38	RO	50.0	-251	43.1	2862.9	0	
39	RO	20.0	-292	35.0	2862.9	0	
40	RO	18.0	-272	39.8	2862.9	0	
41	RO	15.0	87	118.9	2862.9	1.82	

*See Figure 3.2-12.

Table 3.2-VIII.

Nitrogen subsystem process conditions, 2500-lb N_2/hr capacity.

Point*	Fluid	Pressure (psia)	Temperature (°F)	Enthalpy (BTU/lb)	Flow (lb/hr)	Water rate (lb/hr)	Remarks
1	Air	14.7	77	128.5	3353.2	33.8	1. Enthalpy is on dry basis.
2	Air	97.0	242	167.8	3353.2	33.8	2. WA = water.
3	Air	96.2	97	132.8	3353.2	19.3	3. OL = oxygen-rich liquid.
4	WA	96.2	97			14.5	4. NL = nitrogen liquid.
5	Air	96.0	97	132.8	3353.2	19.3	5. NG = gaseous nitrogen.
6	Air	95.5	75	127.5	3353.2	9.7	6. WG = waste gas.
7	WA	95.5	75			9.6	7. RO = electrolyzer O_2.
8	Air	95.0	75	127.5	3353.2	9.7	
9	Air	92.0	90	131.1	3353.2	0	
10	Air	90.0	-252.5	45.92	3353.2	0	
11	Air	89.0	-279	36.43	3353.2	0	
12	OL	89.0	-280	-39.55	2063.2	0	
13	OL	45.0	-293	-44.5	2063.2	0	
14	NL	88.0	-286	-36.6	1290.0	0	
15	NL	22.0	-314	-36.6	29.8	0	
16	NL	88.0	-286	-36.6	1260.0	0	
17	NG	22.0	-314	34.0	0.58	0	
18	NL	87.0	-290	-38.0	1260.0	0	
19	NG	22.0	-314	34.0	2514.4	0	
20	NG	21.0	-291	40.5	2514.4	0	
21	NG	20.0	-274	45.0	2514.4	0	
22	NG	17.0	80	133.8	2514.4	0	
23	NG	325.0	396	211.6	2514.4	0	
24	Air	17.0	75	132.5	1050.0	17.0	
25	Air	21.0	109	136.1	1050.0	17.0	
26	Air	20.5	109/450		1050.0	17.0	
27	Air	18.0	90/300		1050.0	17.0	
28	Air	17.5	97	133.2	1050.0	36.5	
29	WA	17.0	75			19.5	
30	WG	23.5	-289.5	35.0	809.6	0	
31	WG	21.0	-271	39.4	809.6	0	
32	WG	17.0	80	118.0	809.6	0	
33	OL	46.0	-289	-43.1	2063.2	0	
34	RO	290.0	150	128.5	4285.7	31.0	
35	RO	289.0	97	116.7	4285.7	31.0	
36	WA	289.0	97			24.0	
37	RO	48.0	97	120.8	4285.7	7.2	
38	RO	46.0	-255	42.4	4285.7	0	
39	RO	20.0	-292	35.0	4285.7	0	
40	RO	18.0	-274	39.3	4285.7	0	
41	RO	15.0	87	118.9	4285.7	7.2	

*See Figure 3.2-12.

The precooled air is directed through another separator (D-3) which removes condensed water and oil as well as oil vapor from the air stream. The air then passes through an adsorbent type purifier (R-1) for the removal of residual water vapor and carbon dioxide before it goes to the cold section of the plant. The purifier has two adsorbent vessels—one is in service while the other is being reactivated. Reactivation is accomplished by an air loop which includes a reactivation blower (C-4), a reactivation heater (H-1), an air-cooled gas cooler (E-8), and a refrigerated gas cooler (E-9). During the heating phase of the reactivation cycle, ambient air is pumped by means of the blower through the reactivation heater which warms it to a temperature of about 450°F. This hot air is directed through the air purifier vessel being reactivated where it removes the water and carbon dioxide adsorbed while the bed was in service. The warm air containing the water and carbon dioxide is rejected to the atmosphere. After the air purifier bed is thoroughly heated, it is cooled down by circulating air in a closed loop. The reactivation heater is shut off and air from the blower is passed through the air purifier bed for cooling. The warm air leaving the bed is passed through the air-cooled gas cooler (E-8) and the refrigerated gas cooler (E-9) and is cooled to a temperature of about 75°F. After cooling, the air is returned to the suction side of the blower to form a closed loop and prevent the introduction of water and carbon dioxide during the cooling process. After the air purifier vessel is cooled, it is put on stream and the other air purifier vessel is taken out of service for reactivation.

The air leaves the purifier and passes through a dust filter (F-1) for removal of any dust particles carried over from the air purifier bed. It then enters the cold box which houses the equipment operating at subatmospheric temperatures. The air is first cooled in the main heat exchanger (E-1) by means of the product streams from the air fractionation column. Further cooling and partial liquefaction are accomplished in a refrigerated heat exchanger (E-2). Cooling is supplied by the product streams from the column as well as a stream of cold oxygen which provides the low temperature refrigeration required for the process.

The oxygen used for refrigeration is the waste oxygen from the hydrogen production subsystem. This oxygen is at a pressure of 300 psig and a temperature of approximately 167°F. It is piped to the nitrogen generation subsystem and is first cooled in an air-cooled oxygen cooler (E-10) to within 20°F of ambient temperature. After cooling, it passes through a separator (D-6) which removes any condensate. The oxygen from the separator passes through a pressure regulating valve which maintains the proper back pressure on the hydrogen production subsystem.

The oxygen is then directed to the oxygen reversing heat exchanger (E-3) where the incoming oxygen is cooled by the exiting oxygen stream. Flow through the two passages of the exchanger is alternated at frequent intervals so that water deposited from the warm oxygen stream is re-evaporated into the returning cold oxygen. After passing through the exchanger, the oxygen goes through an accumulator (D-4) which is filled with silica gel. The accumulator damps out

pressure pulsations in the oxygen circuit and removes water from the oxygen during the start-up period before the normal temperature gradients are established in the exchanger.

The oxygen from the accumulator is directed to an expansion turbine (X-1) where it is expanded isentropically to low pressure. This results in a large temperature drop for the oxygen and provides the low temperature refrigeration required for the air fractionation process. The cold oxygen from the expander passes through the refrigerated heat exchanger (E-2) to help cool the process air and then returns through the oxygen reversing heat exchanger and rejected to the atmosphere.

The cold process air from the refrigerated heat exchanger (E-2) is fed to the high pressure column (T-1) where it is separated into a nitrogen-rich fraction and an oxygen-rich fraction. Part of the nitrogen overhead from the column is condensed to provide the reflux. The refrigeration required to condense the reflux is provided by the oxygen-rich product which is withdrawn as liquid from the bottom of the column and fed to the low pressure side of the overhead condenser. A part of the nitrogen-rich vapors from the top of the high pressure column and the oxygen-rich vapors from the low pressure side of the condenser are fed to separate reboilers. The reboilers are located in the bottom of the low pressure column (T-2). The vapors are liquefied, subcooled in two subcoolers (E-4) and (E-5), and fed to the low pressure column. The nitrogen-rich liquid enters at the top of the column and the oxygen-rich liquid is introduced at an intermediate point. Final separation of the nitrogen from argon and oxygen is accomplished in the low pressure column. The final nitrogen product is withdrawn from the top of the column and, after passing through the subcoolers (E-4 and E-5) to cool the feed streams, is directed through the refrigerated heat exchanger (E-2) and main heat exchanger (E-1) to help cool the incoming process air. The nitrogen product leaves the air fractionation unit at low pressure and a temperature of approximately 80°F and is delivered to a nitrogen compressor for pressurization to approximately 310 psig. The pressurized nitrogen product is then piped to the ammonia synthesis subsystem.

The results of later work (presented in the Integration and Packaging subsection) indicated that in some cases, it is advantageous to pressurize the nitrogen in the multiservice compressor located on the ammonia skid. Consequently, for the final designs, the nitrogen is compressed in the multiservice ammonia compressor for Systems 1 through 8 and in a separate compressor on the nitrogen skid for Systems 9, 10, and 11.

The oxygen-rich product from the bottom of the low pressure column leaves the nitrogen generation subsystem through exchangers E-2 and E-1 and is then rejected to the surroundings.

Heat is removed from the process in an air-cooled heat exchanger which combines coolers E-6, E-7, E-8, E-10, and E-12. Intercooling and jacket cooling for the nitrogen compressor is accomplished by a recirculating water system (not shown in Figure 3.2-12) which forms the

sixth flow circuit in the air-cooled heat exchanger. A fan (C-5) forces ambient air across the extended surfaces in the air-cooled heat exchanger and removes heat from the process streams.

The general characteristics of the major items of equipment required for the nitrogen generation subsystem are as follows.

- Electric Motors—Five electric motors are required as prime movers for the compressors and blowers. Wherever possible, the speed of the motor was selected to permit use of a direct drive. Motors were selected for use with 460-v, 3-phase, 60-cps power.

- Compressors and Blowers—For the air and nitrogen compressors, the flow rates for all plant sizes dictate the use of reciprocating type machines. Rotary and centrifugal compressors are generally not available for the low flow rates involved.

  The air compressors are two-stage, air-cooled machines and are of a semiradial configuration (one or two rows of cylinders depending on the capacity). In all cases, a direct drive at 1200 rpm is used. These machines are relatively light and small and provide for dependable, efficient, economical service.

  The nitrogen compressor, which operates at a discharge pressure of about 310 psig, requires a different type of machine. Heavy duty, double acting, water-cooled, two-stage compressors were selected. These compressors consist of a closed circuit water cooling system with a circulating pump and an air-cooled exchanger. When selecting the compressors, a tabulation was made of weight, size, and cost for all compressors offered. The machine offering an optimum combination of these factors was chosen.

  The refrigerant compressor (C-3) is a multiple cylinder, relatively high speed unit which is driven directly by an electric motor. Individual cylinder unloading is possible and provides for capacity reduction to 33%. Pressure or liquid temperature are used for automatic capacity control.

  A rotary, positive displacement, lobe type unit was selected for the reactivation air blower. This blower delivers clean, dry air. The unit has two impellers mounted on parallel shafts which rotate in opposite directions. This type of blower is most suitable for applications requiring a discharge pressure on the order of 4 psig and constant airflow. An exhaust fan with an air handling wheel was selected for the air-cooled exchanger which removes heat from several process streams.

- Continuous Drainers—Continuous, float-operated drainers are provided for the automatic removal of condensate at various points in the process. The rise of condensate level acts on a float ball to open a valve and discharge the condensate.

- Tanks—Tanks are used for several purposes in the plant. Two entrainment separators are provided in the process air line to remove condensate. Each is a simple cylindrical vessel provided with filters and settling chambers to remove all liquid from the air. The separator ahead of the purifier is provided with a charcoal cartridge for removal of oil vapor from the air.

  A separator is also provided in the refrigeration oxygen line downstream of the air-cooled exchanger to remove condensed water from the oxygen before it goes to the reversing exchanger. This is a stainless steel cylindrical vessel with mist eliminators and settling chamber.

  The refrigerant receiver is a simple cylindrical pressure vessel used for storage of liquid refrigerant.

  The accumulator ahead of the oxygen expander is a simple cylindrical vessel made of silicon bronze and is filled with silica gel. Screen supports and filling ports are provided.

  After considering and investigating several alternate methods for achieving quick start-up of the plants, it was concluded that the best way was to accumulate liquid nitrogen in a storage tank during normal operation and then feed this nitrogen back into the plant when starting after a shutdown to cool the plant rapidly. The liquid nitrogen storage tank is a vertical, vacuum-perlite-insulated tank. The inner tank is fabricated of stainless steel, and the outer tank is made of carbon steel. The inner tank is suspended in the outer tank by means of a low conductance suspension system.

  The silencer in the discharge line for the refrigeration oxygen is a stainless steel cylindrical vessel designed to eliminate discharge noise.

- Heat Exchangers—For the low temperature heat exchangers in the cold box, it is important to hold weight and size to a minimum to reduce heat leak and start-up time. Two types of heat exchangers have proven suitable for low temperature air fractionation plants—brazed aluminum plate and fin type and helical coil-in-shell type which are usually fabricated of copper or aluminum. For low pressure plants, brazed aluminum heat exchangers are superior because they are smaller and lighter. Brazed aluminum heat exchangers were used for the main air exchanger (E-1), the refrigerated exchanger (E-2), and the oxygen reversing exchanger (E-3). For very small heat transfer areas—viz, the crude oxygen and nitrogen subcoolers, (E-4) and (E-5)—the helical coil type exchanger was used because of cost considerations. These are small cylindrical units made of copper.

  The air-cooled exchanger for removal of high level heat from process streams must be compact and light. The use of an extended surface type unit is almost mandatory. Several

of the available units were investigated and an inner-fin type unit appeared to be the most suitable for this application. The fundamental principle of the inner-fin is the arrangement of the fins inside of a tube in conjunction with outer fins. The inner-fin provides a large inside surface area and a small hydraulic radius, thus permitting a more rapid transfer of heat. One air-cooled heat exchanger assembly incorporates the air compressor aftercooler (E-6), the refrigerant condenser (E-7), the reactivation gas cooler (E-8), the oxygen cooler (E-10), the nitrogen compressor aftercooler (E-12), and the cooling water exchanger for the nitrogen compressor.

There are two R-12 refrigerant cooled exchangers in the plants—one for final precooling of the process air before it goes to the air purifiers and the other for final cooling of the reactivation air during the cooling phase of the cycle. Helical coil-in-shell units were used for these exchangers. They are fabricated of copper and the air is on the shell side.

- Filters—A dust filter is needed downstream of the air purifier unit to remove any dust from the purifiers which might cause damage in the air fractionation section of the plant. This is a commercial filter with replaceable type elements capable of removing all particles greater than 5 microns. Air intake filters are required for the air compressor and the reactivation blower. Because of the frequent movement of the plant and the extremes in environmental conditions, the dry type filter was selected.

- Electric Heaters—An electric heater is used for heating the reactivation air during the reactivation cycle. The heater is designed for a 460-v, 3-phase, 60-cps power supply and is controlled by a temperature operated switch and a cycle timer.

- Air Purifiers—A purifier is used to remove water and carbon dioxide from the process air to avoid obstructions due to freeze-out at the low temperature levels.

  The air purifier consists of two adsorber vessels—one is reactivated while the other is in service. A two-layer adsorber bed is used in each vessel. The process air passes through the activated alumina for removal of water and then through molecular sieve for removal of carbon dioxide. The impurity level is reduced to less than 5 ppm.

  The adsorbers are cylindrical pressure vessels fabricated of stainless steel and provided with screen supports for retaining the adsorbent materials. Fill and dump ports are provided for replacement of the adsorbent material when required.

- Air Fractionation Column—A split double column operating at two pressure levels is used for air fractionation. The high pressure section operates at about 89 psia and the low pressure section at about 23 psia. The results of the computer analysis of the fractionation column, presented in the subsection titled Investigation, indicated that the final nitrogen purity is 99.83% and that the high and low pressure columns have 17 and 30 actual trays, respectively.

The high pressure column is equipped with a liquid sump on the bottom and a condenser on the top. The sump serves as a reservoir for the bottoms product. The overhead condenser is required to generate the liquid reflux for the column.

The low pressure column is equipped with only a reboiler on the bottom. As mentioned earlier, the high pressure column overhead product is condensed and used to provide the reflux required in the rectifying section of the low pressure column. Hence, there is no need for a condensing section on the low pressure column.

As mentioned previously, bubble-cap trays were chosen as the contacting device. The specific tray design specified has proved to operate successfully under shipboard conditions where a permanent pitch up to 5° may exist. Using this design, column performance will be satisfactory even though the skid is a few degrees off level. The trays, called annular trays, will be spaced 2 in. apart.

- Expansion Turbine—An expansion turbine with gas bearings and centrifugal gas compressor loading is used to provide low temperature refrigeration. Depending on the size of plant, the turbine wheel diameters vary from 1.7 to 3.9 in., and the operating speeds range from 87,200 to 28,800 rpm. The expander is very small and light.

In addition to the basic components described, the following items are required.

- Insulation—Insulation is required in the nitrogen generation subsystem to conserve heat or cold and to protect personnel. In the air fractionation section, temperatures to -300°F are encountered. To reduce heat leak, this section is enclosed in a "cold box" which is filled with loose glass wool insulation suitable for these low temperatures.

  Outside of the cold box, low temperature lines such as those for the low temperature refrigerant are covered with foam rubber insulation to prevent sweating. Hot lines, such as compressor discharge lines, are covered with an asbestos base insulation for personnel protection.

- Insulation Case (Cold Box)—The cold box consists of a structural aluminum frame to which aluminum sheet panels are attached. Most of these panels are removable to provide easy access.

- Skid—All plant equipment is mounted on a skid to permit transportation. Aluminum and steel construction materials were compared and aluminum was selected because of its light weight and resistance to corrosion. The basic members of the skid are channels and the frame is covered with plate.

Weight data were obtained by preparing bills of material for each plant size. The weights of major components were determined from vendor information. A summary of the major component and total subsystem weights is given in Table 3.2-IX.

Table 3.2-IX.
Weight breakdown for nitrogen generation subsystem.

Plant size (lb N_2/hr)	Weight (lb) 500	1000	1670	2500
1. Electric motors	2,360	3,820	5,684	7,636
2. Compressors and blowers	6,535	7,710	10,779	21,774
3. Continuous drainers	32	32	32	32
4. Tanks	1,287	2,030	3,663	6,299
5. Heat exchangers	840	1,850	2,805	3,990
6. Filters	60	91	98	142
7. Reactivation heater	72	135	50	75
8. Air purifiers	450	900	1,500	2,224
9. Fractionation columns	680	1,213	2,113	2,880
10. Turbo expander	10	12	15	20
11. Valves	2,297	2,938	3,609	4,835
12. Instruments	269	269	269	269
13. Piping	717	980	1,500	2,310
14. Electrical controls	216	338	2,248	2,320
15. Electric wiring and conduit	143	196	300	460
16. Cold box	451	551	600	762
17. Control cabinet	215	215	215	215
18. Skid	1,761	1,932	3,035	4,660
19. Insulation	790	940	1,130	1,510
20. Miscellaneous	100	130	150	175
Total	19,285	26,282	39,795	62,588

Operation at several alternate off-design conditions is of interest in connection with the design of the nitrogen generation plants. However, the only off-design condition which requires equipment modification is an ambient temperature below the freezing point of water. Under this condition, the automatic drainers and the water separators could freeze. To prevent freezing, these components should be insulated and heated electrically. The required equipment would weigh about 50 lb. The power required would be insignificant.

A change in the frequency of the electrical power source from 60 cps to 400 cps primarily affects the electric motors. The difference in weights and sizes for the two sources is not significant.

Layout drawings were prepared for the four plant sizes and the major pieces of equipment were arranged so that intercomponent connections could be simplified and the equipment made accessible. The layout drawings are shown in Figures 3.2-13 through 3.2-16. The nomenclature used is the same as defined earlier in Table 3.2-IV.

PREDICTED PERFORMANCE

Electric power is the only utility required for the nitrogen generation subsystems. A breakdown of power requirements for the various subsystem sizes is given in Table 3.2-X. The power requirement for the refrigerant compressor varies considerably with the ambient temperature. In very cold weather, auxiliary refrigeration will not be needed and there will be no power required for the refrigerant compressor. However, in very warm weather, the requirements for auxiliary refrigeration will be high and the power required for the refrigerant compressor will increase. Figure 3.2-17 shows this effect. The form of the power requirement curve results from the unloading of compressor cylinders as the refrigeration load decreases.

Table 3.2-X.

Nitrogen generation subsystem power distribution loads at design conditions.

Subsystem capacity (lb N_2/hr)	500		1000		1670		2500	
	Load	Rating	Load	Rating	Load	Rating	Load	Rating
Air compressor- hp	37.2	40.0	69.9	75.0	120.5	125.0	198.0	200.0
Nitrogen compressor (315 psia)—hp	33.6	40.0	77.5	75.0	130.0	150.0	200.0	200.0
Refrigerant compressor—hp	2.5	10.0	3.3	15.0	4.0	15.0	5.0	20.0
Reactivation heater—hp	6.0	6.0	12.0	12.0	20.0	20.0	26.7	26.7
Reactivation blower—hp	2.0	2.0	3.1	3.0	4.5	5.0	6.3	7.5
Cooling air fan—hp	0.61	0.75	0.91	1.0	1.46	1.5	2.76	3.0
Controls--hp	0.50	0.66	0.6	0.66	0.80	1.0	0.9	1.0
Total (horsepower)	87.41	99.41	167.31	131.66	281.36	317.5	439.66	158.2
(kw)	65.4	74.5	125.3	136.0	210.0	287.5	329.0	343.0

DISCUSSION

One important characteristic of the nitrogen generation subsystem is its dependence, as designed, on oxygen from the hydrogen production subsystem. This oxygen is needed to provide low temperature refrigeration in the air fractionation section. Any occurrence affecting the supply of oxygen from the hydrogen production subsystem will have a resultant effect on the nitrogen generation subsystem. This interdependence between subsystems is dictated by savings in power, but does lead to a reduction in flexibility.

In the design and selection of equipment, considerable reliance has been placed on engineering judgement. Portable nitrogen plants of the type investigated are by no means new. This type of plant has been under study for over 20 years—during this period many process variations and equipment designs have been explored and tested. The proposed plant represents the state of the art.

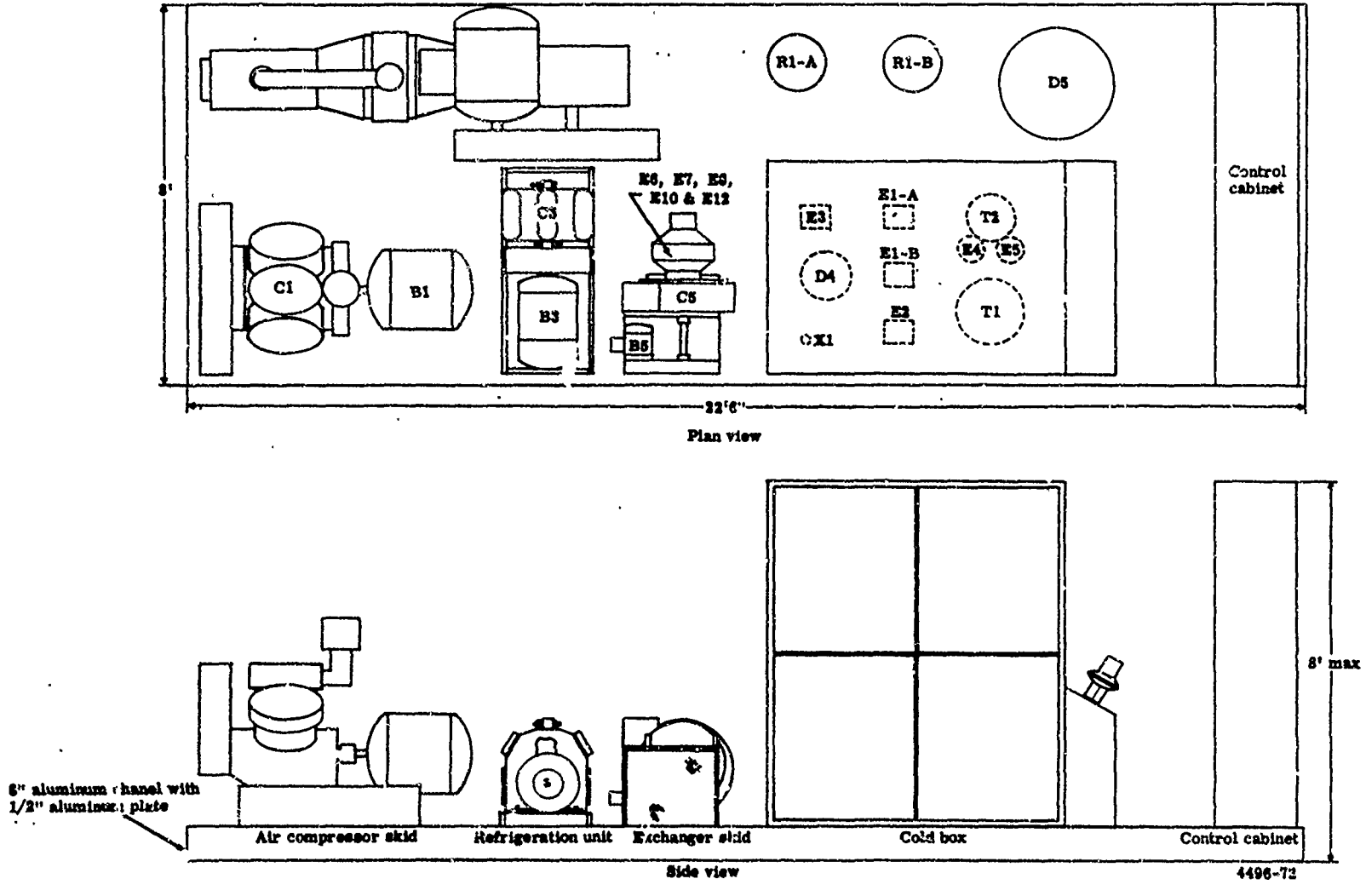


Figure 3.2-13. Nitrogen subsystem skid layout—Class I (500 lb N_2/hr capacity).

Allison

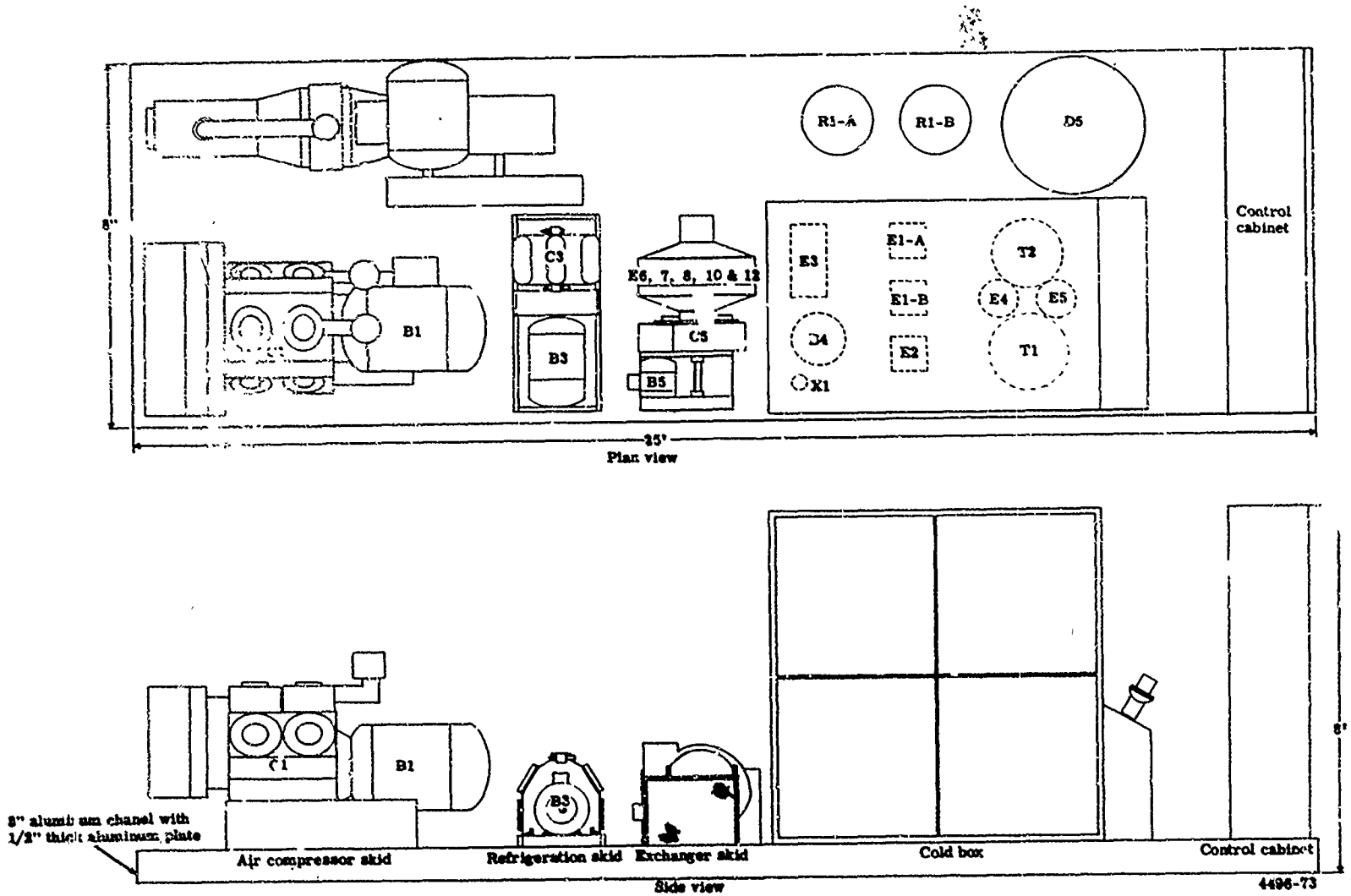


Figure 3.2-14. Nitrogen subsystem skid layout—Class I (1000 lb N_2/hr capacity).

Allison

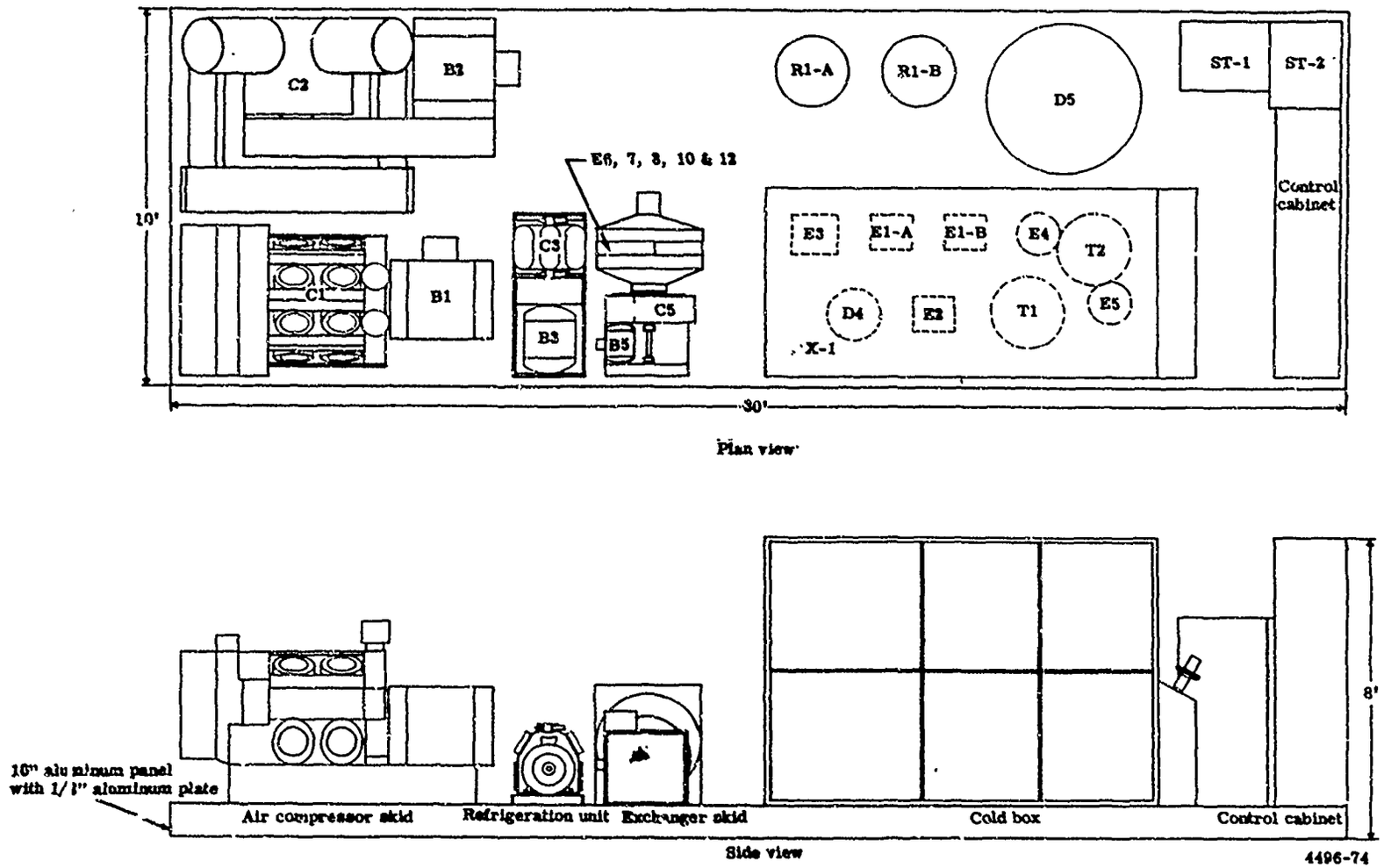


Figure 3.2-15. Nitrogen subsystem skid layout—Class II (1670 lb N_2/hr capacity).

Allison

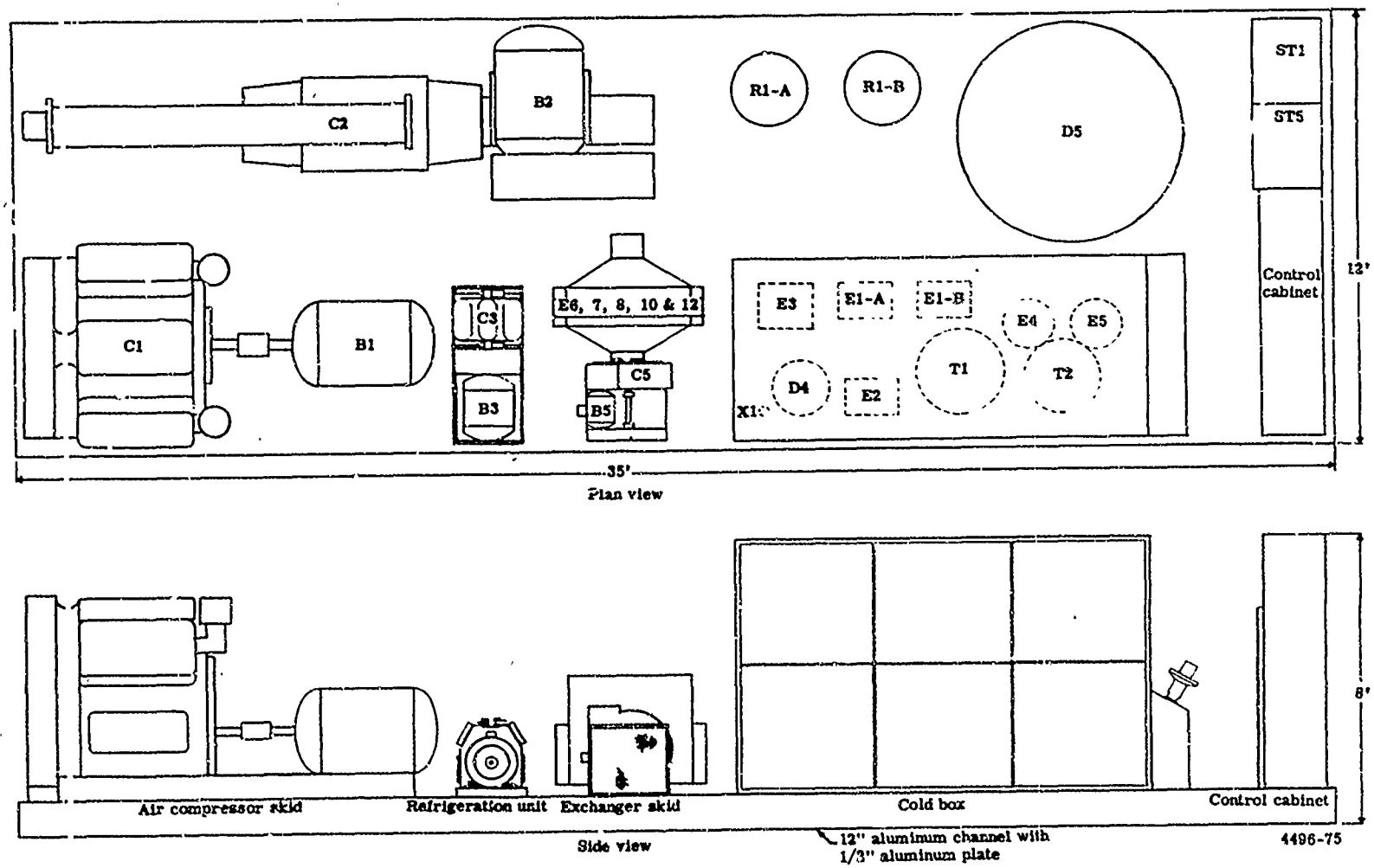


Figure 3.2-16. Nitrogen subsystem skid layout—Class II (2500 lb N_2/hr capacity).

Allison

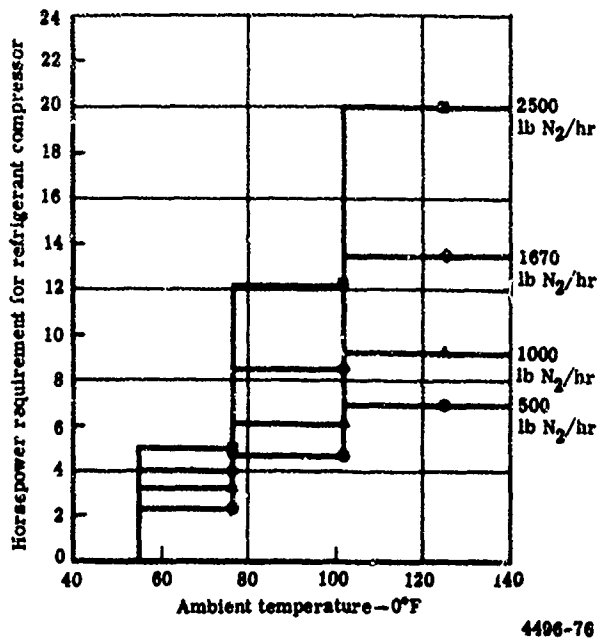


Figure 3.2-17. Horsepower requirement for refrigerant compressor in nitrogen subsystem as a function of ambient temperature (two-step unloading).

Considering power requirement, size, weight, and cost, it is believed that the proposed nitrogen generation subsystem represents the best in type and design for the intended use. Present engineering know-how permits the manufacture of these plants with a minimum amount of development and a high degree of reliability. The plant, as described, would provide a dependable subassembly for integration into the complete Energy Depot system.

Additional study and development effort is recommended as follows before "fixing" the final design of the nitrogen generation subsystem.

- With the current expander designs in the size range of interest for the proposed plants, recovery of useful work from the expander is difficult, if not impossible. The air compressed by the centrifugal compressor used for loading the turbine might find some use in the plant. Further study of the loading characteristics of turboexpanders is recommended. It is also recommended that the possibility of recovering mechanical or electrical work from the expander be investigated more thoroughly.
- Part of the water in the process air is removed by refrigeration and condensation and the remainder by adsorption. The amount of water removed in each step depends on the temperature to which the air is cooled before it goes to the adsorber. For the present study, a temperature of 75°F was used since experience indicates that this temperature gives reasonable designs for the refrigerated exchanger and the adsorber. It is recommended that the effect of varying this precooling temperature on equipment design and power requirements be studied to determine the optimum value. The present plant designs will permit adjustment of this temperature to the optimum point.

CALCULATIONS

- Derivation of Liquid Nitrogen Draw-off Rate Equation

Let W_A = process airflow, lb/hr
W_N = liquid nitrogen flow, lb/hr
q_L = heat leak, BTU/lb process air
θ = minimum operating time
θ' = average shutdown time
Q_L = accumulated heat leak into system during shutdown, BTU
N_R = liquid nitrogen required to cool down plant after a shutdown, lb
Δh_N = available refrigeration per pound liquid nitrogen, BTU/lb

Then $Q_L = W_A q_L \theta'$

Assume that liquid nitrogen for cooling enters the system as saturated liquid at 85 psia and leaves the system at 15 psia and 80°F:

$\Delta h_N = 133.8 + 39 = 172.8$
$N_R = W_A q_L \theta'/172.8$
$W_N = W_A q_L \theta'/(172.8)(\theta)$
$\theta/\theta' = 3$
$W_N = W_A q_L/(172.8)(3)$
$= W_A q_L/518.4$

- Sample Process Calculation (Process N)

The process points for Process N are shown in Figure 3.2-18.

- Fixed conditions

Assume nitrogen and waste gas product streams are discharged at 17 psia to allow some pressure drop for transfer and control.

$P_9 = P_{10} = 17$ psia
$P_5 = 17 + 5 = 22$ psia

Assume P through low pressure column is 1.5 psi. Then, $P_6 = P_5 + 1.5 = 22 + 1.5 = 23.5$ psia (condenser pressure).

Now determine the condenser temperature. The condenser liquid is in equilibrium with waste gas which contains 6% nitrogen. From equilibrium data, liquid composition is 1.2% nitrogen. At 23.5 psia, the temperature of the liquid is -289°F.

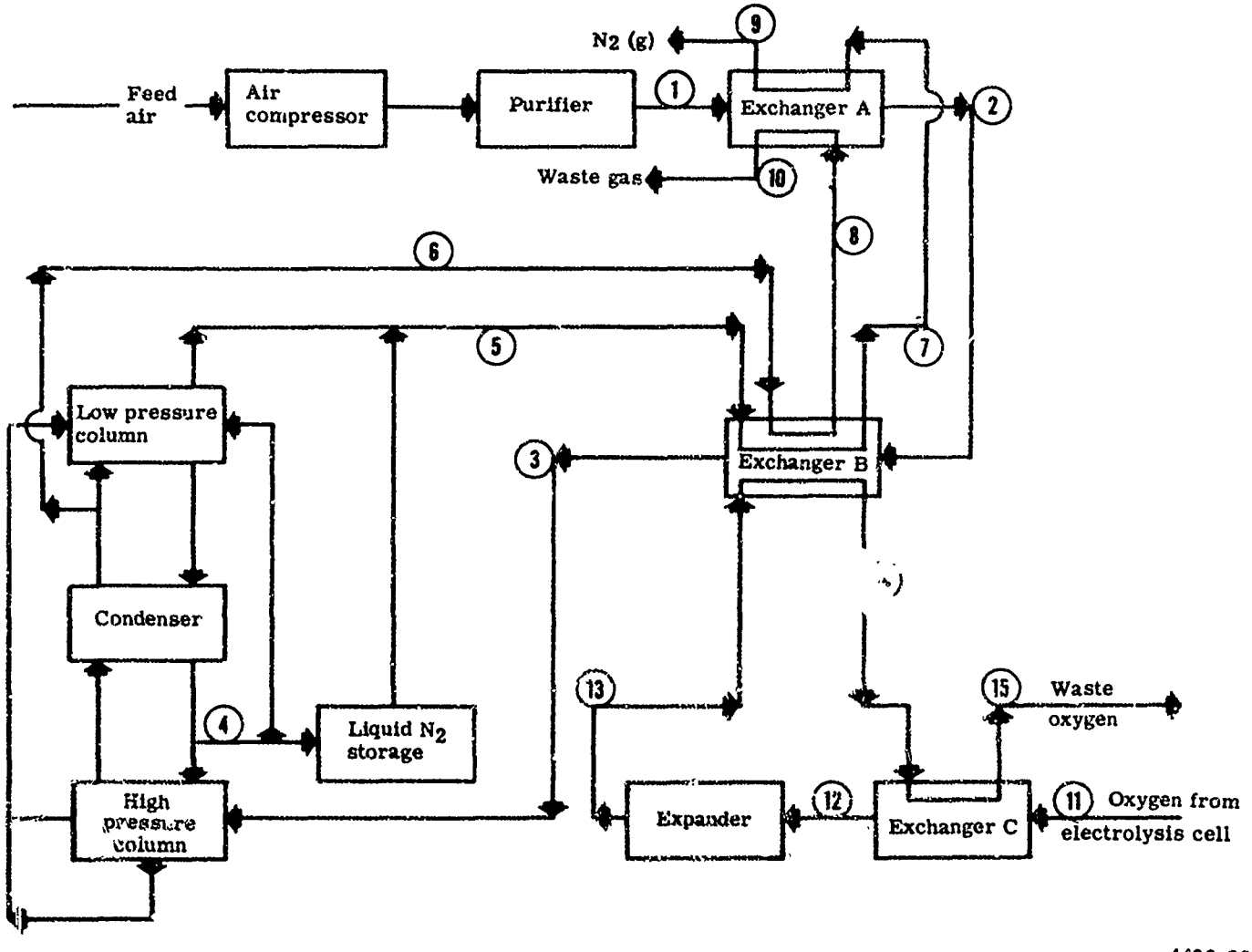


Figure 3.2-18. Nitrogen subsystem Process N process points.

Allison

The temperature of the condensing nitrogen is -289 + 3 = -286°F.
The pressure of the condensing nitrogen is 88 psia.
Allow 1-psi pressure drop through high pressure column.
The pressure of feed to the column is P_3 = 88 + 1 = 89 psia.
The inlet air pressure is P_1 = 89 + 3 = 92 psia.
The pressure of air to the liquefier is P_2 = 92 - 2 = 90 psia.
The pressure of nitrogen from the liquefier to the exchanger is P_7 = 22 - 2 = 20 psia.
The pressure of waste gas from the liquefier to the exchanger is $P_8 = P_6 - 2.5$ = 23.5 - 2.5 = 21 psia.

Determine the enthalpy of waste gas leaving the system.
h for oxygen at 17 psia and 80°F = 117.3 BTU/lb
h for air at 17 psia and 80°F = 129.2 BTU/lb
h for nitrogen at 17 psia and 80°F = 133.8 BTU/lb
From plot of data, h for waste gas (6% N_2) = 118.0 BTU/lb

- Overall nitrogen balance

Let A = moles of process air
N_L = moles liquid nitrogen
N_G = moles of gaseous nitrogen product
G = moles of waste gas
$0.79A = (0.995)(1.0) + (0.995) N_L + (0.06)(A-1-N_L)$
$N_L = Aq_L/518.4 \times (28)/(28.84)$
q_L = 5 BTU/lb process air
$N_L = (A)(5)(28.4)/(518.4)(28) = 0.00977A$
$0.79A = 0.995 + 0.00973A + (0.06)(0.99023A-1)$
$0.72077A = 0.935$
A = 1.297 moles
$N_L = (0.00977)(1.297) = 0.0127$ moles
G = 1.2970 -1-0.0127 = 0.2843 moles
N_G = 1.00

- Overall heat balance

Heat in, $Q = W_1h_1 + W_1q_L$
$W_1 = 1.297 \times 28.84 = 37.4$
h_1 = 131.1
q_L = 5.0
$Q = (37.4)(131.1 + 5.0) = 5090$

Heat out, $Q' = W_q h_q + W_{10} h_{10} + W_4 h_4 + Q_R$

$W_q = 28$

$h_q = 133.8$

$W_4 = 0.0127 \times 28 = 0.355$

$h_4 = 36.3$

$W_{10} = 37.4 - 28.0 - 0.355 = 9.045$

$h_{10} = 118.0$

$Q' = (28)(133.8) + (9.045)(118.0) + (0.355)(-36.3) + Q_R = 4789 + Q_R$

$Q = Q'$

$Q_R = 5090 - 4799 = 291$ BTU

- Heat leak distribution

  Total heat leak = $5 \times 37.4 = 187$ BTU

  Main exchanger = $187 \times 0.15 = 28$

  Refrigerated exchanger = $187 \times 0.15 = 28$

  Columns = $187 \times 0.65 = 122$

  Liquid N_2 storage tank = $187 \times 0.05 = 9$

  $\overline{187}$ BTU

- Column heat balance

  $P_5 = 22.0$

  $h_5 = 34.0$ (saturated nitrogen vapor)

  $P_6 = 23.5$

  Determine h_6 by interpolation

  For saturated N_2 at 23.5 $h = 34.6$

  For saturated air at 23.5 $h = 34.6$

  For saturated O_2 at 23.5 $h = 35.0$

  For 6% nitrogen $h_6 = 35.0$

  Heat balance on the column is:

  $Q = W_3 h_3 + 122 + 9$

  $W_3 = W_1 = 37.4$

  $Q = 37.4\, h_3 + 131$

  $Q' = W_4 h_4 + W_5 h_5 + W_6 h_6$

  $= (0.355)(-36.3) + (28.0)(34.0) + (9.045)(35.0) = 1255$

  $Q = Q'$

  $h_3 = (1255 - 131)/37.4 = 30.1$

- Heat balance on refrigerated exchanger (liquefier)

  Oxygen available from electrolyzer is 1.5 moles or $1.5 \times 32 = 48$ lb

Take P_{13} = 20 psia, T_{13} = -292°F, and h_{13} = 34.6.

$Q = W_2h_2 + W_5h_5 + W_6h_6 + W_{13}h_{13} + 28$

$W_2 = 37.4$

$W_5 = 28$

$h_5 = 34.0$

$W_6 = 9.045$

$h_6 = 35.0$

$W_{13} = 48$

$h_{13} = 34.6$

$Q = 37.4\ h_2 + (28)(34.0) + (9.045)(35.0) + (48)(34.6) + 28 = 37.4\ h_2 + 2956$

From overall heat balance, Q_R = 291 BTU.

$Q_R = W_{14}h_{14} - W_{13}h_{13}$

$W_{13} = W_{14} = 48.0$ lb

$48\ (h_{14} - 34.6) = 291$

$h_{14} = 6.06 + 34.6 = 40.66$

$t_{14} = -268°F$

Assume $T_7 = T_8 = T_{14} = -268°F$. Then, $h_7 = 45.7$. Obtain h_8 by interpolation. At -268°F and 21 psia:

h for N_2 = 46.6

h for Air = 45.0

h for O_2 = 40.8

For 6% nitrogen, $h_8 = 40.7$.

$Q' = W_3h_3 + W_7h_7 + W_8h_8 + W_{14}h_{14}$

$W_3 = 37.4$

$h_3 = 30.1$

$W_7 = 28.0$

$h_7 = 45.7$

$W_8 = 9.045$

$h_8 = 40.7$

$W_{14} = 48.0$

$h_{14} = 40.66$

$Q' = (37.4)(30.1) + (28.0)(45.7) + (9.045)(40.7) + (48.0)(40.66) = 4754$

$Q = Q'$

$37.4\ h_2 + 2956 = 4754$

$h_2 = 48.0$ and $t_2 = -245°F$ (at 90 psia)

Check temperature gradients in exchanger for possible violation of the second law of thermodynamics.

Dew point of air at 90 psia = -278°F.

Enthalpy at dew point = 38.0.

Total heat transferred = (37.4) (48 - 30.1) = 670 BTU.

Latent heat transferred = (37.4) (38.0 - 30.1) = 296 BTU.

Oxygen enters at -292°F and leaves at -268°F.

Waste gas enters at -289°F and leaves at -268°F.

Nitrogen gas enters at -314°F and leaves at -268°F.

Air enters at -245°F.

Temperature gradients are plotted in Figure 3.2-19.

Minimum temperature difference will be 2 to 3°F at worst, so design is satisfactory, i.e., no violations of the second law of thermodynamics.

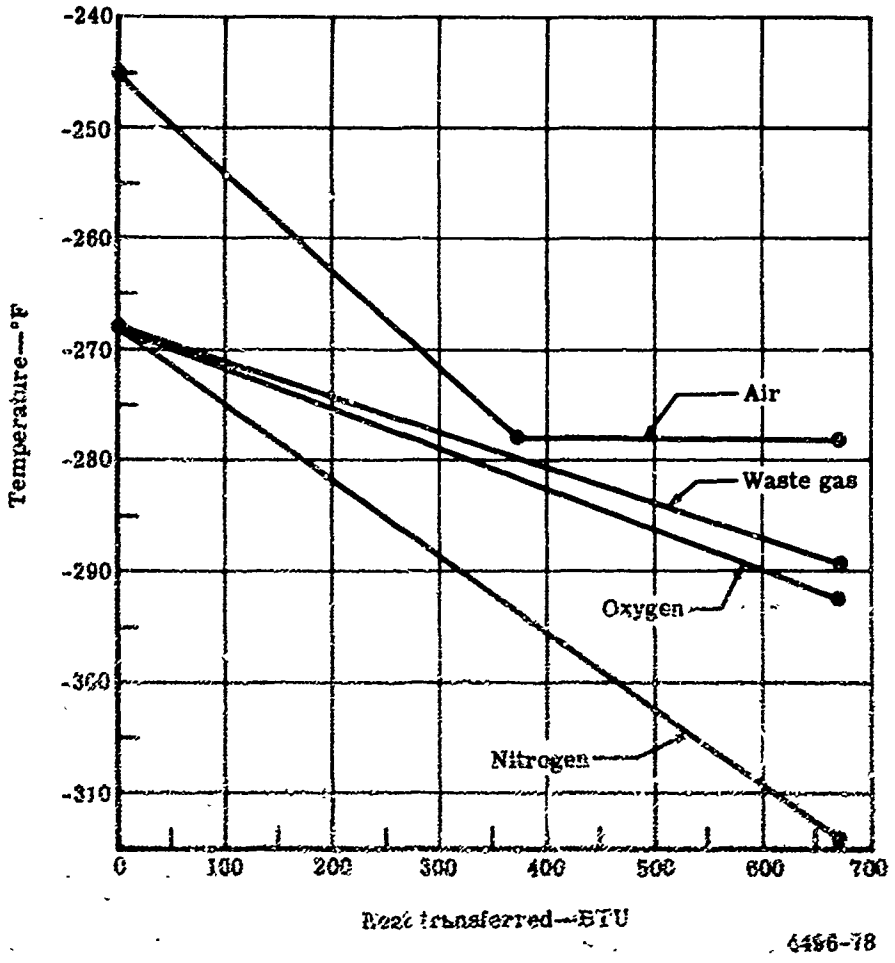


Figure 3.2-19. Temperature gradients in nitrogen subsystem refrigerated exchanger (liquefier).

- Heat balance on main exchanger:

$Q = W_1h_1 + W_7h_7 + W_8h_8 + 28$

$= (37.4)(131.1) + (28.0)(46.7) + (9.045)(40.7) + 28 = 6614$

$Q' = W_2h_2 + W_qh_q + W_{10}h_{10}$

$= (37.4)(48.0) + (28)(113.8) + (9.045)(118.0) = 6607$

Q should equal Q', Q - Q' = 7 which is satisfactory agreement.

- Heat balance on oxygen exchanger:

Assume heat leak = 2 BTU/lb oxygen

$q_L = 2 \times 48 = 96$ BTU

$Q = W_{11}h_{11} + W_{14}h_{14} + 96$

$W_{11} = W_{14} = 48$

$h_{11} = f(P_{11})_{90°}$

$h_{14} = 40.66$

$Q = (48)(h_{11}) + (48)(40.66) + 96 = 48\,h_{11} + 2049$

$Q = W_{12}h_{12} + W_{15}h_{15}$

$W_{12} = W_{15} = 48$

$h_{15} = 117.3$

$h_{12} = \phi(P_{12})$

$Q' = 48\,h_{12} + (48)(117.3) = 48\,h_{12} + 5640$

$P_{11} = P_{12} + 3$

Determine value of P_{11} which will make Q = Q'.

$\phi(P_{12})$ depends on the expander.

Expander efficiency = 75%.

Take P_{12} = 70 psia. Then, P_{11} = 73 psi.

Expander data:

Inlet T	h_{12}	h_s	Δh_s	$e\Delta h_s$	h_{13}
-250°F	42.4	29.2	13.2	9.9	32.5
-245°F	43.6	30.2	13.4	10.05	33.55
-240°F	44.9	31.2	13.7	10.28	34.62

h_{13} should be 34.50.

From plot of data:

$h_{12} = 44.8$

$h_{11} = 119.1$

$Q' = (48)(44.8) + 5640 = 7790$

$Q = (48)(119.1) + 2049 = 7789$

Repeating for P_{12} = 65 psia and P_{11} = 68 psia, obtain:

h_{12} = 44.05 Q' = (48) (44.05) + 5640 = 7758

h_{11} = 119.1 Q = (48) (119.1) + 2049 = 7769

By plotting Q and Q' versus P_{12}, find Q = Q' when P_{12} = 67 psia.

P_{11} = 67 + 3 = 70 psia

- Power Requirements

Compressor discharge pressure = P_1 + 5 = 92 + 5 = 97 psia.

Overall pressure ratio = 97/14.7 = 6.6.

Single-stage compressor is allowable.

The theoretical work of compression for air as a function of pressure ratio is plotted in Figure 3.2-20.

For pressure ratio of 6.6, work = 2680 BTU/lb mole.

Moles of air = 1.297.

Compressor efficiency = 70%.

Inlet temperature = 77°F.

Air compressor work = (2680) (1.297)/(0.70) = 1960 BTU.

For nitrogen compressor, pressure ratio = 3000/15 = 200.

Five stages of compression are required.

Pressure ratio per stage = 200 = 2.89.

For first stage, inlet temperature = 80°F.

Work of compression = (1350) (540)/(537) = 1358 BTU/lb mole.

For other stages, inlet temperature = 97°F.

Work of compression = (1350) (557)/(537) = 1400.

Total compressor work = 1358 + (4) (1400) = 6958.

Actual nitrogen compressor work = 6958/0.70 = 9950.

Total work input = 4960 + 9950 = 14,910.

- Cooling Power Requirements

For preliminary process calculations, the power required for cooling was estimated using the method for air-cooled exchangers described by Smith.[6] As a basis for calculations, assume 1,000,000 BTU/hr of heat to be removed.

Let:

Inlet fluid temperature, T_1 = 300°F

Outlet fluid temperature, T_2 = 97°F

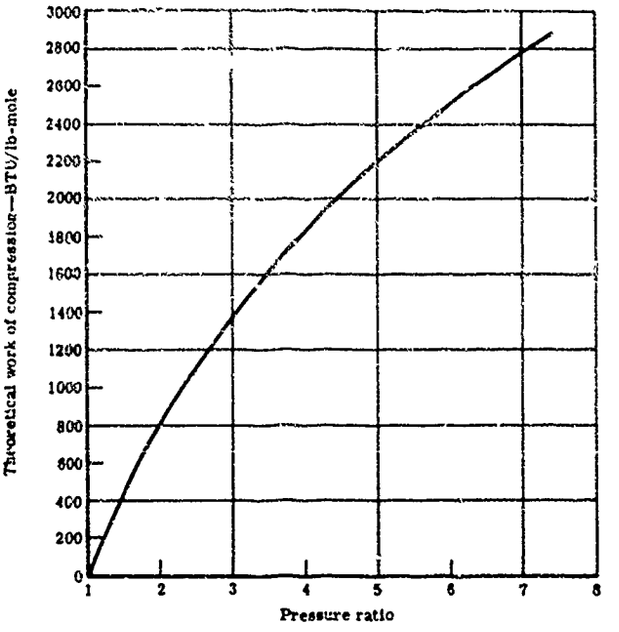
Ambient air temperature, t_1 = 77°F

Assume U = 20. Then,

$T_1 - t_1$ = 300 - 77 = 223

$(T_1 - t_1)/U$ = 223/20 = 11.15

Figure 3.2-20. Theoretical work of compression of air as a function of pressure ratio.



Note: Work is based on an inlet temperature of 77°F = 537°R. For any other temperature, t, multiply by (t + 460)/537

4496-79

Optimum bundle depth = 12 (Figure 1, Reference 5).
Unit weight = 147 lb/ft^2 FA (face area).
Typical face velocity, FV = 405 ft/min.
Surface area = 15.2 ft^2/ft^2 FA.
Assume $t_2 - t_1 = 50°F$

$$FA = Q/(t_2 - t_1)\,(FV)\,(1.08)$$
$$= 10^6/(50)\,(405)\,(1.08) = 45.5 \text{ ft}^2$$
$$T_2 - t_1 = 97 - 77 = 20$$
$$T_1 - t_2 = 300 - 127 = 173$$
$$\Delta T_{l.m.} = 71.0$$

Surface area = $10^6/(20)\,(71.0) = 704$ ft^2.
FA = 704/15.2 = 46.4 ft^2.
Surface per fan horsepower = 150 (Figure 2, Reference 6).
Power required = 704/150 = 4.7 hp.
4.7 hp × 2544 = 12,000 BTU/hr.
This gives 12.0 BTU/3000 BTU of cooling.

- Power for Purification

Relate power to quantity of air to be purified. For solid adsorbent type purifiers, the following values are typical:

Process air (lb/hr)	Heater power (hp)	Blower power (hp)	Total power (hp)	Power (BTU/lb air)
676	6.0	2.0	8.0	30.2
1347	12.0	3.1	15.1	28.6
2243	20.0	4.5	24.5	27.8
3353	26.7	6.3	33.0	25.1

Use a mean value of 28 BTU/lb process air for comparison of processes.

- Power for Auxiliary Refrigeration

To supplement air cooling, some auxiliary refrigeration is required. Following are typical power requirements for R12 refrigeration units to cool air from 97° to 75°F.

Process air (lb/hr)	Refrigeration unit power (hp)	Power (BTU/lb air)
676	2.5	9.45
1347	3.3	6.25
2243	4.0	4.53
3353	5.0	3.81

For comparing processes, a value of 6 BTU/lb of process air was used for auxiliary refrigeration.

- Heat Leak as a Function of Plant Size

Assume the heat leak equation is of the general form

$$q_A = A + B/\sqrt{F}$$

where

q_A = heat leak, BTU/lb process air
A = minimum heat leak
B = a constant of proportionality
F = process air rate, lb/hr

As air fractionation plants become very large, q_A approaches a minimum of 1, therefore,

$$A = 1$$

For a plant processing air at a rate of 800 lb/hr, q_A is approximately 5 BTU/lb of process air. Substituting these known values into the heat leak equation to obtain the value of B gives,

$$5 = 1 + B/\sqrt{800}$$
$$B = 113$$

Hence,

$$q_A = 1 + 113/\sqrt{F}$$

An example of the use of this equation follows:
For a plant producing nitrogen at a rate of 2500 lb/hr, the required airflow rate (F) is

$$F = 2500 \times 37.4/28.0 = 3340 \text{ lb/hr}$$

The corresponding heat leak then is

$$q_A = 1 + 113/\sqrt{3340}$$
$$= 2.98 \text{ (This is a reasonable value.)}$$

For the oxygen refrigeration circuit, use the same form of the equation

$$q_o = C + D/\sqrt{X}$$

where

q_o = heat leak, BTU/lb oxygen
X = oxygen rate, lb/hr

Assume q_o approaches a minimum of 0.5 with increasing plant size.
Assume $q_o = 2.0$ when X = 1000 lb/hr.
Substituting $2 = 0.5 + D/\sqrt{1000}$, D = 47.5.
Therefore, $q_o = 0.5 + 47.5/\sqrt{X}$.

- Computer Analysis of Low Pressure Column Performance (Problem No. 12)

A series of computer analyses were made for the fractionation columns to determine the effects of the number of stages and operating conditions on the amounts and purities of the products. The computer input data and calculated results for problem 12 which follow illustrate the data used and obtained in the computer analysis of column performance.

THE SERVICE BUREAU CORPORATION
DISTILLATION SERVICE

GAS EQUIPMENT ENGR. CORP.
LOW PRESSURE AIR SEPARATION COLUMN
ESTIMATE NO.525172 JOB NO.1110 DATE11/12/65

UNITS

FLOWRATE - MOLES/HOUR
ENTHALPY - BTU/LB. MOLE
TEMPERATURE - DEGREES FAHRENHEIT
PRESSURE - P.S.I.A.

COMPONENT IDENTIFICATION

COMP. NO.	IDENTIFICATION
1	NITROGEN
2	OXYGEN
3	ARGON

NUMBER OF COMPONENTS = 3

TOTAL NUMBER OF STAGES (INCLUDING THE REBOILER AND CONDENSER) = 26

FEED LOCATIONS

FEED NO. 1 ENTERS ABOVE STAGE 7
FEED NO. 2 ENTERS ABOVE STAGE 25

LOCATION OF PRODUCTS (TOP, BOTTOM AND SIDE STREAMS)

PRODUCT NO.	LOCATION
1	BOTTOM PRODUCT
2	TOP PRODUCT

FEED AMOUNTS ,MOLE FRACTIONS ,AND TEMPERATURES
(FROM LOWEST TO HIGHEST FEED)

FEED NO. 1 TEMPERATURE = -3.07000E 02

COMP. NO.	LIQ. AMT.	LIQ. X	VAP. AMT.	VAP. Y
1	3.304340E-01	6.119148E-01	5.044000E-02	8.406667E-01
2	2.004600E-01	3.712222E-01	9.000000E-03	1.500000E-01
3	9.106000E-03	1.686296E-02	5.600000E-04	9.333333E-03

S.B.C.. DISTILLATION SERVICE PAGE 2
INPUT DATA FOR PROBLEM NO. 12 CASE NO. 1

TOTAL	5.400000E-01	1.000000E 00	6.000000E-02	1.000000E 00

FEED NO. 2 TEMPERATURE = -3.13000E 02

COMP. NO.	LIQ. AMT.	LIQ. X	VAP. AMT.	VAP. Y
1	3.457884E-01	9.970830E-01	5.314700E-02	9.990038E-01
2	6.875200E-04	1.982468E-03	3.200000E-05	6.015038E-04
3	3.240800E-04	9.344867E-04	2.100000E-05	3.947368E-04
TOTAL	3.468000E-01	1.000000E 00	5.320000E-02	10.000000E-01

TABULATED EQUILIBRIUM DATA

TEMPERATURE	-3.130000E 02	-3.110000E 02	-3.080000E 02	-3.050000E 02
COMP. NO.				
1	9.800000E-01	1.080000E 00	1.300000E 00	1.550000E 00
2	3.000000E-01	3.350000E-01	3.950000E-01	4.640000E-01
3	4.250000E-01	4.630000E-01	5.300000E-01	6.150000E-01

TABULATED EQUILIBRIUM DATA CONTINUED

TEMPERATURE	-3.020000E 02	-2.980000E 02	-2.940000E 02	-2.900000E 02
COMP. NO.				
1	1.820000E 00	2.220000E 00	2.670000E 00	3.200000E 00
2	5.360000E-01	6.450000E-01	7.600000E-01	9.000000E-01
3	7.100000E-01	8.600000E-01	1.030000E 00	1.220000E 00

TABULATED ENTHALPY DATA FOR LIQUID

TEMPERATURE	-3.130000E 02	-3.060000E 02	-2.980000E 02	-2.900000E 02
COMP. NO.				
1	-1.378000E 03	-1.290000E 03	-1.185000E 03	-1.062000E 03
2	-2.040000E 03	-1.956000E 03	-1.855000E 03	-1.750000E 03
3	-2.130000E 03	-2.050000E 03	-1.967000E 03	-1.885000E 03

TABULATED ENTHALPY DATA FOR VAPOR

TEMPERATURE	-3.130000E 02	-3.060000E 02	-2.980000E 02	-2.900000E 02
COMP. NO.				
1	9.520000E 02	9.790000E 02	1.008000E 03	1.039000E 03
2	9.880000E 02	1.028000E 03	1.072000E 03	1.120000E 03

```
S.B.C.: DISTILLATION SERVICE                                        PAGE   3
INPUT DATA FOR PROBLEM NO.  12 CASE NO.  1


TEMPERATURE -3.130000E 02  -3.060000E 02  -2.980000E 02  -2.900000E 02

COMP. NO.
     3       7.800000E 02   8.160000E 02   8.580000E 02   9.000000E 02


STARTING ESTIMATES

REBOILER TEMPERATURE =       -290.00000
CONDENSER TEMPERATURE =       -313.00000
REFLUX AMOUNT =          0.
TOP PRODUCT AMOUNT =          0.76400


TOLERANCES

E1 =   0.10000E-04
ERROR LIMIT ON SUMMATION OF MOLE FRACTIONS =   1.00000E-03


THE REBOILER PRESSURE IS   0.23100E 02
THE CONDENSER PRESSURE IS   0.22200E 02
THE PRESSURE DROP PER STAGE UP THE COLUMN IS   0.36000E-01
THE REFERENCE PRESSURE FOR THE EQUILIBRIUM CONSTANT IS    0.23000E 02


ADDITIONAL VARIABLES SET FOR THIS PROBLEM

THE FRACTION LIQUID IN THE BOTTOM PRODUCT =  0.

THE FRACTION LIQUID IN THE TOP PRODUCT =  0.

THE AMOUNTS OF ALL FEEDS ARE FIXED

THE TOP PRODUCT
WAS TO BE HELD AT  0.76400E 00 WITHIN A TOLERANCE OF -0.

THE REFLUX
WAS TO BE HELD AT  0.          WITHIN A TOLERANCE OF -0.
```

S.B.C.. DISTILLATION SERVICE PAGE 4
CALCULATED INPUT DATA FOR PROBLEM NUMBER 12

RESULTS OF FLASH WITH SET VAPOR RATIO
FOR FEED NUMBER 1

ITERATIONS = 3

VAPOR RATIO (V/F) SET = 1.000000E-01

TEMPERATURE OF FLASHED FEED = -307.237

ENTHALPY OF FLASHED LIQUID = -8.451286E 02

ENTHALPY OF FLASHED VAPOR = 5.879169E 01

COMPONENT AMOUNTS OF LIQUID AND VAPOR

COMP. NO.	LIQ. AMT.	LIQ. X	VAP. AMT.	VAP. Y
1	3.306423E-01	6.123006E-01	5.023165E-02	8.371941E-01
2	2.002517E-01	3.708365E-01	9.208294E-03	1.534716E-01
3	9.105942E-03	1.686286E-02	5.600576E-04	9.334292E-03
TOTAL	5.400000E-01	10.000000E-01	6.000000E-02	10.000000E-01

*****THE ABOVE DATA IS THE CORRECT SOLUTION OF THE FLASH*****

NO FLASH WAS ATTEMPTED ON FEED 2 ITYPE =-0

FEED ENTHALPIES

FEED NO.	LIQUID	VAPOR
1	-8.451286E 02	5.879169E 01
2	-4.785892E 02	5.064394E 01
SUM	-1.214282E 03	

TOTAL FEED

COMP. NO.	AMOUNT
1	7.798094E-01
2	2.101795E-01
3	1.001108E-02
SUM	10.000000E-01

S.B.C.C. DISTILLATION SERVICE PAGE 5
OUTPUT DATA FOR PROBLEM NO. 12 CASE NO. 1

ITERATION NUMBER 23

TOTAL PRODUCT AMOUNTS (IN ORDER OF LOWEST TO HIGHEST WITHDRAWN)

PROD. NO.	LIQUID	VAPOR
1	0.	0.23600E-00
2	0.	0.76400E 00

PRODUCT STREAM RECOVERY FRACTIONS

COMP. NO.	BOTTOM PRODUCT		TOP PRODUCT	
	LIQUID	VAPOR	LIQUID	VAPOR
1	0.	0.21698E-01	0.	0.97806E 00
2	0.	0.99835E 00	0.	0.25036E-02
3	0.	0.92370E 00	0.	0.77133E-01

PRODUCT STREAM COMPOSITIONS (MOLE FRACTIONS)

COMP. NO.	BOTTOM PRODUCT		TOP PRODUCT	
	LIQUID	VAPOR	LIQUID	VAPOR
1	0.21425E-01	0.71696E-01	0.99543E 00	0.99830E 00
2	0.94776E 00	0.88912E 00	0.22472E-02	0.68876E-03
3	0.30810E-01	0.39163E-01	0.23205E-02	0.10107E-02

LIQUID PLATE COMPOSITIONS

STAGE	COMP. 1	COMP. 2	COMP. 3
CONDENSER	0.99543E 00	0.22472E-02	0.23205E-02
STAGE 25	0.99531E 00	0.22679E-02	0.24266E-02
STAGE 24	0.99319E 00	0.27137E-02	0.40932E-02
STAGE 23	0.99064E 00	0.34135E-02	0.59490E-02
STAGE 22	0.98748E 00	0.45094E-02	0.80077E-02
STAGE 21	0.98350E 00	0.62181E-02	0.10279E-01
STAGE 20	0.97837E 00	0.88674E-02	0.12764E-01
STAGE 19	0.97161E 00	0.12941E-01	0.15453E-01
STAGE 18	0.96256E 00	0.19131E-01	0.18312E-01
STAGE 17	0.95035E 00	0.28383E-01	0.21271E-01
STAGE 16	0.93390E 00	0.41890E-01	0.24211E-01
STAGE 15	0.91208E 00	0.60978E-01	0.26946E-01
STAGE 14	0.88396E 00	0.86795E-01	0.29228E-01
STAGE 13	0.84901E 00	0.12016E-00	0.30829E-01
STAGE 12	0.80734E 00	0.16111E-00	0.31551E-01
STAGE 11	0.76095E 00	0.20783E-00	0.31214E-01

S.B.C.. DISTILLATION SERVICE PAGE 6
OUTPUT DATA FOR PROBLEM NO. 12 CASE NO. 1

STAGE	COMP. 1	COMP. 2	COMP. 3
STAGE 10	0.71317E 00	0.25702E-00	0.29811E-01
STAGE 9	0.66783E 00	0.30464E-00	0.27526E-01
STAGE 8	0.62832E 00	0.34706E-00	0.24613E-01
STAGE 7	0.59474E 00	0.38337E-00	0.21888E-01
STAGE 6	0.54440E 00	0.42957E-00	0.26028E-01
STAGE 5	0.44982E-00	0.51756E 00	0.32629E-01
STAGE 4	0.31119E-00	0.64897E 00	0.39838E-01
STAGE 3	0.16981E-00	0.78736E 00	0.42837E-01
STAGE 2	0.71696E-01	0.88912E 00	0.39183E-01
REBOILER	0.21425E-01	0.94776E 00	0.30810E-01

STAGE VARIABLES

STAGE	TEMPERATURE	LIQUID FLOW	VAPOR FLOW	HEAT INPUT
CONDENSER	-0.31325E 03	0.	0.76400E 00	-0.14486E-00
STAGE 25	-0.31322E 03	0.34617E-00	0.71080E 00	-0.40320E-05
STAGE 24	-0.31316E 03	0.34601E-00	0.71017E 00	0.22160E-04
STAGE 23	-0.31309E 03	0.34582E-00	0.71001E 00	0.12083E-04
STAGE 22	-0.31302E 03	0.34556E-00	0.70982E 00	0.18117E-04
STAGE 21	-0.31294E 03	0.34523E-00	0.70956E 00	0.30178E-04
STAGE 20	-0.31284E 03	0.34478E-00	0.70923E 00	0.52270E-04
STAGE 19	-0.31272E 03	0.34418E-00	0.70878E 00	0.92384E-04
STAGE 18	-0.31256E 03	0.34334E-00	0.70818E 00	0.12836E-03
STAGE 17	-0.31236E 03	0.34220E-00	0.70734E 00	0.25023E-03
STAGE 16	-0.31209E 03	0.34066E-00	0.70620E 00	0.38936E-03
STAGE 15	-0.31174E 03	0.33861E-00	0.70466E 00	0.65668E-03
STAGE 14	-0.31128E 03	0.33606E-00	0.70261E 00	0.94301E-03
STAGE 13	-0.31077E 03	0.33291E-00	0.70006E 00	0.15427E-02
STAGE 12	-0.31022E 03	0.32911E-00	0.69691E 00	0.20392E-02
STAGE 11	-0.30958E 03	0.32500E-00	0.69311E 00	0.25361E-02
STAGE 10	-0.30889E 03	0.32090E-00	0.68900E 00	0.27988E-02
STAGE 9	-0.30819E 03	0.31714E-00	0.68490E 00	0.27694E-02
STAGE 8	-0.30754E 03	0.31397E-00	0.68114E 00	0.25509E-02
STAGE 7	-0.30696E 03	0.84918E 00	0.61797E 00	0.55575E-02
STAGE 6	-0.30607E 03	0.83933E 00	0.61318E 00	0.70092E-02
STAGE 5	-0.30421E 03	0.82223E 00	0.60333E 00	0.60036E-02
STAGE 4	-0.30080E 03	0.80035E 00	0.58623E 00	0.17984E-02
STAGE 3	-0.29611E 03	0.78314E 00	0.56435E 00	-0.87385E-03
STAGE 2	-0.29170E 03	0.77558E 00	0.54714E 00	-0.82751E-03
REBOILER	-0.28888E 03	0.	0.53958E 00	0.22030E 04

STAGE	M. R. UNBAL	LIQ. MOL. H	VAP. MOL. H
CONDENSER	0.94610E-05	-0.13844E 04	0.95089E 03
STAGE 25	0.45565E-04	-0.13840E 04	0.95101E 03
STAGE 24	0.32306E-04	-0.13849E 04	0.95112E 03
STAGE 23	0.93316E-05	-0.13859E 04	0.95123E 03

S.B.C.. DISTILLATION SERVICE PAGE 7
OUTPUT DATA FOR PROBLEM NO. 12 CASE NO. 1

STAGE	M. R. UNBAL	LIQ. MOL. H	VAP. MOL. H
STAGE 22	0.16241E-04	-0.13873E 04	0.95137E 03
STAGE 21	0.77462E-05	-0.13891E 04	0.95153E 03
STAGE 20	0.21376E-04	-0.13914E 04	0.95176E 03
STAGE 19	0.31255E-04	-0.13946E 04	0.95208E 03
STAGE 18	0.43187E-04	-0.13988E 04	0.95253E 03
STAGE 17	0.56931E-04	-0.14047E 04	0.95319E 03
STAGE 16	0.71581E-04	-0.14125E 04	0.95416E 03
STAGE 15	0.83322E-04	-0.14228E 04	0.95554E 03
STAGE 14	0.67714E-04	-0.14359E 04	0.95747E 03
STAGE 13	0.76680E-04	-0.14529E 04	0.95979E 03
STAGE 12	0.72663E-04	-0.14738E 04	0.96249E 03
STAGE 11	0.57038E-04	-0.14966E 04	0.96580E 03
STAGE 10	0.34118E-04	-0.15196E 04	0.96957E 03
STAGE 9	0.20598E-04	-0.15409E 04	0.97353E 03
STAGE 8	0.20471E-04	-0.15589E 04	0.97737E 03
STAGE 7	0.32512E-05	-0.15738E 04	0.98089E 03
STAGE 6	0.35225E-04	-0.15968E 04	0.98547E 03
STAGE 5	0.68123E-04	-0.16366E 04	0.99523E 03
STAGE 4	0.74672E-04	-0.16865E 04	0.10148E 04
STAGE 3	0.89697E-04	-0.17207E 04	0.10467E 04
STAGE 2	0.83237E-04	-0.17283E 04	0.10837E 04
REBOILER	0.54[illegible]E-07	-0.17248E 04	0.11121E 04

OVERALL COMPONENT UNBALANCE

COMP. NO.	COMP. UNBAL
1	-0.24048E-03
2	0.85261E-03
3	0.83047E-03

THE TOP PRODUCT
WAS TO BE HELD AT 0.76400E 00. WITHIN A SPECIFIED TOLERANCE
OF -0. CALCULATED VALUE WAS 0.76400E 00

THE REFLUX
WAS TO BE HELD AT 0. . WITHIN A SPECIFIED TOLERANCE
OF -0. CALCULATED VALUE WAS 0.

THE AMOUNTS OF ALL FEEDS ARE FIXED

CASE NUMBER 1 OF PROBLEM NUMBER 12 HAS BEEN SOLVED

REFERENCES

1. Weedman, J. P. and Dodge, B. F. "Performance of Column Packings in Air Separation." Ind. Eng. Chem. Vol. 39 (1947) p. 732.

2. Lobo, W. E. and Williams, B. Liquid Air Fractionation. OSRD Report No. 3768, 13 June 1944.

3. Ruhemann, M. The Separation of Gases. Chapter IX, Oxford Press.

4. Wilson, Silverberg, and Zellner. Argon-Oxygen-Nitrogen Three Component System Experimental Vapor-Equilibrium Data. USAF APL TDR 64-64, April 1964.

5. Smith, E. C. "Air-Cooled Heat Exchangers." Chemical Engineering. (17 November 1958).

Hydrogen Production Subsystem

DESCRIPTION

The hydrogen required for the synthesis of ammonia is obtained by decomposing water into hydrogen and oxygen. The subsystem is shown schematically in Figure 3.3-1. The decomposition is performed in a water electrolyzer which consumes electrical power and water. The electrolyzer is composed of several modules which are arranged in banks of three modules each. These modules consist of approximately 110 electrolysis cells and are described in the Design section. The electric power consumed during electrolysis is supplied to the modules by the rectifiers and appropriate electrical busses. In addition, switch gear and cabling are included to feed the auxiliary loads.

The oxygen and hydrogen produced in the modules leave the electrolyzer in separate streams. Each of these exit streams is a mixture of gas and liquid electrolyte (an aqueous solution of potassium hydroxide) forming two-phase mixtures. Each of the two-phase mixtures is separated in the separator-reservoir, and the hydrogen and oxygen gases are withdrawn.

In addition to separating the two liquid-gas streams, the separator-reservoir also accommodates variations in electrolyte level. These level changes occur during start-up and shutdown and are caused by gas generation within the subsystem.

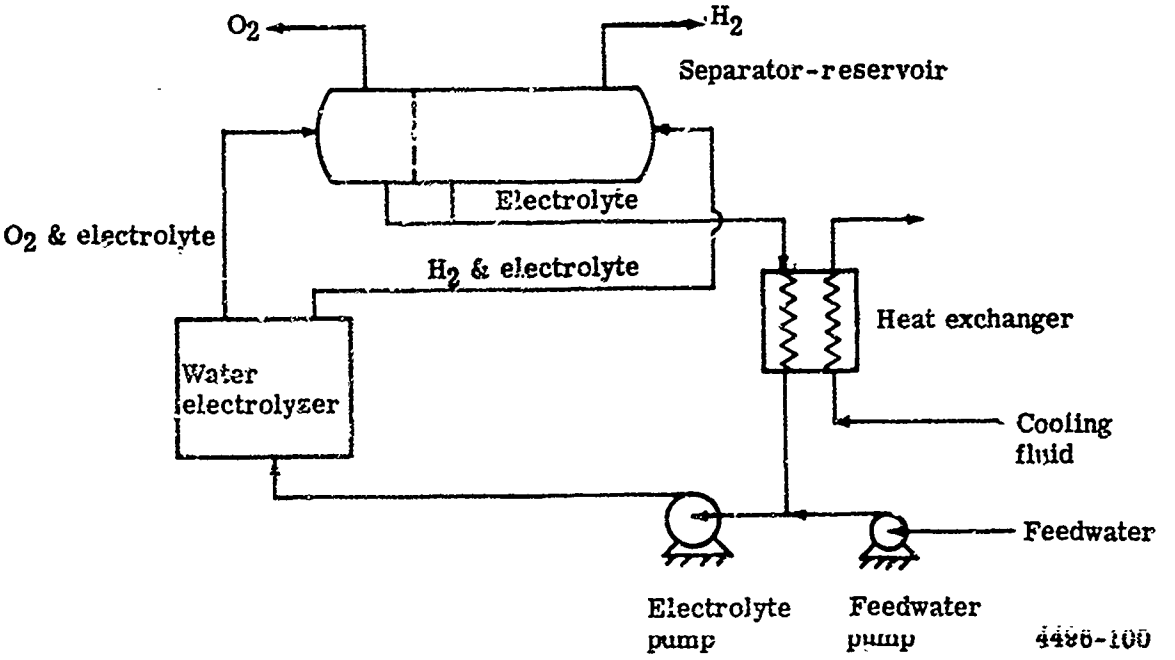


Figure 3.3-1. Simplified diagram of hydrogen production subsystem.

The liquid electrolyte is collected in the separator-reservoir and is then passed through a heat exchanger which is used to control the electrolyte temperature. On emerging from the heat exchanger, feedwater is added to the electrolyte and the mixture is then pumped back to the electrolyzer.

Necessary controls and instrumentation are provided to monitor and regulate the subsystem. The assembly is mounted on a rugged lightweight structure suitable for transporting the entire subsystem.

The hydrogen production subsystem consists of the following major components: electrolyzer modules, heat rejection equipment, pumps and piping, separator-reservoir, electrical distribution system, and skid and structure. In this section of the report, the term "hydrogen production subsystem" refers to this combination of equipment.

INVESTIGATION

The purpose of this investigation was to design 11 hydrogen production subsystems corresponding to the 11 prescribed mobility and power level classifications shown in the Introduction (Figure 2-2) and repeated here for convenience (Figure 3.3-2). In the investigation, subsystem weights and production rates were selected on the basis of a preliminary cost optimization study. A digital computer program was used to prepare subsystem designs which closely approximated the optimum subsystem configurations.

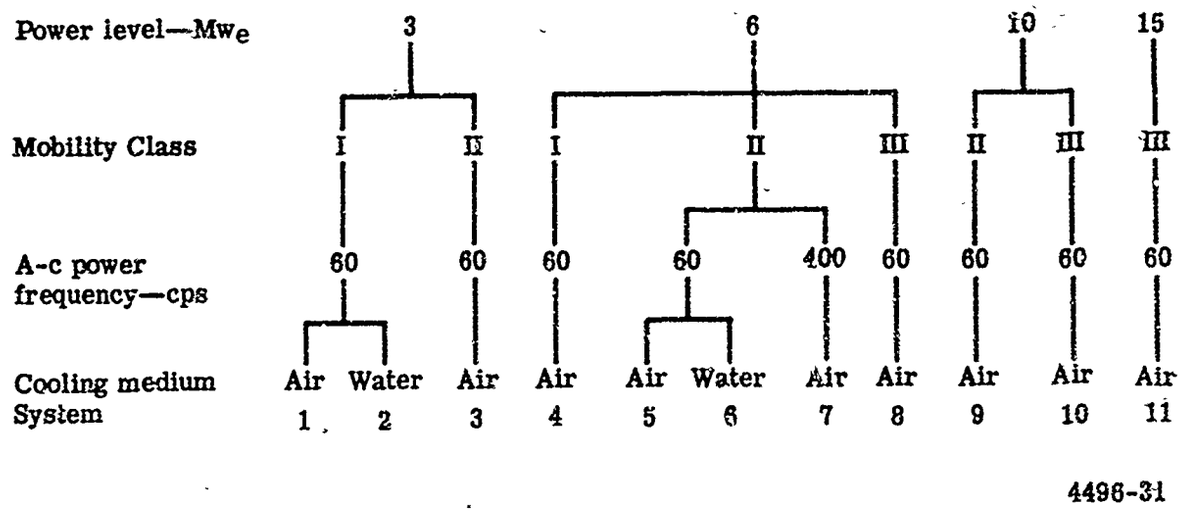


Figure 3.3-2. Ammonia production systems definition.

One of the first steps in the investigation was the analysis of the 11 subsystem designs using the computer program. The program was previously developed by Allison and is discussed in Reference 1. The program utilizes laboratory test data and weight information based on current electrolyzer designs. Briefly, the major input data used by the program are (1) maximum subsystem weight, (2) electrical power, (3) operating temperature and pressure, and (4) weight of subsystem components. Using these input data, the program iterates until the proper combination of system components is found which maximizes the hydrogen production, subject to the input constraints. When the iteration converges, the details of the design are printed as output summarizing the subsystem design.

Of the input data stated in the preceding paragraph, the first three are obtained directly from the contract specifications. (See the subsequent Design Parameter section.) The component weight data were not readily available and had to be obtained from component vendors, electrolyzer design studies, and concurrent packaging studies.

Component Data

The hydrogen production subsystem was considered in terms of the following major components:

- Electrolyzer modules
- Heat rejection equipment
- Pumps and piping
- Separator-reservoir
- Electrical distribution equipment
- Skid and structure
- Controls and instrumentation

Weight and performance data for each of these components were also determined.

Electrolyzer Modules

In the following analysis, the weight of each module was considered in two portions—fixed weight and variable weight. The fixed weight represents those parts of the module whose weight does not vary as the number of cells is changed. For example, two end plates are required whether there are 100 or 120 cells in each module. Examples of variable weight parts are the cells and the required tie bolts.

Commensurate with this distinction between fixed and variable weight, Table 3.3-I was used for electrolyzer modules.

Table 3.3-I shows that a module composed of 113 cells weighs approximately 1062 lb when filled with electrolyte. Of this total, 154.7 lb is independent of the number of cells. The

remaining 907.3 lb varies as the number of cells varies. The distinction between fixed and variable weight is necessary because the number of cells per module is not constant for all of the 11 designs. In the design of the 11 subsystems, the input power level and frequency, allowable plant weight, and cooling medium are varied. Variations in these parameters result in differing module and auxiliary equipment weights. In general, these weight changes are not linear functions. Hence, for the 11 designs, weight trade-offs between the modules and the auxiliary equipment are necessary. As these trade-offs are made, variations in electrolyzer current density and cell voltage drop occur. Hence, to secure a voltage drop consistent with the output of the power conditioning equipment, the number of cells per module must also be varied.

Table 3.3-I.
Electrolyzer module weight tabulation.*
(113 cell module)

Item (No. required per module)	Fixed wet weight** (lb)	Variable wet weight** (lb)
Separator frame (112)		490.4
Seal assembly (112)		9.5
Manifold plate (1)	19.9	
Manifold plate assembly (1)	0.2	
Electrode, end (anode) (1)	12.0	
Electrode, end (cathode) (1)	11.8	
Electrode, bipolar (111)		352.5
Membrane separator (112)		27.9
Spool, insulating (48)		5.5
Tie bolt (48)		19.9
Washer, insulating (48)	0.1	
End plate (anode) (1)	54.6	
End plate (cathode) (1)	54.2	
Eyelet, electrode reinforcement (5328)		1.6
Washer, flat (96)	1.0	
Nut 1/4-28 (48)	0.6	
	154.7	907.3
Total wet weight	1062.0	

*The items listed are discussed in the following Design section and are illustrated in Reference 1.

**Wet weight of 1062.0 lb includes 409.8 lb electrolyte inventory of which 401.7 lb is included in the 490.4 lb separator frame weight.

Heat Rejection Equipment

For the hydrogen production subsystem designs, the heat exchangers were sized with suitable values of electrolyte velocity and pressure drop, coolant velocity and pressure drop, and fan power. The weight of these designs were then expressed as a function of heat load as shown in Figure 3.3-3. These data were obtained from vendors and are representative of stainless steel units capable of operating at pressures of 300 psia. The final heat exchanger designs are discussed in the Design section.

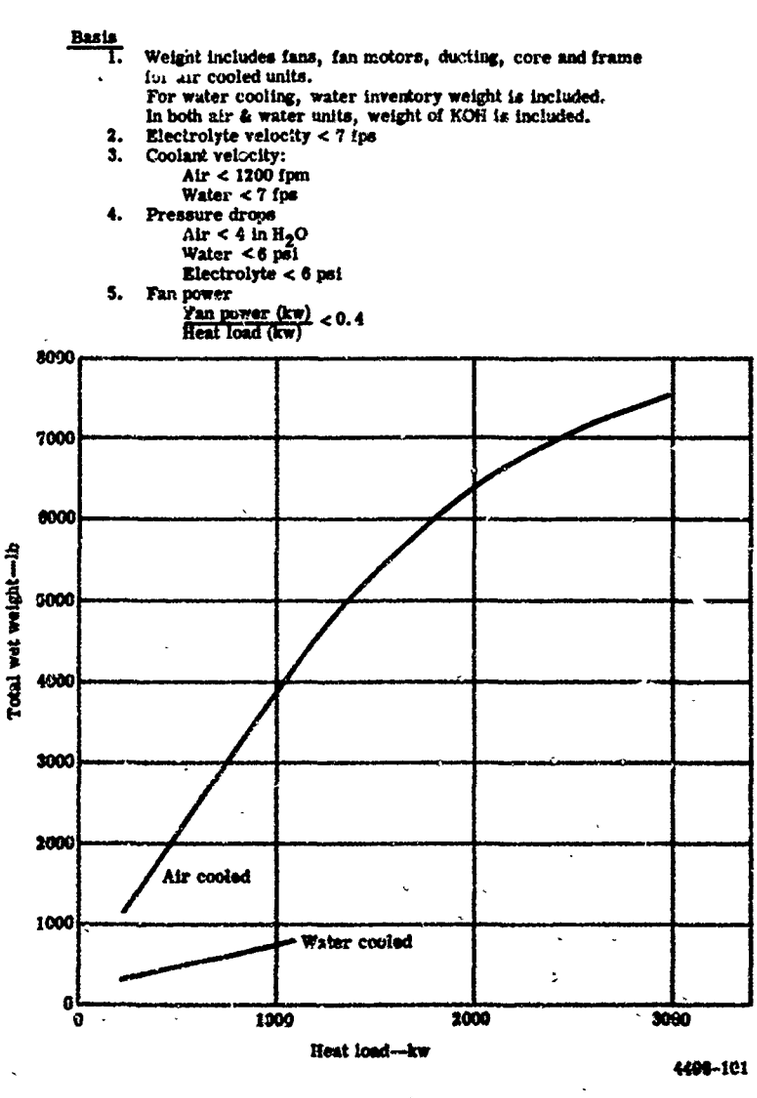


Figure 3.3-3. Heat exchanger weight versus heat load.

Electrical Distribution Equipment

The electrical distribution equipment for the hydrogen production subsystem consists of the following items:

- Rectifier—A rectifier is used to convert the 460-v a-c power available at the transformer secondary to approximately 590 v d-c power suitable for electrolyzer modules.
- Shielding and d-c busses—Bus bars are used to conduct the d-c power from the rectifier to the modules. Where required to protect personnel, these busses are guarded by plastic shielding.
- Motor controllers—Controllers are used to start and stop individual pump and fan motors.
- A-C cables—Suitable cables are used to connect the fan and pump motors to the controllers and subsystem power supply.

Rectifier

Rectifier weight is shown in Figure 3.3-4 as function of the number of module banks and the power level. The data shown are for rectifiers capable of operating at full power with 25% of the module banks shutdown. The discontinuities in the data represent changes in the rectifier designs. As the number of banks is decreased for a given power level, a point is reached where the number of diodes must be increased to accommodate the greater currents imposed on the remaining banks. The point where this transition occurs is illustrated in Figure 3.3-4 by the dotted lines. Further information on the rectifier is presented in the Power Conditioning section.

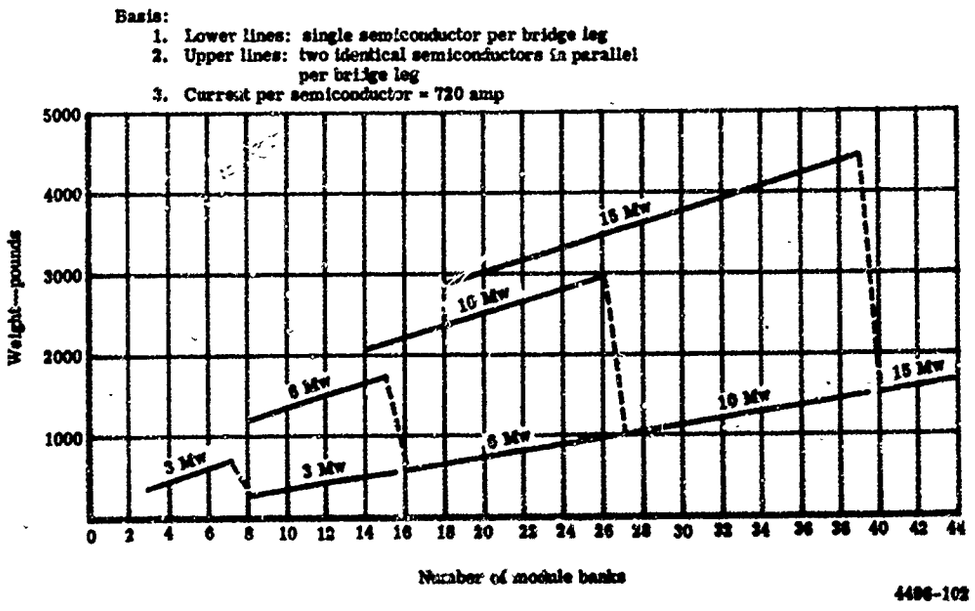


Figure 3.3-4. D-C rectifier weight versus number of module banks for various power levels.

D-C Busses and Shielding

The weight of the bus bars depends on the number of modules, the current, and the position of the modules on the skid. The last consideration requires the selection of a suitable packaging layout. Bus bar weights used in the final analysis were computed after the layout studies were completed. Similarly, shielding weight estimates were developed from the final layout drawings.

Motor Controllers

The size and weight of the motor controllers depends on the characteristics of the various major components. For example, the capacity of the fan motors must await definition of the radiator heat load. As the 11 subsystem designs became defined, vendor data were used to determine the controller weights.

A-C Cables

A-C cable weight estimates were determined in the same manner as the bus bars and motor controllers.

Pumps and Piping

The hydraulic system requires two types of pumps—a large electrolyte circulation pump and a small feedwater pump. The entire hydraulic system is interconnected with piping, and flow within the system is regulated by appropriate valves.

Electrolyte Pump

The variation in pump weight with electrolyte flow rate is presented in Figure 3.3-5. Values used as parametric computer input data are indicated by solid lines. At low flow rates, relatively lightweight compact pumps are currently available and were used in the designs. At higher flow rates, i.e., above 1500 gpm, compact lightweight pumps are not currently available commercially so data for heavier pumps were used. In general, no penalty was paid for using the heavier pumps since they were used on the larger subsystem designs which, in general, are not weight sensitive. The dotted line in Figure 3.3-5 indicates the range of lightweight pumps currently being developed. These pumps could be utilized when they become available.

Feedwater Pump

Weight data for positive displacement feedwater pumps are tabulated in Table 3.3-II.

Basis:

1. Electric drive motor included in weight
2. Drive motors are totally enclosed fan cooled, 460 v, 3 ph
3. Bedplate weight not included
4. Type of pump: centrifugal
5. Direct drive

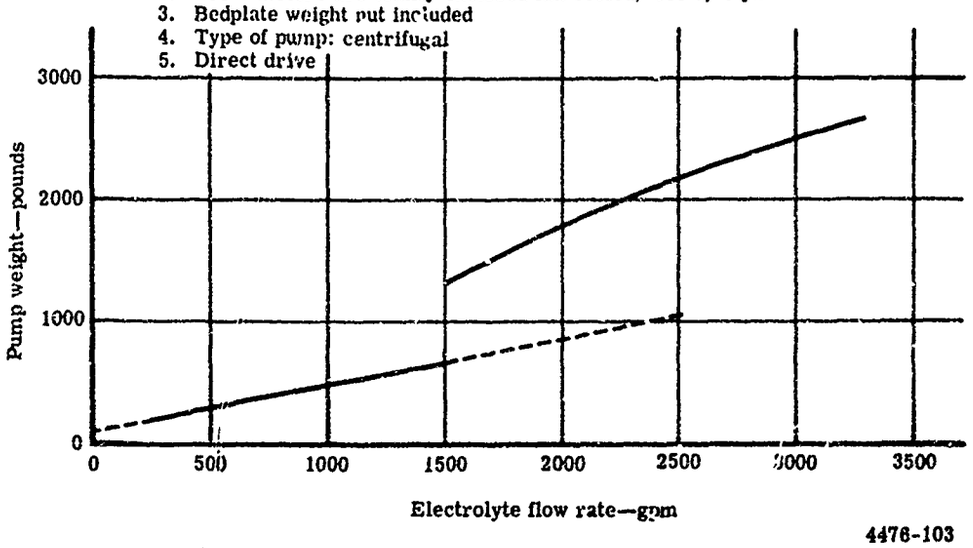


Figure 3.3-5. Pump weight versus electrolyte flow rate.

Table 3.3-II.
Feedwater pump data.

Flow rate (gpm)	Pump and motor weight (lb)
2	50
4	60
6.5	60
10	75

Piping and Valves

Piping and valve sizes and weights were computed from plant layout drawings.

Separator-Reservoir

The weight of the separator-reservoir is a function of the following variables: (1) electrolyte flow rate, (2) gas to liquid volume ratio in the module effluent streams, and (3) electrolyte inventory.

These parameters varied as the analysis was conducted. Final separator weights and configurations were based on plant layout drawings and the three parameters. Final results are tabulated in the Design section.

Skid and Structure

The weight of the skid and structure was taken as a function of the equipment location and weight. Final weight estimates were computed from the plant layout drawings.

Controls and Instrumentation

Controls and instrumentation includes such components as an electrolyte flow indicator, pneumatic valve actuators, differential pressure and level controls, etc. (See Control Subsystem—subsection 3.6—of this report for further information.) The number of these various components depended on the final subsystem design and control system. Hence, final control and instrumentation weights were determined after the layout drawings and control system were completed.

Subsystem Data

After the basic component data had been compiled, it was possible to utilize the computer program discussed previously to determine hydrogen production subsystem weight and production rate.

Parametric data, generated by the computer program, are shown in Figure 3.3-6 for air-cooled subsystems and in Figure 3.3-7 for water-cooled subsystems over the range of interest for this study.

In general, for a specified power input, an increase in the number of module banks (a bank is three modules connected in series electrically and hydraulically) increases subsystem efficiency and, therefore, increases the design hydrogen production rate. This effect is most pronounced in the subsystem designs with large power inputs. Similarly, with a specified number of module banks, decreasing the power input to the system (by increasing the number of cells per module) improves subsystem efficiency and, therefore, hydrogen production rate does not decrease as rapidly as power input. It is interesting to note that the decrease in weight of the heat exchanger and other auxiliaries, as power input is decreased and subsystem efficiency is increased, tends to offset the increase in module bank weight. This effect is more pronounced in the air-cooled subsystem (Figure 3.3-6) where heat exchanger weight is a more significant part of subsystem weight.

The data in Figures 3.3-6 and 3.3-7 are based on hydrogen production system designs which increase subsystem availability by permitting operation at full electrical power with up to 25% of the module banks out-of-service. Therefore, with 619 v dc (the design maximum voltage capability of the power conditioning subsystem) supplied to the electrolyzer banks and one-fourth of the banks out-of-service, the electrical power input will be the same as indicated in Figures 3.3-6 and 3.3-7. However, efficiency will be lower and output will be less than indicated. With

all electrolyzer banks in service, the d-c voltage across the banks will be approximately 590 v for the power input and hydrogen production rate indicated.

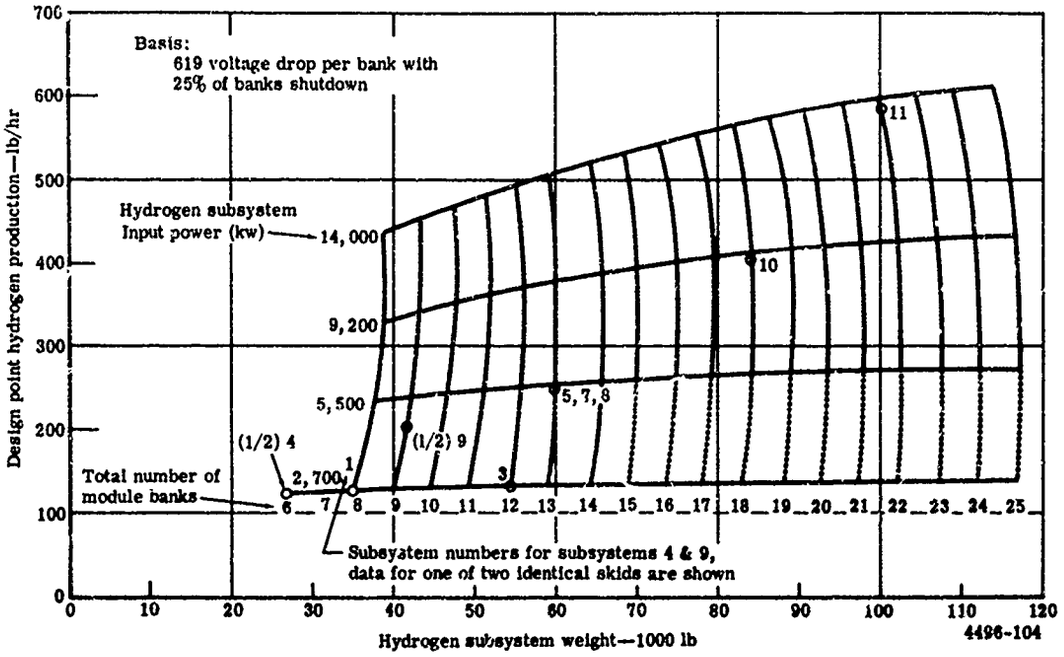


Figure 3.3-6. Production rate versus hydrogen subsystem weight for air-cooled subsystems.

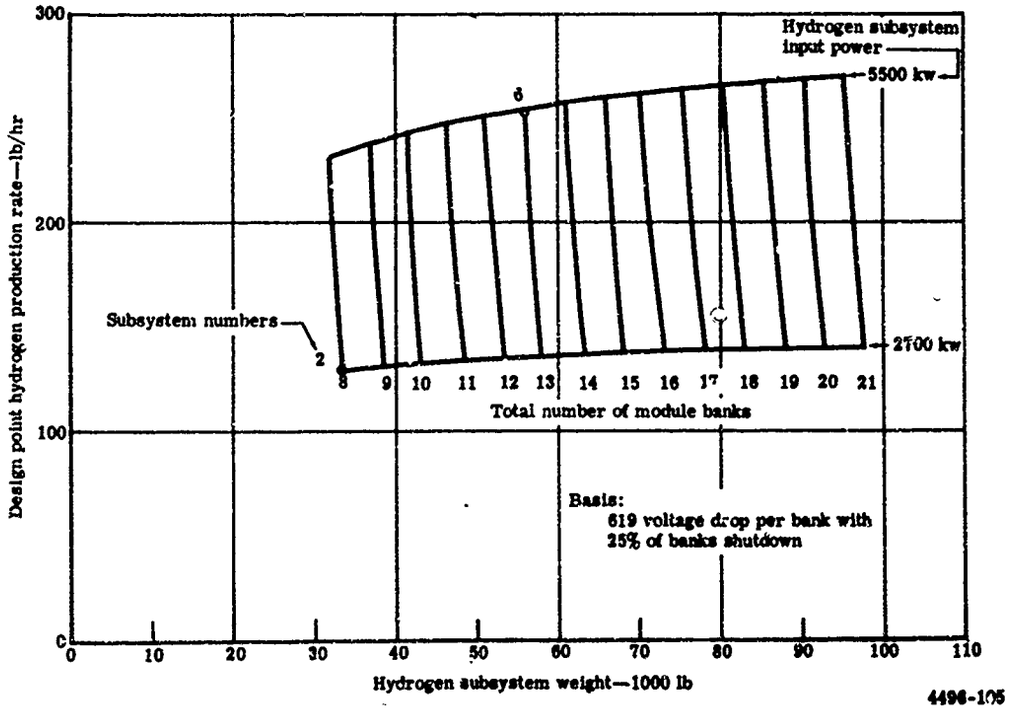


Figure 3.3-7. Production rate versus hydrogen subsystem weight for water-cooled subsystems.

The design points for the 11 subsystem designs, which are developed in following sections of this report, are indicated on Figures 3.3-6 and 3.3-7 for future reference.

Subsystem Analysis

After the basic component and subsystem data had been gathered, it was then possible to begin developing the 11 subsystem designs. These designs were evolved in the following steps.

- Optimum Electrolyzer Weight—A preliminary cost optimization was conducted to obtain estimates of the optimum electrolyzer weight for each of the 11 system designs.
- Selection of Number of Module Banks—After the optimum electrolyzer weight had been estimated, the problem became one of selecting a suitable number of module banks—three modules connected in series hydraulically and electrically. The selection was made by establishing the voltage drop per bank. It was then possible, using an iterative computer calculation involving the number of banks and cells per module, to compute the required number of banks and the number of cells per module.
- Final Design Specification—When the number of module banks was determined, a final computer run was made to define the design point performance.

Optimum Electrolyzer Weight

A cost optimization study based on preliminary cost estimates was conducted to determine the optimum weight and cost of the electrolyzer for each power level. In the analysis, the electrolyzer current density was varied with the power level fixed. The variations in current density were produced by altering the module weight by varying the number of cells which in turn affected the electrode area. As the current density and module weight were changed, the electrolyzer efficiency also changed. As the module weight (and cost) was increased, the efficiency was increased and more hydrogen and ammonia were produced. As the ammonia output was increased through the addition of more modules, the cost of the system also increased. The optimum electrolyzer weight was the weight that resulted in the minimum specific fuel cost. The costs also include nuclear powerplant and ammonia fuel production system costs. The optimum electrolyzer weights computed from the preliminary cost analysis are shown in Figure 3.3-8 as a function of input power. The validity of the resulting design point was checked using final cost estimate data (see Cost Analysis subsection 3.10).

Selection of Number of Module Banks

The procedure used to select the number of module banks for each of the 11 hydrogen production subsystem designs is illustrated in Figure 3.3-9 for subsystem design 10—a 10-Mw_e, Mobility Class III, air-cooled system. The procedure illustrated is typical of the procedure used in other designs. The data in Figure 3.3-9 were obtained using the previously discussed computer

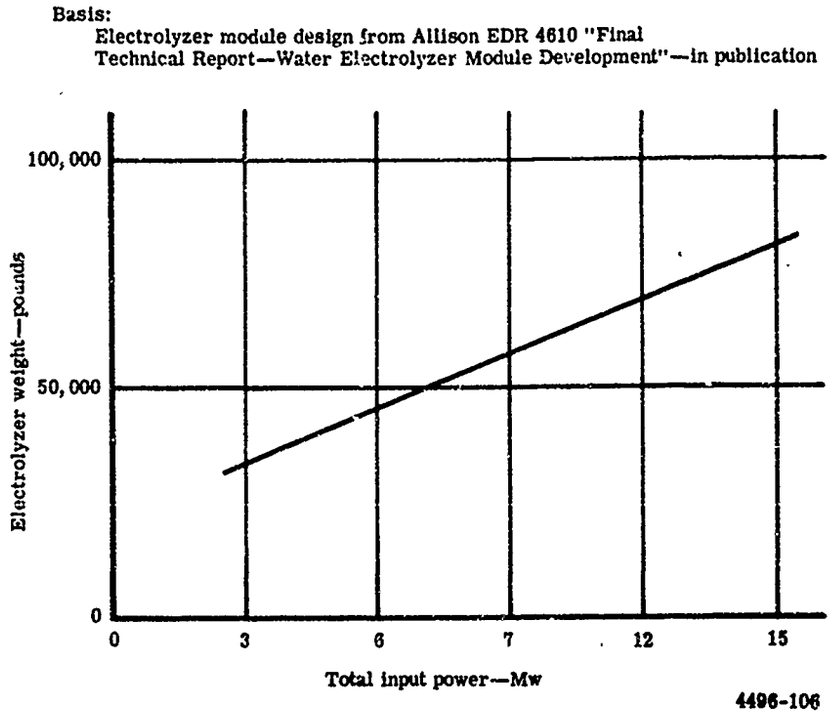


Figure 3.3-8. Optimum electrolyzer weight versus total input power.

program. The optimum electrolyzer weight for 10-Mw_e systems (see Figure 3.3-8) is 60,500 lb. By entering Figure 3.3-9 with a module weight of 60,500 lb, the hydrogen production subsystem weight is within the range of 80,000 to 100,000 lb for 8 to 30 module banks, respectively. Moving to the right on Figure 3.3-9, the optimum number of module banks was determined by fixing the maximum voltage* per bank at 619 v. For hydrogen production subsystem 10, the optimum number of module banks is approximately 20.

At this point in the analysis, it is necessary to consider possible packaging arrangements. The modules will be arranged in banks of three modules (see Design subsection). Further, to utilize the full height of the skid and at the same time limit the required floor area, the banks will be stacked in columns of three. This arrangement results in bank configurations that are multiples of three, e.g., ... 15, 18, 21, 24, etc, banks. Since there are three modules per bank, the corresponding number of modules would be 45, 54, 63, and 72, respectively. (These arrangements are illustrated in the subsequent Design subsection.) Since approximately 20 banks are optimum, it appears that 18 banks would result in the simplest packaging arrangement. The penalty in reduced hydrogen output for 18 banks opposed to 20 was computed. The output

*The maximum voltage per bank is defined as the d-c voltage which, when applied to the electrolyzer with 25% of the banks out of service, results in full design power to the system at design ambient design conditions.

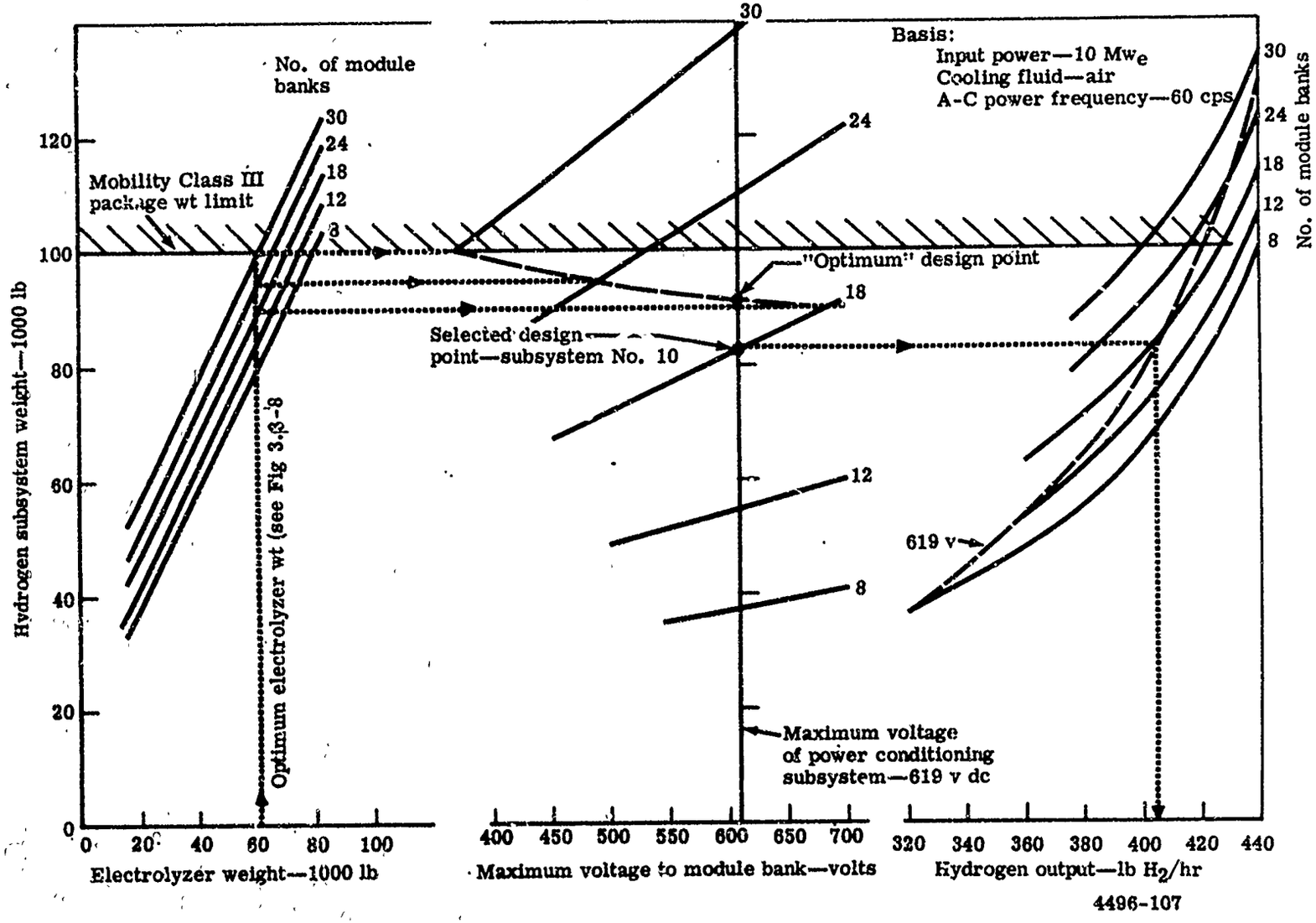


Figure 3.3-9. Hydrogen production subsystem procedure for determination of the number of electrolyzer modules.

for 18 banks is 405.6 lb H_2/hr. The output for 20 banks was found to be 413 lb H_2/hr. By using 20 banks instead of 18, the output would be increased only approximately 1.8% while the weight would increase 11%. Since the potential gain is small, 18 banks were selected.

It is also worthwhile to note the implicit trade-offs that occur during the previous iterative computation. The right side of Figure 3.3-9 shows production rate versus subsystem weight for various numbers of module banks. For a given number of module banks, it can be seen that the production rate drops rapidly as the subsystem weight is reduced. This degradation in performance is caused by a reduction in process efficiency. The situation is aggravated by a necessary weight increase of the auxiliary equipment. Assume that a given subsystem is to have its weight reduced by 20% while the input power is fixed. As a first step, assume that the weight of the electrolyzer modules is reduced by 20%. Since the module weight is decreased, the electrode area is reduced while the current density is increased. Accordingly, the electrolyzer efficiency and production rate are reduced and the heat load increased above its normal design point value. Since the heat load is increased, the size and weight of the heat exchanger must be increased. The weight of the modules and other auxiliary equipment must be further reduced beyond 20% to offset the increased size of the heat exchanger. Hence, the degradation in performance is compounded by the increased weight of the heat exchanger.

Final Design Specification

Once the number of module banks had been determined, a final computer run was made to verify the iterative calculations and establish hydrogen production rate, heat load, electrolyte flow rate, etc. A typical computer run is presented in Appendix A. After each of the 11 subsystems had been sized, packaging layouts were completed and are discussed in the Design subsection.

DESIGN PARAMETERS

The hydrogen subsystem analysis was conducted subject to the following design parameters:

- Design and off-design point (see Table 2-I in Introduction)
- Water requirements
  - For those plants that are to be water-cooled, the water available will be apportioned to the subsystems according to the relative size of the various individual plant heat loads. The total flow is not to exceed 12,000 and 24,000 gph for the 3- and 6-Mw_e plants, respectively.
  - Feedwater requirements will be approximately 125 gph for the 3-Mw_e plant and are not expected to exceed 700 gph for the largest plant.
- Limiting package dimensions and weight (see Table 2-II in Introduction)

- Module operating conditions
  - Pressure = 300 psig
  - Electrolyte temperature = 167°F
- Module weight and dimension data (see Reference 2)
- Module performance data (see Reference 1)
- The maximum voltage drop per module bank is 619 v

DESIGN

Each of the hydrogen productions subsystems is designed to be mounted on a skid with each piece of equipment supported by suitable superstructure. The entire plant is enclosed in an aluminum cover to protect equipment from weather and to help maintain subsystem temperature. Entrance and exit openings are provided for personnel, electrical, and hydraulic connections and air flow for the air-cooled systems. In general, the hydrogen subsystem is designed to be mounted on one skid. Two exceptions are subsystems 4 and 9. In these cases, two packages are required for each plant. In both cases, however, the two packages have the same dimensions, consume the same amount of power, and contain the same equipment.

Figure 3.3-10 is a process flow diagram representing the 11 hydrogen production subsystems. All of the hydrogen production subsystems share the same basic equipment. For this reason, the subsystems may be considered in terms of the following components:

- Electrolyzer modules
- Heat exchanger
- Liquid-gas separator-reservoir
- Pumps
  - Electrolyte
  - Feedwater
- Electrical distribution system
- Piping
- Structure

The components are discussed in the following paragraphs; the number and type required for each subsystem are included. The skid location and orientation of each component are indicated in Figures 3.3-11 through 3.3-21.

Electrolyzer Modules

The basic electrolyzer module design and configuration used in this analysis conforms to the module design developed on a previous contract.[2] Minor variations (about 10%) in the number of cells per module were allowed to obtain suitable voltage drops for each of the 11 subsystem designs.

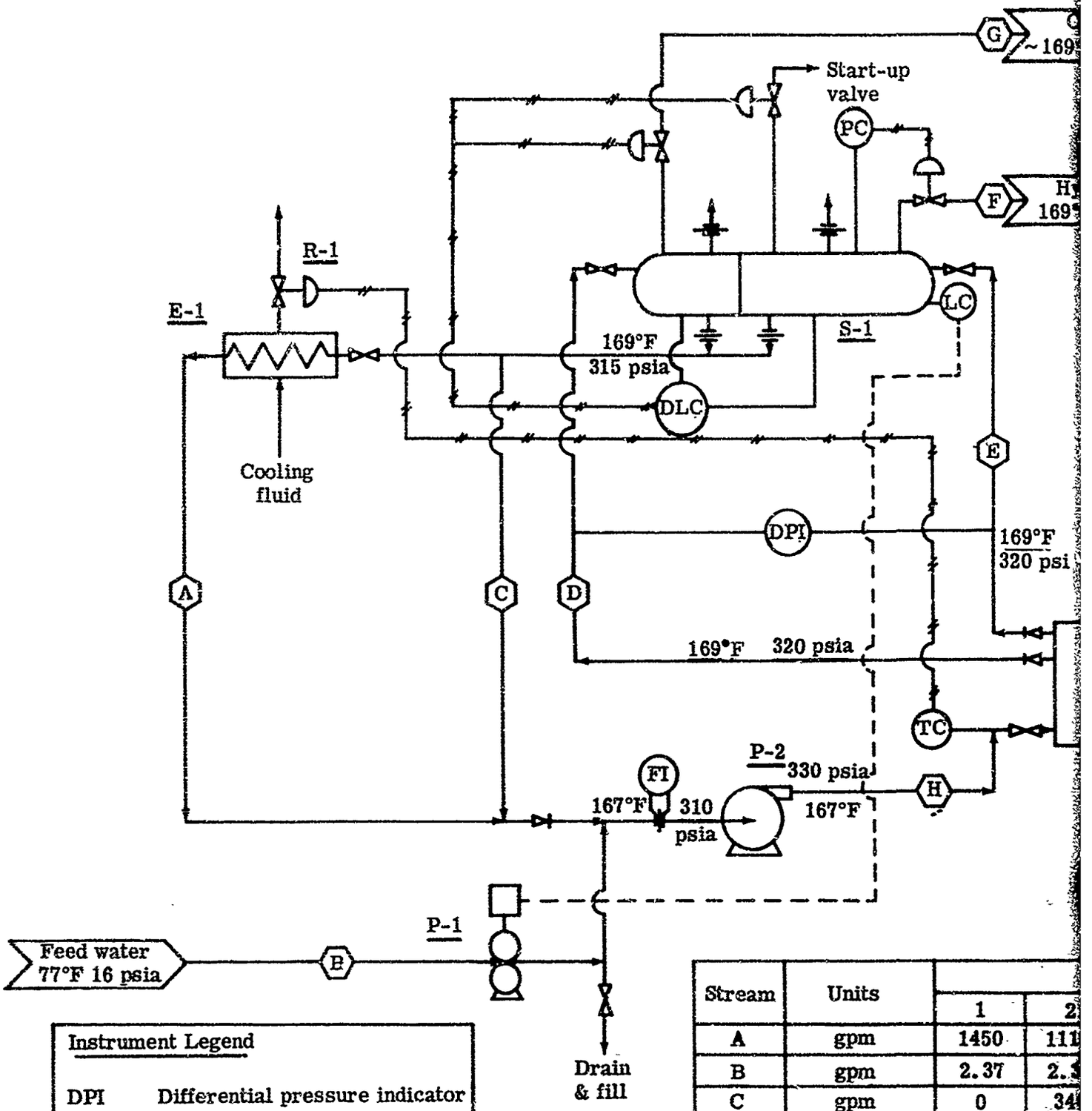
The exterior of a typical module is shown in Figure 3.3-22. The electrolyte inlet port can be seen at the bottom of the end plate. The outlet parts through which the two effluent streams (hydrogen gas mixed with liquid electrolyte and oxygen gas mixed with electrolyte) can be seen near the top of the end plate. The electrodes and separator frames are stacked between the end plates, and the entire assembly is held together by tie bolts.

Figure 3.3-23 illustrates the flow distribution within an electrolyzer module. The incoming electrolyte is distributed into the anodic and cathodic chambers where the process of electrolysis occurs. The electrolyte, mixed with gas, then flows into the passages at the top of the cell and is carried out of the module. Internal manifold passages are formed when individual separator-frame assemblies and electrodes are stacked. Each separator-frame assembly has a series of openings at the top and bottom. There are twelve passages at the bottom of the module which serve to distribute electrolyte throughout the module. Of the 12 passages at the top of the module, four collect the oxygen-loaded electrolyte and eight collect the hydrogen-loaded electrolyte. These various passages are collected together by a manifolding plate sandwiched between the end electrode and the end plate of the module.

In all of the designs developed in this analysis, the modules are connected in banks of three. Early in the study, other methods for arranging the modules were considered[3] (see Figure 3.3-24). At that time, it was decided that the three modules in a bank should be hydraulically connected in a modified series-parallel arrangement rather than in parallel as had previously been the case.[1] This change in arrangement eliminated the need for intermodule flow interrupters and insulated pipe. Only a small increase in pressure drop is expected with modules connected in series.

A third module configuration is shown in Figure 3.3-24(c). This design appears to offer potential reductions in system weight, complexity, and pressure drop. At the same time, however, undefined problems (frame creepage and complicated assembly) have yet to be investigated. Since the design appears to offer potential advantages, it warrants additional study in the future. For the present analysis, however, the module arrangement shown in Figure 3.3-24(b) was employed because of its simplification over the previous parallel arrangement and known reliability.

In Table 3.3-III, electrolyzer module data are tabulated for each of the 11 systems. Two different types of packaging arrangements were used to layout the module banks. In the first arrangement, the three modules in a bank are positioned along the length of the skid. This arrangement is suitable for small numbers of module banks and is illustrated in Figure 3.3-14. From Table 3.3-III, it can be seen that this design, subsystem 4, has only 6 module banks. Because the width of the skid is limited to eight feet, large numbers of banks could not be accommodated. If the banks had been positioned behind one another, access for maintenance would be difficult. In the present arrangement, a module bank can be disconnected and removed from the skid by opening an access panel at the right of the skid and sliding the modules out.

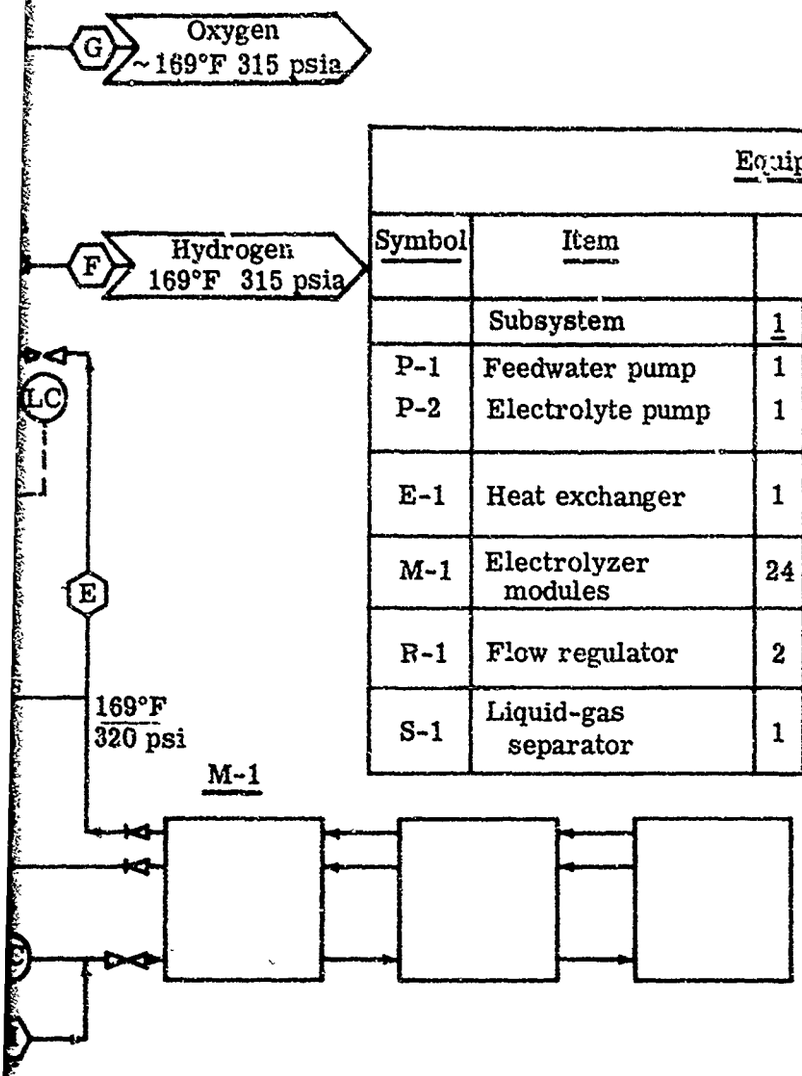


Instrument Legend

DPI	Differential pressure indicator
DLC	Differential level control
FI	Flow indicator
LC	Level control
PC	Pressure control
TC	Temperature control

Stream	Units	1	2
A	gpm	1450	111
B	gpm	2.37	2.3
C	gpm	0	34
D	gpm	569	56
E	gpm	1138	113
F	lb/hr	129.1	129
G	lb/hr	1032	103
H	gpm	1450	145

*Two identical plants required



Equipment List

Symbol	Item	Number of units required										
	Subsystem	1	2	3	4	5	6	7	8	9	10	11
P-1	Feedwater pump	1	1	1	1	2	2	2	2	2	2	3
P-2	Electrolyte pump	1	1	1	1	1	1	1	1	1	2	2
E-1	Heat exchanger	1	1	1	1	2	1	2	2	2	2	2
M-1	Electrolyzer modules	24	24	36	18	39	39	39	39	27	54	66
R-1	Flow regulator	2	1	2	2	4	1	4	4	4	6	6
S-1	Liquid-gas separator	1	1	1	1	3	2	3	3	2	3	3

Plant design number										
1	2	3	4*	5	6	7	8	9*	10	11
1450	1110	1240	1400	2380	400	2380	2380	1890	3630	3266
2.37	2.37	2.46	2.24	4.59	4.63	4.59	4.59	3.73	7.45	10.7
0	340	0	0	0	1940	0	0	0	0	1633
569	569	503	549	959	947	959	959	763	1475	2020
1138	1138	1006	1098	1918	1894	1918	1918	1526	2950	4040
129.1	129.3	134.2	122.0	250.1	252.1	250.1	250.1	203	405.6	585
1032	1032	1073	975	2006	2013	2006	2006	1623	3241	4675
1450	1450	1240	1400	2380	2340	2380	2380	1890	3630	4900

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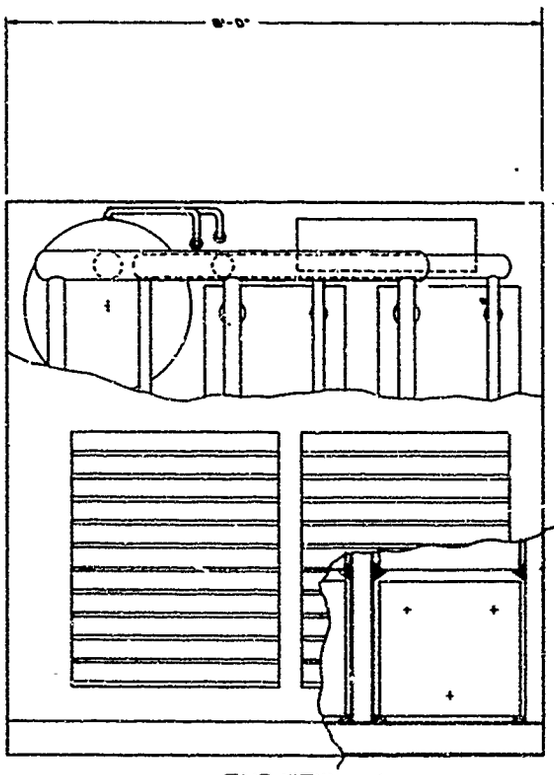
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Figure 3.3-10. Process flow diagram for hydrogen production subsystems.

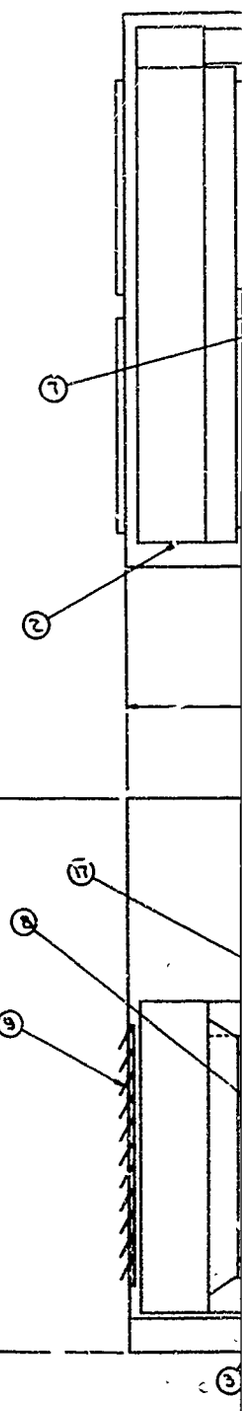
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LEGEND

1. ELECTRODE-TYPE BOILER (6 UNITS OF 3 - 1.5 TON)
2. RADIATOR
3. ELECTROLYTE PUMP AND MOTOR
4. SEPARATOR
5. LIQUID GAS SEPARATOR
6. RECTIFIER
7. FEEDWATER PUMP AND MOTOR
8. FAN MOTOR (2-REQ'D)
9. RADIATOR AIR LOUVERS
10. ELECTROLYTE SHUT-OFF VALVE (2-REQ'D)
11. OXYGEN OUTLET CHECK VALVE (2-REQ'D)
12. HYDROGEN OUTLET CHECK VALVE (2-REQ'D)
13. OXYGEN SHUT-OFF VALVE
14. HYDROGEN SHUT-OFF VALVE
15. PNEUMATICALLY OPERATED OXYGEN EXHAUST VALVE
16. PNEUMATICALLY OPERATED HYDROGEN EXHAUST VALVE
17. CHECK VALVE



END VIEW



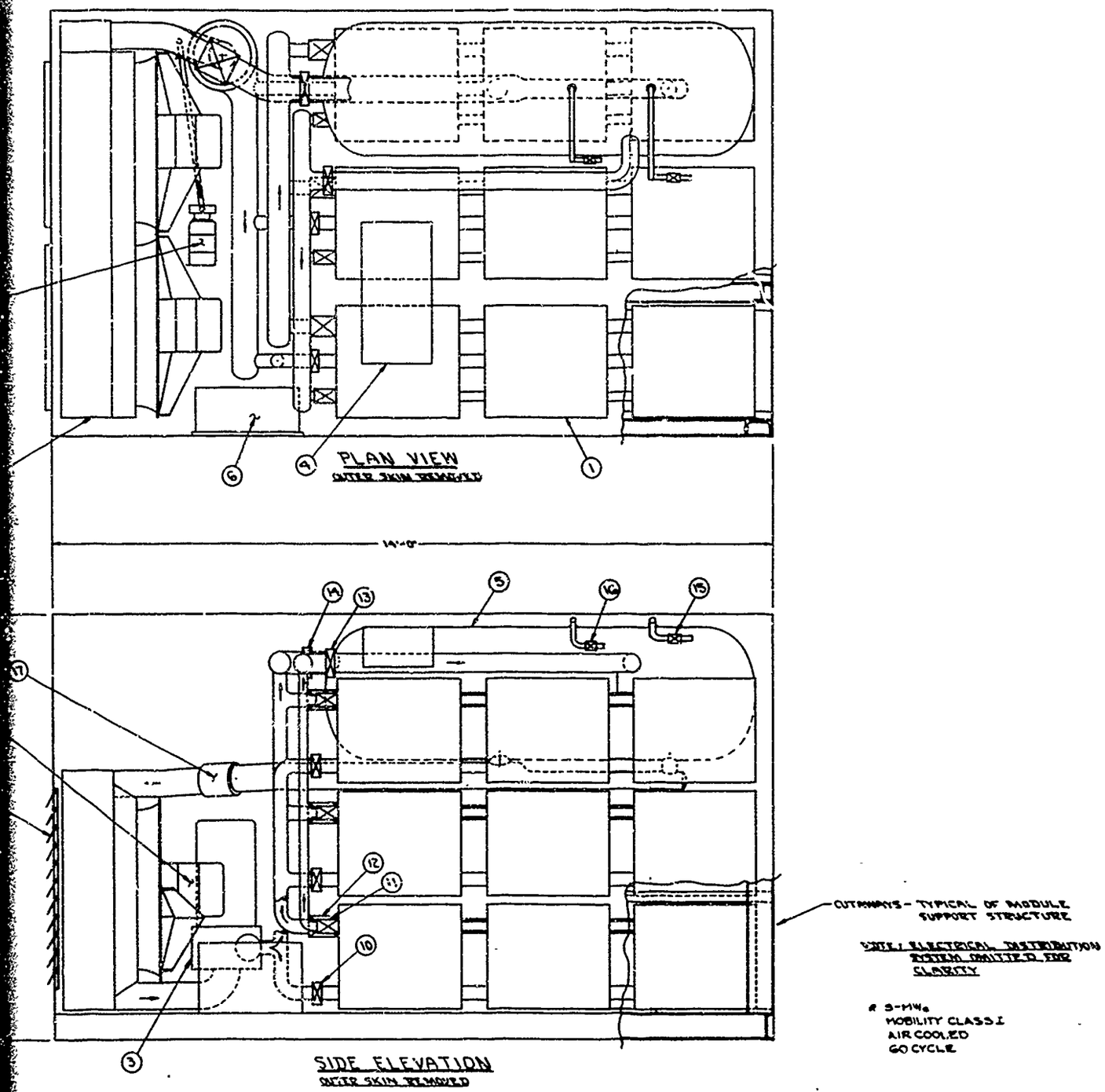
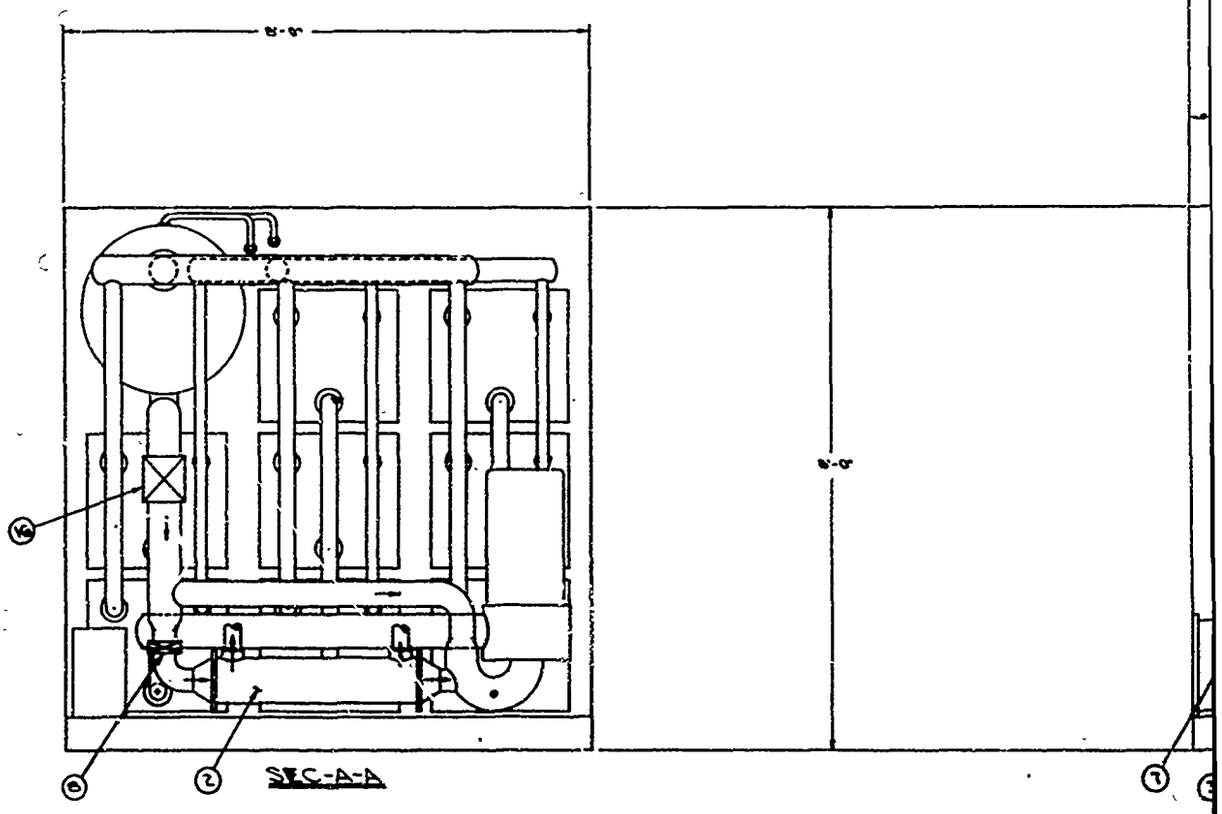


Figure 3.3-11. Hydrogen Subsystem No. I layout.

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LEGEND

1. ELECTROLYZER MODULE
   (SEE DRAWING NO. X-XX-XXXX)
2. HEAT EXCHANGER (TUBULAR TYPE)
3. ELECTRONIC POINT BOARD ASSEMBLY
4. TWITCH-GEARS
5. LIQUID GAS SEPARATOR
6. RECEIVER
7. FEED-WATER PUMP AND MOTOR
8. BY-PASS VALVE
9. ELECTRICALLY SHUT-OFF VALVE
   (SEE DWG NO. XXXXXX)
10. OXYGEN P[illegible] VALVE
    (SEE DWG NO. XXXXXX)
11. HYDROGEN P[illegible] CHECK VALVE
    (SEE DWG NO. XXXXXX)
12. OXYGEN SHUT-OFF VALVE
13. HYDROGEN SHUT-OFF VALVE
14. PNEUMATICALLY OPERATED
    OXYGEN VENTING VALVE
15. PNEUMATICALLY OPERATED
    HYDROGEN VENTING VALVE
16. CHECK VALVE



SEC-A-A

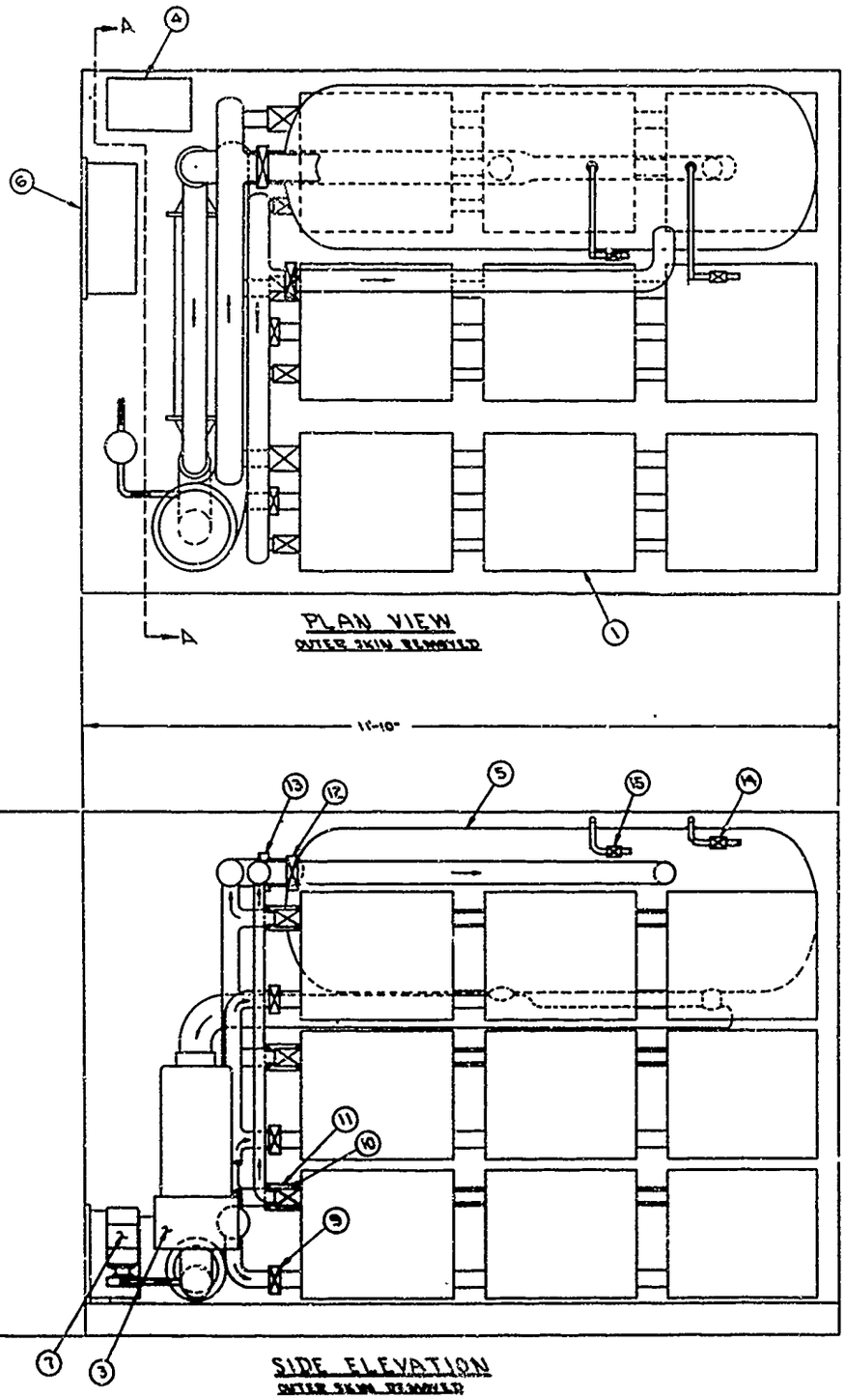
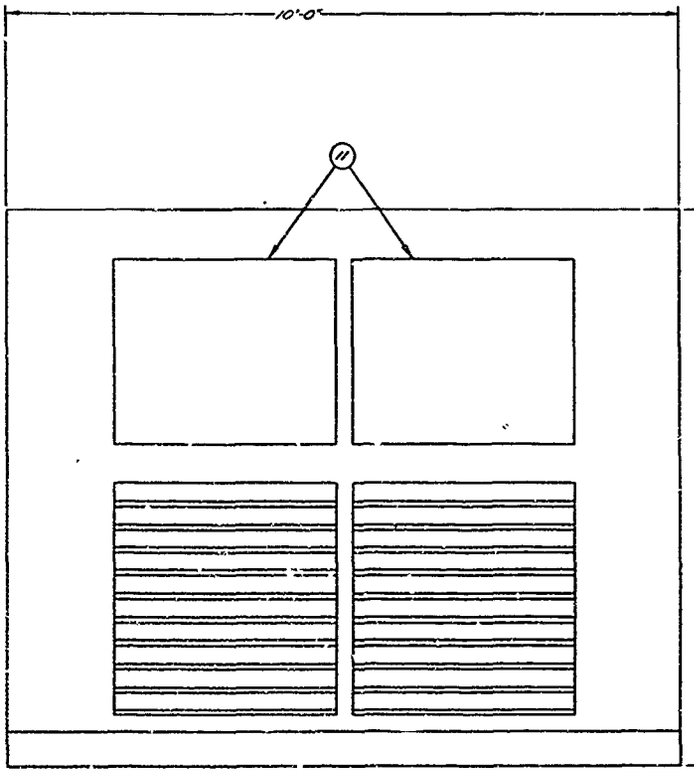


Figure 3.3-12. Hydrogen Subsystem No. II layout.

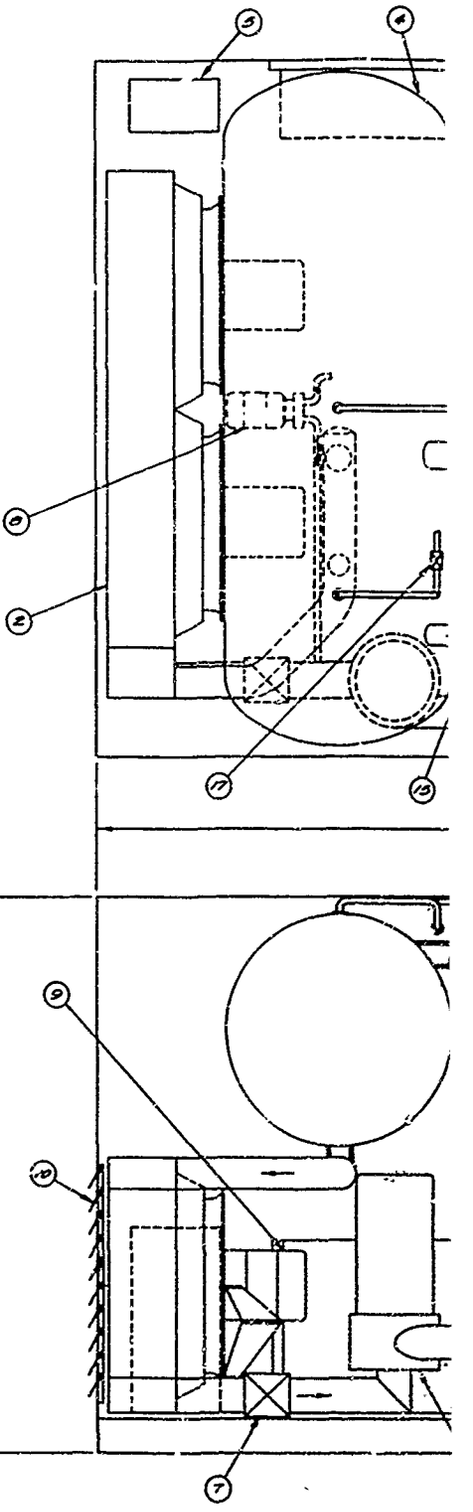
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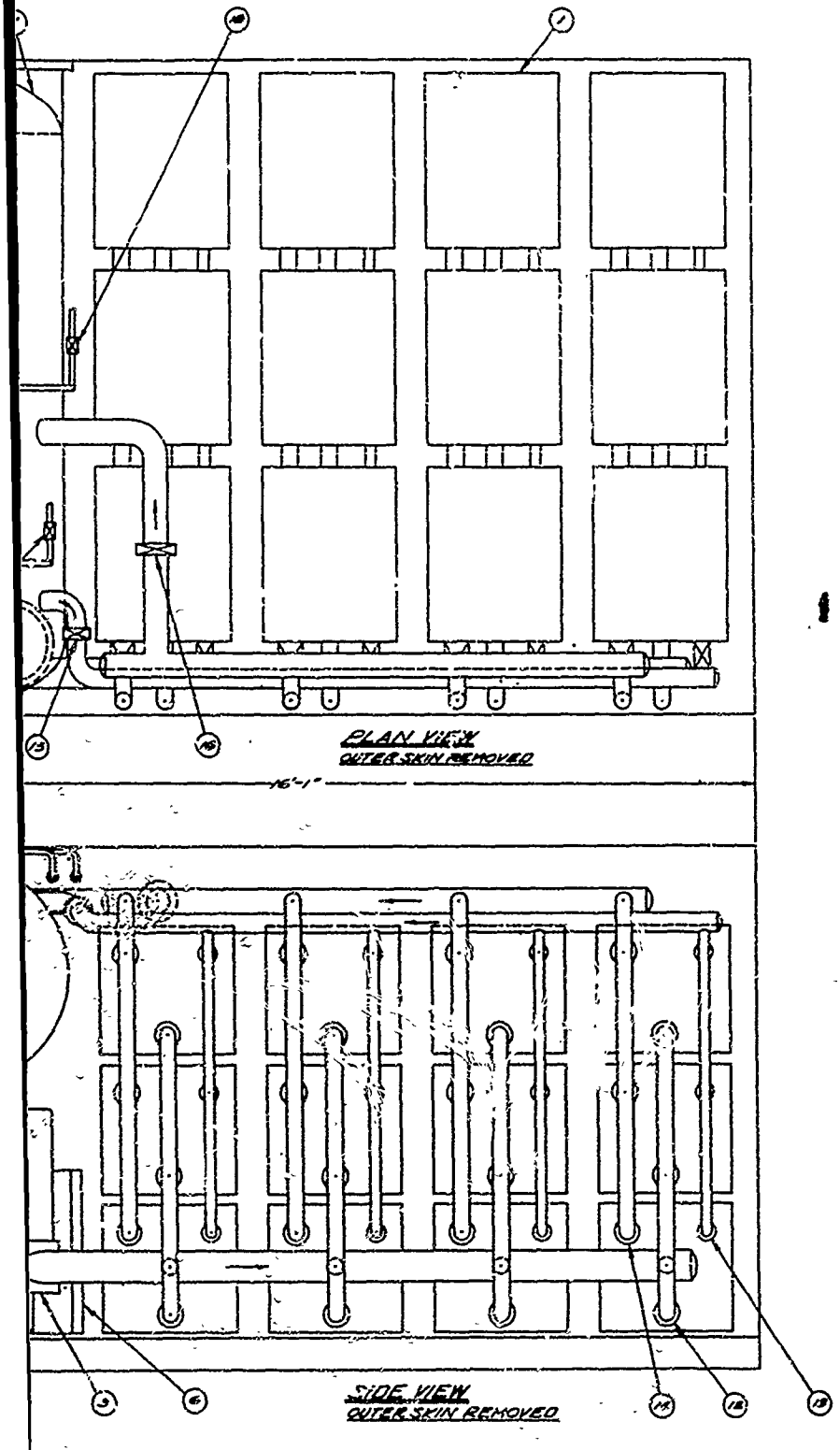
LEGEND

1 ELECTROLYZER MODULE (12 BANKS OF 3-36 TOTAL)
2 RADIATOR
3 ELECTROLYTE PUMP & MOTOR
4 LIQUID GAS SEPARATOR
5 SWITCH GEAR
6 RECTIFIER
7 CHECK VALVE
8 FEED-WATER PUMP & MOTOR
9 FAN MOTOR (2-REQ)
10 RADIATOR AIR LOUVERS
11 TYPICAL AIR INLET DOORS
12 ELECTROLYTE SHUT-OFF VALVE (1-PER BANK)
13 OXYGEN OUTLET CHECK VALVE (1-PER BANK)
14 HYDROGEN OUTLET CHECK VALVE (1-PER BANK)
15 OXYGEN SHUT-OFF VALVE
16 HYDROGEN SHUT-OFF VALVE
17 PNEUMATICALLY OPERATED OXYGEN EXHAUST VALVE
18 PNEUMATICALLY OPERATED HYDROGEN EXHAUST VALVE



END VIEW





NOTE: ELECTRICAL DISTRIBUTION SYSTEM OMITTED FOR CLARITY

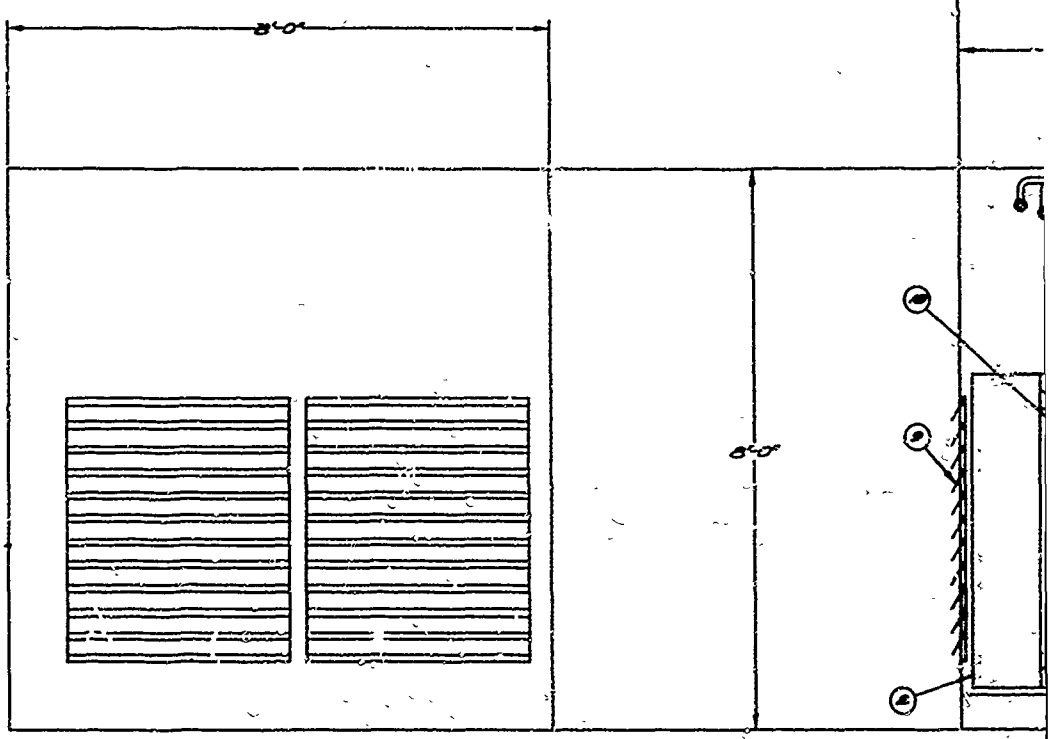
* 3-MW$_e$
MOBILITY CLASS II
AIR COOLED
60 CYCLE

Figure 3.3-13. Hydrogen Subsystem No. III layout.

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LEGEND

1. ELECTROLYZER MODULE (6 BANKS OF 3-18 TOTAL)
2. RADIATOR
3. ELECTROLYTE PUMP & MOTOR
4. SWITCH GEAR
5. LIQUID GAS SEPARATOR
6. RECTIFIER
7. FEED-WATER PUMP & MOTOR
8. TRANSFORMER
9. RADIATOR AIR LOUVERS
10. ELECTROLYTE SHUT-OFF VALVE (1-PER BANK)
11. OXYGEN OUTLET CHECK VALVE (1-PER BANK)
12. HYDROGEN OUTLET CHECK VALVE (1-PER BANK)
13. OXYGEN SHUT-OFF VALVE
14. HYDROGEN SHUT-OFF VALVE
15. PNEUMATICALLY OPERATED OXYGEN EXHAUST VALVE
16. PNEUMATICALLY OPERATED HYDROGEN EXHAUST VALVE
17. CHECK VALVE
18. FAN MOTORS (2 REQ)



END VIEW

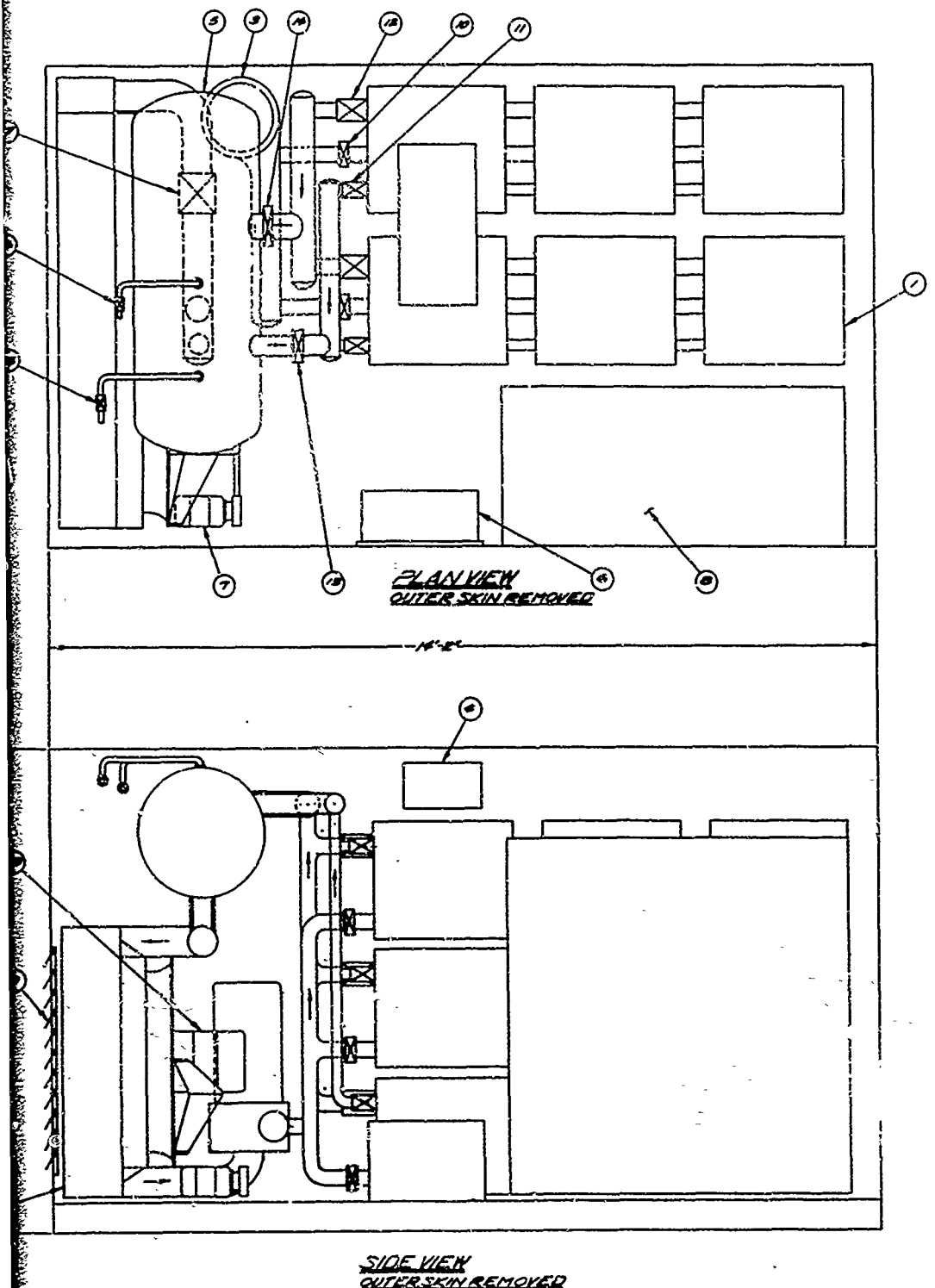
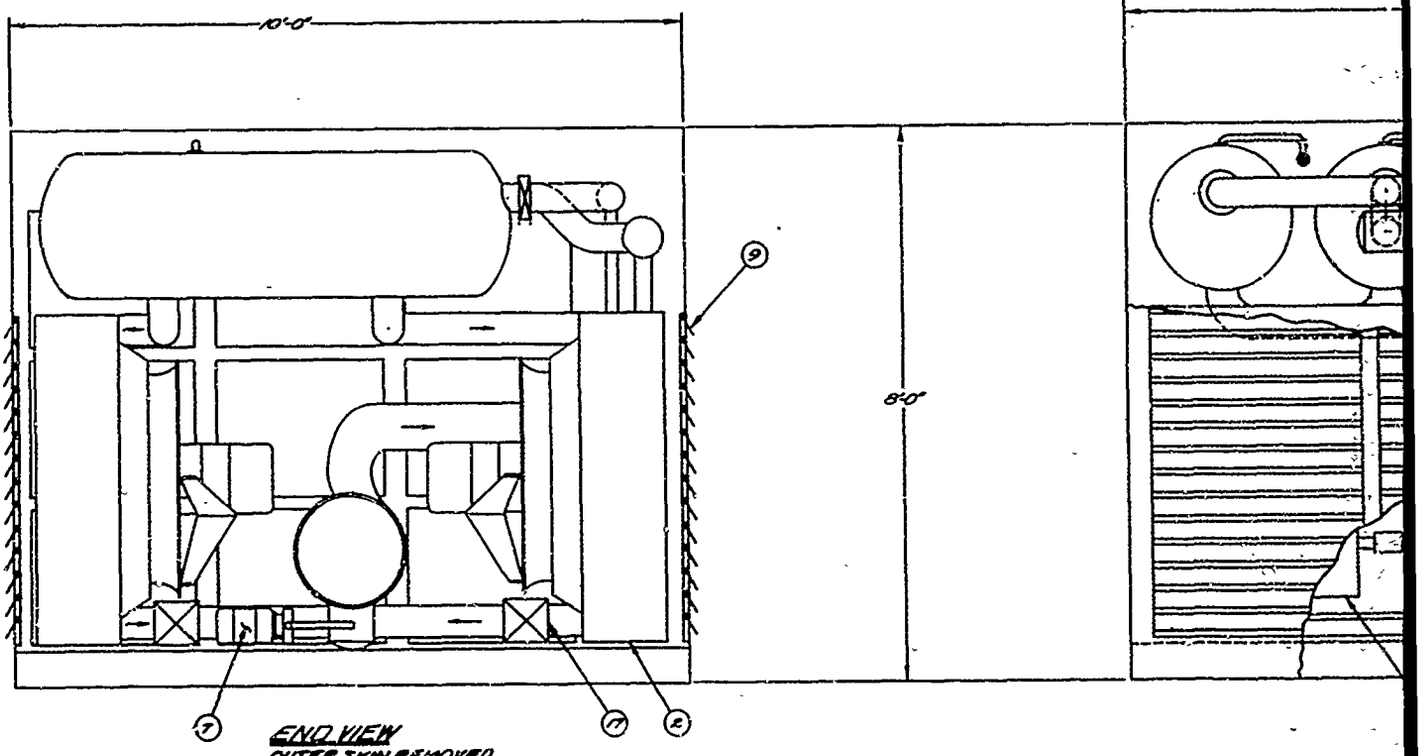


Figure 3 3-14. Hydrogen Subsystem No. IV layout.

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LEGEND

1. ELECTROLYZER MODULE (19 BANKS OF 3 - 99 TOTAL)
2. RADIATORS (2 REQ)
3. ELECTROLYTE PUMP & MOTOR
4. LIQUID GAS SEPARATORS, (a) O_2 - 1 REQ., (b) H_2 - 2 REQ.
5. SWITCH GEAR
6. RECTIFIER
7. FEED-WATER PUMP & MOTOR
8. CONTROL CABINET
9. RADIATOR AIR LOUVERS
10. ELECTROLYTE SHUT-OFF VALVE (1 PER BANK)
11. OXYGEN OUTLET CHECK VALVE (1 PER BANK)
12. HYDROGEN OUTLET CHECK VALVE (1 PER BANK)
13. OXYGEN SHUT-OFF VALVE
14. HYDROGEN SHUT-OFF VALVES (2 REQ)
15. PNEUMATICALLY OPERATED OXYGEN EXHAUST VALVE
16. PNEUMATICALLY OPERATED HYDROGEN EXHAUST VALVE
17. CHECK VALVES (2 REQ)
18. FAN MOTORS (2 REQ)



END VIEW
OUTER SKIN REMOVED

NOTE: ELECTRICAL DISTRIBUTION SYSTEM OMITTED FOR CLARITY

* GM116
MOBILITY CLASS II
AIR COOLED
60 CYCLE

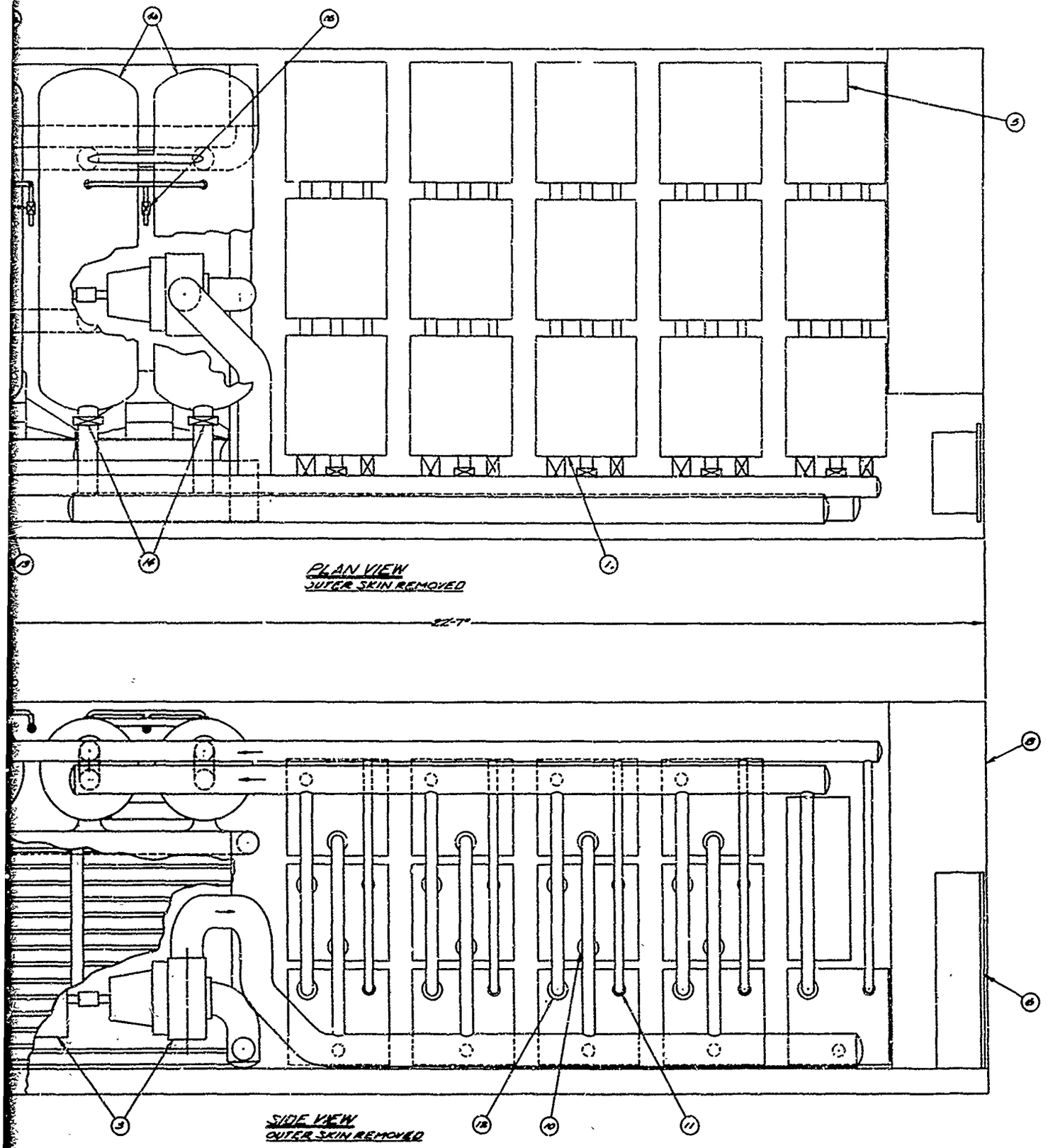
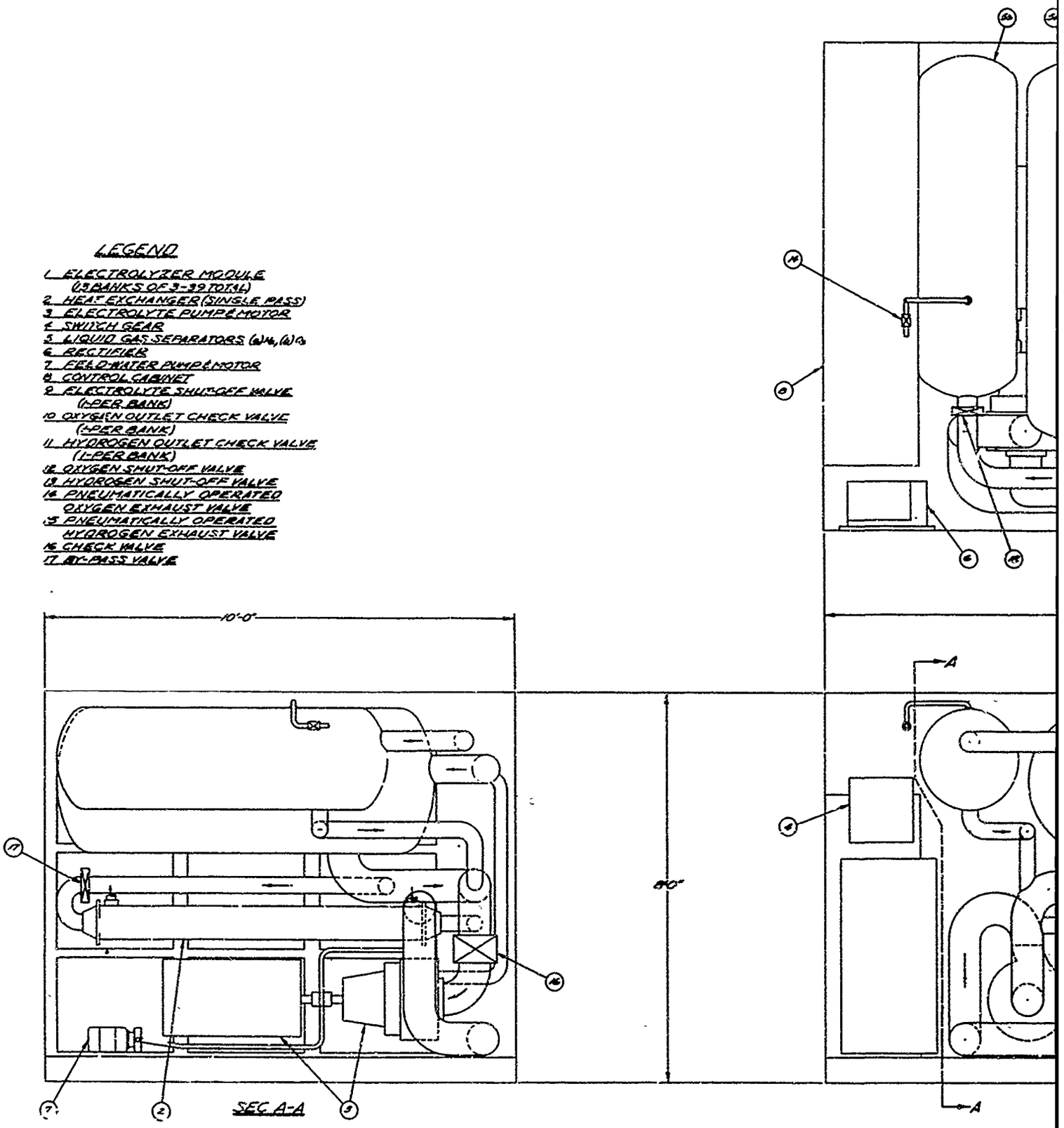
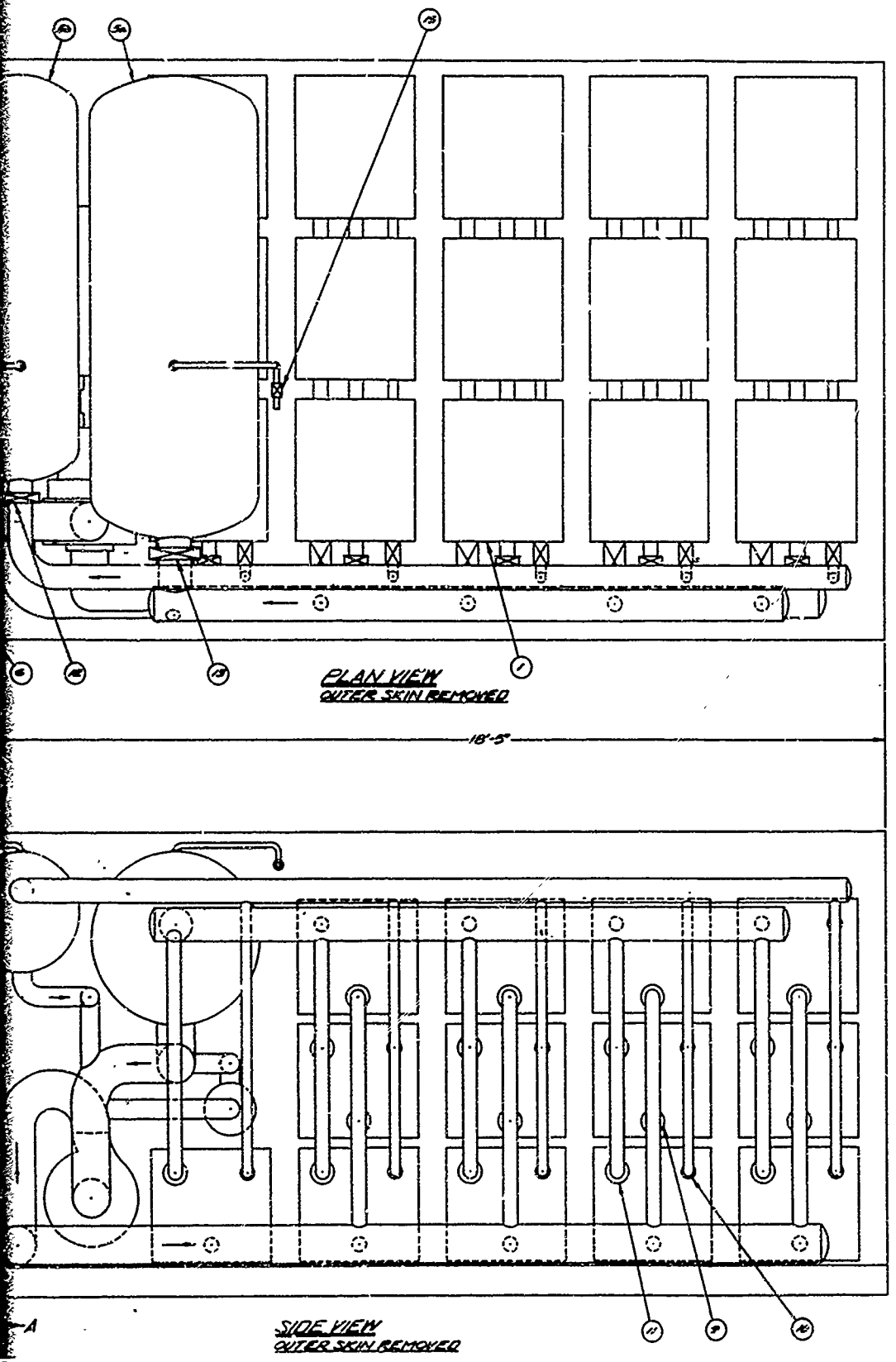


Figure 3.3-15. Hydrogen Subsystem No. V layout.

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NOTE: ELECTRICAL DISTRIBUTION SYSTEM OMITTED FOR CLARITY

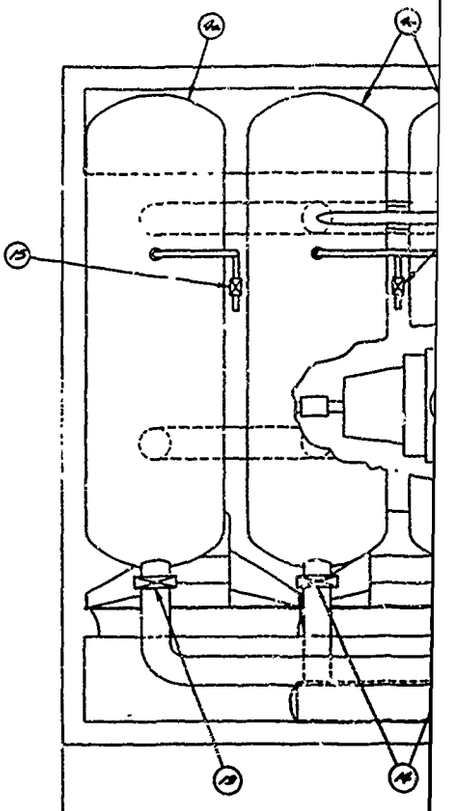
* GMW
MOBILITY CLASS II
WATER COOLED
60 CYCLE

Figure 3.3-16. Hydrogen Subsystem No. VI layout.

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LEGEND

1. ELECTROLYZER MODULE (12 BANKS OF 3-36 TOTAL)
2. RADIATORS (2 REQ)
3. ELECTROLYTE PUMP & MOTOR
4. LIQUID GAS SEPARATORS, (1) O_2-1REQ; (1) H_2-2REQ
5. SWITCH GEAR
6. RECTIFIER
7. FEED-WATER PUMP & MOTOR
8. CONTROL CABINET
9. RADIATOR AIR LOUVERS
10. ELECTROLYTE SHUT-OFF VALVE (1 PER BANK)
11. OXYGEN OUTLET CHECK VALVE (1 PER BANK)
12. HYDROGEN OUTLET CHECK VALVE (1 PER BANK)
13. OXYGEN SHUT-OFF VALVE
14. HYDROGEN SHUT-OFF VALVES (2 REQ)
15. PNEUMATICALLY OPERATED OXYGEN EXHAUST VALVE
16. PNEUMATICALLY OPERATED HYDROGEN EXHAUST VALVE
17. CHECK VALVES (2 REQ)
18. FAN MOTORS (4 REQ)
19. TRANSFORMER (400 CYCLE)



END VIEW
OUTER SKIN REMOVED

NOTE: ELECTRICAL DISTRIBUTION SYSTEM OMITTED FOR CLARITY

* 6 TYPE
MOBILITY CLASS II
AIR COOLED
400 CYCLE

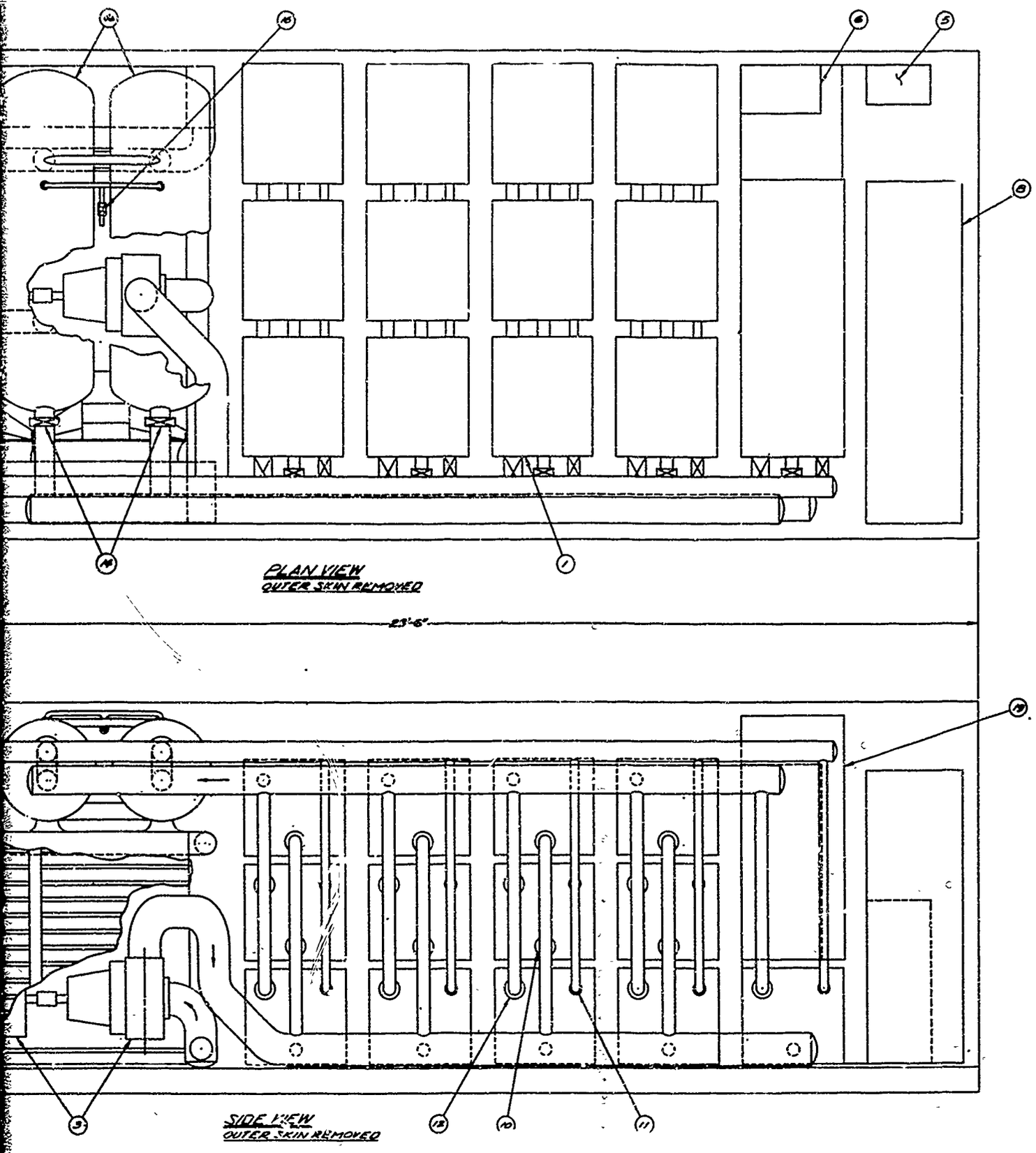
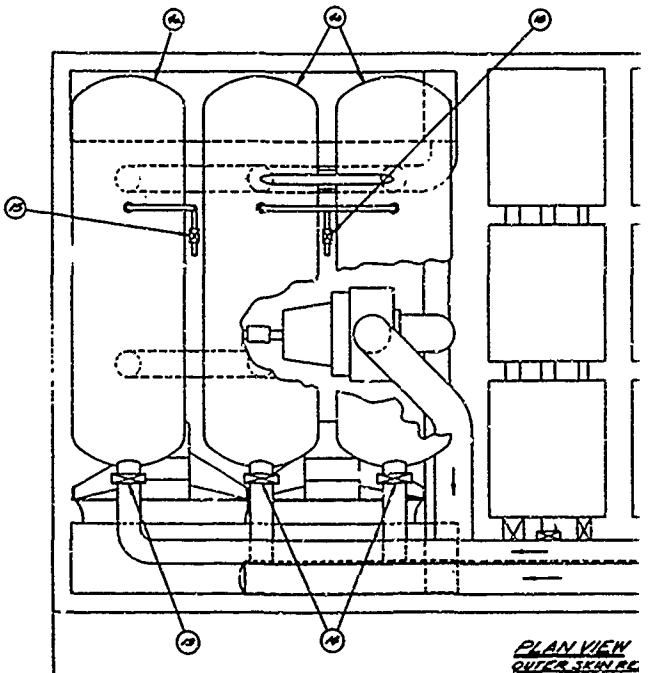


Figure 3.3-17. Hydrogen Subsystem No. VII layout.

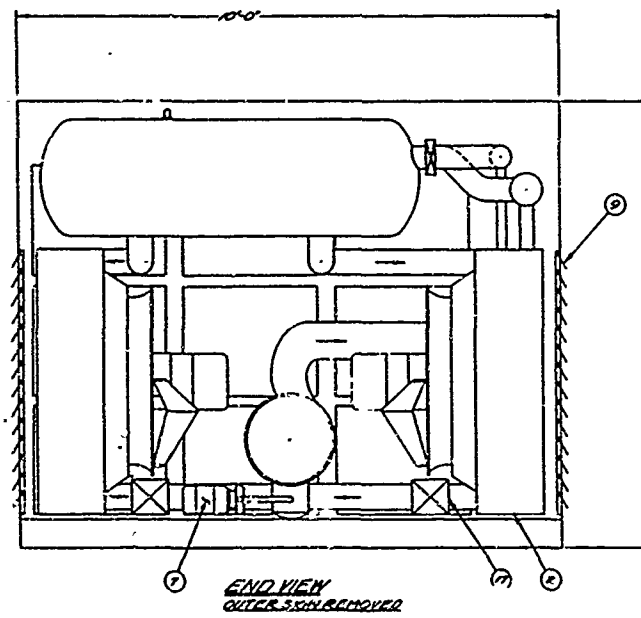
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LEGEND

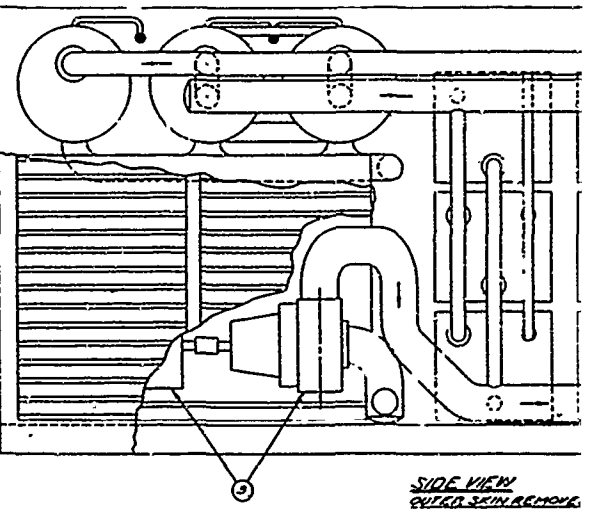
1. ELECTROLYTER MODULE (13 BANKS OF 3 = 39 TOTAL)
2. RADIATORS (2 REQ'D)
3. ELECTROLYTE PUMP & MOTOR
4. LIQUID GAS SEPARATORS, (A) O_2 - 1 REQ., (B) H_2 - 2 REQ.
5. SWITCH GEAR
6. RECTIFIER
7. FEED-WATER PUMP & MOTOR
8. CONTROL CABINET
9. RADIATOR AIR LOUVERS
10. ELECTROLYTE SHUT-OFF VALVE (1 PER BANK)
11. OXYGEN OUTLET CHECK VALVE (1 PER BANK)
12. HYDROGEN OUTLET CHECK VALVE (1 PER BANK)
13. OXYGEN SHUT-OFF VALVE
14. HYDROGEN SHUT-OFF VALVES (2 REQ'D)
15. PNEUMATICALLY OPERATED OXYGEN EXHAUST VALVE
16. PNEUMATICALLY OPERATED HYDROGEN EXHAUST VALVE
17. CHECK VALVES (2 REQ'D)
18. FAN MOTORS (2 REQ'D)
19. TRANSFORMER



PLAN VIEW
OUTER SKIN RE



END VIEW
OUTER SKIN REMOVED



SIDE VIEW
OUTER SKIN REMOVE

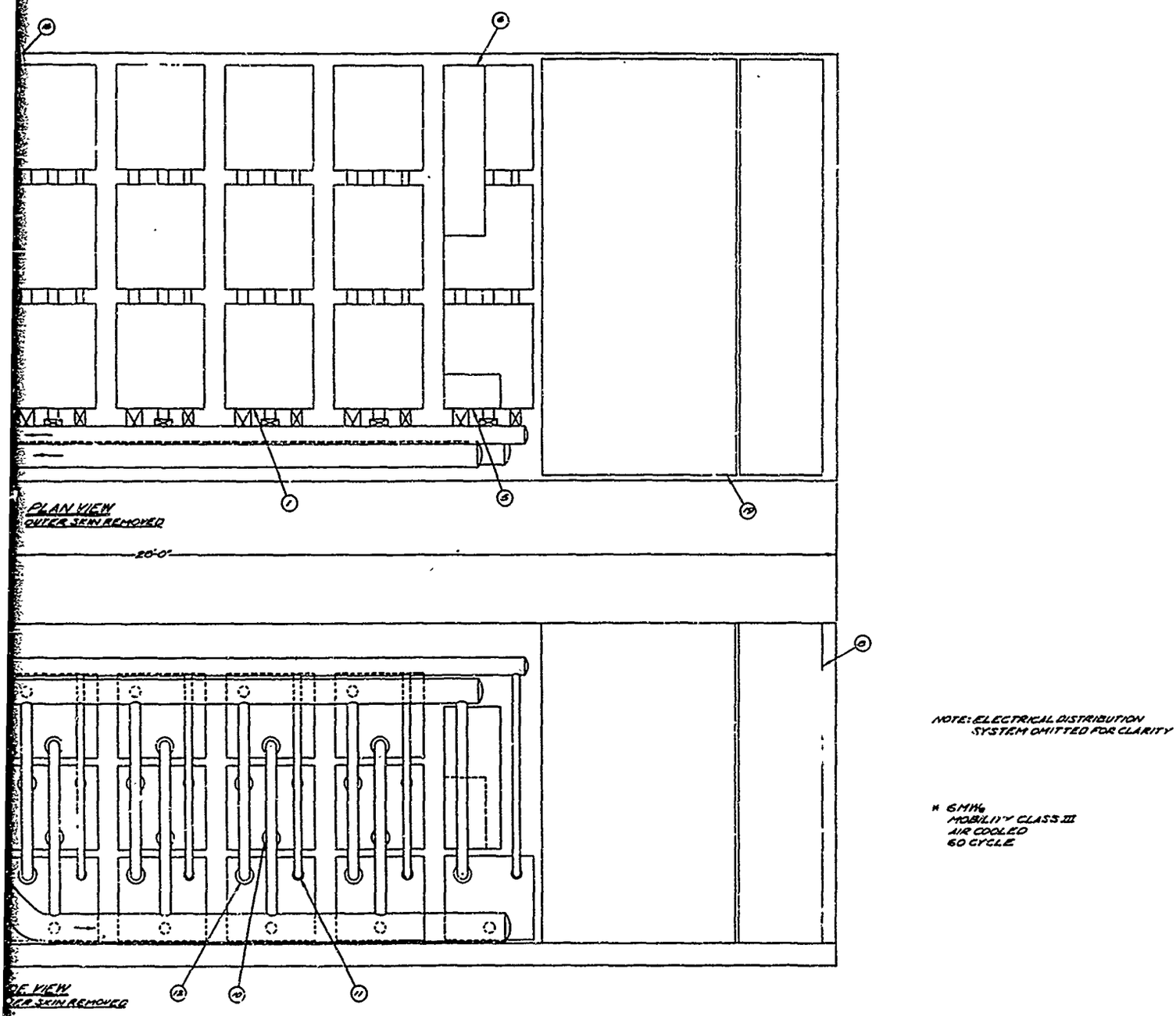
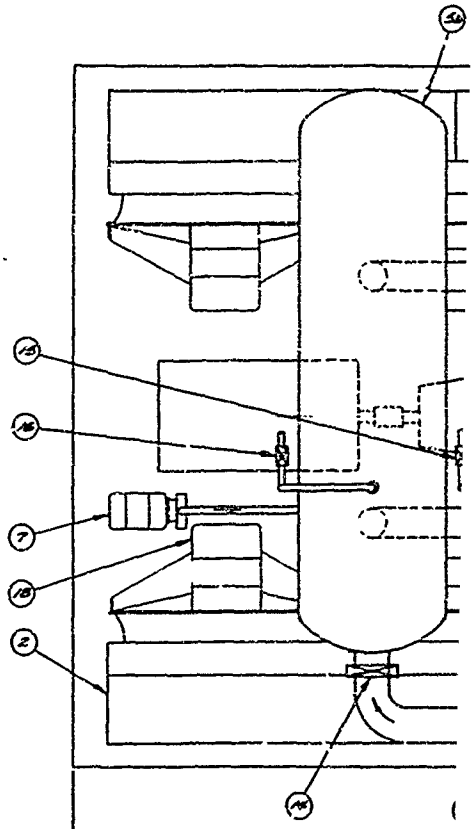


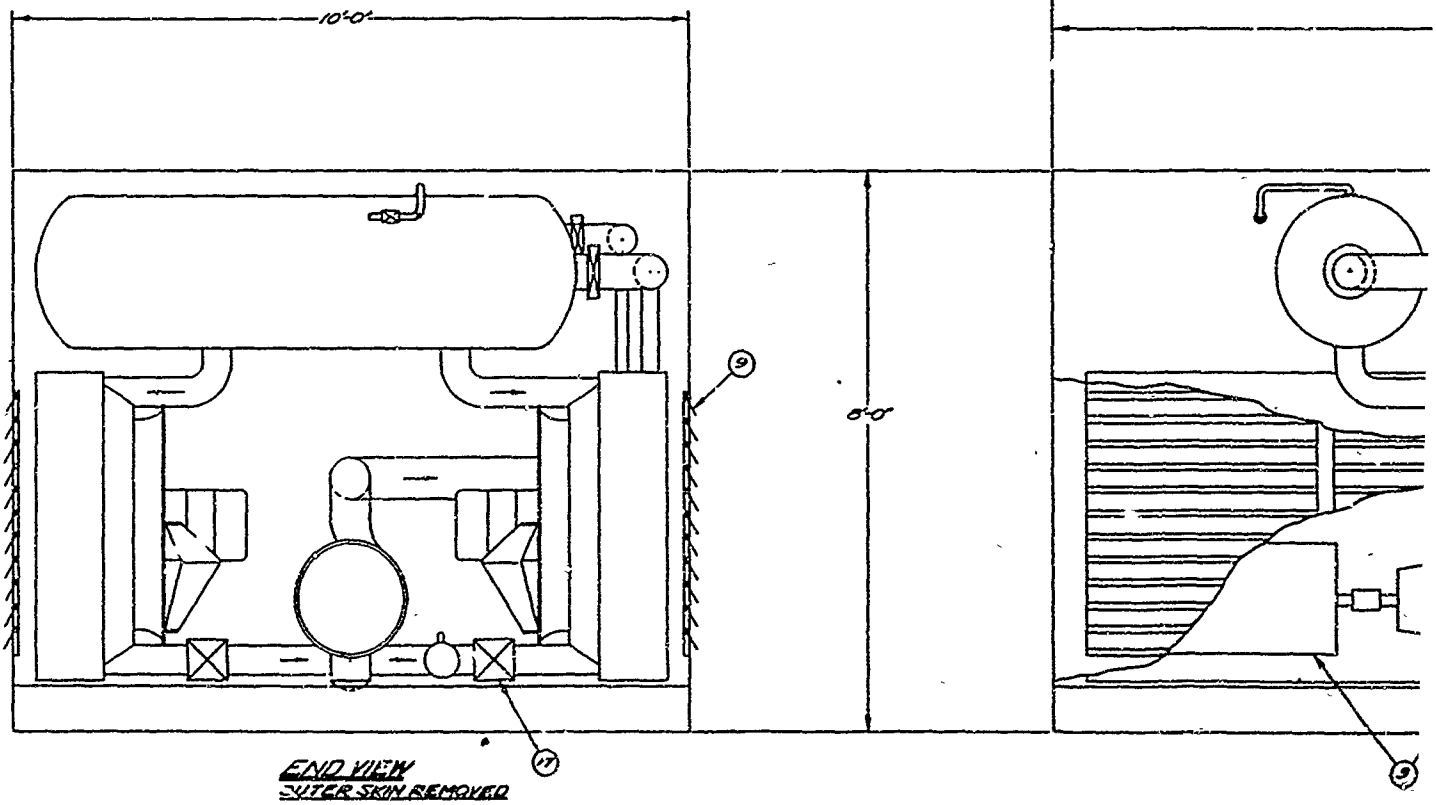
Figure 3.3-18. Hydrogen Subsystem No. VIII layout.

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LEGEND

1. ELECTROLYZER MODULE (2 BANKS OF 9-ET TOTAL)
2. RADIATOR (2 REQ)
3. ELECTROLYTE PUMP & MOTOR
4. SWITCH GEAR
5. LIQUID GAS SEPARATORS, (a) O_2, (b) H_2
6. RECTIFIER
7. FEED-WATER PUMP & MOTOR
8. TRANSFORMER
9. RADIATOR AIR LOUVERS
10. ELECTROLYTE SHUT-OFF VALVE (1-PER BANK)
11. OXYGEN OUTLET CHECK VALVE (1-PER BANK)
12. HYDROGEN OUTLET CHECK VALVE (1-PER BANK)
13. OXYGEN SHUT-OFF VALVE
14. HYDROGEN SHUT-OFF VALVE
15. PNEUMATICALLY OPERATED OXYGEN EXHAUST VALVE
16. PNEUMATICALLY OPERATED HYDROGEN EXHAUST VALVE
17. CHECK VALVE (2 REQ)
18. FAN MOTORS (4 REQ)





END VIEW
OUTER SKIN REMOVED

NOTE: ELECTRICAL DISTRIBUTION SYSTEM OMITTED FOR CLARITY

* 10MW TOTAL-
2 SKIDS REQ @ 5MW
MOBILITY CLASS II
AIR COOLED
60 CYCLE

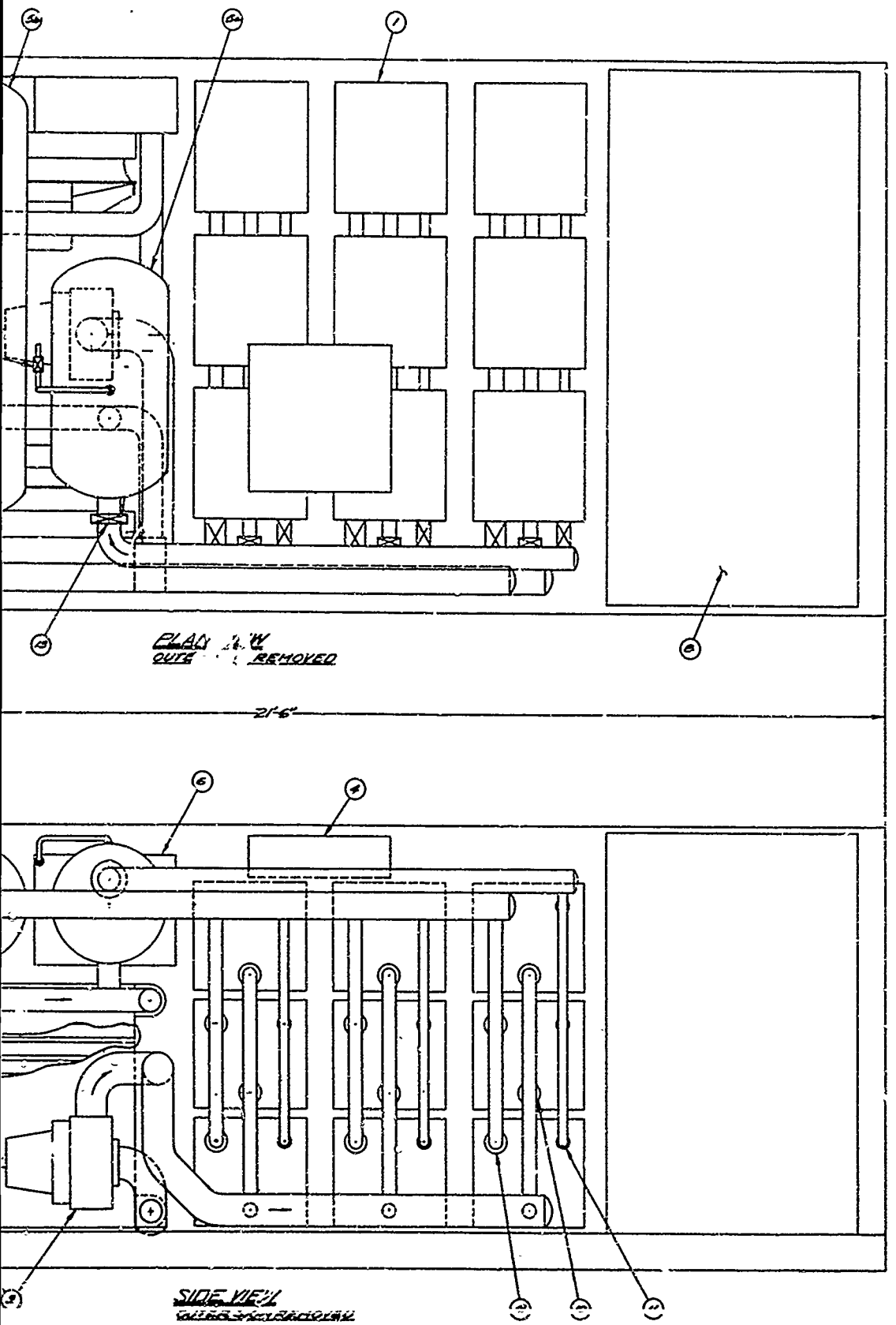
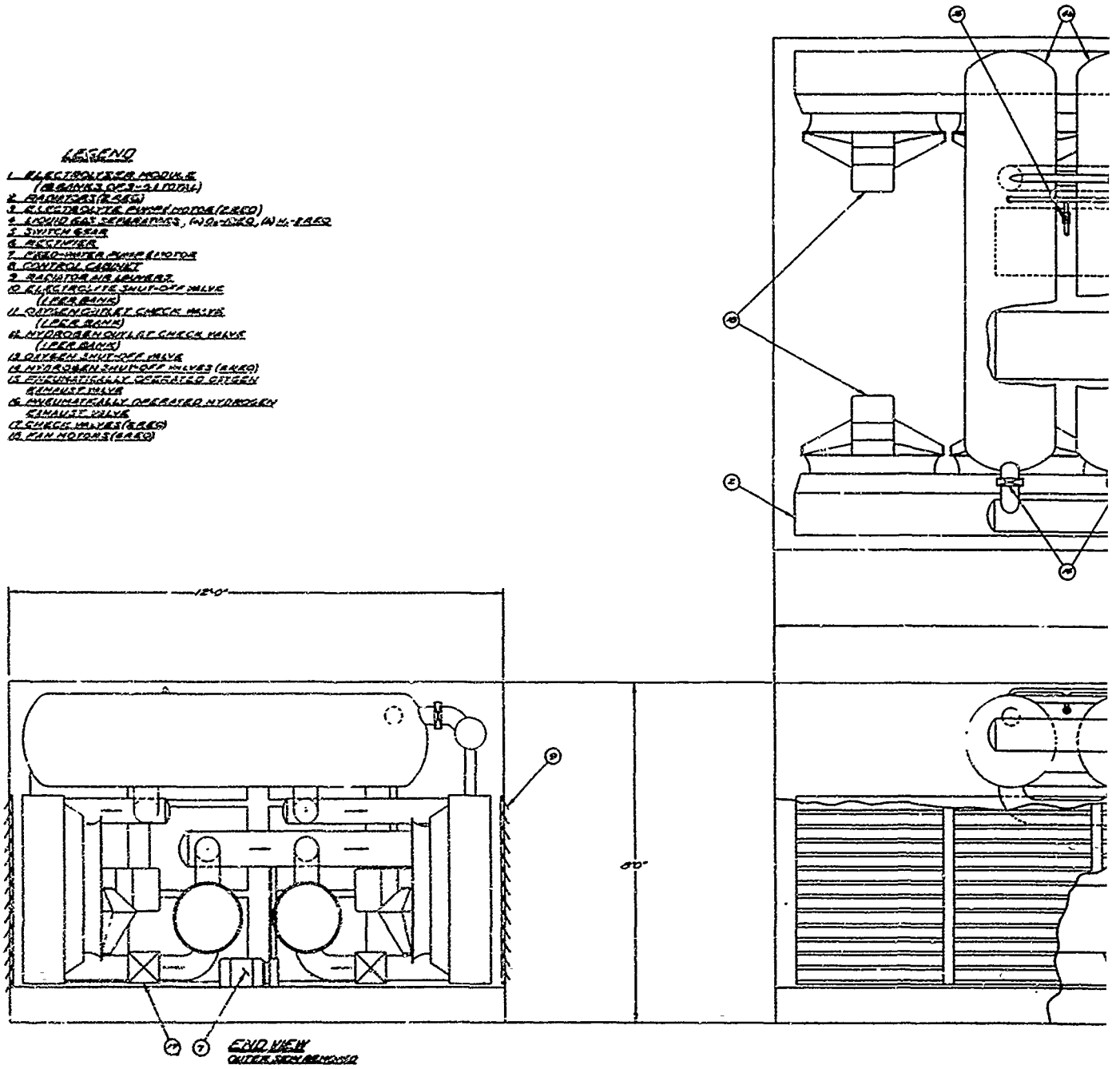


Figure 3.3-19. Hydrogen Subsystem No. IX layout.

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LEGEND

1. ELECTROLYZER MODULE (8 BANKS OF 3-CELLS EA.)
2. RADIATORS (8 REQ.)
3. ELECTROLYTE PUMP & MOTOR (2 REQ.)
4. LIQUID GAS SEPARATORS, H2 O2 (2 REQ. EA.)
5. SWITCH GEAR
6. ACCUMULATOR
7. H2O-WATER PUMP & MOTOR
8. CONTROL CABINET
9. RADIATOR AIR LOUVERS
10. ELECTROLYTE SHUT-OFF VALVE (1 PER BANK)
11. HYDROGEN OUTLET CHECK VALVE (1 PER BANK)
12. HYDROGEN OUTLET CHECK VALVE (1 PER BANK)
13. OXYGEN SHUT-OFF VALVE
14. HYDROGEN SHUT-OFF VALVES (8 REQ.)
15. PNEUMATICALLY OPERATED OXYGEN EXHAUST VALVE
16. PNEUMATICALLY OPERATED HYDROGEN EXHAUST VALVE
17. CHECK VALVES (8 REQ.)
18. FAN MOTORS (8 REQ.)



END VIEW
OUTER SKIN REMOVED

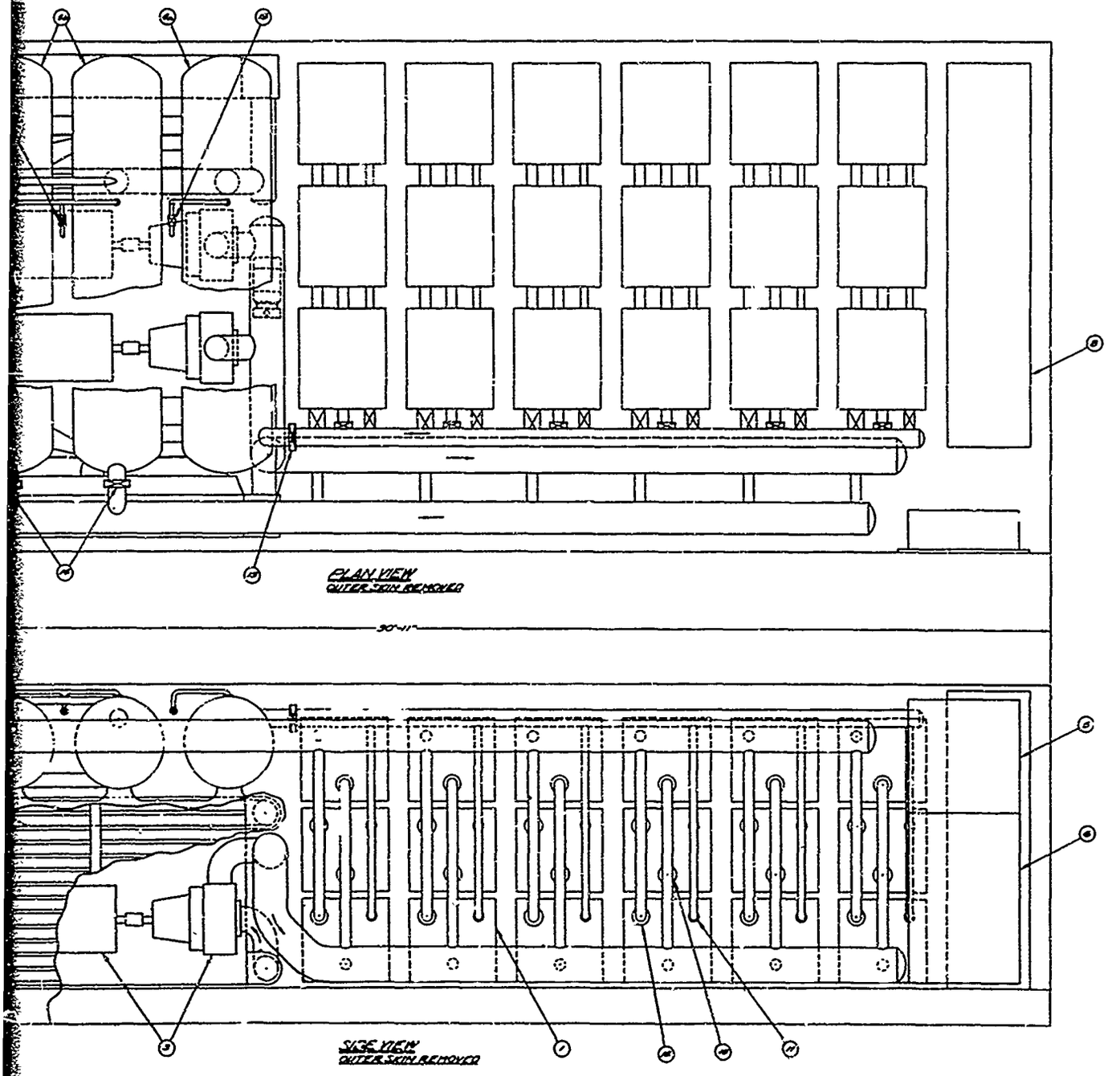
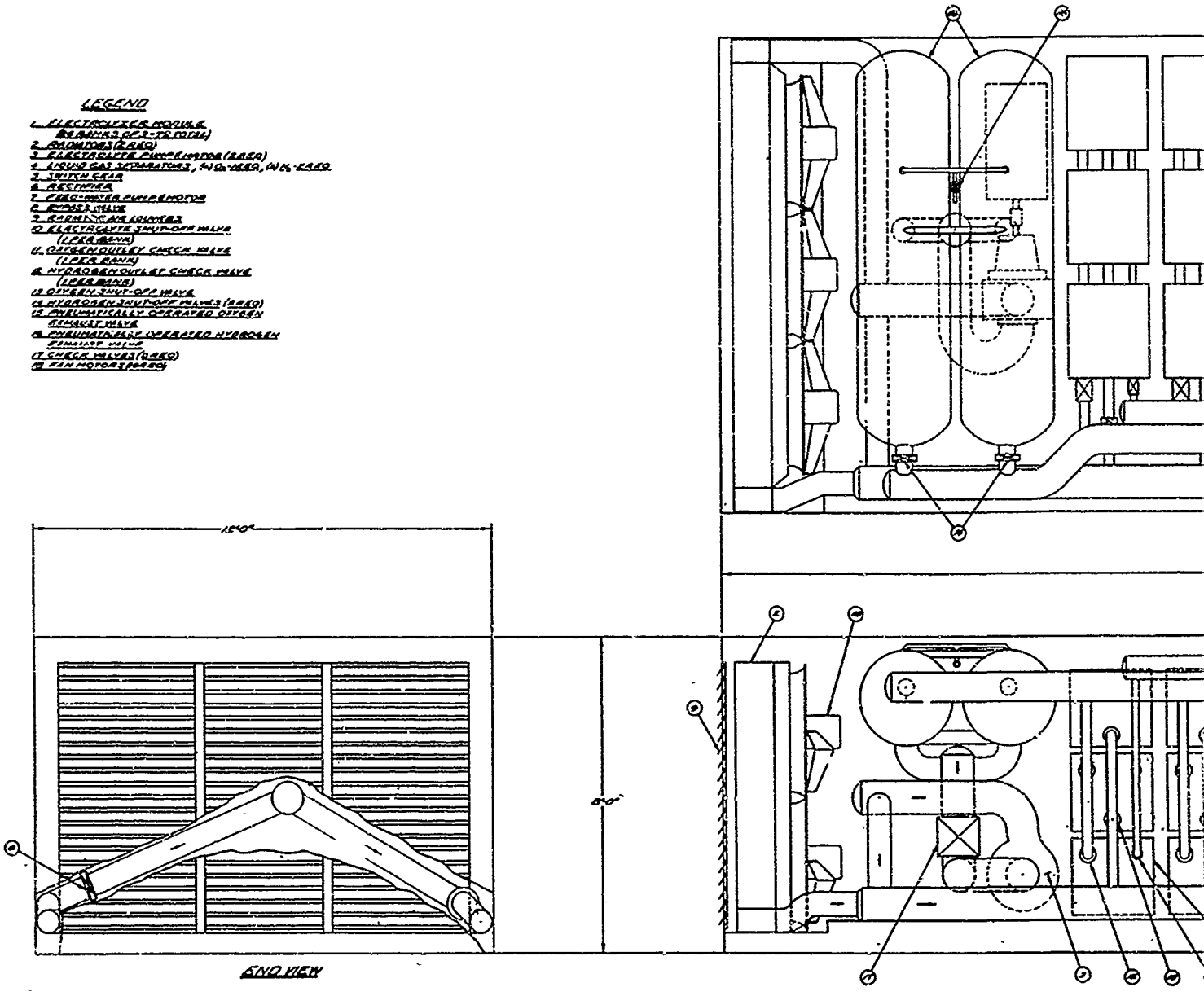


Figure 3.3-20. Hydrogen Subsystem No. X layout.

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LEGEND

1. ELECTROLYZER MODULE (6 BANKS OF 2-12 CELLS)
2. MANIFOLDS (2 REQ)
3. ELECTROLYTE PUMP/MOTOR (2 REQ)
4. LIQUID GAS SEPARATORS, H2 (2 REQ), O2 (2 REQ)
5. SWITCH GEAR
6. RECEIVER
7. CIRC-WATER PUMP/MOTOR
8. BYPASS VALVE
9. KARMAN NOZZLES
10. ELECTROLYTE SHUT-OFF VALVE (1 PER BANK)
11. OXYGEN OUTLET CHECK VALVE (1 PER BANK)
12. HYDROGEN OUTLET CHECK VALVE (1 PER BANK)
13. OXYGEN SHUT-OFF VALVE
14. HYDROGEN SHUT-OFF VALVES (2 REQ)
15. PNEUMATICALLY OPERATED OXYGEN BYPASS VALVE
16. PNEUMATICALLY OPERATED HYDROGEN BYPASS VALVE
17. CHECK VALVES (2 REQ)
18. FAN MOTORS (2 REQ)



END VIEW

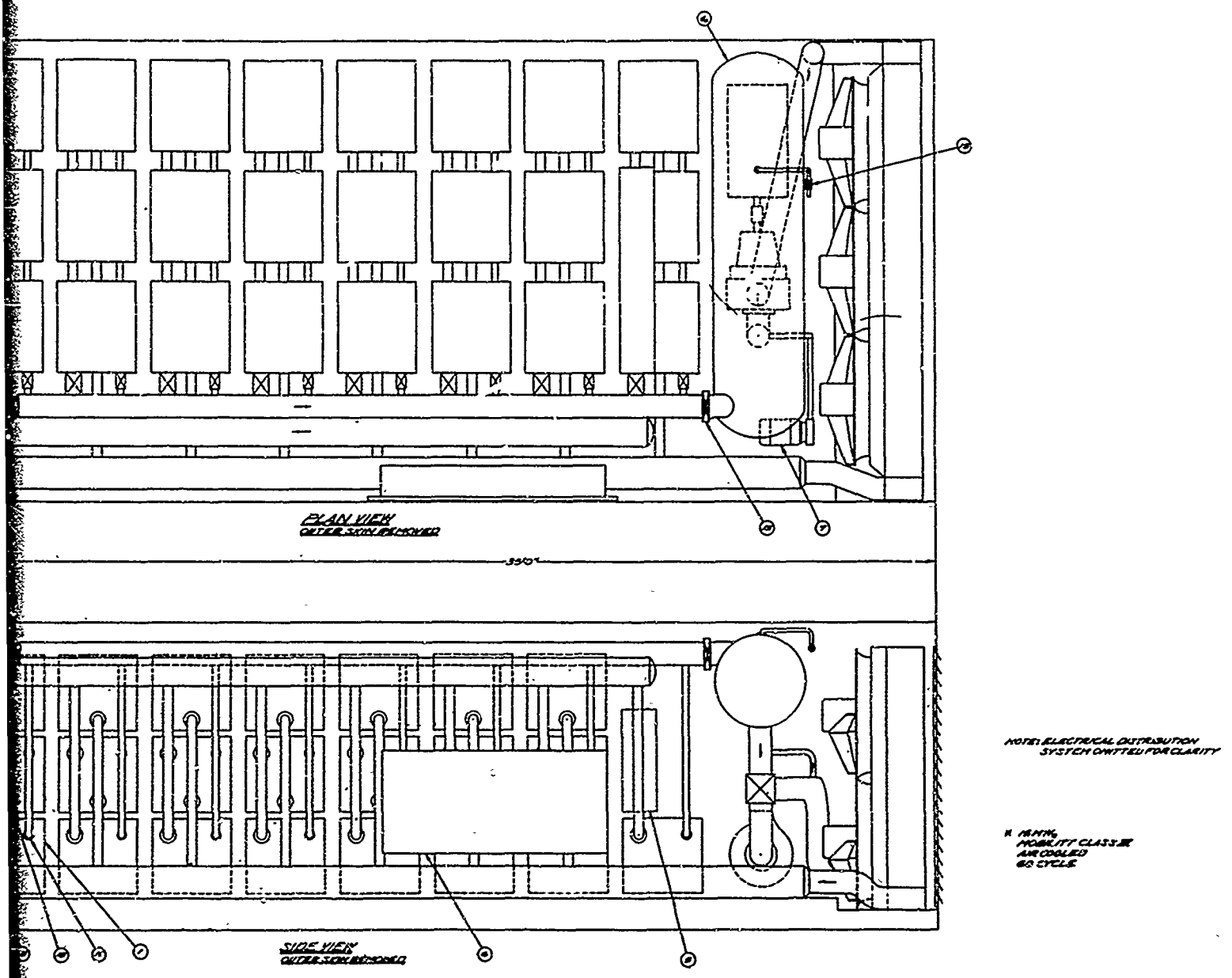


Figure 3.3-21. Hydrogen Subsystem No. XI layout.

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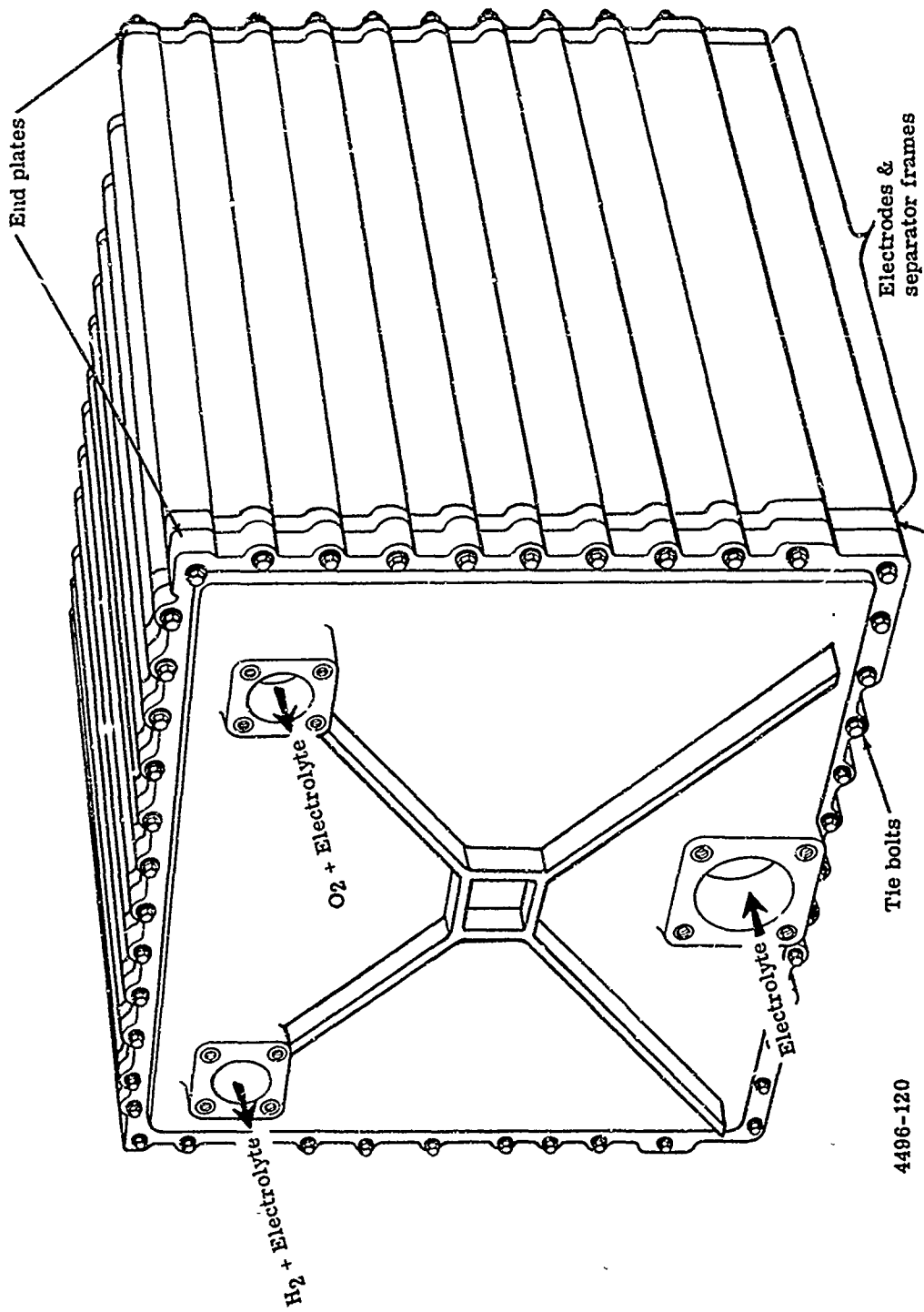


Figure 3.3-22. Typical electrolyzer module.

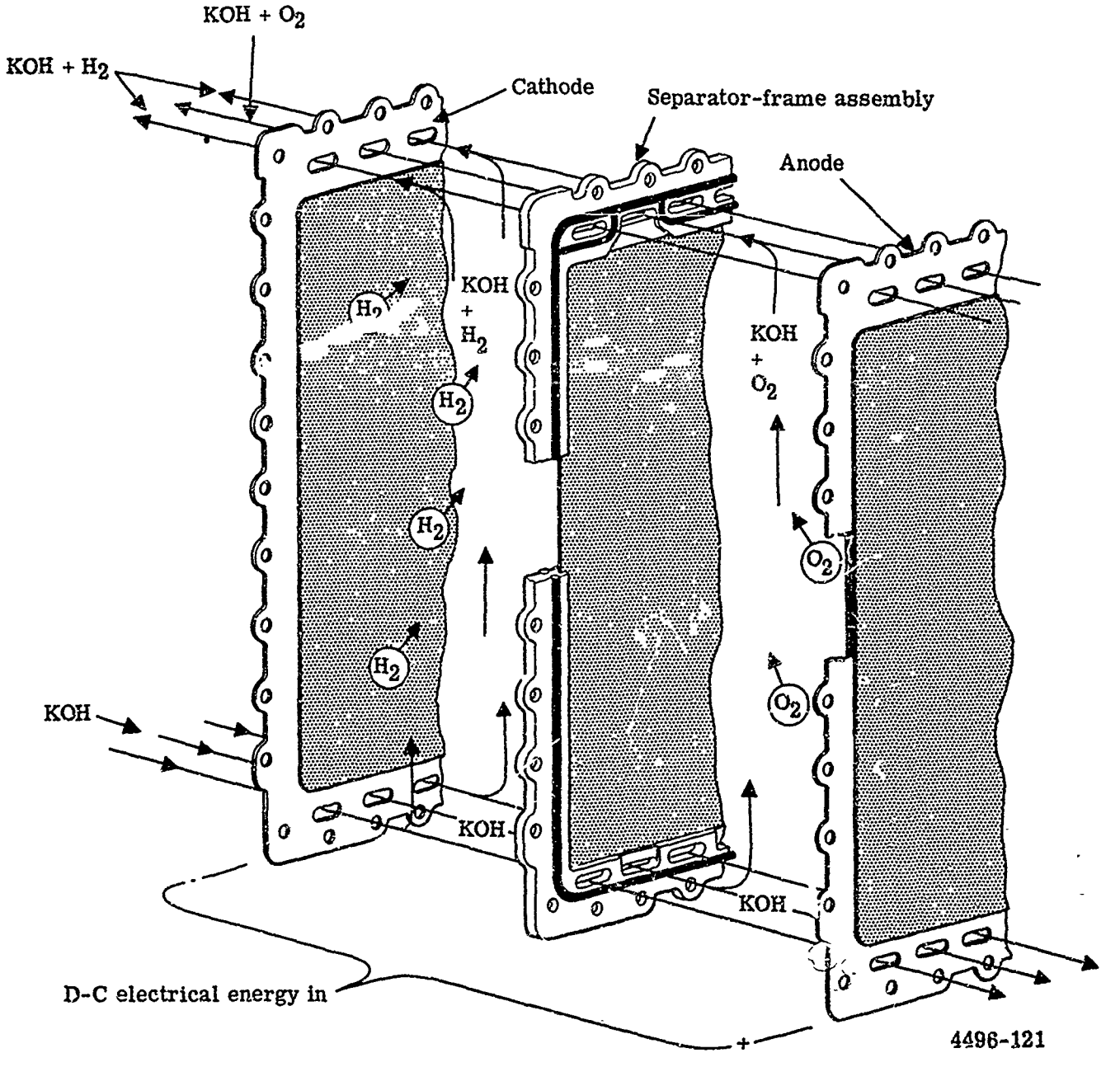
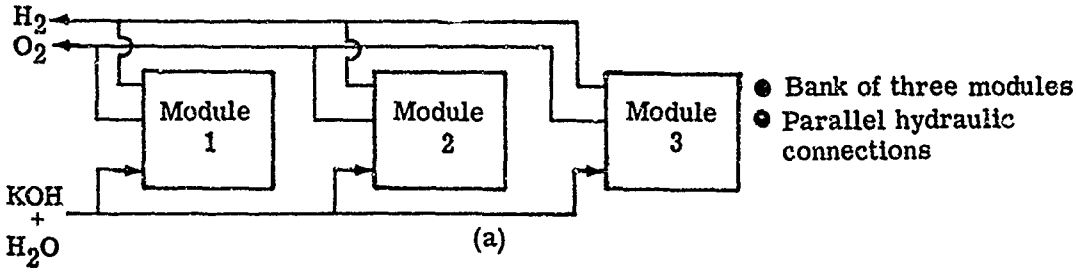
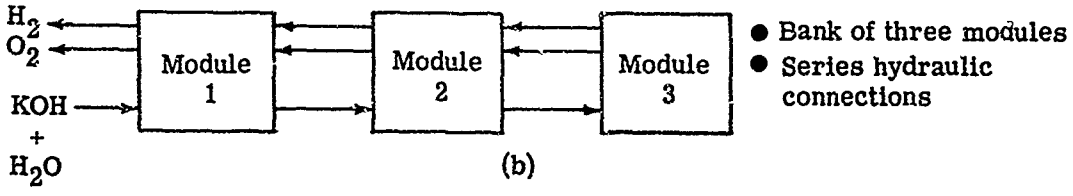


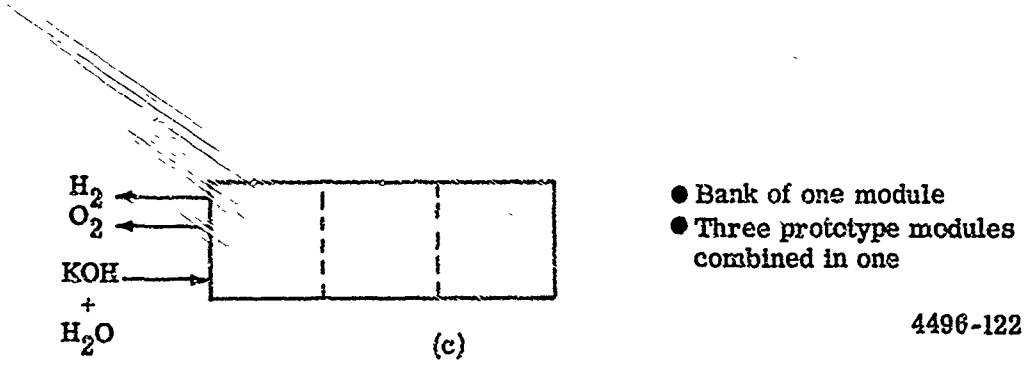
Figure 3.3-23. Flow distribution within an electrolyzer module.



Design proposed during energy depot electrolysis system study



Current proposed design



Alternate design

Figure 3.3-24. Various methods of module arrangement.

Table 3.3-III.

Hydrogen production subsystem electrolyzer data.*

Entry	Parameter	Subsystem 1	2	3	4**	5	6	7	8	9**	10	11
1	Number of modules	24	24	33	18	39	39	39	39	27	54	66
2	Number of banks	8	8	12	6	13	13	13	13	9	18	22
3	Total electrolyzer wt (wet)—lb	26,210	26,210	41,417	18,765	41,160	41,160	41,160	41,160	27,506	55,012	64,094
4	Total cell area—ft^2	7,680	7,680	12,220	5,450	11,990	11,990	11,990	11,990	7,960	15,920	18,300
5	Cells per module	116	116	124	110	112	112	112	112	107	107	101
6	Current density—ma/cm^2	216	216	142	290	272	272	272	272	330	330	413
7	Voltage drop per cell—volts	1.704	1.704	1.621	1.767	1.750	1.750	1.750	1.750	1.813	1.813	1.898
8	Cell electrolyte velocity—fps	0.043	0.043	0.023	0.059	0.045	0.045	0.045	0.045	0.052	0.052	0.061

*Electrode dimensions (for all plants) 18 in. × 22 in.
Cell spacing (for all plants) 0.446 cm.

**Two packages are required. Data tabulated are for one package.

In the second packaging arrangement, the three modules in each bank are positioned across the width of the skid. An example of this arrangement is design 11. From Table 3.3-III, it can be seen that this design has 22 module banks. The packaging layout is shown in Figure 3.3-21. This arrangement, which is generally preferable compared to the first, is limited only by the length of the skid and allows more module banks to be used. The second arrangement also has simpler piping layouts and easier access for maintenance.

Heat Exchangers

Subsystem heat exchanger data is tabulated in Table 3.3-IV. For design conditions, the inlet cooling fluid temperature and pressure were 77°F and 14.7 psia, respectively. The heat exchangers are designed to dissipate 8% more heat than the amount tabulated in Item 3. This increase in capacity is necessary to accommodate the increased heat load that occurs when a subsystem plant is operated at lower efficiency—with 25% of its module banks shutdown.

Normally, the heat exchangers were sized to accommodate full electrolyte flow. In three cases, designs 2, 6, and 11, however, part of the electrolyte flow was bypassed around the heat exchangers.

In designs 2 and 6, a bypass was used because of the small size of the heat exchangers. Unlike the air-cooled units, high film coefficients were possible in both the electrolyte and cooling fluid streams. As a result, the amount of heat transfer surface area required to dissipate the heat load was reduced from that of comparable air-cooled units. Hence, the size of the heat exchanger was reduced to take full advantage of the improved thermal performance. The resulting compact heat exchanger designs lacked sufficient electrolyte flow area, and partial flow was bypassed.

Table 3.3-IV.

Hydrogen production subsystem heat exchanger data.

Item	Parameter	Subsystem No. 1	2	3	4*	5	6	7	8	9*	10	11
1	Cooling fluid	Air	Water	Air	Air	Air	Water	Air	Air	Air	Air	Air
2	Number of heat exchangers per plant	1	1	1	1	2	1	2	2	2	2	2
3	Heat rejected from electrolysis—kw	338	338	233	422	625	825	825	825	804	1,608	2,898
4	Weight - wet—lb	1,555	386	1,290	1,767	3,443	694	3,443	3,443	2,045	5,890	7,611
5	Height—ft	4.5	0.92**	3.8	4.5	4.6	0.92**	4.6	4.6	4.3	4.3	6.8
6	Length—ft	6.7	3.8	6.2	6.7	6.5	7.5	6.5	6.5	6.7	10.2	10.7
7	Number of fans per radiators	2	-	2	2	2	-	2	2	2	3	6
8	Fan diameter—in.	36	-	30	36	36	-	36	36	36	36	36
9	Fan power - each—kw	4.1	-	4.5	7.7	20.2	-	20.2	20.2	15.7	48.3	51.3
10	Cooling fluid flow rates—Air: acfm—water: gph	12,500	3,960	9,400	18,300	21,000	9,900	21,000	21,000	16,400	36,800	50,000
11	Cooling fluid velocity—fpm	560	105	625	820	975	250	975	975	805	1,130	820
12	Cooling fluid outlet temperature —°F	155	110	154	154	146	110	146	146	149	147	156

*Two packages required. Data tabulated are for one package.

**Diameter for water cooled units.

Table 3.2-II.
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In design number 11 (Figure 3.3-21), the piping and separator-reservoir layout contains two heat exchangers. The electrolyte flow in the hydrogen and oxygen separator-reservoirs is split in a volume ratio of 2 to 1. Rather than designing two separate heat exchangers for differing flow rates, it was decided that two identical heat exchangers would be simpler and cheaper. To obtain equal flow rates, a bypass valve was used on the heat exchanger associated with the hydrogen separator-reservoir.

The quantity of flow diverted around each of the heat exchangers in systems 2, 6, and 11 would be fixed when the plants are assembled by a "tuning" valve in the bypass line. Once the valve position is set, it would not be necessary to vary its setting as a continuous control fuction. The flow split would be such that when the two streams are reunited downstream from the heat exchanger, the resulting temperature would be 167°F.

In the case of air-cooled systems, the units are single pass for both the electrolyte and air. The electrolyte flows through 5/8-in. OD stainless steel tubes with aluminum fins. For off-design point operation, the heat dump capability would be varied by either turning one or more of the fans on or off or pneumatically actuated air louvers. For operation above 77°F, all of the fans would be operated and the louvers would be fully opened. In cold weather, the fan motors could be shut off and/or the louvers could be partially closed.

For the water-cooled systems, heat exchangers are of the single pass shell and tube design. The electrolyte flows in 3/8-in. OD stainless steel tubes. The cooling water would flow outside the tubes inside the 11-in. steel shell. The heat load dissipated by the unit would be controlled by varying the cooling water flow rate.

Separator-Reservoir

The separator-reservoirs used in this analysis consist of stainless steel cylindrical tanks in which stainless steel filter cartridges are located. The arrangement is shown in Figure 3.3-25. The tanks are mounted horizontally. Inside the tank, a perforated deck is positioned parallel to the horizontal axis.

The filter cartridges, made of corrugated stainless steel screen, are mounted over the holes in the deck. The two-phase mixture to be separated enters the vessel and impinges on the cartridge screening. The liquid passes through the screening and is withdrawn and sent to the heat exchanger. By virtue of the mesh size, the gas is prevented from passing through the screen and is separated from the liquid. Before leaving the vessel, both gas streams pass through demisters to eliminate entrainment.

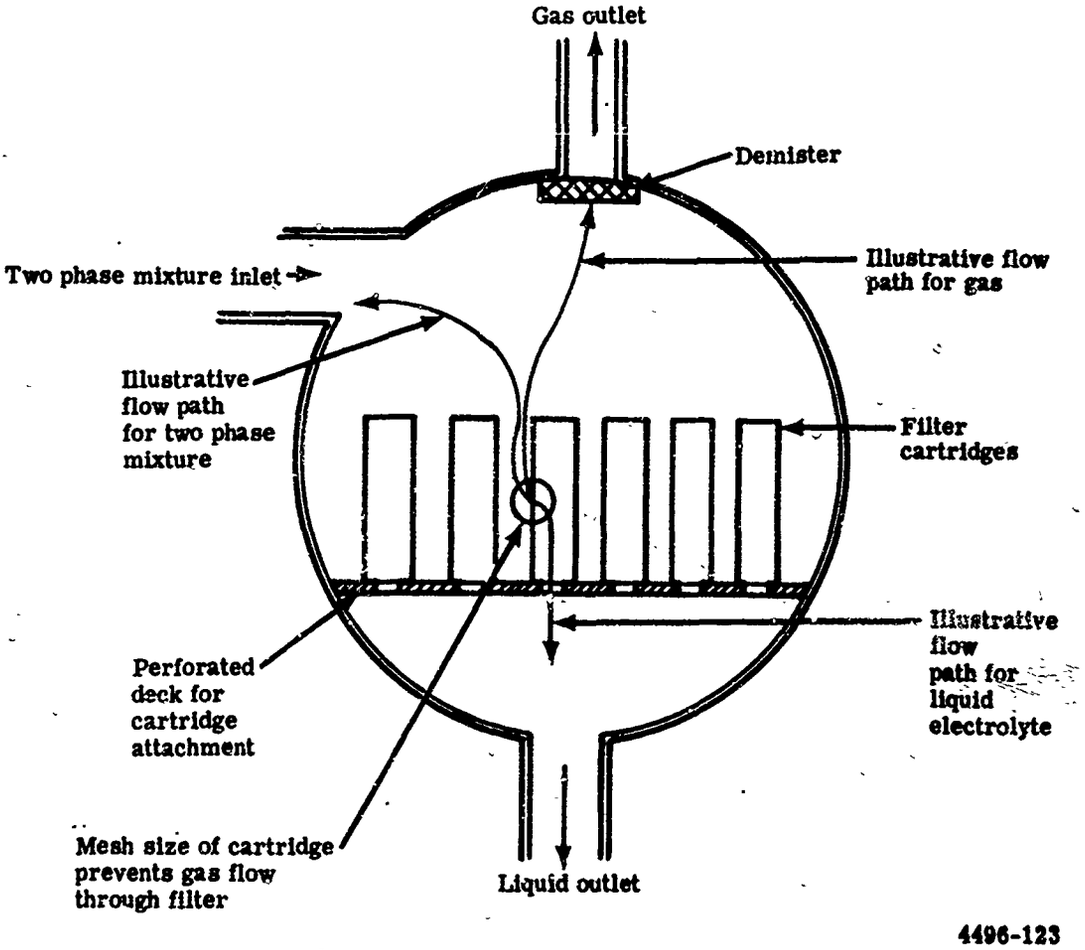


Figure 3.3-25. Separator-reservoir arrangement.

The oxygen-electrolyte and hydrogen-electrolyte are kept separated in different parts of the same tank or in separate tanks. In those cases where one tank is used, (item 5, Table 3.3-V), the tank is divided into the hydrogen and oxygen regions by a disk perpendicular to the axis of the cylinder. In cases where the mixtures are separated in different vessels (systems 5 through 11), two or more tanks are provided.

Table 3.3-V.

Hydrogen production subsystem liquid-gas separator-reservoir data.

Case	Parameter	Subsystem No. 1	2	3	4*	5	6	7	8	9*	10	11
1	Number of separator-reservoirs per plant	1	1	1	1	3	2	3	3	2	3	3
2	Approximate tank volume—ft^3	35	35	65	28	21	21 & 44	21	21	13 & 27	30	64
3	Separator-reservoir weight, dry, per package—lb	870	870	1575	633	1594	1621	1594	1594	1098	2269	2834
4	Approximate number of filter cartridges	140	140	125	135	240	240	240	240	100	370	500
5	Type of separation tanks**	S	S	S	S	D	D	D	D	D	D	D

*Two packages required. Data tabulated are for one package.
**S—O_2 and H_2 separated in different parts of the same tank.
D—O_2 and H_2 separated in different tanks.

The method of using screens to separate two-phase mixtures has been previously investigated and is reported in Reference 1. Data indicates that separation efficiencies near 100% will be achieved. Further investigation is needed, however, before data suitable for detailed design purposes are available.

Pumps

Pump data for each of the 11 subsystems are given in Table 3.3-VI.

Electrolyte Pump

As shown in Table 3.3-VI, most of the systems use one electrolyte pump. In systems 10 and 11, two pumps were used to reduce the quantity of electrolyte handled by each pump. In system 11, because of packaging considerations, one pump handles twice the flow of the other pump. This flow split occurs because the pumps are matched to the liquid-gas separators. Since the hydrogen separator has twice as much flow as the oxygen separator, the pumps were sized accordingly with a resulting simplification in piping.

As shown in item 2, lightweight canned rotor pumps are used on systems 1 through 4. For the systems with larger flow rates where compact, lightweight canned rotor pumps are not currently in production, centrifugal pumps with balanced seals are used. These latter pumps

Table 3.3-VI.

Hydrogen production subsystem pump data.

Item	Parameter	Subsystem No. 1	2	3	4*	5	6	7	8	9*	10	11
	Electrolyte pump data											
1	Number of pumps required	1	1	1	1	1	1	1	1	1	2	2
2	Type of pump	CR**	CR	CR	CR	C***	C	C	C	C	C	C
3	Nominal pressure rise—psi	15	15	20	15	20	20	20	20	20	25	30
4	Flow rate—gpm	1450	1450	1240	1400	2380	2340	2380	2380	1890	3630	4900
5	Pump power—kw	14.6	14.6	16.6	14.1	31.8	31.3	31.8	31.8	25.3	60.6	98.2
	Feedwater pump data											
6	Number of pumps required	1	1	1	1	2	2	2	2	2	2	3
7	Type of pump	P****	P	P	P	P	P	P	P	P	P	P
8	Nominal pressure rise—psi	295	295	295	295	295	295	295	255	295	295	295
9	Flow rate—gpm	2.4	2.4	2.5	1.2	4.8	4.6	4.6	4.6	3.7	7.5	10.7
10	Pump power—kw	0.55	0.55	0.55	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1

*Two packages required. Data tabulated are for one package.
**CR denotes single-speed, canned-rotor, centrifugal pump.
***C denotes single-speed, centrifugal pump.
****P denotes piston pump.

employ a small amount of feedwater to balance the static pressure of the electrolyte. By maintaining the water at a slightly higher pressure than the electrolyte, any leakage through the seal would be into the electrolyte. At the same time, since the fluid wetting the seal is water, vendors indicate that the possibility of abrasive KOH crystal growth on the seal surfaces is eliminated.

Feedwater Pump

A positive displacement piston type pump is used to supply feedwater. The pump will draw water from the purifier storage tank and supply it at the operating pressure of the electrolysis plant. During normal operation, the pump will operate intermittently because the flow of feedwater will be governed by high and low level limits in the liquid-gas separator. A positive displacement pump was selected to prevent backflow when the pump is turned off.

As presented in item 8, a nominal pressure rise of 295 psi is required (Figure 3.3-10). All the pumps used, however, have the capability of providing higher pressures if required.

As presented in item 6 of Table 3.3-VI. systems 5 through 11 use more than one pump. In the multiple pump systems, it is planned that the pumps be driven by one motor with the pumps arranged in double or triple configurations. Therefore, the two or three pumps, as the case may be, will be operated as one.

Electrical Distribution System

A tabulation of data pertaining to the electrical distribution system is presented in Tables 3.3-VII and 3.3-VIII. Further information of the major electrical components is presented in the Power Conditioning section.

Table 3.3-VII.

Hydrogen production subsystem electrical distribution system data.

Item	Parameter	Subsystem 1	2	3	4*	5	6	7	8	9*	10	11
	Rectifier data											
1	Design point voltage—volts dc	593	593	603	585	588	588	588	588	582	582	575
2	Design point current—ampere	558	558	364	744	693	696	693	693	848	843	1055
3	Weight—lb	300	300	449	592	1576	1581	1578	1576	1128	2304	3168
	Bus data											
4	Weight—lb	270	270	423	193	565	567	565	565	340	1125	1679
	Switch gear											
5	Number of contactor sets	4	2	4	4	6	2	6	6	6	8	14
6	Weight—lb	157	75	192	157	299	135	299	299	264	610	680

*Two packages required. Data presented are for one package.

Table 3.3-VIII.

Hydrogen production subsystem electric load distribution—kilowatts.

Load	Subsystem 1	2	3	4*	5	6	7	8	9*	10	11
Modules	2,643.1	2,647.6	2,630.1	2,601.2	5,298.7	5,339.1	5,298.7	5,298.7	4,432.3	8,849.5	13,341
Fans	4.1	0	4.5	7.7	20.2	0	20.2	20.2	15.7	48.3	51.3
Electrolyte pump	14.6	14.6	18.6	14.1	31.8	31.3	31.8	31.8	25.3	60.6	98.2
Feedwater pump	0.55	0.55	0.55	0.55	1.1	1.1	1.1	1.1	1.1	1.1	2.2
Rectifier losses	13.3	13.3	13.2	13.1	26.6	26.8	26.6	26.6	44.5	44.5	66.7
Total	2,675.6	2,676.0	2,665.0	2,636.6	5,378.4	5,388.3	5,378.4	5,378.4	4,518.9	9,004.0	13,559

*Two packages required. Data presented are for one package.

The module busses used are 3 in. × 0.25 in. aluminum bars. Two bus bars connect the rectifier to each module bank. The bus bars are connected at opposite ends of each bank. Hence, there are two busses for each module bank. In addition, there are two intermediate connecting busses joining the three modules in each bank. In areas where the bars are exposed to personnel, they are guarded by a plastic shroud. In other areas, the bars are unprotected. For structural strength, the bars are bound together into bundles and insulated from one another by plastic. Wherever possible, the bars are mounted with the 3 in. dimension positioned vertically to promote heat dissipation by natural convection.

The switch gear entry in Table 3.3-VII represents motor starters. The number of separate starters used in each system is tabulated in item 5. When mounted in the plants, all of the starters are located in the same compartment (item 4, Figure 3.3-11) and are actuated from the control cabinet.

Piping and Valves

The piping and valve weights for each of the 11 hydrogen subsystems are presented in Table 3.3-IX.

Table 3.3-IX.
Hydrogen production subsystem piping and valve weight.

System No.	1	2	3	4	5	6	7	8	9	10	11
Weight (wet) —lb	1315	1245	2189	1228	2926	2791	2926	2926	1506	4856	5781

As an illustration of a typical piping arrangement, consider system 3 shown in Figure 3.3-13. The hydrogen and oxygen leave each module bank in separate pipes. The two streams emerge from each bank at the position indicated by arrows 13 and 14; two check valves are located at these points. It is the purpose of these valves to prevent backflow into the bank, if the bank should develop a serious leak. Downstream from the check valves, the two streams join a vertical manifold pipe. These pipes in turn join separate horizontal collection pipes located near the top of the plant. In these pipes, the two streams are transported to the separator-reservoirs. Manual shutoff valves are located immediately upstream from the separator tank. These valves are closed only during shutdown to prevent the transfer of electrolyte from the modules to the separator. Such a transfer could occur if the plant was sharply tilted or if the check valves opened during transit by sudden changes in acceleration.

The electrolyte is collected at the bottom of the separator and transferred to the heat exchanger. Immediately downstream of the heat exchanger, a check valve is located. This valve prevents flow reversal during normal operation. In addition, it prevents electrolyte level equilization in the modules and the separator-reservoir during shutdown.

The electrolyte pump is located downstream from the check valve. The pump discharges into a horizontal manifold pipe. At intervals along this pipe, vertical distribution pipes carry the flow to the module banks. Immediately upstream of each bank a manual shutoff valve is positioned. In the event of a leak, the shutoff valve would be used to isolate the defective bank from the rest of the system. The three modules in each bank are hydraulically connected in series as previously stated.

The piping in the hydrogen subsystem is stainless steel. Flexible hangers are used to accommodate strain in the system resulting from temperature changes or shock loads.

Structure

The estimated structure weight for each of the 11 systems appears in Table 3.3-X.

Table 3.3-X.
Hydrogen production subsystem structure weight.

Subsystem No.	1	2	3	4	5	6	7	8	9	10	11
Weight—lb	3215	3120	5550	3190	5000	4850	5290	7700	5890	8125	9250

Each of the subsystems is mounted on a skid made of "I" beams and channels. Above the skid, a rigid superstructure is used to house the electrolyzer modules and support auxiliary equipment (Figure 3.3-11). Similar superstructure is used on the other plants. As shown in Figure 3.3-11, the modules in each bank are positioned within a structural network. When maintenance is necessary, the modules in a given bank may be disconnected and pulled out. Once in position, the modules are clamped in place.

The weight of the heat exchanger is supported directly by the skid and the vertical columns located at the end of the structure. The separator-reservoir tank is hung between vertical columns and can move slightly to accommodate the flexing of the piping.

Because of its numerous interconnecting members, the superstructure stiffens the entire structure and helps the skid support the weight of the equipment. To minimize weight, aluminum members are used wherever possible.

Design Summary

A weight breakdown for each of the 11 subsystems is presented in Table 3.3-XI. The dimensions of each subsystem are presented in Table 3.3-XII. In addition, Table 3.3-XII indicates which hydrogen packages contain extra equipment such as the overall plant control cabinet and transformer. As nearly as practical, each subsystem design contains the number of modules indicated by the cost optimization analysis (see Investigation).

PREDICTED PERFORMANCE

The design point power consumption and hydrogen production rates are presented in Table 3.3-XIII.

Table 3.3-XI.

Hydrogen production subsystem weight tabulation.*

Item	Weight (pounds)										
Subsystem design number	1	2	3	4**	5	6	7	8	9**	10	11
Electrolyzer	26,210	26,210	41,417	18,765	41,160	41,160	41,160	41,160	27,442	55,012	64,084
Heat exchanger, including fans and motors	1,555	386	1,290	1,767	3,443	694	3,443	3,443	2,945	5,860	7,611
Separator-reservoir	870	870	1,575	633	1,594	1,621	1,594	1,594	1,698	2,259	2,834
Electrolyte pump	615	615	580	595	2,132	2,106	2,132	2,132	1,797	2,930	3,863
Transformer	-	-	-	7,400	-	-	3,200	22,200	20,200	-	-
Rectifier	300	300	449	592	1,576	1,581	1,576	1,576	1,126	2,304	3,168
Bus bars	270	270	423	193	565	567	565	565	340	1,125	1,670
Structure	3,215	3,120	5,550	3,190	5,000	4,850	5,290	7,700	5,880	8,125	8,250
Piping and valves	1,315	1,245	2,189	1,228	2,926	2,791	2,925	2,926	1,506	4,856	5,981
Switch gear	157	75	192	157	299	135	299	299	264	610	680
Feedwater pump	50	50	50	50	60	60	60	60	60	60	75
Miscellaneous (cables, control and safety equipment)	443	549	635	430	595	595	595	595	540	710	785
Total hydrogen subsystem package weight—lb	35,000	33,700	54,350	35,000	59,350	56,160	62,840	84,250	63,210	83,960	100,000

*Weights included electrolyte inventory which typically is 30 to 35% of the subsystem weight.
**Data for subsystem 4 and 9 is for one of two identical packages required.

Table 3.3-XII.

Hydrogen production subsystem package dimensions.

Parameter	Subsystem dimensions (ft-in.)										
	1	2	3	4**	5	6	7	8	9**	10	11
Package length	14	11-10	16-1	14-2	22-7	18-5	23-6	25	21-6	30-11	35
Package width	8	8	10	8	10	10	10	10	10	12	12
Transformer on skid				*			*	*	*		
Control cabinet on skid					*	*	*	*		*	

*Indicates equipment on skid, all packages have d-c rectifier.
**Two packages required, same dimensions for each skid.

The data tabulated for subsystems 4 and 9 represent totals for both hydrogen skids.

From the specific power column in Table 3.3-XIII, it can be seen that the power required to produce each pound of hydrogen rises as the size of the subsystems is increased. This trend is due to the higher current densities at which the larger subsystems operated. The greater current densities are related to lower process efficiencies dictated by cost optimization studies. In subsystem 3, the increased package weight limit was translated into higher process efficiency through the use of more module banks.

Table 3.3-XIII.
Hydrogen subsystem power consumption and production rates.

Subsystem No.	Power* input (kw)	Hydrogen output (lb/hr H_2)	Specific power (kwh/lb H_2)
1	2,675.6	129	20.7
2	2,676.0	129	20.7
3	2,665.0	134	19.9
4	5,273.2	244	21.6
5	5,378.4	250	21.5
6	5,388.3	252	21.4
7	5,378.4	250	21.5
8	5,378.4	250	21.5
9	9,037.8	406	22.3
10	9,004.0	406	22.2
11	13,559.0	585	23.2

*Includes power to electrolyte and feedwater pumps, modules, fans, and rectifier losses (0.5% of power to modules).

DISCUSSION

As shown in Figures 3.3-6 and 3.3-7 at the beginning of this subsection, production rate is in general not a linear function of power or plant weight. As a rule, the output can be increased by either adding more weight or more power. However, the rate of increase in production diminishes as the number of modules is increased. For a given power level, the hydrogen output can be increased by adding more modules. For small plants, the addition of one bank increases the output relatively more than the addition of one bank to a large plant of the same electrical capacity.

The proposed liquid-gas separator-reservoir design appears to offer a satisfactory solution for the problem of extracting the product gases. However, detailed performance data is not presently available. Further work is required to confirm the suitability of the design.

Two different types of electrolyte pumps were used in the plant designs. Because of the good reliability and operating experience gained on the electrolyzer module development,[2] canned rotor pumps are preferred. However, because compact lightweight canned rotor pumps for relatively large flow rates are not currently available, centrifugal pumps were used on the larger systems. To make the centrifugal pumps more suitable for this application, the pump seal would be balanced with feedwater to prevent electrolyte leakage and inhibit crystal growth.

When the lightweight canned rotor pumps currently being developed for larger flow rates are available, they will be used on subsequent designs.

As presented in Table 3.3-II, most subsystems are on one package which weighs less than the maximum allowable for the applicable mobility class. Each plant was designed as indicated in the Investigation section. In several cases, package weights in excess of the allowable limit were indicated. In these cases, the possibility of using additional skids was investigated. Subsystems 4 and 9 use two identical packages. However, the additional package results in significantly more interconnections which increased the time required for start-up and shutdown. In other cases, the number of module banks was reduced and the package weight was decreased. The number of banks removed was determined from packaging studies and the weight limit.

The total weight of the subsystems could be reduced by using larger electrolyzer modules. The module configuration that appears to result in a minimum weight plant is shown in Figure 3.3-24(a). Whether such a module could be built and operated satisfactorily is not yet clearly known. However, the potential gain appears attractive and the concept warrants further study.

The use of heaters to prevent electrolyte freezing in extremely low temperatures can be avoided by increasing the KOH content of the electrolyte. An increase from 27% to 30% depresses the freezing point below -65°F and results in an insignificant weight change.

CALCULATIONS

See Appendix A for calculations and sample computer run.

REFERENCES

1. Schade, C.W., et al. Final Report Energy Depot Electrolysis Systems Study. Allison Division, General Motors. EDR 3714, Vol. I. June 1964.

2. Schade, C.W., et al. Final Technical Report Water Electrolyzer Module Development. Allison Division, General Motors. EDR 4610. (in publication).

3. Schade, C.W. and Thomson, B.N. Interim Technical Report Water Electrolyzer Module Development. Allison Division, General Motors. EDR 4447. November 1965.

Water Purification Subsystem

DESCRIPTION

The water purification subsystem purifies the raw water obtained on site to provide makeup water for the electrolyzer in the hydrogen production subsystem. The raw water must be purified to minimize buildup of impurities in the closed electrolyte loop, to prevent electrode contamination, and to maintain electrolyzer efficiency.

A flow schematic of the vapor compression process[1] selected for water purification is shown in Figure 3.4-1. The unit operates by evaporating pure steam vapor from impure water by the application of heat. The pure steam vapor is withdrawn from the evaporation chamber, pressurized to increase its condensation temperature, and then condensed on the outside of the evaporator tubes. Raw feedwater is heated as it is pumped through the inside of the tubes of the vertical tube evaporator using the latent heat given up by the steam condensing at a higher temperature and pressure on the outside of the tubes. The evaporator chamber is operated at subatmospheric pressure, so part of the heated feedwater flashes to vapor as it leaves the tubes. Since these units operate at high vacuum, operating temperatures are low. The low temperature eliminates scale formation and also reduces corrosion and heat losses to the atmosphere. The hot distilled water and the blowdown are pumped through the feedwater heater where most of their heat is recovered in preheating the feedwater. Thus, the feedwater is nearly at the boiling point when it reaches the evaporator. A small side stream of cold feedwater is diverted and used to condense additional pure water from the vent gas stream. The vapor compression unit operates continuously and, therefore, needs a constant supply of raw feedwater. The unit produces a continuous output of pure distilled water and blowdown or concentrated solution. The subatmospheric pressure is maintained by using a vacuum pump located in the vent gas line. For start-up, or for feedwater temperatures below about 75°F, immersion heaters are used to supply additional required thermal energy to the unit.

Systems 2 and 6 are water-cooled and therefore require a supply of cooling water. It was considered unnecessary and impractical to purify the cooling water. However, the cooling water requires filtration to remove solid particles. Consequently, a simple sand bed water filtering unit was devised. The unit consists of a pump, two sand bed filters, and suitable valves and connecting piping.

INVESTIGATION

Several water purification techniques were examined and compared, viz, reverse osmosis, ion exchange, regenerative distillation, electrodialysis, and vapor compression. The comparison was made using criteria germane to the Energy Depot, e.g., it was considered advantageous for a process to have a low power demand and capital cost and to be lightweight, compact, reliable, simple, and rugged.

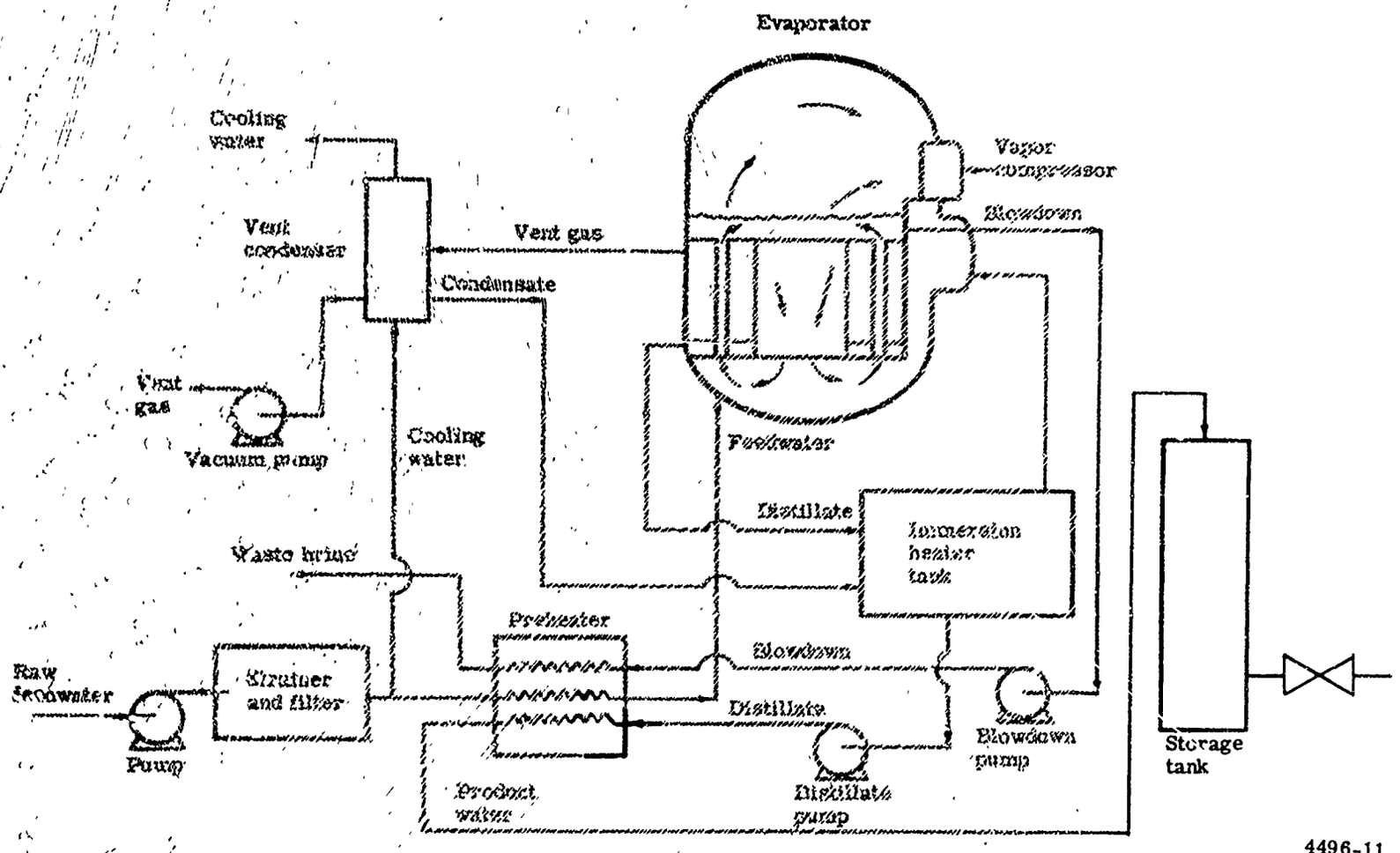


Figure 3.4-1. Schematic of vapor compression water purification process.

Allison

With the exception of the reverse osmosis technique, the water purification processes studied have been tested and proved acceptable. There are several prominent companies that regularly manufacture and supply water purification systems. To obtain data for the present Energy Depot applications, leading manufacturers of systems for each process were contacted. The data supplied by these manufacturers were then studied and compared to select the one most advantageous for the Energy Depot. A general description of each of the major water purification processes studied is presented in the following paragraphs along with the process data supplied by the vendors. All the systems are supplied with sand bed filters to strain and filter the raw feedwater before delivery to the purification subsystem to remove solid particles suspended in solution.

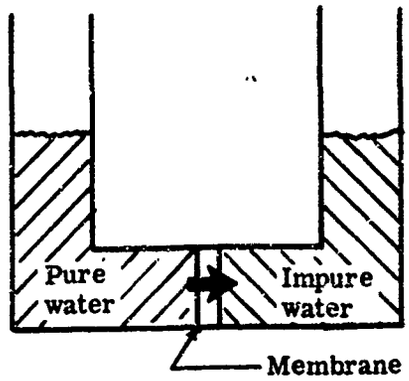
Reverse Osmosis

The scientific principle of osmosis is used to describe the spontaneous flow of water from a dilute solution through a membrane into a concentrated solution. When fluids of different concentrations are separated by a membrane, the dilute solution will flow through the membrane into the concentrated solution until an equilibrium is established. The pressure difference between the two liquid levels is known as the osmotic pressure. If a pressure in excess of the osmotic pressure is applied to the concentrated solution, the fluid will flow in the opposite direction, i.e., from the concentrated solution to the dilute solution. This process is known as reverse osmosis.[2] The phenomena are shown schematically in Figure 3.4-2.

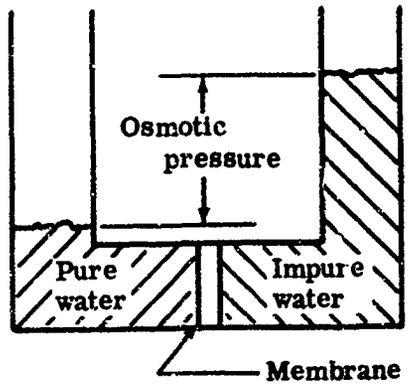
The key to the successful use of reverse osmosis is the separating membrane. Membranes for water purification are currently fabricated of modified cellulose acetate with aqueous magnesium perchlorate and acetone. In one process, a resin bonded continuous Fiberglas filament tube is used with the membrane contained as an integral layer of the tube. A schematic of a typical reverse osmosis water purification scheme is shown in Figure 3.4-3. Impure feedwater is introduced into the tube at several hundred pounds pressure. The high pressure forces pure water through the membrane and Fiberglass support tube. The waste water leaving the tube can be passed through a turbine where part of its energy will be recovered. The product water from one stage is used as feed to the next stage with as many stages included as required to meet the final water purity specifications. In general, the higher the feedwater salinity, the greater the number of stages needed to reach a given purity.

The reverse osmosis process was investigated through correspondence with the leading company in this field--Havens Industries, San Diego, California. Unfortunately, the reverse osmosis process is still in the development stage and it was predicted that production models would not be available for two to five years. Nonetheless, the data obtained suggest that the reverse osmosis process exhibits many advantages over other conventional water purification techniques. For example, the power required to produce 135 gph of 1-ppm purity water from 5000-ppm feedwater is approximately 3 kw. This is about one-half the power required for the lowest

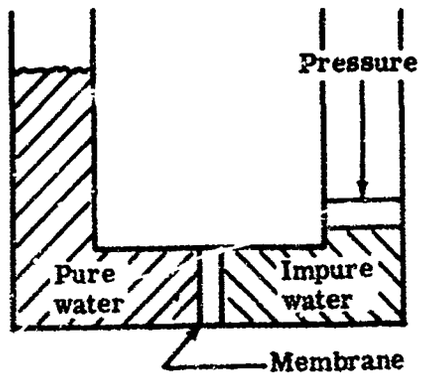
Osmosis – When fluids of different concentrations are separated by a membrane, the dilute solution will flow through the membrane into the concentrated solution.



Osmotic pressure – The level of the dilute solution drops and the level of the concentrated solution rises until an equilibrium is reached. The pressure difference between these two levels is the osmotic pressure.



Revenue osmosis – If a pressure in excess of the osmotic pressure is applied to the concentrated solution, the flow is reversed from the concentrated solution to the dilute solution. This is reverse osmosis.



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Figure 3.4-2. Osmosis principles.

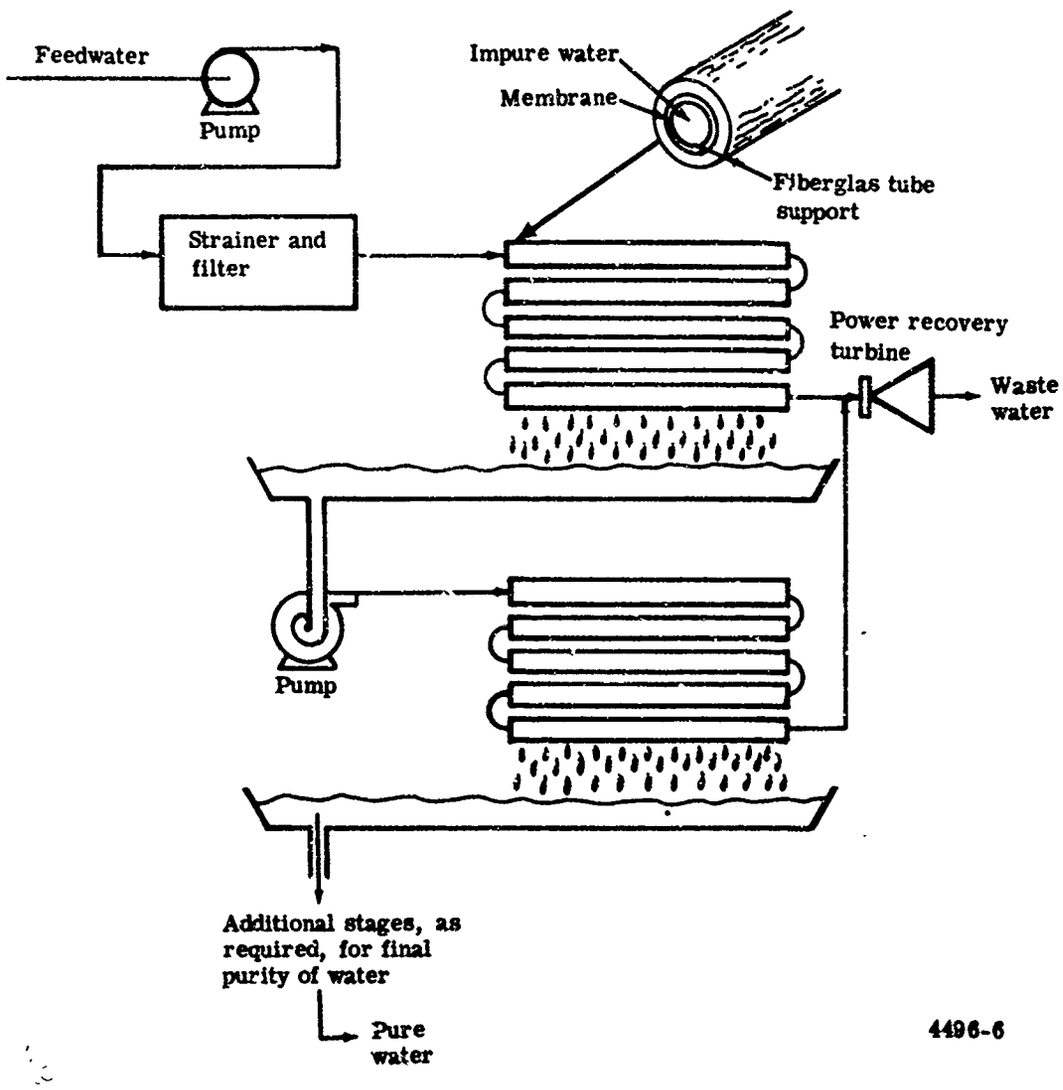


Figure 3.4-3. Schematic of reverse osmosis water purification process.

power-consuming distillation technique—vapor compression. Although firm data are not yet available, it is believed that commercial reverse osmosis plants will be considerably smaller, lighter, and lower in capital cost than other water purification plants currently envisioned. Reverse osmosis plants are conceptually simple and rugged, but the reliability of the process remains to be demonstrated, particularly the lifetime of the membranes.

In view of the developmental nature of the reverse osmosis process, sufficient data are not available to consider it for the Energy Depot. However, the process has not exhibited any characteristics that would make it unsuitable for Energy Depot applications. Since it is probable that the process will be sufficiently advanced by the time a prototype Energy Depot is constructed, it is recommended that it be reevaluated at that stage of the program.

Ion Exchange

Ion exchange processes involve producing pure water by removing salt and mineral impurities through ion substitution or exchange. This is possible since salt and mineral impurities dissolved in water are ionized. The investigation of ion exchange water purification processes was performed with the aid of the Rohm and Haas Co, Philadelphia, Pennsylvania. Their Kunin desalination process is reported to exhibit more advantages than any other ion exchange process currently available.[3] Consequently, the major emphasis was placed on this process.

The basic chemistry of the Kunin process is shown schematically in Figure 3.4-4 using NaCl as an example of the impurity. The first exchanger contains a weakly basic resin which exchanges chloride, sulfate, and nitrate ions for bicarbonate ions and also reduces the organic content of the water. The second exchanger contains a weakly acidic resin which exchanges the incoming cations (e.g., Na^+) for hydrogen ions. The net chemical result of these two columns is the production of carbon dioxide and water. The third exchanger contains a second weakly basic resin which removes the carbon dioxide. The product from the third exchanger is pure water.

When the exchanger resins are nearly exhausted, they are regenerated chemically as follows. The feedwater supply is either turned off or diverted to a second array of exchangers. Ammonia is passed through the first exchanger where the following reaction occurs:

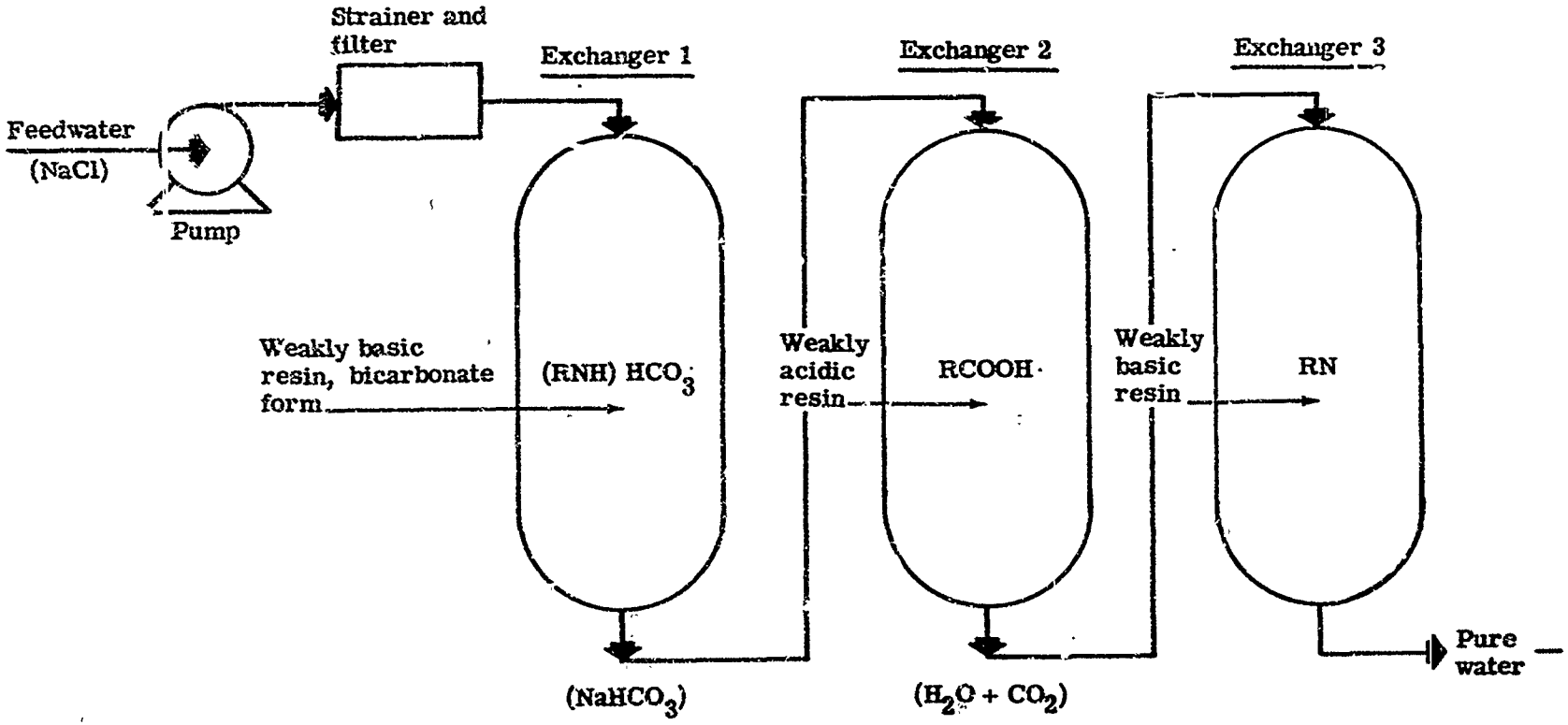
$$RNHCl + NH_3 = RN + NH_4Cl$$

The ammonium chloride is flushed from the exchanger leaving the regenerated RN resin. In the second exchanger, sulfuric acid is introduced to give the following reaction:

$$2\ RCOONa + H_2SO_4 = 2\ RCOOH + Na_2SO_4$$

The sodium sulfate is flushed from this exchanger leaving the regenerated RCOOH resin. Theoretically, no regeneration should be required in the third exchanger since the resin is converted to the form shown in Exchanger 1 ($RNHHCO_3$) during normal operation. In practice, however, the conversion does not proceed to completion and it is usually necessary to add some makeup carbon dioxide to finish the reaction.

At this point, the resins have been regenerated, but they are now in reverse order, i.e., Exchanger 3 contains the resin form previously present in Exchanger 1, and vice versa. Consequently, for the next ion exchange cycle, the feedwater is introduced into this third exchanger and the product water leaves the first exchanger. The complete ion exchange-regeneration cycle is then repeated in the alternate fashion outlined. For the continuous operation required in the present application, at least two exchanger vessels of each type are required so that one is undergoing the regeneration step while the other is on stream.



R = organic molecule

Exchanger 1: $(RNH)HCO_3 + NaCl = RNH\,Cl + NaHCO_3$

Exchanger 2: $RCOOH + NaHCO_3 = RCOO\,Na + H_2O + CO_2$

Exchanger 3: $RN + CO_2 + H_2O = RNHHCO_3$

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Figure 3.4-4. Schematic of Kunin ion exchange water purification process.

Allison

Electrical power is required in the Kunin process to pump the various fluids and to operate the system instruments and controls. For the four Energy Depot plant capacities studied, the power requirement ranges from 2.5 to 11.2 kw.

Resin requirements were determined by the vendor based on the capacity of the resin to remove ions, the desired product water flow rate, and the length of time the resin must be on stream. For these requirements, an 8-hr duty cycle was assumed for the exchangers, i.e., one set operating for 8 hr while a second set undergoes regeneration. Thus, one complete ion exchange-regeneration cycle is carried out during each 8-hr work shift. Therefore, six exchanger vessels are required. For sizing purposes, the vessels were assumed to be 7.5 ft in height. The amount of chemical regenerants needed was obtained assuming that the water purifier will operate 24 hr per day.

A summary of the Kunin ion exchange water purification plant data is given in Table 3.4-I. While the power requirements are attractive, the process does exhibit several disadvantages. The plants are relatively large and heavy and there is the necessity to maintain sizeable inventories of regenerant chemicals, e.g., about 1400 lb/day are needed for the largest plant. The regenerants shown must be supplied daily or space must be provided to store sufficient amounts for several days. In either event, the need to periodically replenish the supply of regenerants imposes an undesirable constraint on the system.

Another disadvantage of the ion exchange method is the fact that this technique is not generally suited to treating feedwater containing more than 3000 ppm impurities. The vendor indicated that the 5000-ppm design condition would be difficult to handle even using the Kunin technique. Data are presented later for other water purification techniques which compare favorably with ion exchange data. Consequently, due to the disadvantages outlined, ion exchange processes were eliminated from further consideration for Energy Depot applications and no additional data were obtained.

Table 3.4-I.
Summary of Kunin ion exchange water purification plant data.

Product water rate (gph)	Power required (kw)	Total plant weight (lb)	Total plant size (ft^3)	Regenerant chemicals needed (lb/day)		
				NH_3	H_2SO_4	CO_2
125	2.5	8,400	150	47	145	55
250	4.6	15,800	280	94	290	110
500	8.6	29,200	550	188	580	220
700	11.2	55,200	770	260	820	310

Regenerative Distillation

In general, regenerative distillation[4,5] schemes use the thermal energy available in leaving product streams to provide heat to incoming feed streams. In this context, the vapor compression technique falls into the regenerative distillation category, but since the product vapors are externally pressurized, it is considered a special case. The two usual examples of regenerative distillation are multiple-effect distillation and multistage flash evaporation. In ordinary or single-effect distillation, water is vaporized and condensed one time. In contrast, multiple-effect distillation repeats the vaporization-condensation cycle several times to reuse the latent heat of vaporization. This essentially multiplies the amount of distillate produced from a given energy input by the number of distillation steps used. These three processes are discussed in the following paragraphs.

Single-Effect Distillation

Initial vendor* data received for single-effect distillation showed that the process consumes enormous quantities of power compared to the other water purification techniques. For the 125-gph system the power demand is 325 kw—higher than for reverse osmosis or ion exchange by more than two orders of magnitude. Consequently, single-effect distillation was eliminated in favor of examining multiple-effect distillation.

Multiple-Effect Distillation

A schematic of a typical multiple-effect distillation process is shown in Figure 3.4-5. Steam is produced at high temperature and atmospheric pressure in Effect 1. This steam condenses in Effect 2 yielding pure liquid water by releasing its latent heat. The release of this latent heat vaporizes another similar amount of water in Effect 2 since the pressure and, consequently, boiling temperature are maintained by pumps (not shown) at a lower level. The cycle repeats through successive effects, with each effect operating at progressively lower pressure and temperature. The vapor produced in the final effect is condensed using cooling water or air in a heat exchanger. A practical limit on the number of effects that can be used is imposed by the need to maintain a finite temperature difference at each step to achieve heat transfer. Data obtained from the vendor—Barnstead Still and Sterilizer Co.—showed that the power required for the multiple-effect process was lower than that for the single-effect process but was still too high. For example, for the 125-gph system, a double-effect process requires about 150 kw, a triple-effect process requires about 110 kw, and a four-effect process requires about 85 kw. In addition, two other serious drawbacks exist according to the vendor. First, it is doubtful if even a triple-effect unit could be constructed small enough to meet the Energy

*American Sterilizer Co. (AMSCO). Erie, Pennsylvania; Consolidated Stills and Sterilizers. Boston, Massachusetts; and Barnstead Still and Sterilizer Co, Boston, Massachusetts.

Depot size constraints. Second, multiple effect units are not normally adapted for electrical heating. As a result, the vendor would be required to undertake a development program to satisfy the Energy Depot criteria. Consequently, multiple-effect distillation processes were eliminated from contention.

Multistage Flash Evaporation

A schematic diagram of a typical multiple-stage flash evaporation process is shown in Figure 3.4-6. With this technique, the impure feedwater is progressively heated and then introduced into a large chamber (Stage 1) where pumps (not shown) maintain the pressure just below the boiling point of the hot feedwater. When the hot feedwater enters this chamber, the reduced pressure causes part of the liquid to flash into steam. The remaining brine is passed through the series of stages at successively higher vacuums where the flash process is repeated at progressively lower temperatures. Progressive heating of the impure feedwater is accomplished by piping the incoming feedwater through the flash chambers, starting at the low temperature end (Stage 3 in Figure 3.4-6). The flashed vapor condenses as it gives up its latent heat to the cooler feedwater. Final heating of the feedwater, before flashing, takes place in an auxiliary heater.

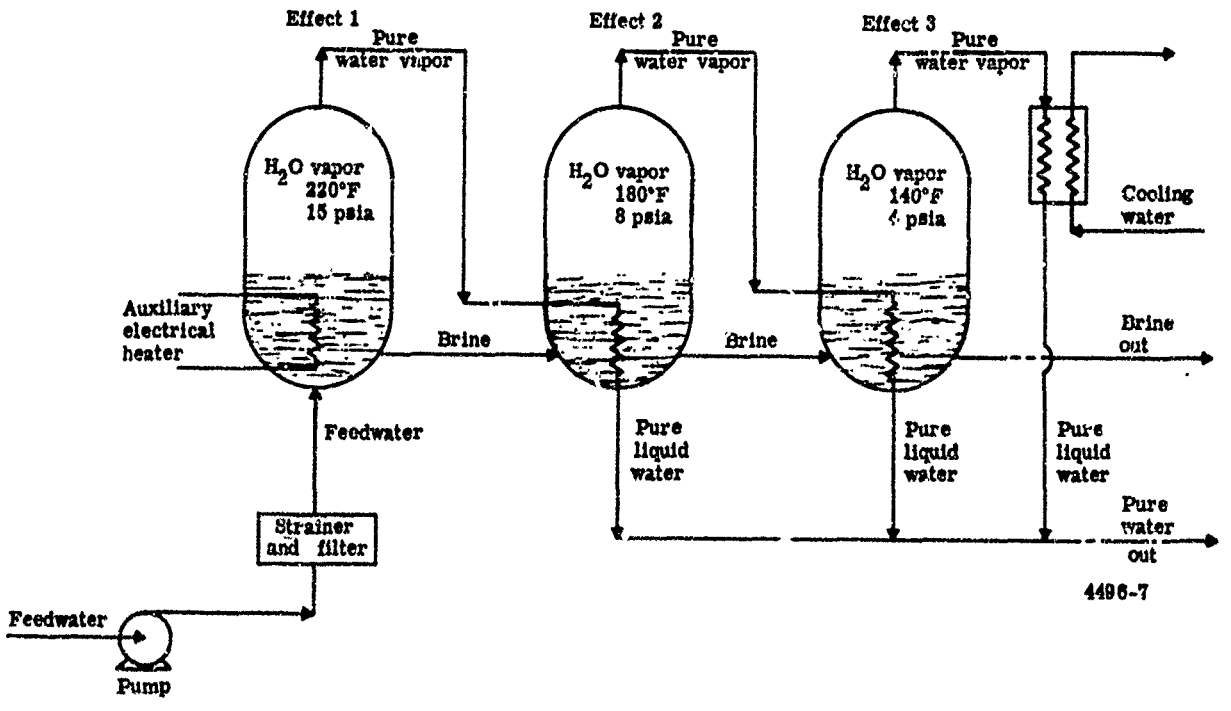


Figure 3.4-5. Schematic of typical multiple-effect distillation water purification process.

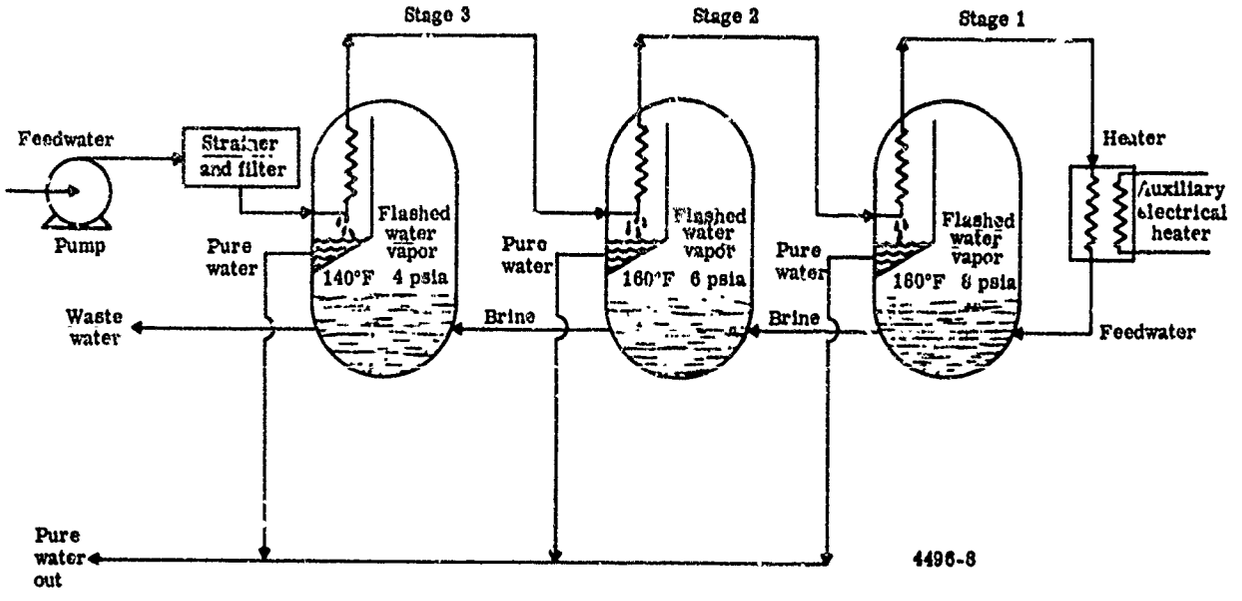


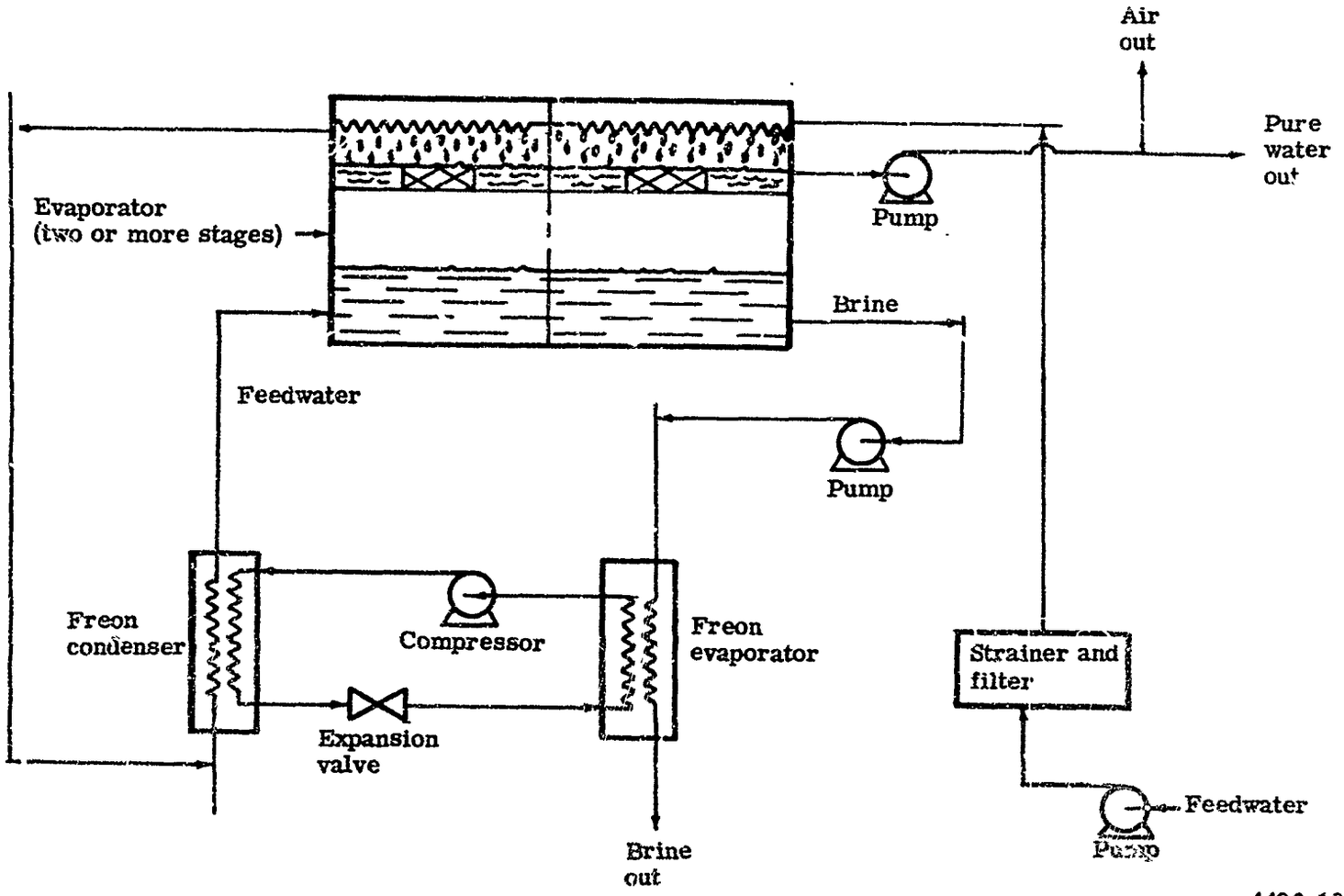
Figure 3.4-6. Schematic of typical multistage flash evaporation water purification process.

The source of data and information for this process was the Maxim Division of the Cuno Engineering Corporation, a subsidiary of the American Machine and Foundry Company, Meriden, Connecticut. The Maxim Division has developed an efficient multistage flash evaporator which uses a heat pump to deliver thermal energy to the feedwater. A schematic of this process for two evaporation stages is shown in Figure 3.4-7. The main difference in this process from other flash evaporation processes is that a closed-loop Freon system is used to transfer heat to the feedwater.

Some pertinent data for the multistage flash evaporation plant are given in Table 3.4-II. These data are for standard commercial units produced by Maxim, with the exception of the weights. Maxim estimated that they could reduce the weights of their commercial units to the levels shown in Table 3.4-II. The plant sizes and weights given in Table 3.4-II are reasonably attractive. However, the power requirements are undesirably high. Also, the chemicals must be supplied to the system daily to remove the scale that builds up on the heat exchange surfaces. For these reasons, the multistage flash evaporation process was eliminated from further consideration.

Electrodialysis

In electrodialysis* electric power removes impurities from the feedwater by forcing the impurities through selective membranes. Salts and minerals dissolved in water are ionized and are free to wander. Consequently, when a direct current is passed through this water, positively charged ions move toward the negative electrode and negatively charged ions move



Figur 3.4-7. Schematic of Maxim multistage flash evaporation water purification process.

Allison

toward the positive electrode. In an electrodialyzer, plastic, semipermeable membranes are inserted into the solution. Half of the membranes are permeable to positive ions and the other half are permeable to negative ions. These membranes are arranged in an alternate array. A schematic of a typical electrodialysis process is shown in Figure 3.4-8. When the electrical potential is applied to the cell electrodes, impurity cations are attracted toward the negative electrode (cathode) and impurity anions are attracted toward the positive electrode (anode). However, the alternate array of selective membranes prevents the ions from traveling the complete distance between electrodes. Thus, impurities are concentrated in one cell while pure water is isolated in the adjacent cells. The products leaving the cells are collected in either a waste water or a pure water manifold, whichever is appropriate. Periodically, the electrode potential is reversed, i.e., the cathode becomes the anode and vice versa to eliminate the deposits that naturally build up on the electrode surfaces. Electrodialysis systems can be designed for either continuous or batch operation.

Table 3.4-II.
Summary of multistage flash evaporation plant data.

Product water rate (gph)	Power required (kw)	Total plant weight (lb)	Total plant size (ft^3)
125	32	1800	110
250	57	2600	150
500	100	3700	190
700	135	4400	220

The electrodialysis water purification system was studied with the cooperation of Ionics, Inc, Watertown, Massachusetts, who supplied the data and specifications. There is considerable latitude in the design of electrodialysis plants to satisfy a given application. For example, if plant weight is not a problem, designs can usually be developed to yield a maximum production efficiency. Conversely, production efficiency generally is lower for plants designed specifically for light weight. Consequently, electrodialysis data were obtained based on compromised design characteristics, i.e., judgment was used to arrive at reasonable plant sizes and efficiencies. In all cases, an attempt was made to stay as close to standard commercial designs as possible to prevent a need for extensive development efforts.

The amount of power required for purifying water by electrodialysis is a function of the temperature of the feedwater and the desired product output rate. Power requirements were obtained for both parameters.

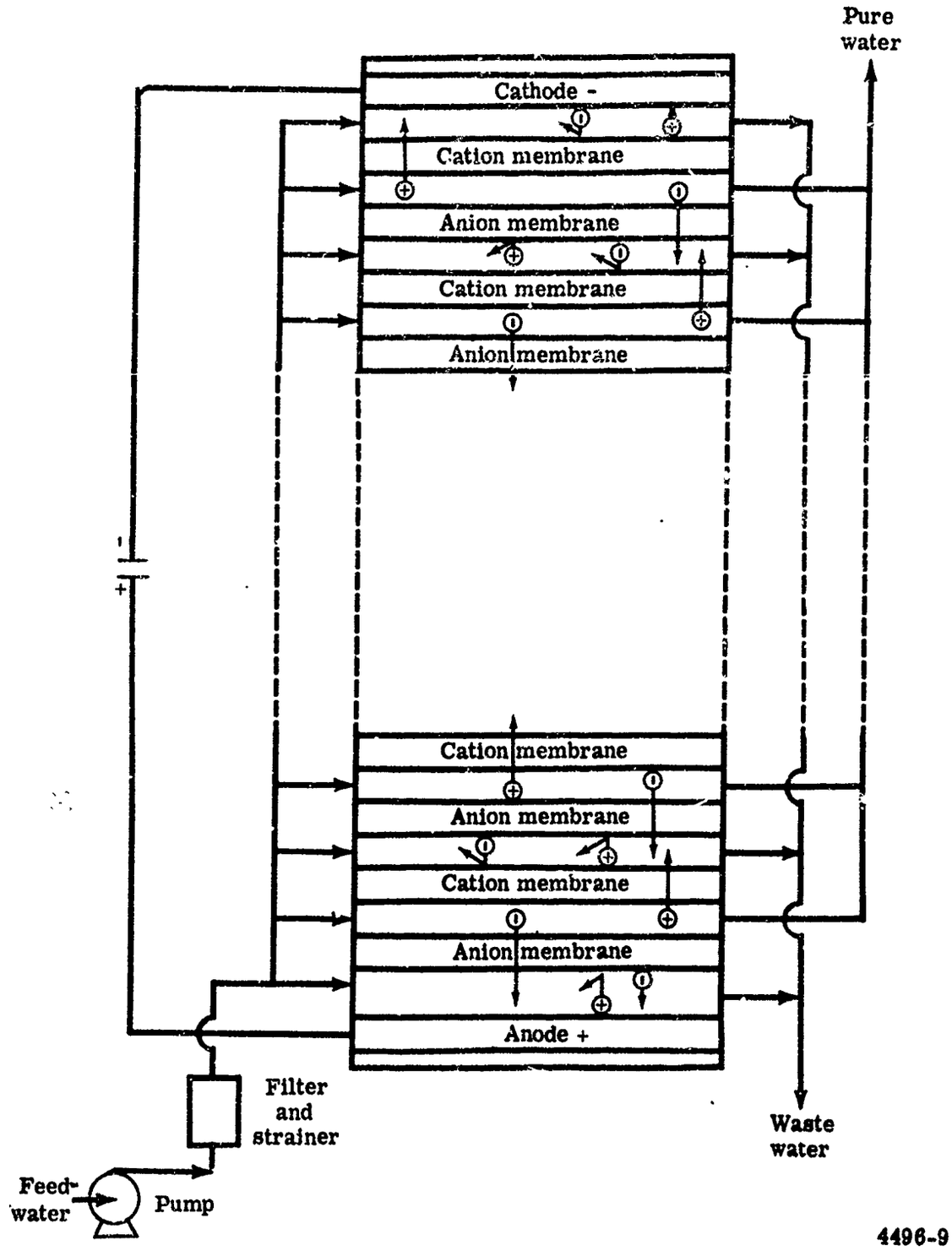


Figure 3.4-8. Schematic of typical electrodialysis water purification process.

Electrodialysis plant size and weight values were obtained from the vendor for two different plant designs—commercial units and a unit developed for Navy shipboard applications. The Navy units were specially designed to be compact and lightweight and appeared to satisfy the Energy Depot size and weight criteria.

A summary of the power, size, and weight data for the two electrodialysis systems studied is given in Table 3.4-III. Comparing these data with the data presented earlier for other water purification processes reveals that electrodialysis has several advantages. The power requirements compare favorably with the lowest power consuming technique currently available—ion exchange. For a 125-gph capacity plant at design conditions, for example, ion exchange requires 2.5 kw (Table 3.4-I) and electrodialysis requires 3.3 kw. The commercial plant sizes are nearly equivalent for the smallest capacity plants. However, as the capacities increase, the electrodialysis unit sizes increase much less rapidly than the ion exchange sizes—e.g., for 700-gph capacity, the commercial electrodialysis system occupies 240 ft^3 compared to 770 ft^3 for the ion exchange system. For the compact electrodialysis system, the discrepancy is even greater. Electrodialysis also exhibits a marked advantage over ion exchange in plant weight. The ratio of ion exchange system weight to electrodialysis system weight ranges from about 10:1 to about 20:1 over the range of capacities studied.

The multistage flash evaporation process (Table 3.4-II) was eliminated based on a data comparison with the electrodialysis process. The two systems do not differ markedly in size, but electrodialysis plants weigh only about one-half as much and consume only about one-tenth as much power as the multistage flash evaporation plants.

Table 3.4-III.

Summary of electrodialysis plant data.

Feedwater temperature (°F)	Product water rate (gph)	Power required (kw)	Plant size (ft^3)		Plant weight (lb)	
			Commercial	Compact	Commercial	Compact
35		5.0				
77	125	3.3	145	45	800	650
105		2.2				
77	250	6.5	150	50	1,100	950
77	500	12.9	185	66	2,050	1,800
77	700	18.0	240	85	2,500	2,200

Feedwater temperature (°F)	Purchase price from vendor ($/unit)						Development and test costs ($)	
	Commercial			Compact				
	1 unit lot	10 unit lots	100 unit lots	1 unit lot	10 unit lots	100 units lots	Commercial	Compact
35								
77	24,000	18,000	13,500	30,000	23,000	17,500	125,000	175,000
105								
77	40,000	31,000	24,000	50,000	38,000	30,000	140,000	190,000
77	67,000	53,000	41,000	84,000	65,000	51,000	170,000	220,000
77	86,000	68,000	54,000	107,000	84,000	66,000	190,000	240,000

Data were obtained simultaneously for all the water purification processes under consideration. As the data were received and analyzed, it became apparent that electrodialysis and vapor compression processes were superior to the other processes investigated. To make the final process choice more valid, therefore, cost information was also obtained from the vendors for these two processes. For convenience, the electrodialysis cost data are included in Table 3.4-III.

As a result of the favorable characteristics discussed, the electrodialysis process was retained for comparison with the vapor compression process.

Vapor Compression

The vapor compression water purification process was investigated using data and information supplied by the Mechanical Equipment Co, Inc, New Orleans, Louisiana. As was the case with electrodialysis, considerable latitude can be exercised in the design of vapor compression plants to satisfy a given application. Accordingly, judgment was used to reach compromises with standard commercial designs. System power requirements are dependent on both the feedwater temperature and the plant capacity. Both parameters were considered in obtaining power data. To conserve weight, aluminum was substituted where possible for other materials normally used in commercial plants. Plant size was reduced by considering stacking and integrating standard components where operation or maintenance would not suffer as a result. As mentioned in the electrodialysis discussion, the need for obtaining cost data from the vendor was also manifested.

The data received from the vendor for the vapor compression process are summarized in Table 3.4-IV. The data show that the compact design affords the greatest saving in plant weights, i.e., the compact designs weigh less than half of the standard commercial designs.

Table 3.4-IV.
Summary of vapor compression plant data.

Feedwater temperature (°F)	Product water rat. (gph)	Power required (kw)	Plant size (ft^3)		Plant weight (lb)		Purchase price from vendor ($/unit)			Development and test costs ($)
			Commercial	Compact	Commercial	Compact	1 unit lot	10 unit lots	100 unit lots	
35		10.3								
77	125	6.8	240	175	5,300	2,250	24,000	13,500	11,500	0
105		5.9								
77	250	12.7	420	300	9,300	3,900	30,000	16,500	14,000	0
77	500	24.2	730	510	16,000	6,800	36,000	20,000	17,000	0
77	700	33.4	950	660	21,000	8,800	40,000	22,000	19,000	0

Electrodialysis plant size and weight values were obtained from the vendor for two different plant designs—commercial units and a unit developed for Navy shipboard applications. The Navy units were specially designed to be compact and lightweight and appeared to satisfy the Energy Depot size and weight criteria.

A summary of the power, size, and weight data for the two electrodialysis systems studied is given in Table 3.4-III. Comparing these data with the data presented earlier for other water purification processes reveals that electrodialysis has several advantages. The power requirements compare favorably with the lowest power consuming technique currently available—ion exchange. For a 125-gph capacity plant at design conditions, for example, ion exchange requires 2.5 kw (Table 3.4-I) and electrodialysis requires 3.3 kw. The commercial plant sizes are nearly equivalent for the smallest capacity plants. However, as the capacities increase, the electrodialysis unit sizes increase much less rapidly than the ion exchange sizes—e.g., for 700-gph capacity, the commercial electrodialysis system occupies 240 ft^3 compared to 770 ft^3 for the ion exchange system. For the compact electrodialysis system, the discrepancy is even greater. Electrodialysis also exhibits a marked advantage over ion exchange in plant weight. The ratio of ion exchange system weight to electrodialysis system weight ranges from about 10:1 to about 20:1 over the range of capacities studied.

The multistage flash evaporation process (Table 3.4-II) was eliminated based on a data comparison with the electrodialysis process. The two systems do not differ markedly in size, but electrodialysis plants weigh only about one-half as much and consume only about one-tenth as much power as the multistage flash evaporation plants.

Table 3.4-III.

Summary of electrodialysis plant data.

Feedwater temperature (°F)	Product water rate (gph)	Power required (kw)	Plant size (ft^3)		Plant weight (lb)	
			Commercial	Compact	Commercial	Compact
35		5.0				
77	125	3.3	[illegible]	45	800	650
105		2.2				
77	250	6.5	150	50	1,100	950
77	500	12.9	185	66	2,050	1,800
77	700	18.0	240	85	2,500	2,200

Feedwater temperature (°F)	Purchase price from vendor ($/unit)						Development and test costs ($)	
	Commercial			Compact				
	1 unit lot	10 unit lots	100 unit lots	1 unit lot	10 unit lots	100 units lots	Commercial	Compact
35								
77	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
105								
77	40,000	31,000	24,000	50,000	38,000	30,000	140,000	190,000
77	67,000	53,000	41,000	84,000	65,000	51,000	170,000	220,000
77	86,000	68,000	54,000	107,000	84,000	66,000	190,000	240,000

Process Selection

Some of the more pertinent data presented previously for the water purification processes are given in Table 3.4-V. Significant plant data are given for each process for a feedwater temperature of 77°F and a water production rate from 125 to 700 gph. As mentioned earlier, reverse osmosis was eliminated because of its development status; ion exchange was eliminated because of its size and requirement for a significant supply of regenerative chemicals; single- and multiple-effect distillation processes were eliminated because of size and power requirements; and multistage flash evaporation was eliminated because of high power requirements and the need for regenerative chemicals. Electrodialysis was retained because of its attractive technical qualities, e.g., low power consumption and size and weight. (See Table 3.4-V.)

Comparing the vapor compression process data with the data for the other processes shows that vapor compression is not outstanding in any single category. However, the power requirements, sizes, and weights are not unreasonably high. At design conditions, the 125-gph electrodialysis plant requires 3.3 kw and the vapor compression plant requires 6.9 kw of electrical power. Thus, the power difference is not large. A comparison of plant sizes and weights for electrodialysis and vapor compression shows that the former exhibits a significant advantage in both areas. Since the weight and size of the vapor compression plant is only a small fraction of the total Energy Depot weight and size, this advantage is marginal.

Comparing the vendor-furnished purchase price data for electrodialysis and vapor compression reveals a significant advantage in favor of the latter. For the compact designs considered, the single unit electrodialysis to vapor compression purchase price ratio varies from 1.25 to 2.68 over the range of plant capacities. These cost values represent only a part of the overall system cost, however, since development and test costs are not included. As shown in Table 3.4-V, the vapor compression process requires no development effort and test costs are included in the purchase price estimates. For the electrodialysis process, however, development effort is required and the development and test costs are estimated to be very substantial, as shown in Table 3.4-V.

Based on the vendor-supplied data, the electrodialysis process exhibits a lower power demand and the vapor compression process exhibits a lower system purchase price. Consequently, a brief trade-off study was made of these two items.

For the 3-Mw_e Energy Depot, the power saved by using electrodialysis rather than vapor compression is about 3.6 kw or 0.12% of the total plant power. The capital cost of the 3-Mw_e nuclear powerplant is estimated to be about $\$3.1 \times 10^6$. Thus, the capital cost saving for vapor compression is about $\$6 \times 10^3 / 3.1 \times 10^6$ or 0.19% of the powerplant cost. For the 15-Mw_e plant, the power saving for electrodialysis is about 0.10%, and the capital cost saving for vapor compression is about 0.83%. Clearly, as the plant capacity increases, the trade-off favors vapor compression. Further, the large difference in development and test costs not included in the trade-off increases the advantage for vapor compression.

Table 3.4-V.

Comparative water purification process data.

Process	Product water rate (gph)	Power required (kw)	Total plant weight (lb)	Total plant size (ft^3)	Total regenerants needed (lb/day)	Purchase price from vendor ($/unit) 1 unit lot	10 unit lots	100 unit lots	Development and test cost ($)	Remarks
Reverse osmosis	125	3								Developmental.
Ion exchange	125	2.5	8,400	130	247					Feedwater may be too concentrated.
	700	11.2	55,200	770	1,390					
Single-effect distillation	125	325								
Multiple-effect distillation										
Two effects	125	150								Doubtful if units can be made sufficiently small.
Three effects	125	110								
Four effects	125	85								
Multistage flash evaporation	125	32	1,800	110						Regenerants required, but amount not determined.
	700	135	4,400	220						
Electrodialysis (commercial)	125	3.3	800	135		24,000	18,000	13,500	125,000	Small ion exchange unit may be required.
	700	18.0	2,500	240		86,000	68,000	54,000	190,000	
Electrodialysis (compact)	125	3.3	650	45		30,000	23,000	17,500	175,000	Small ion exchange unit may be required.
	700	18.0	2,200	85		107,000	84,000	66,000	240,000	
Vapor compression (commercial)	125	6.9	5,300	240		24,000	13,500	11,500	0	Some brackish waters may cause slight scaling.
	700	33.4	21,000	950		40,000	22,000	19,000	0	
Vapor compression (compact)	125	6.9	2,250	175		24,000	13,500	11,500	0	Some brackish waters may cause slight scaling.
	700	33.4	8,800	660		40,000	22,000	19,000	0	

Note: Data shown are for 77°F feedwater temperature.

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Several other important items were considered as part of the vapor compression-electrodialysis comparison, viz, plant simplicity, reliability, ruggedness, and flexibility. On the first three items, both processes appeared to be comparable, but vapor compression is superior with respect to flexibility. The vapor compression process is capable of purifying all water purities from brackish through sea water. Electrodialysis, on the other hand, is apparently never used for highly saline waters such as sea water because of the excessive power requirements.

As discussed previously, ion exchange and regenerative distillation processes require regenerative chemicals. The vapor compression process does not impose this constraint on the system because it operates at sufficiently low temperatures to make scaling virtually nonexistent.

Based on the results of the comparisons outlined, vapor compression was judged to be the most advantageous of the water purification processes studied. Consequently, it was selected for Energy Depot water purification.

DESIGN PARAMETERS

The electrodialysis vendor—Ionics, Inc, Watertown, Massachusetts—is known to have one of the most complete files of worldwide brackish water salinity data in existence. Information was obtained from Ionics which showed that about 90% of the world's inland waters contain less than 3500 ppm total dissolved solids. Hence, to be conservative, the well feedwater salinity was chosen to be 5000 ppm for design purposes.

The purity of the water leaving the purification system was fixed at 1 ppm to prevent the accumulation of deleterious impurities in the water electrolyzer. Standard water purification system designs are not capable of handling this high purification load, and hence, each system was specially designed to meet this specification.

The feedwater was assumed to be available at the water purification skid as a liquid at temperatures from 35 to 105°F and at pressures from 10 to 15 psia.

The water purification subsystem was designed to produce the makeup water at rates of 125 gph for the 3-Mw_e plant, 250 gph for the 6-Mw_e plant, 500 gph for the 10-Mw_e plant, and 700 gph for the 15 Mw_e plant.

It was considered impractical and unnecessary to provide 1-ppm purity cooling water for the Energy Depot designs requiring cooling water. However, provisions were made to filter and strain the cooling water in the water purification subsystem before delivery to the other subsystems.

DESIGN

The specific design information for the vapor compression water purification units was supplied by the Mechanical Equipment Co, Inc, New Orleans, Louisiana. The data developed were based on standard commercial design techniques used for production models. A process schematic and control diagram supplied by the vendor for the four plant capacities is shown in Figure 3.4-9. The schematic nomenclature is given in Table 3.4-VI. The control system is designed for automatic operation. The feedwater is introduced to the preheater at a rate equivalent to four times the product pure water draw-off rate. The feedwater is first heated by the waste water and pure water streams leaving the plant and then delivered to the bottom of the evaporator vessel. The water vapor produced in the evaporator is compressed and sent back into the evaporator where it condenses and evaporates additional feedwater. The condensed pure water flows from the evaporator to the immersion heater tank. If operating conditions require, some of the pure water is vaporized in the immersion heater tank and directed to the evaporator to supply additional heat for evaporating the feedwater. The pure water is withdrawn from the immersion heater tank, pumped through the feedwater preheater, and sent to the storage tank for ultimate delivery to the electrolyzer. The blowdown or waste water is removed from the evaporator chamber, pumped through the feedwater preheater, and discarded. The noncondensed gases are withdrawn from the evaporator chamber and sent through a heat exchanger where the pure water carried over is condensed and sent to the immersion heater tank. The noncondensed gases leaving the heat exchanger are delivered to the vacuum pump which maintains the required pressure in the evaporator chamber.

Table 3.4-VI.
Vapor compression process schematic nomenclature.

	Flowmeter		Check valve
	Pneumatic control valve		Solenoid valve
	Liquid-level control valve		Temperature indicator
	Hand valve		Temperature controller

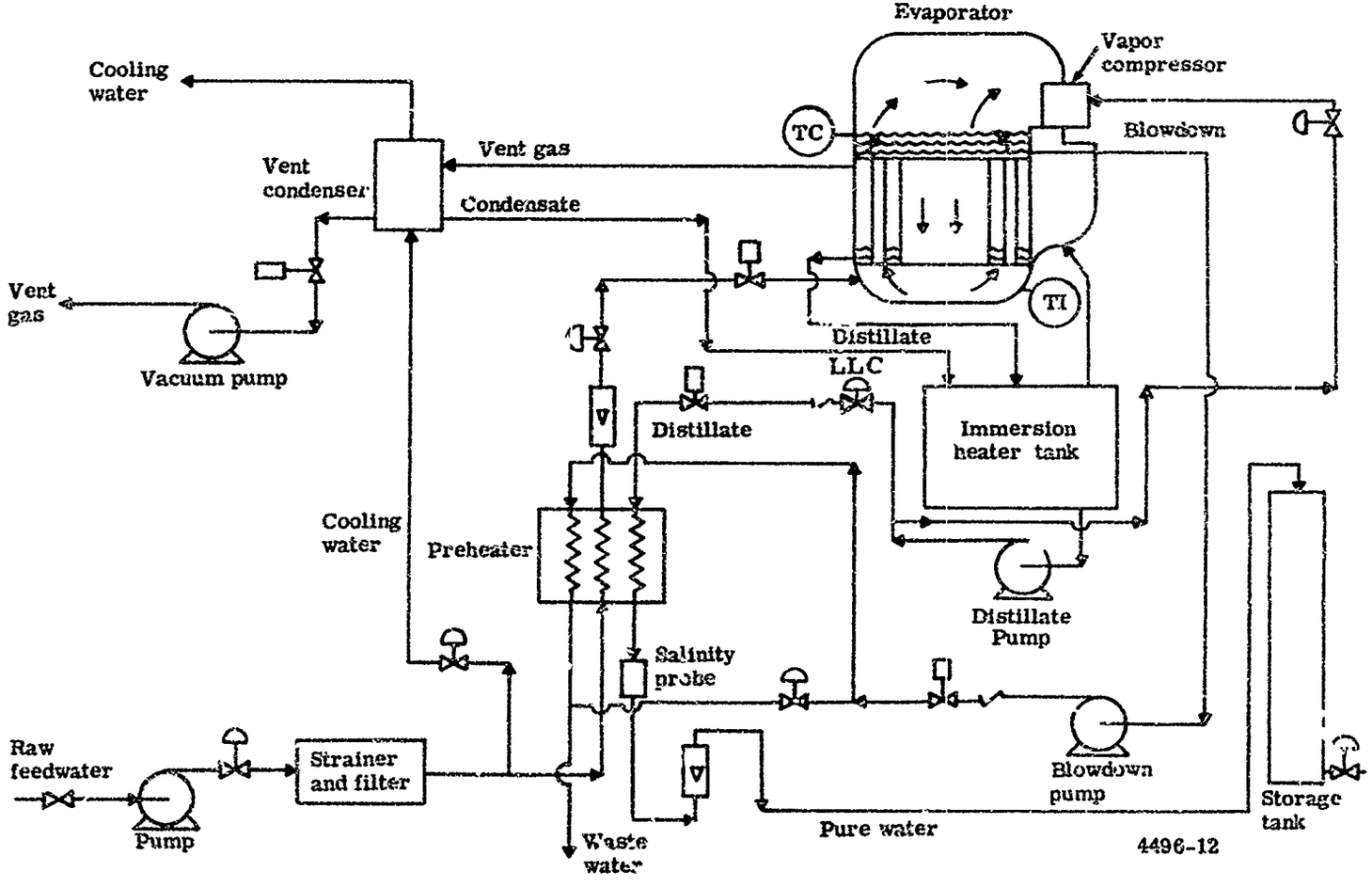


Figure 3.4-9. Process and control schematic of vapor compression water purification process.

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A photograph of a typical vapor compression unit manufactured by the vendor is shown in Figure 3.4-10. This unit is electrically driven and has a nominal product water capacity of 420 gph. A typical plant layout for a similar unit supplied by the vendor is shown in Figure 3.4-11. The overall plant dimensions shown in Figure 3.4-11 are defined in Table 3.4-VII for the eleven sets of Energy Depot design conditions. Table 3.4-VII also gives the plant product water capacities and total weights, including the skid used to assemble the equipment.

The vendor indicated that vapor compression unit cooling loads are minimal and are always handled using water. Introducing air cooling would be of no benefit and would result in slightly increased plant sizes and weights. Consequently, provisions were not made to provide air cooling. The vendor also indicated that 400-cps electrical power could be accommodated readily and no changes in the plant specifications presented earlier would be required.

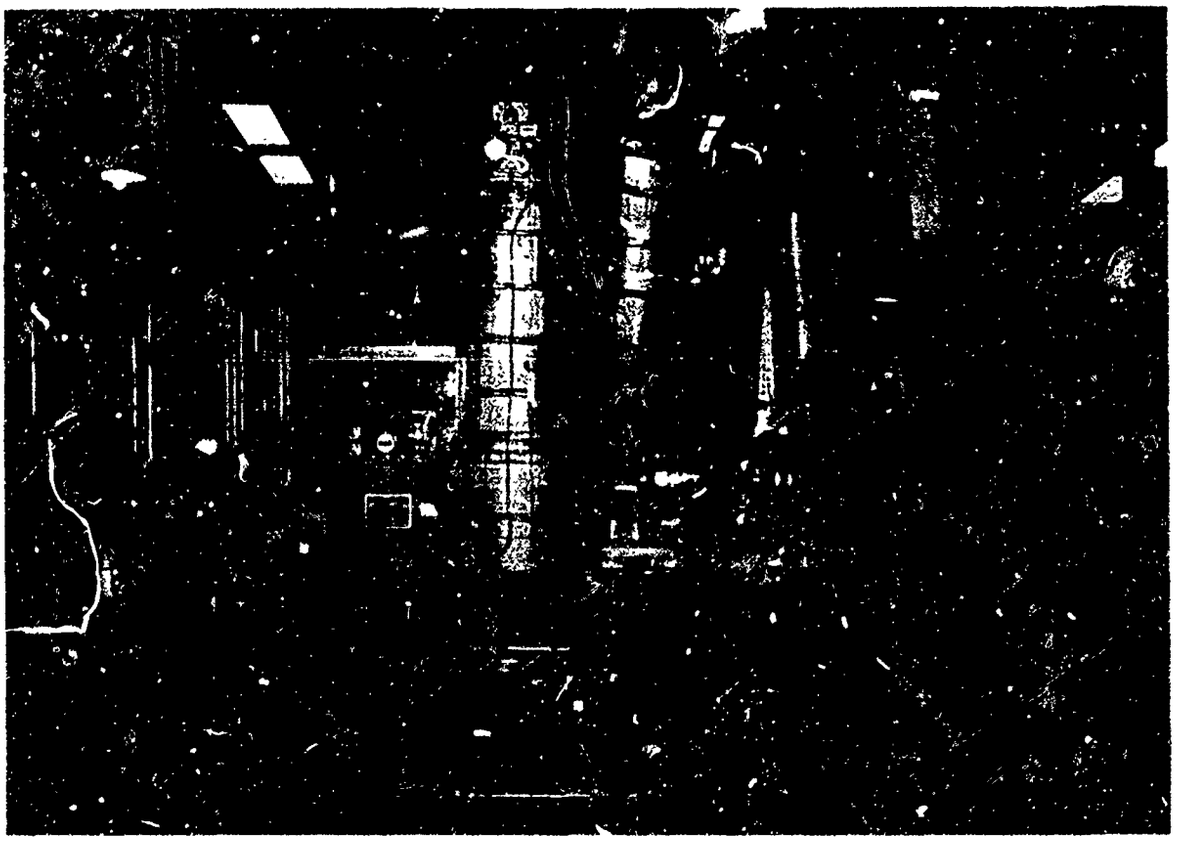
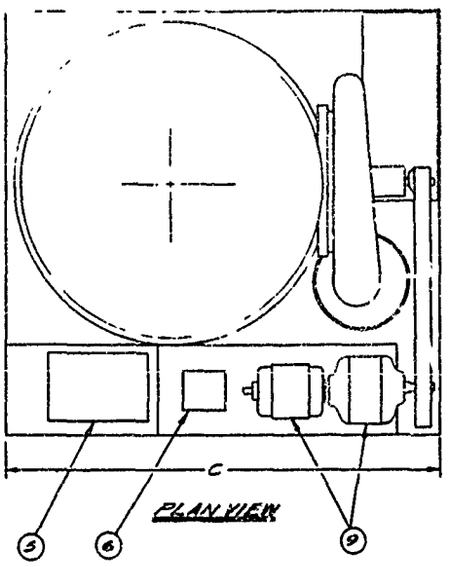
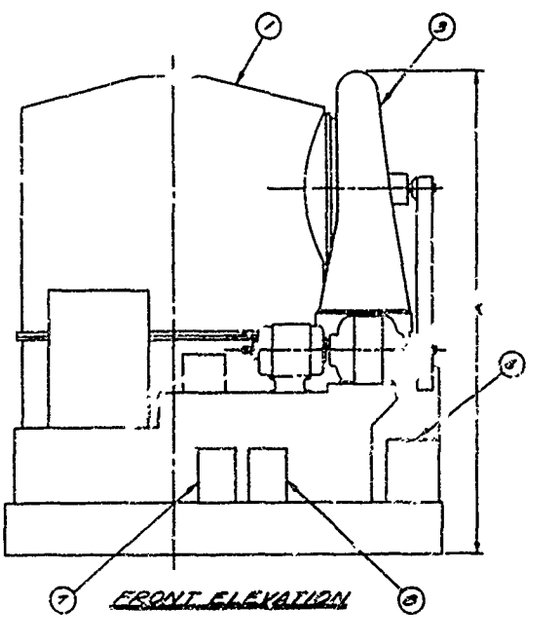


Figure 3.4-10. Typical vapor-compression wate purification unit.



LEGEND
1. EVAPORATOR VESSEL
2. HEAT EXCHANGER
3. COMPRESSOR
4. IMMERSION HEATER TANK
5. VACUUM PUMP
6. FEED-WATER PUMP
7. DISTILLATE PUMP
8. BLOWDOWN PUMP
9. MOTORS



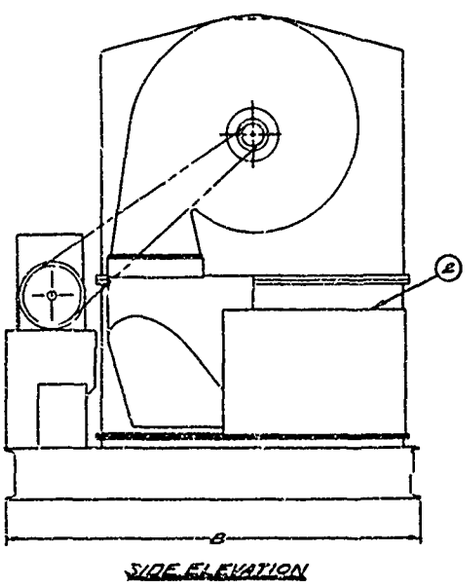


Figure 3.4-11. Water purification subsystem layout.

Table 3.4-VII.
Water purification subsystem vapor compression plant size and weight data.

System	Product water rate (gph)	Mobility Class	Estimated overall dimensions (ft) A. Height	B. Width	C. Length	Total subsystem weight (lb)
1	125	I	8	4.5	5	2250
2	125	I	8	4.5	5	2250
3	125	II	8	4.5	5	2250
4	250	I	8	6	6.5	3900
5	250	II	8	6	6.5	3900
6	250	II	8	6	6.5	3900
7	250	II	8	6	6.5	3900
8	250	III	8	6	6.5	3900
9	500	II	8	7	9.5	6800
10	500	III	8	7	9.5	6800
11	700	III	8	8	10.5	8800

Systems 2 and 6 require cooling water. To avoid plugging of pipes and heat exchanges, provisions were included in the water purification subsystem for filtering the cooling water and delivering it to the other subsystems. A sand filter system commonly used on electrodialyzers was selected for handling cooling water filtration. The necessary filtration unit data were obtained from Ionics, Inc, Watertown, Massachusetts. A flow schematic of the filtration system is shown in Figure 3.4-12. Two filter vessels are required—while one is being cleaned of impurities the other is on stream. For example, when Filter I is on stream, valves V1, V4, V6, and V8 are open and valves V2, V3, V5, and V7 are closed. Each filter vessel is on stream for 15 min and then backwashed. The filtered water is delivered to the pump inlet at an estimated 10-psia pressure where it is pressurized to about 50 psia for delivery to the subsystems.

The cooling water required was estimated to be about 11,000 gph for System 2 and 22,000 gph for System 6. To allow for backwash water, the total amount of water requiring filtration was estimated to be 12,000 and 24,000 gph, respectively. For the 12,000-gph design, each filter will weigh about 200 lb and will be 1 ft in diameter and 2 ft long. For the 24,000-gph design, each filter will weigh about 400 lb and will be 1.5 ft in diameter and 2 ft long.

Cooling water pump data were obtained from the Aurora Pump Division of the New York Air Brake Co, Aurora, Illinois. The pumps operate at 3600 rpm and require 7.5- and 15-hp motors, respectively. For System 2, the pump dimensions are 12 X 12 X 20 in. and the weight is 130 lb. For System 6, the pump dimensions are 16 X 12 X 21 in. and the weight is 175 lb. Pump motor

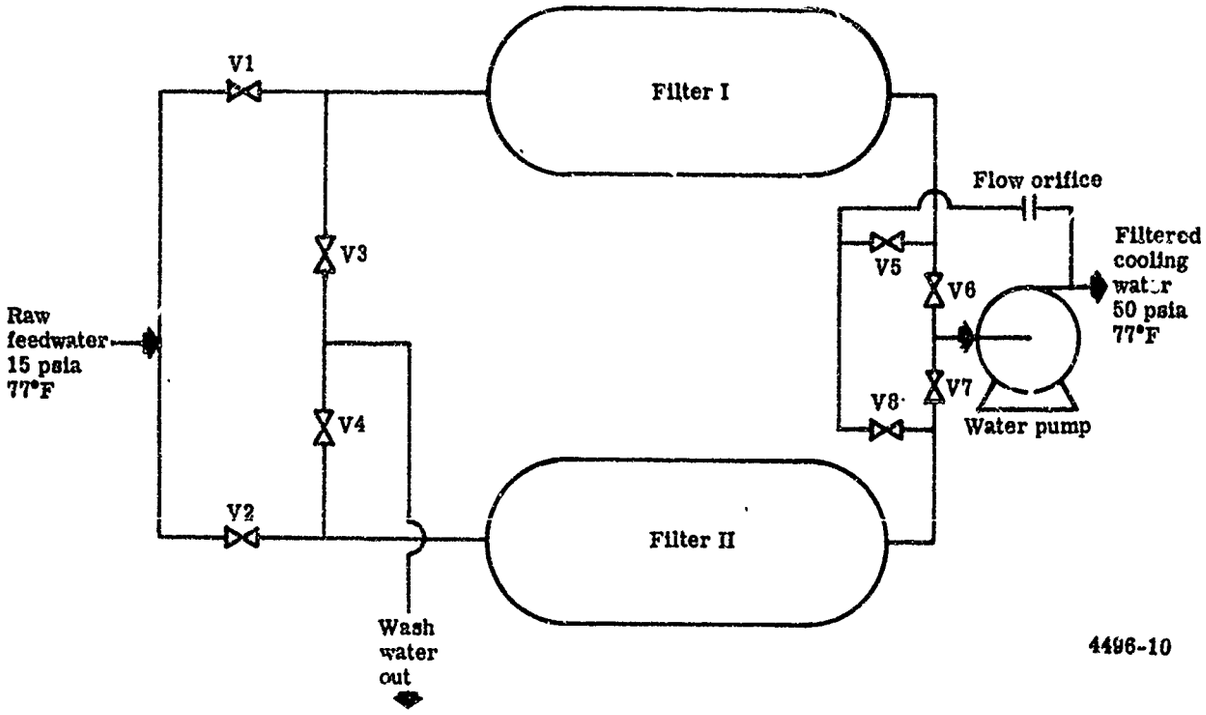


Figure 3.4-12. Schematic of cooling water filtration unit.

data were obtained from General Electric Co, Schenectady, New York, based on their lightweight design. The motor for system 2 weighs 140 lb and is 14.5 in. long and 11.25 in. in diameter. The motor for system 6 weighs 240 lb and is 18 in. long and 13.5 in. in diameter.

The overall weight and size of the cooling water filtration system was obtained by combining the components previously described. For system 2, the total weight is about 800 lb, and the total volume is about 18 ft^3. For system 6, the total weight is about 1300 lb, and the total volume is about 30 ft^3.

PREDICTED PERFORMANCE

The vapor compression water purification subsystem chosen for the fuel production system is a low-temperature, nonscaling unit and is fully automatic in starting as well as in operation. At design point conditions (77°F and 1 atm absolute pressure), the systems designed will deliver 125-, 250-, 500-, and 700-gph of product water containing approximately 1 ppm of impurities. No auxiliary ion-exchange units are required to reach this purity level regardless of the purity of the raw feedwater; the system is capable of purifying all impure waters including sea water. If the feedwater temperature is less than 77°F, immersion heaters are automatically energized to supply the additional thermal energy required to maintain the production rate of purified water. Also, the design point production rate is maintained even if the ambient air temperature is 125°F and the feedwater temperature is 105°F. Likewise, if the ambient

pressure decreases below 14.7 psia, the design point production rate will automatically be maintained. This is accomplished by oversizing the vent condenser and vacuum pump and by providing bypass lines around the feedwater preheater.

An additional off-design condition studied was operation of the entire Energy Depot at half-power. Three design philosophies were considered for accommodating this condition. First, dual electrical components could be used, e.g., instead of using a single distilled water pump, two half-capacity pumps would be used. For half-power operation, only one of the two pumps would be in operation. Second, dual speed electrical components could be specified. Third, a distilled water storage tank could be installed in the outlet stream so that the unit would operate at full power, but only intermittently. The latter design required the simplest modifications and was therefore selected. This design has the advantage that the makeup KOH required in the electrolyzer could easily be added in the distilled water storage tank. During half-power operation, it was assumed that approximately one-half of the design water flow rate would be required and that the vapor compression unit could cycle on and off over a range of time periods. The vendor indicated that the operating vacuum would remain relatively constant for about 2 hr if no product water was withdrawn. Therefore, an operating cycle of 2 hr on at full power and 2 hr off at no power was chosen. To satisfy the storage capacity required by this time cycle, storage vessels were sized to hold approximately 150 gal for the 3-Mw_e plant, 300 gal for the 6-Mw_e plant, 600 gal for the 10-Mw_e plant, and 850 gal for the 15-Mw_e plant.

To assemble the plant for operation requires connection of the feedwater hose, the waste water drain hose, the electrical power line, and the pure water hose leading from the storage tank to the electrolyzer skid. To disassemble the plant, these same lines are disconnected.

Before starting the vapor compression plant, the oil reservoirs in the compressor and the pumps should be checked and, if necessary, filled to the proper specified levels. Then the feed valve switch is placed in the "automatic" position and the salinity control switch is turned to "on." The float switch in the immersion heater tank or the salinity control switch then starts and stops the unit automatically.

The cold distilling unit requires a short period of time for starting, depending on the temperature of the unit components and the feedwater. When first started, the system contains air which is removed by the vacuum pump. Simultaneously, the evaporator is filled to its operating level. When the vacuum in the evaporator reaches about 24 in. Hg, the compressor starts automatically. When the operating temperature is reached, the circulating water in the bottom head of the evaporator will be 107°F. When the operating pressure is reached, the vacuum in the evaporator will be 27.6 in. Hg. When these two conditions are approximately correct, the unit will begin distilling water automatically. These two conditions are normally met within 20 to 30 min.

Distilled water is retained in the immersion heater tank for starting purposes. The amount retained varies from about 5 gal for the 125-gph plant to about 30 gal for the 700-gph plant. When starting the unit for the first time, distilled water must be supplied to fill the immersion heater tank. When locations are changed, the unit can be moved with the tank full. After start-up, the immersion heaters are only cycled on and off to provide the minimum auxiliary heating required. Warm feedwater (77°F or above) requires electric heating for starting only, not for operation. As the feedwater temperature is reduced, the immersion heaters are cycled on and off more frequently.

The unit is stopped automatically by placing the control selector switch in the "off" position. The float switch in the immersion heater tank normally stops the unit when the tank is full; the salinity control system stops the unit if the product purity becomes unacceptable. If the unit is to be exposed to freezing conditions, the water should be drained from the evaporator, immersion heater tank, preheater, and pumps.

The cooling water filtration system described previously is very simple. Before start-up, the feedwater hose, electrical power line, and subsystem delivery hose are connected to the system. The cooling water delivery pump is then started, and filtered water is available immediately. This sequence is merely reversed for shutdown. For less than design flow rates, the filtered water is recycled to the inlet side of the pump. The filtering cycle is controlled automatically by a timed control system which opens and closes the appropriate valves for each filter-backwash sequence.

DISCUSSION

The two processes receiving the most attention for performing water purification were vapor compression and electrodialysis. While final data were being assimilated and compared, communications were established with the Sanitary Sciences Division of the U.S. Army.† This division is responsible for examining water purification techniques suitable for Army applications. It was found that the Army has tested both electrodialysis and vapor compression techniques under actual field conditions. The vapor compression technique has demonstrated superior operating characteristics and was therefore selected for Army field use. It was found, however, that the electrodialysis system does exhibit desirable qualities—e.g., low weight, small size, and low power consumption—but the system is not presently sufficiently flexible in operation to handle the ranges of feedwater purities required for Army applications. Electrodialyzers are still being tested by the Army periodically, but even the improved units appear to have limited applications. The results of these Army communications served to strengthen the choice of vapor compression for Energy Depot applications.

In addition to problems already mentioned, two other relatively minor problems exist with the electrodialysis process. First, the selective membranes must be kept wet at all times, otherwise they will crack and become useless. Second, a small ion exchange system may be required

downstream of the electrodialyzer to achieve 1-ppm purity output levels. Such an ion exchange system would require a supply of regenerative chemicals or disposable cartridges. In either case, a logistics constraint is imposed.

According to the vapor compression vendor, Mechanical Equipment Co, a problem could exist with this system also. Occasionally, a brackish feedwater which exhibits peculiar scaling properties will be encountered. Apparently these properties are not well defined or always predictable. Consequently, the possibility exists that scale-removing chemicals may be required for proper operation of the vapor compression unit in such environments.

The many expected advantages of the reverse osmosis water purification technique were described earlier. This process appears to be the most potentially attractive of the water purification techniques studied for Energy Depot. Therefore, progress in reverse osmosis technology should be monitored closely for Energy Depot applications. In addition, advances made in reverse osmosis membrane technology may be applicable to the electrodialysis process, in which case it is conceivable that either reverse osmosis or electrodialysis could become more attractive than vapor compression.

All water purification schemes are receiving considerable attention, both from industry and government agencies. The results of this work should be monitored closely. There appears to be little to be gained by performing additional research and development work specifically for handling the Energy Depot water purification task.

CALCULATIONS

A major portion of the data reported for the water purification subsystem was obtained from vendors. The only components requiring calculations were the distilled water storage tanks for each plant. As mentioned previously, the tank storage capacities needed were assumed to be approximately 150 gal for the 3-Mw_e plant, 300 gal for the 6-Mw_e plant, 600 gal for the 10-Mw_e plant, and 850 gal for the 15-Mw_e plant. The following calculation is for the tank of the 3-Mw_e plant. Similar calculations were performed for the other plants.

To minimize the space required for the storage tank, the tank height was fixed at 7.5 ft, conforming to the height constraint. The volume of the tank is

$$V = 150 \text{ gal} \times 0.1337 \frac{\text{ft}^3}{\text{gal}} = 20 \text{ ft}^3$$

$$V = \frac{\pi}{4} d^2 h = 0.785 \, d^2 h$$

$$d^2 = \frac{20}{(0.785)\ (7.5)} = 3.4\ ft^2$$

$$d = 1.84\ ft$$

where

d = tank diameter, ft
h = tank height, ft

The maximum pressure the tank must w[illegible]tand is that due to a 7.5-ft head of water, viz,

$$P_{max} = 62.3\ \frac{lb}{ft^3} \times 7.5\ ft + 15\ \frac{lb}{in.^2} \times 144\ \frac{in.^2}{ft^2} = 2626\ lb/ft^2$$

The simple equation describing the material thickness required to accommodate this p[illegible]sure in a cylindrical vessel is

$$t = \frac{P\ D}{2\ S}$$

where

t = wall thickness, in.
P = pressure, psi
D = internal cylinder diameter, in.
S = material yield strength, psi

The tank will be constructed of Type 304 SS to eliminate corrosion problems and to allow KOH additions to be made, if desired. The allowable yield strength for Type 304 SS is 20,000 psi.[8] The minimum wall thickness is

$$t = \frac{(2626\ lb/ft^2)\ (1.84\ ft)}{2\ (20{,}000\ \frac{lb}{in.^2} \times 144\ \frac{in.^2}{ft^2})} = 8.4 \times 10^{-4}\ ft = 0.0101\ in.$$

The vessel weight was calculated using a standard 1/32-in. wall thickness as follows

Volume of metal = 2 × volume of ends + shell volume

or

$$V = 2 \times 0.785\,(1.84)^2 \times \frac{0.03125}{12} + 4 \times 0.785 \times 7.5 \times \frac{0.03125}{12}\,(1.84 + \frac{0.03125}{12})$$

$$= 1.27 \times 10^{-1}\ ft^3$$

The vessel weight is simply the volume of metal times the density

$$W = 1.27 \times 10^{-1}\ ft^3 \times 500\ \frac{lb}{ft^2}$$

$$= 64\ lb$$

Similar calculations were made for the water storage tanks required for the other three plant sizes. A summary of the relevant resulting data is presented in Table 3.4-VIII.

Table 3.4-VIII.
Summary of water storage tank data.

Plant size (Mw_e)	Storage capacity (gal)	Tank dimensions (ft)			Volume (ft^3)	Tank weight (lb)	
		Dia	Wall thickness	Height		Empty	Full
3	150	1.8	2.6×10^{-3}	7.5	20	64	1300
6	300	2.6	2.6×10^{-3}	7.5	40	94	2600
10	600	3.7	2.6×10^{-3}	7.5	80	140	5200
15	850	4.4	2.6×10^{-3}	7.5	114	180	7300

REFERENCES

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6. Electrochemical Processes for Water, Food, and Chemicals. Brochure prepared by Ionics, Inc, Watertown, Massachusetts.

7. Private communications between Mr. R. D. Tyler, Section Chief, Allison Division of General Motors, Indianapolis, Indiana, and Mr. Richard P. Schmitt, Chief, Sanitary Sciences Division, U.S. Army Engineering Research and Development Laboratories, Fort Belvoir, Virginia.

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Power Conditioning Subsystem

DESCRIPTION

The power conditioning subsystem transforms, distributes, and rectifies the 4160-v, 60-cps (or 400-cps), 3-phase input electrical power as required by the electrolyzer modules and the various motors, heaters, and control equipment used throughout the several subsystems of the fuel production system. It also provides protection for the electrical equipment from potential damaging over-voltages and over-currents which conceivably could occur.

A simplified electrical circuit diagram of the power conditioning subsystem is shown in Figure 3.5-1. The electrical power is received at the high voltage terminals. Surge arrestors provide lightning protection; primary fuses provide fault protection for the transformer and 460-v bus; and current transformers provide the electrical load signal for plant control. The 4160-v input voltage is reduced to 460 v by a 3-phase, forced-circulation oil-cooled transformer. Cables distribute the power from the transformer secondary to the several subsystems—the major portion (approximately 90%) of this power goes to the hydrogen production subsystem. A circuit breaker with overload protection is provided for each motor and heater bank.

In the hydrogen production subsystem, the 460-v, 60-cps power is supplied to voltage control-rectifier modules. These modules rectify the a-c power to 600-v, d-c power with a full-wave bridge network; provide d-c voltage control by use of silicon control, rectifier phase control in the bridge circuit; and sectionalize the d-c load by turning off the silicon control rectifiers. A shunt provides an indication of the electrical load of each d-c circuit for monitoring.

INVESTIGATION

The reference design for the power conditioning subsystem investigation was based on the 3-Mw_e power delivered at 60 cps, 3 phase, and 4160 v.[1] In conjunction with the Integration and Packaging, Control Analysis, and Performance Analysis studies, the reference design was revised to meet the requirements of the eleven different systems indicated in Section II. The variations in these eleven systems, which have a major influence on the design of the power conditioning subsystem, are the four input power levels—3, 6, 10, and 15 Mw—and the change from 60-cps to 400-cps frequency for the input power of system 7.

Parametric data were developed for the components of the power conditioning subsystem. These data were required for the system integration and packaging and system optimization studies.

Data indicating the variation in transformer weight with rating were obtained from electrical equipment suppliers for 60- and 400-cps lightweight, oil-cooled transformers. See Figure

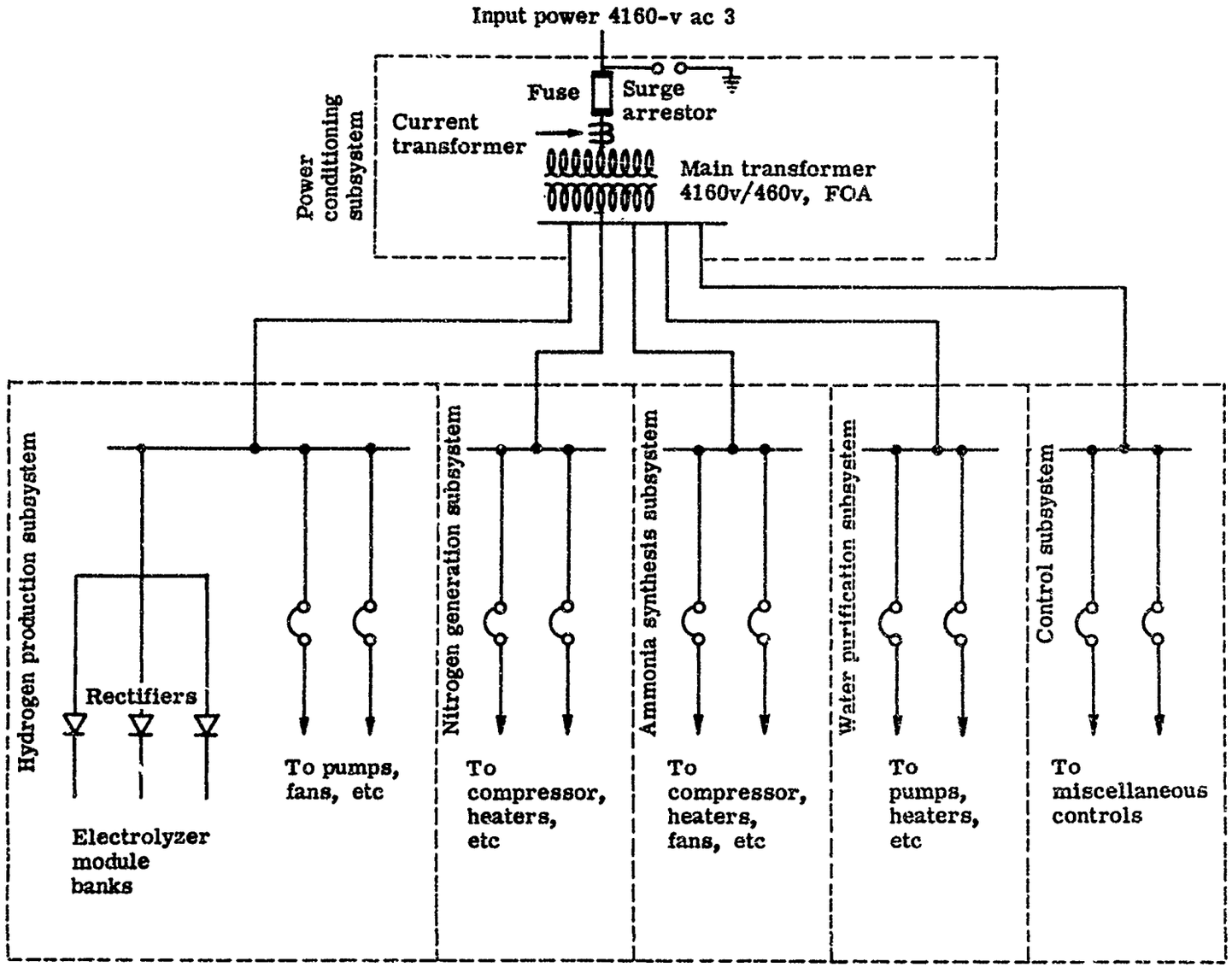


Figure 3.5-1. Simplified electrical circuit diagram of power conditioning subsystem.

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3.5-2. These data are based on the lightest weight transformers considered practical for the application, a further reduction in weight is possible but would cause significant reductions in system efficiency.

The improvement in transformer efficiency which can be achieved with an increase in weight is shown in Figure 3.5-3. Since the power available for fuel production is directly related to the efficiency of the transformer, this variation was of interest for plant optimization in the system integration and packaging studies.

The optimization studies indicated that, where additional weight could be accommodated, transformer efficiency should be increased to approximately 99%. At this efficiency level, conventional transformer core designs can be used. However, these transformer designs differ from conventional designs in that the transformer case is designed with lightweight structural materials and with a minimum cooling oil volume. In addition, the lightweight transformer ratings are based on five-year insulation life at a hot spot temperature established by full power and 77°F ambient temperature.

The variation in load carrying capability of the transformer with an ambient temperature variation is indicated in Figure 3.5-4. At an ambient temperature of 125°F the transformer capability is reduced to 77% of the rating at the design point ambient temperature of 77°F.

Parametric data on other electrical equipment were obtained from vendors. Weight and size data for 1200- and 3600-rpm lightweight electric motors are shown in Figure 3.5-5. Weight and size data for aluminum bus bars are shown in Figures 3.5-6, 3.5-7, and 3.5-8.

High voltage, 4160-v supply to the a-c power loads was investigated for the larger plants. Low voltage, 460-v supply was retained for the following reasons: the a-c load is a relatively small portion of the total load, approximately 12%; the a-c load consists of many small loads (the ammonia synthesis subsystem compressors, which are the largest individual a-c loads, do not exceed 350 hp each); and although not considered in the study, the input power supply to the largest system will probably be higher than 4160 v.

A simplified schematic of a voltage-controlled rectifier is shown in Figure 3.5-9. Three of the six legs of the full wave bridge contain silicon control rectifiers while the remaining legs contain diodes. The rectifiers block current flow until they are fired by a short current pulse at the gates. The silicon control rectifiers will continue to conduct until the end of the half cycle or until the voltage at the anode is no longer positive with respect to the cathode. The diode will conduct as long as the anode is positive with respect to the cathode.

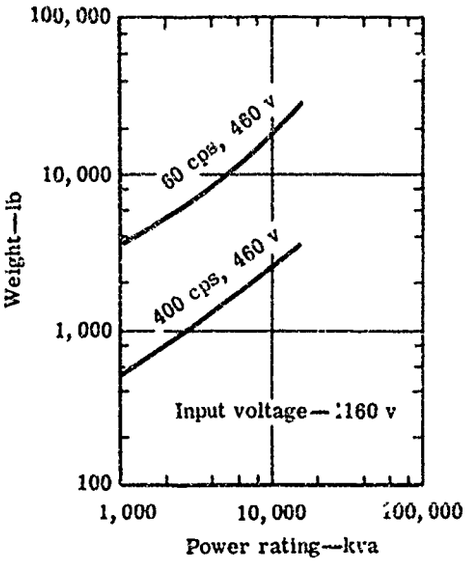


Figure 3.5-2. Transformer weight vs power rating data.

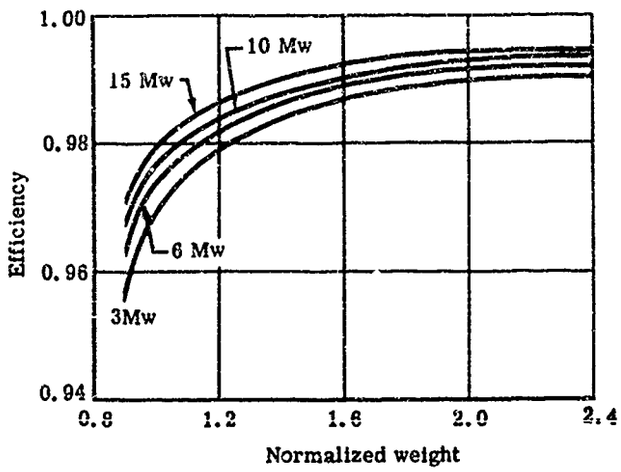


Figure 3.5-3. Transformer efficiency.

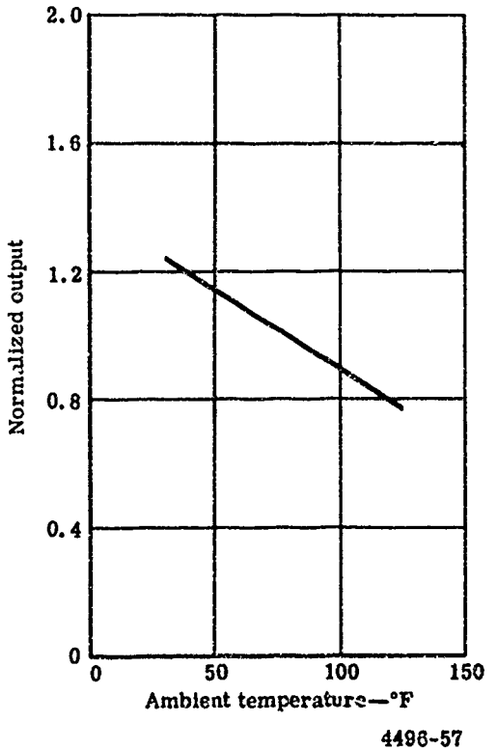


Figure 3.5-4. Transformer capability versus ambient temperature.

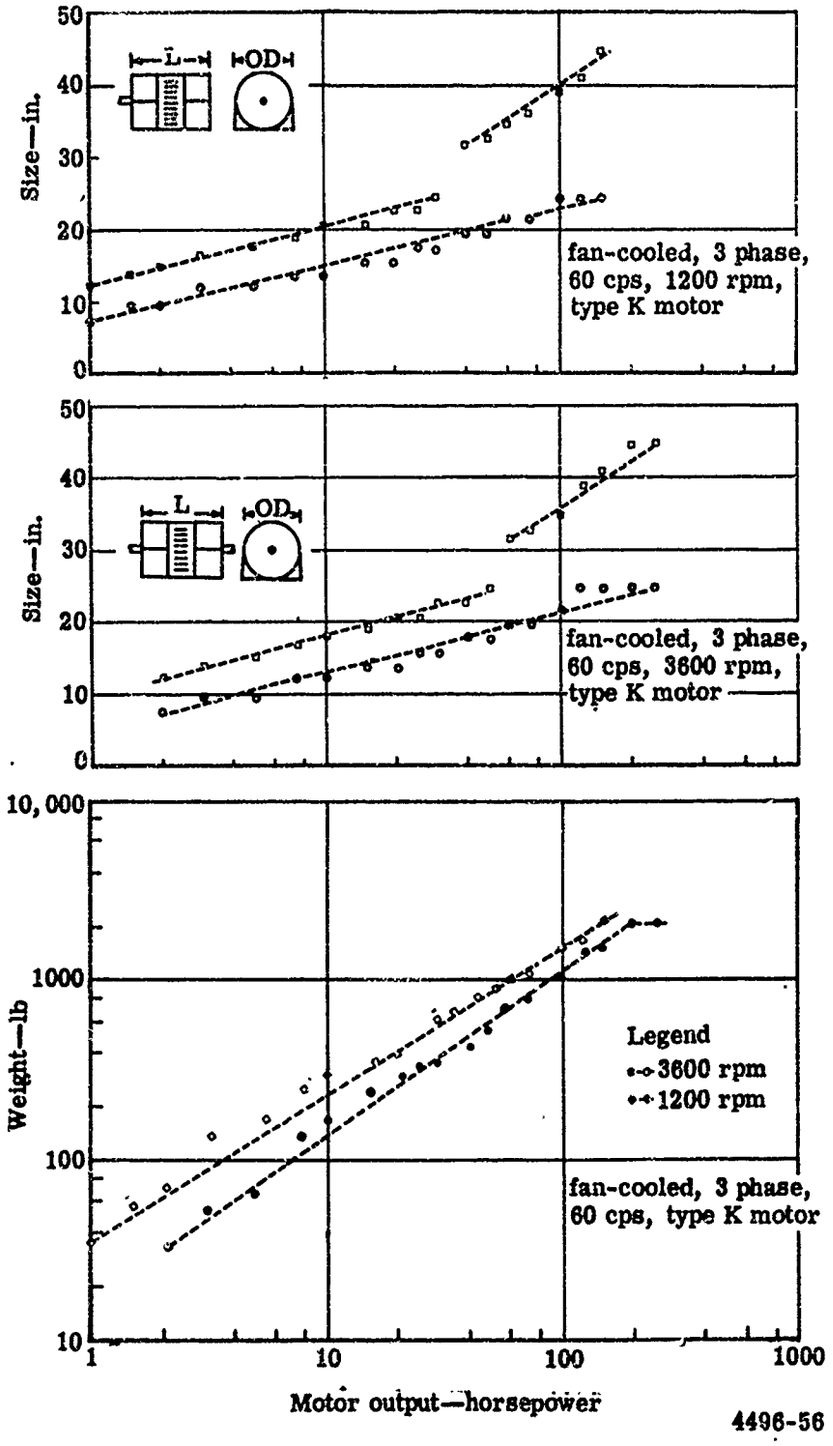


Figure 3.5-5. Size and weight data for totally enclosed, fan-cooled, 3-phase, type-K motors.

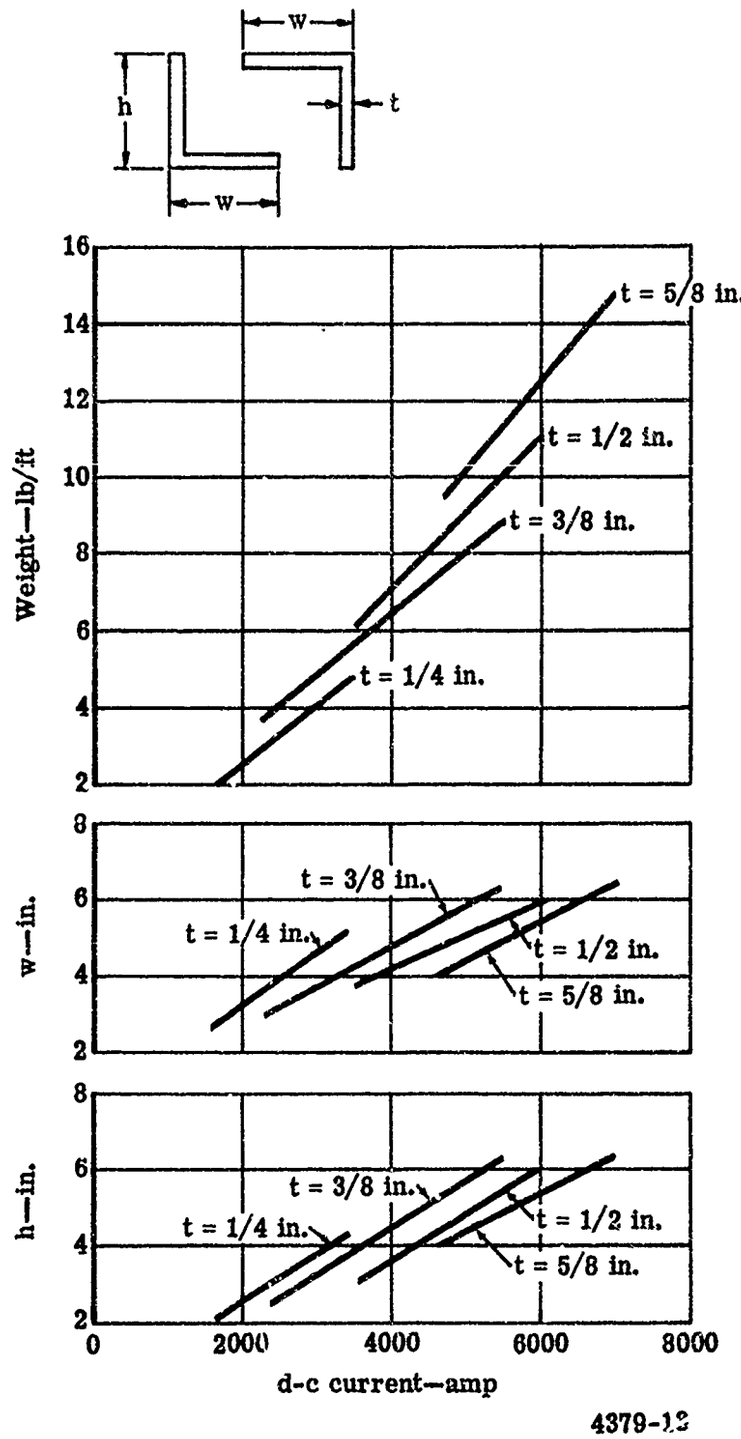


Figure 3.5-6. D-C current-carrying capacities of aluminum channel bus for 35°C rise.

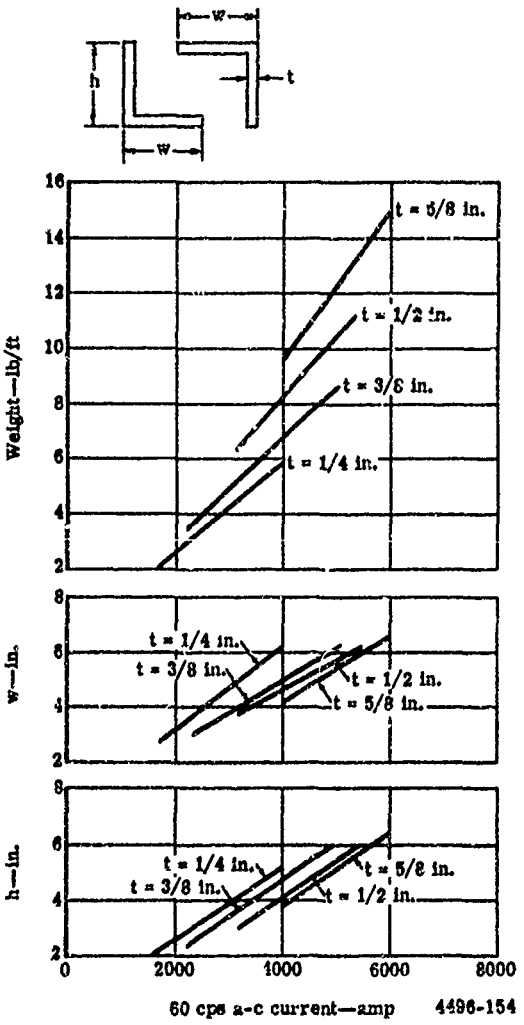


Figure 3.5-7. A-C current-carrying capacities of aluminum channel bus bars for 35°C rise.

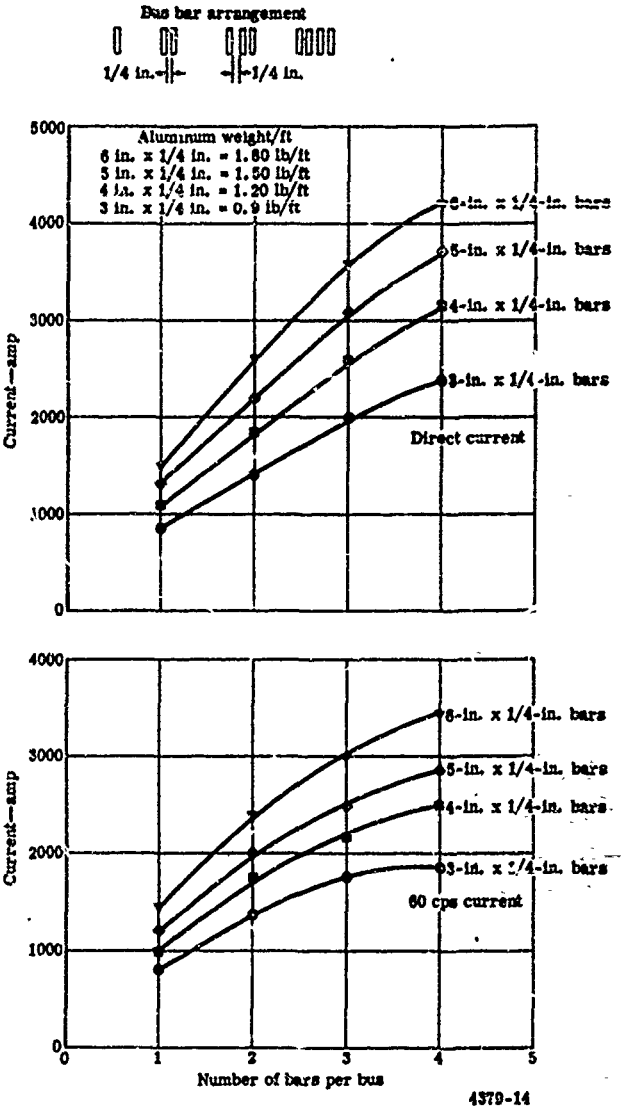


Figure 3.5-8. Current-carrying capacities of aluminum bus bars for 35°C—dc and 60-cps current.

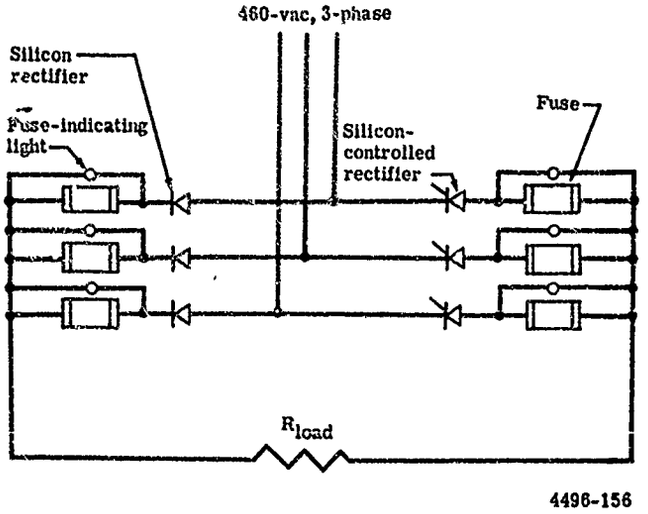


Figure 3.5-9. 3-phase, full-wave, voltage-controlled rectifier.

Alternate methods of varying the d-c voltage to the electrolyzer modules were considered. These methods include multitap transformers, autotransformers, and saturable reactors on the a-c voltage portion of the electrical system. The a-c voltage is then rectified by a conventional, solid state power supply. Multitap transformers tend to have complex mechanical "make and break" equipment which reduce reliability. Autotransformers are not recommended because they are limited to applications where the voltage transformer ratios do not exceed four-to-one. Saturable reactors would perform the voltage control function but they are magnetic devices and, consequently, are large and bulky. Also, heavy-duty line contactors as well as rectifiers would be required on each electrolyzer module.

The d-c voltage control method shown in Figure 3.5-9 requires a firing circuit and low power control relays in conjunction with the bridge rectifiers. The required voltage and current ratings are high for the diodes and silicon control rectifiers but are not beyond the state of the art. The ruggedness and reliability of these units for military use are well known and no problems are anticipated. Since other d-c voltage control methods offer no cost or weight advantage, the solid state, voltage controlled power supply was selected.

DESIGN PARAMETERS

The design parameters are as follows:

- Input power
  - 4160-v; 60-cps; 3-phase; and 3-, 6-, 10-, and 15-Mw_e
  - 4160-v; 400-cps; 3-phase; and 6 Mw_e

- Output power
  Approximately 90% is 600-vdc and the remainder is 460-vac
- Ambient temperature
  Design point—77°F
  Range— -65° to + 125°F

DESIGN

A one-line circuit diagram of the power conditioning subsystem is shown in Figure 3.5-10. Table 3.5-I lists the pertinent characteristics for the major components.

The transformer design for each system is three-phase transformer with 4160-v primary and 460-v secondary (phase to phase) voltage. The transformer is cooled by forced oil circulation; the oil is cooled in an oil-to-air heat exchanger. Systems 4 and 9 each include two 50% capacity transformers to facilitate better weight division between packages. With the exception of system 4, transformer efficiency is approximately 99%. The two 50% capacity transformers used in system 4 are weight restricted to meet package weight limitations; they have an efficiency of approximately 97%.

With the exception of system 4 approximately 88.5% of the total input power to each system is supplied to the electrolyzer modules as variable, 575- to 619-vdc power. In system 4, approximately 90.5% of the input power is supplied to the electrolyzer module.

The largest a-c load in the system is the ammonia synthesis subsystem compressor motor(s). This load is approximately 7% of the input power in the 3- and 6-Mw systems. In the large systems—10 and 15 Mw_e—the nitrogen compressor is driven by a separate motor; the ammonia compressor load is approximately 5.5% of the input power.

A diagram of the d-c voltage supply to the electrolyzer module banks is shown in Figure 3.5-11. The voltage-control rectifier design is based on the concept discussed previously in the Investigation subsection and shown in Figure 3.5-9.

One 3-phase, full-wave, voltage-controlled rectifier is provided for each electrolyzer module bank. A single semiconductor element is used in each of the rectifier bridge legs of the 3 Mw_e system. Two semiconductor elements are used in each rectifier bridge leg for the 6-, 10-, and 15-Mw_e systems. A fast-acting, current-limiting fuse is provided for protection of each semiconductor.

The rectifiers and the d-c bus conductors are designed to permit operation at the maximum conditions indicated in Table 3.5-I. These conditions are approached as a limit when one or more electrolyzer module banks are shut down and the power to the remaining electrolyzer modules banks is increased.

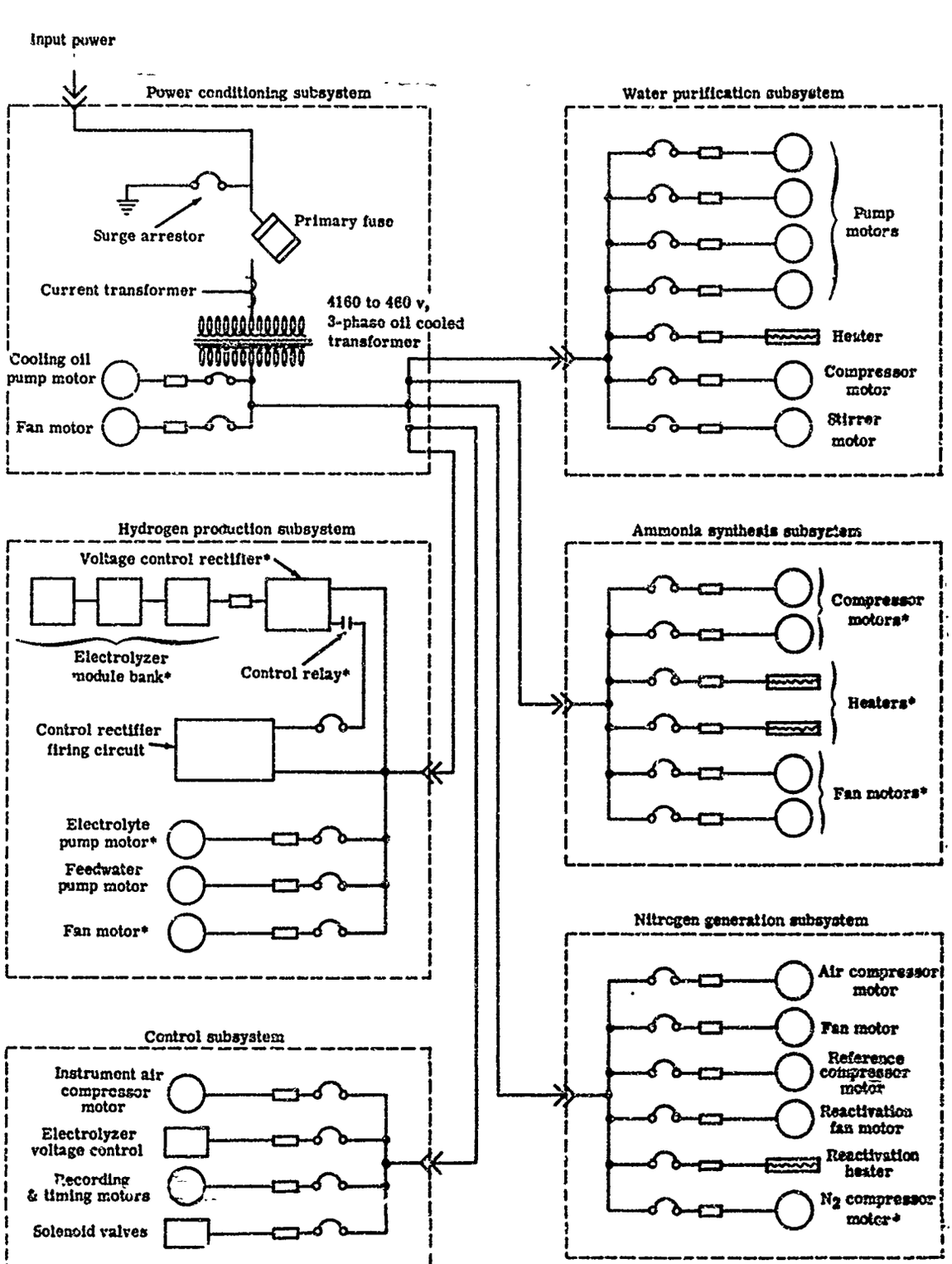


Figure 3.5-10. Electrical distribution diagram.

Table 3.5-I.

Ratings of major components for the power conditioning subsystem.

System	1	2	3	4	5	6	7	8	9	10	11
Input Power											
Power—kw	3,000	3,000	3,000	6,000	6,000	6,000	6,000	6,000	10,000	10,000	15,000
Frequency—cps	60	60	60	60	60	60	400	60	60	60	60
Voltage—v	4160	4160	4160	4160	4160	4160	4160	4160	4160	4160	4160
Transformer—3-phase, 2-winding, FOA, 4160 v/460 v											
Rating—kva	3,333	3,333	3,333	3,333	6,667	6,667	6,667	[illegible]37	5,556	11,111	16,667
Efficiency—%	98.9	98.9	98.9	97.0	99.0	99.0	99.0	99.0	99.0	99.1	99.1
Number of transformers	1	1	1	2	1	1	1	1	2	1	1
Weight—lb/transformer	13,500	13,500	13,500	7,400	22,200	22,200	3,600	22,200	20,200	35,200	49,600
Hydrogen Production Subsystem											
Direct current load—% of input power	88.2	88.2	87.8	90.5	88.3	88.7	88.3	88.3	88.8	88.[illegible]	89.0
Design point (normal operation)											
Voltage—v	593	593	603	585	[illegible]3[illegible]	588	588	588	595	582	575
Current per circuit—amp	558	558	364	774	693	696	693	693	848	848	1055
Number of circuits	8	8	12	12	13	13	13	13	18	18	22
Design point (maximum conditions)											
Maximum voltage—v	619	619	619	619	619	619	619	619	619	619	619
Maximum current—amp	714	714	473	976	878	878	878	878	1067	1060	1310
Alternating Current Loads											
Electrolyte pump motor											
Number	1	1	1	2	1	1	1	1	2	2	2
Total power requirements—kw	14.6	14.6	16.6	28.2	31.8	31.3	31.8	31.8	50.6	60.6	98.2
Feedwater pump motor											
Total power requirements—kw	0.55	0.55	0.55	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1
Number	1	1	1	1	2	2	2	2	2	2	3
Fan motor											
Total power requirements—kw	4.1	—	4.5	15.4	20.2	—	20.0	20.2	31.4	48.3	51.3
Number	2	—	2	4	4	—	4	4	8	6	12
Water Purification Subsystem Power Requirements—kw	7.7	7.7	7.9	13.3	13.6	13.7	13.6	13.6	21.1	21.1	29.3
Nitrogen Generation Subsystem											
Air compressor motor power requirements—kw	30	30	30	56	56	56	56	56	93	93	150
Nitrogen compressor motor power requirements—kw	—	—	—	—	—	—	—	—	112	112	150
Miscellaneous power requirements—kw	15	15	17	27	30	29	30	30	111	111	176
Ammonia Synthesis Subsystem											
Compressor motor											
Total power requirements—kw	220.8	220.2	239.1	420.7	445.6	437.4	445.6	445.6	652.6	552.6	765.2
Number	1	1	1	2	2	2	2	2	3	3	4
Fan motor											
Total power requirements—kw	4	—	4	16	12	—	12	12	12	12	28
Number	1	—	1	4	3	—	3	3	3	3	7
Start-up heaters											
Total power requirements—kw	100	100	100	150	150	150	150	150	200	200	250
Number	2	2	2	2	2	2	2	2	2	2	2

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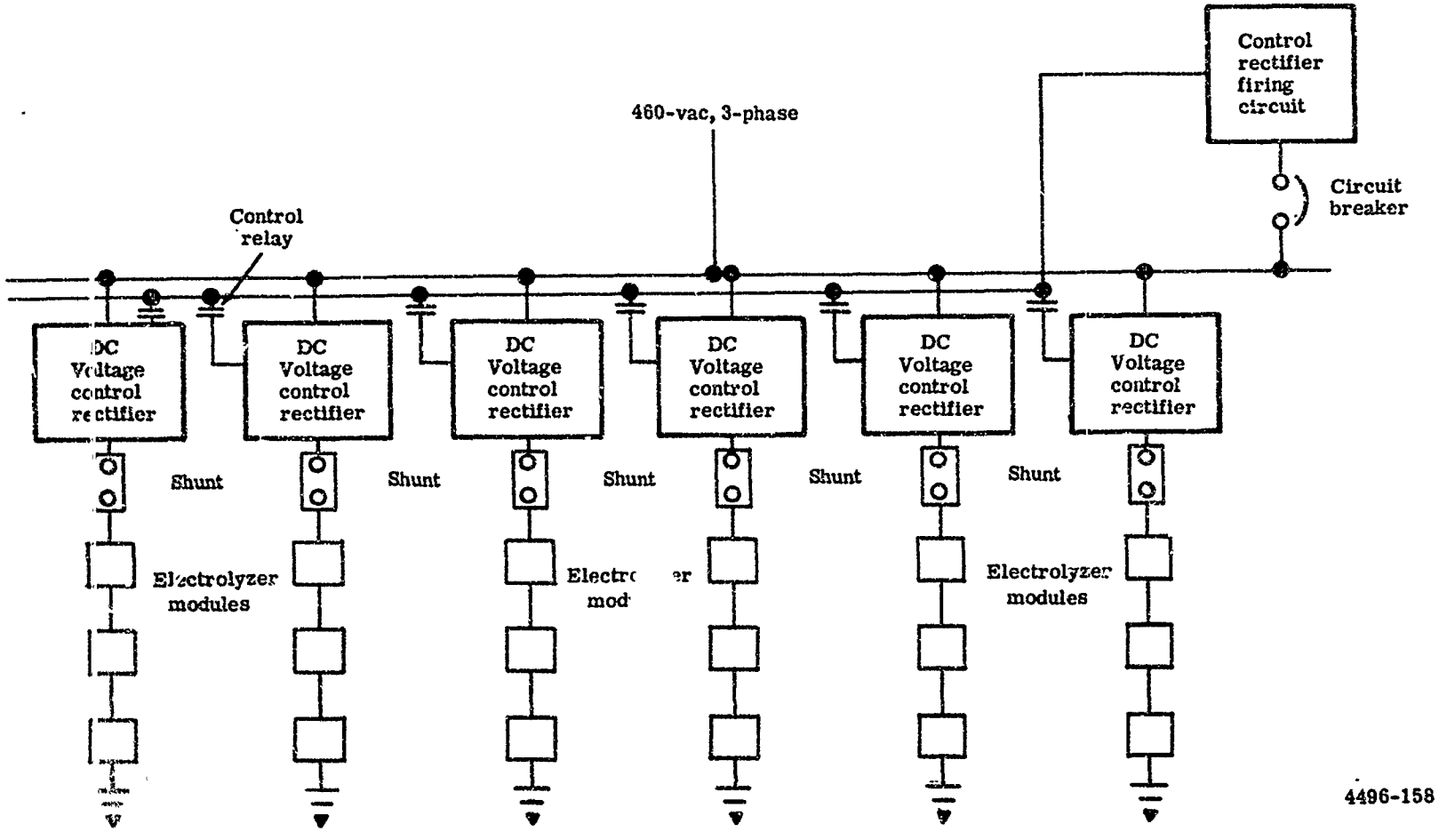


Figure 3.5-11. Electrical power supply to electrolyzer modules.

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The power to the individual electrolyzer module banks is increased by increasing the d-c voltage. The silicon control rectifiers (Figure 3.5-9) block current flow until they are fired by a short circuit pulse at the gate. These rectifiers will continue to conduct after the gate signal is removed until the end of the half cycle or until the anode is no longer positive with respect to the cathode. The diode will conduct whenever the anode is positive with respect to the cathode.

The d-c voltage to the electrolyzer module banks, E_{dc}, is varied by delaying the firing of the short circuit pulse at the gate. This phase delay, ϕ_F, is provided by a firing circuit which supplies gate signals in the form of pulses common to all three silicon control rectifiers. Because the anode must be positive with respect to the cathode and only one phase at a time may be conducted, only the appropriate rectifier will be fired at the right time by the common pulses. For $\phi_F = 0$, $E_{dc} = 1.35 \times 460$ v which is the d-c voltage of a conventional 3-phase, full wave bridge. The wave form is shown in Figure 3.5-12. Figure 3.5-12 also shows the wave form for 10° increases in ϕ_F. Figure 3.5-13 is a schematic of a typical firing circuit for use with the power supply—the input and output wave forms are shown. The circuit is made up of three zero-crossing threshold detectors (only one shown) which product a pulse output each time the line voltage changes sign. Each detector is tied to a phase of the line voltage and output pulses are gated such that six pulses are generated per cycle. The pulses are 60° apart and occur at 0°, 60°, 120°, etc. A delay circuit causes the pulses to be delayed in time from 0 to 1.4 msec (0° to 30°) as determined by the delay control signal. In this fashion, pulses synchronized to the line frequency with a 0° to 30° phase delay capability are generated to fire the silicon control rectifiers. Figure 3.5-14 indicates variation in d-c voltage and a-c power factor with variation in silicon control rectifier phase delay.

PREDICTED PERFORMANCE

The power conditioning subsystem is designed to operate at rated power at an ambient design temperature of 77°F. The load carrying capacity of limiting components at off-design conditions is described in the Performance Analysis subsection.

The power conditioning, hydrogen production, and control subsystems are designed to permit continued operation of the entire system at full power input with a maximum of one-fourth of the electrolyzer module banks out of service. The maximum d-c operating voltages and currents are indicated in Table 3.5-I.

DISCUSSION

The components used in the power conditioning subsystem, with the exception of the main transformer, are of a design which is in use for other applications with similar requirements. The transformer design provided by an electrical equipment vendor is unique in that lightweight materials are used for the transformer casing and the design minimizes the cooling oil inventory. No development work is considered necessary.

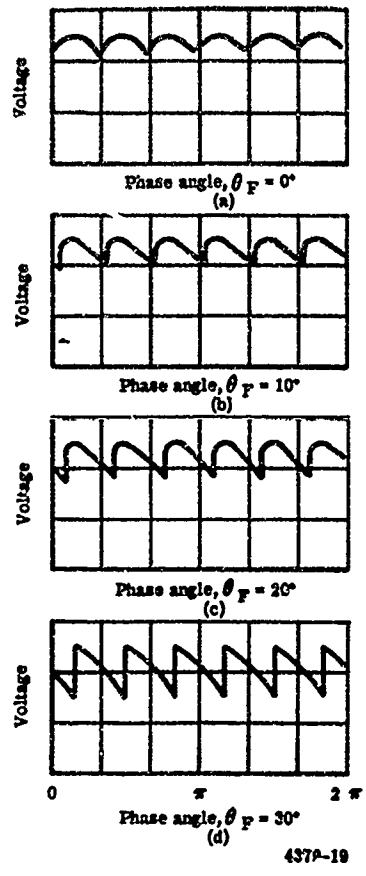


Figure 3.5-12. Rectified d-c predicted waveforms.

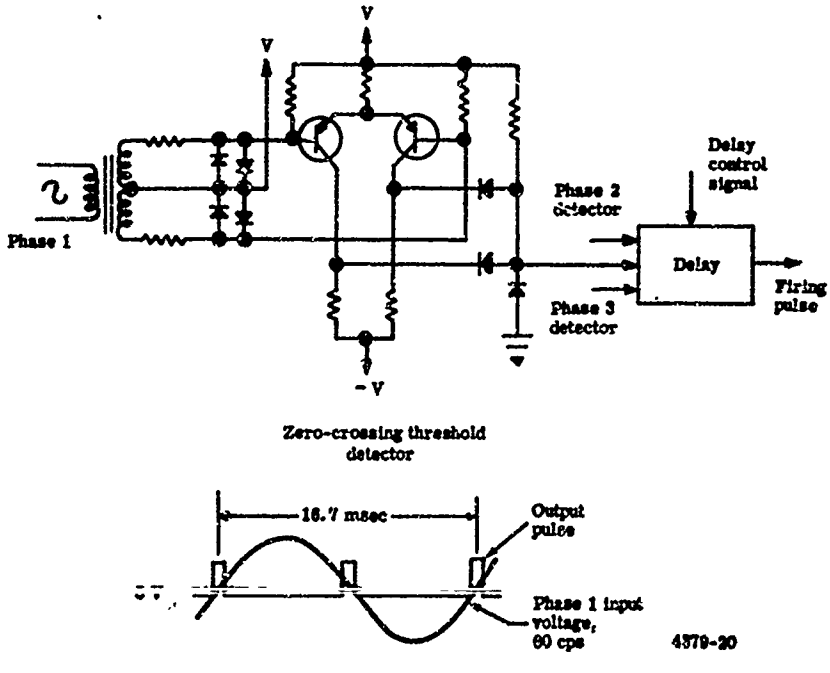


Figure 3.5-13. Firing circuit.

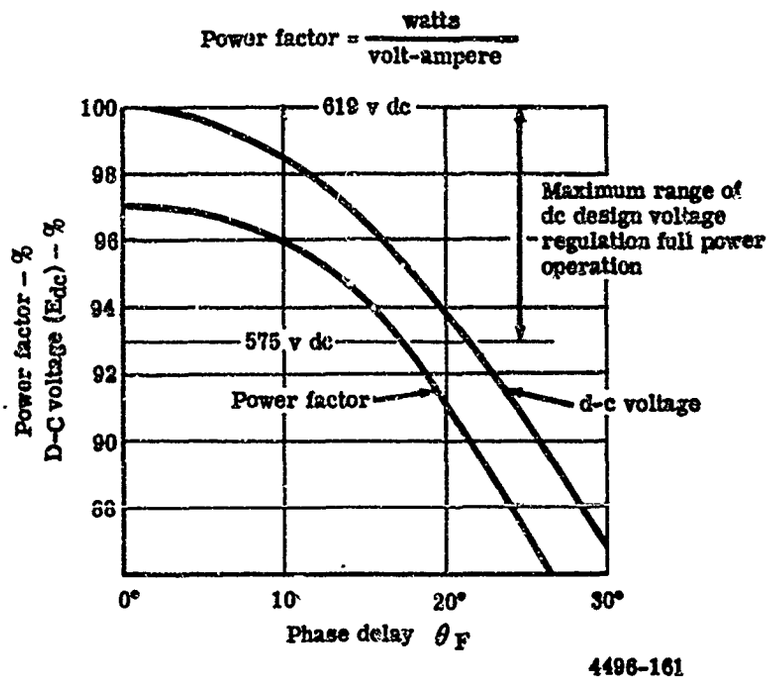


Figure 3.5-14. D-C voltage regulation and a-c power factor as function of silicon control rectifier phase delay.

A considerable weight saving was achieved with 400-cps power input to system 7. The proposed transformer size, 6667 mva, is beyond the state of the art. However, progress can be expected in 400-cps equipment prior to fabrication of the first fuel production system.

CALCULATIONS

All parametric data and other information for the power conditioning subsystem were obtained from vendor and handbook data.

REFERENCE

1. Final Report Energy Depot Electrolysis Systems Study. Allison Division of General Motors. EDR 3714, 1 June 1964.

Control Subsystem

DESCRIPTION

The control subsystem regulates and coordinates the subsystems during start-up, automatic operation, and shutdown. The control subsystem consists of the instrumentation, signal modification equipment, actuators, switches, and timing devices required to regulate and coordinate the operation of the ammonia production system. The instrumentation measures system temperatures, pressures, flows, voltage, current, and gas purity. The signal modification equipment performs the scaling, integration, and differentiation required to produce the proper type actuator input signals. The actuators position valves and switches to perform the regulatory functions required in system limitation and coordination.

Since the hydrogen production subsystem uses the majority of the available power, a control mode was selected in which the hydrogen production rate is established by limits of the system—normally the electrical power available. The other subsystems are coordinated to maintain or utilize the hydrogen production.

The overall control mode design is shown in Figure 3.6-1. This control mode provides the maximum ammonia output from the power available and coordinates subsystem operation so that buffer storage is held to a minimum. Description of the control mode operation follows.

The d-c voltage to electrolyzer module banks in the hydrogen production sybsystem is increased until the power consumed by the entire ammonia production system matches the power available to it. The amount of hydrogen production is regulated by the amount of power available to the hydrogen production subsystem. Hydrogen flow is measured and the nitrogen supplied to the ammonia synthesis subsystem is controlled to provide the proper hydrogen-nitrogen ratio. The water purification process rate is controlled by the water level in a buffer storage tank. The synthesis pressure in the ammonia subsystem increases with an increase in synthesis gas flow rate resulting in an increased synthesis rate. The ammonia production rate is thus regulated to make use of all the synthesis gas supplied to the ammonia subsystem.

Hydrogen Production Subsystem Control

Although the hydrogen production rate is determined by power available to the hydrogen production subsystem, the subsystem operating parameters must be regulated to ensure proper operation. The major parameters are minimum hydrogen pressure, differential liquid level in the liquid-gas separator-reservoirs, the average liquid levels in the liquid-gas separator-reservoirs, and the electrolyte temperature.

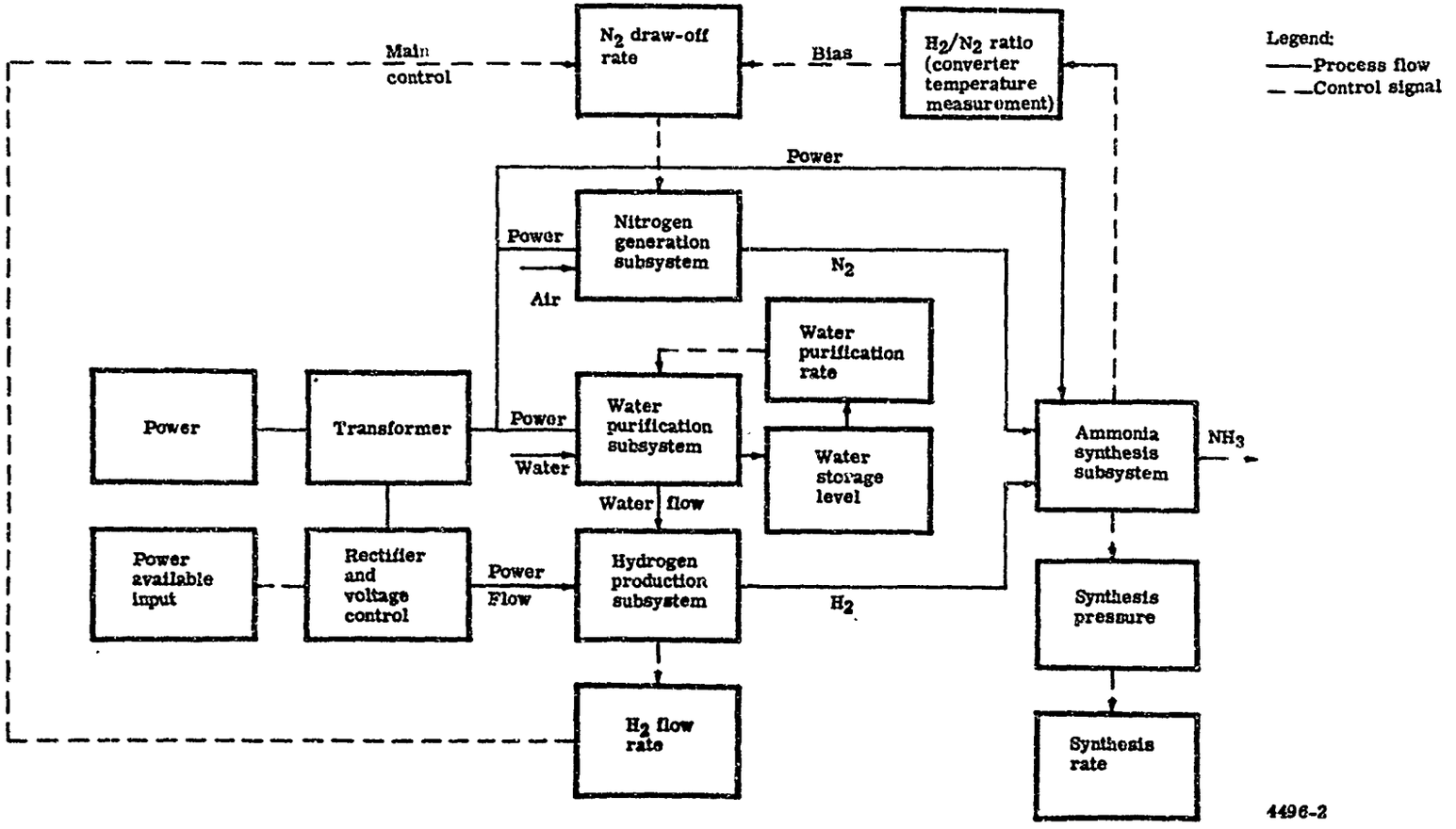


Figure 3.6-1. Overall fuel production system control.

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Hydrogen production subsystem pressure is controlled at 315 psia by regulating the hydrogen efflux. Subsystem pressure is maintained by readmitting high pressure discharge gas from the synthesis gas compressor into the compressor inlet as required by the hydrogen pressure controller. Differential liquid level in the electrolyte-oxygen and the electrolyte-hydrogen separator-reservoirs is controlled by regulating oxygen efflux. Liquid levels in the separator-reservoirs are maintained by off-on feedwater pump control.

The void fraction (gas bubbles in the electrolyte) increases with increase in current to the hydrogen production subsystem. This causes liquid level in the separator-reservoirs to increase as power increases. To minimize the size of the separator-reservoir tanks, a bias is included on the liquid level controller which is proportional to the power supplied for electrolysis. This bias adjusts the limits for feedwater addition to avoid overfilling the separator-reservoirs when the electrolyzer is operating at partial power levels.

The temperature of the electrolyte is controlled by the flow of air (or water) through an electrolyte-to air (or water) heat exchanger. The temperature of the electrolyte entering the electrolyzer is controlled to within a few degrees of 167°F.

Nitrogen Generation Subsystem Control

The nitrogen generation subsystem is controlled to supply enough nitrogen to provide a stoichiometic hydrogen-nitrogen ratio for feed gases to the ammonia synthesis subsystem. Level, pressure, temperature, and purity controls are provided.

The nitrogen product drawoff rate control is the only controller in the nitrogen system that must be coordinated with the remainder of the subsystems. Several other parameters are controlled; specifically, the liquid levels in the low pressure and the high pressure air fractionation column reboilers are controlled to compensate for changes in the drawoff rate or ambient conditions. The low pressure column liquid level is controlled by varying the amount of oxygen from the hydrogen subsystem which bypasses the expansion engine thereby controlling the amount of refrigeration. An increased portion of the oxygen stream bypasses the expansion engine to raise the temperature in the column and lower the liquid level. The liquid level of the high pressure column is controlled by varying the quantity of liquid withdrawn and sent to the high pressure column condenser.

Water Purification Subsystem Control

Since water storage is provided, the water purification subsystem need not be controlled to keep close pace with the fuel manufacturing system. However, variables within the subsystem must be controlled to ensure proper operation. The variables controlled include temperature, pressures, and flows throughout the subsystem.

Power Conditioning Subsystem Control

The major function of the power conditioning subsystem control is to match the electrical power input to the fuel production system to the power available. The power input to the fuel production system is controlled by increasing or decreasing the d-c voltage applied to the electrolyzer module banks, thereby varying power input to the modules.

Ammonia Synthesis Subsystem Control

The ammonia synthesis subsystem requires control of the hydrogen and nitrogen feed rates as well as subsystem pressures, liquid levels, and synthesis temperature.

The ammonia subsystem is controlled to utilize all of the hydrogen produced by the hydrogen production subsystem to synthesize ammonia. In this requirement, the ammonia synthesis subsystem is self-regulating. When more hydrogen is available (and the nitrogen input is controlled to provide a stoichiometric ratio in the feed gas stream), the pressure in the ammonia synthesis loop increases to the point that the synthesis rate equals the feed rate of the synthesis gases.

To provide proper operation of the plant, it is necessary to control the liquid levels in the primary, secondary, and product ammonia separators.

The ammonia synthesis rate can be optimized by controlling the molar ratio of hydrogen to nitrogen contained in the synthesis loop to approximately 2.5 to 1 (this is not to be confused with the feed ratio which must average 3 to 1). This is accomplished by measuring the temperature difference across one of the converters at time t, repeating the measurement at time (t + x) and adjusting the nitrogen flow. A greater temperature difference in the converter implies a greater ammonia conversion rate. The nitrogen flow is biased at each time interval in the direction of greater temperature and higher ammonia synthesis rate until optimum is reached.

INVESTIGATION

Two general types of controllers were considered for the fuel manufacturing plant. The first was a pneumatic control system composed of pneumatic instrumentation, pneumatic control computation, and pneumatic actuators for the valves. The second was an electric and electronic control system in which all of the measurement, computation, and actuation are done with electrical power. A variation of this system using electrical-hydraulic valve actuators was also considered.

Both pneumatic and electrical systems are capable of controlling the individual subsystems and the plant as a whole. A major part of the chemical process industry uses pneumatic control systems. Their systems have proven to be both versatile and reliable. The transducer which couples an air signal to a mechanical motion is simple and trouble-free; on the other hand, the coupling of an electrical control system to a digital computer is simpler than it is with a pneumatic system. No digital programmer was considered in this study; therefore, this advantage is not important.

The pneumatic system was selected because of the availability of pneumatic control components and because plants similar to the subsystems have been built using pneumatic controls. In addition, valve pneumatic drivers require very little power to position them. In hydraulic systems the viscosity of the hydraulic fluid at extremely low temperatures would present a problem during start-up.

An exception to the use of pneumatic control is the control of electrical power to the electrolysis unit. This part of the control subsystem consists of electronically driven silicon control rectifiers.

DESIGN PARAMETERS

The range of ambient conditions for which the fuel manufacturing subsystem must operate was considered in the specification of the control subsystem. Other parameters considered were the temperature pressures, and flows that must be controlled during start-up and at power operation.

DESIGN

The overall function of the control subsystem is presented in the Description subsection. The control subsystem provides regulation and coordination of flows and pressures of the fluids that flow from one subsystem to another. In addition to this coordination task, the control subsystem regulates temperatures, pressures, flows, and levels so that each subsystem will operate properly.

Pneumatic controllers and pneumatic valve actuators which are used in the control subsystem are designed to operate satisfactorily over the full range of ambient conditions to be encountered by the fuel manufacturing plant. The connectors used in the air lines will be of the type that seal when the connection is broken, thus preventing the accumulation of moisture during the transport of the system. Moisture would cause difficulty at subfreezing temperatures.

Each controller is adjustable and may be set at a gain that is acceptable for all ambient conditions. Care must be taken in optimizing the control subsystem gains at design conditions to be sure that these control gains produce adequate control for off-design operation.

A small control air compressor and air dryer-purifier are included at the control console. About 60 lb/hr of control air is required for the 3 Mw systems. This increases to about 100 lb/hr of control air for the 15 Mw systems due to an increase in the number of controllers and a slight increase in actuator sizes. The air dryer-purifier is included to remove moisture and impurities from control air.

If system operation is changed by reduced power operation or by off-ambient temperature or pressure operation, controls are provided to ensure adequate control. An example of such a control is the reduction of cooling air to the heat exchanger on the nitrogen generation subsystem air compressor discharge during subfreezing ambient operation.

Figure 3.6-2 is a control diagram of the fuel production system. The major instrumentation and control points are shown. The letter designations that are used for control and instrumentation and called out on the control system schematic are defined in Table 3.6-I. Major control points are identified for each subsystem and referenced in the discussion that follows.

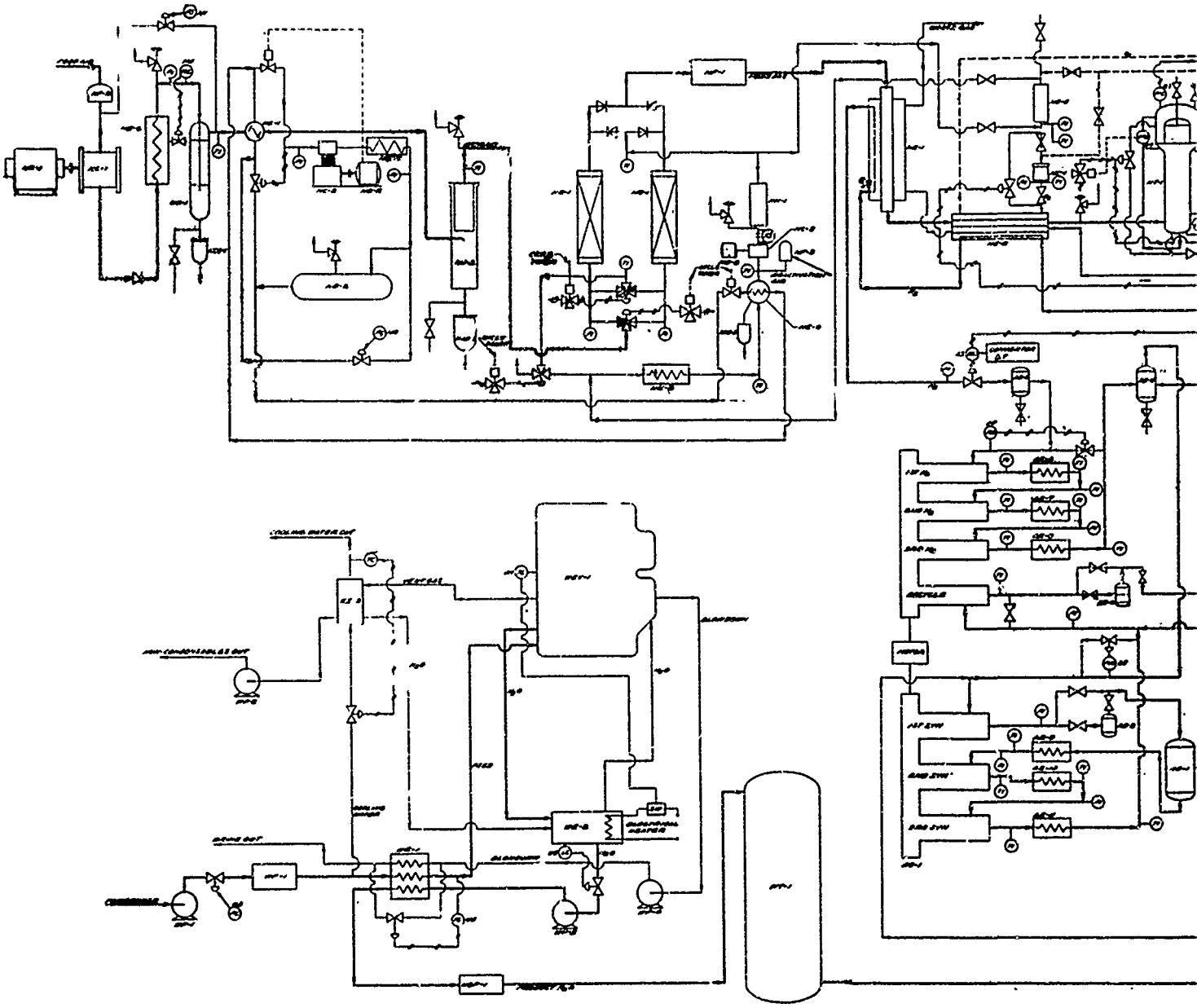
Table 3.6-I.
Letter designations for controls and indicators shown in the control schematic (Figure 3.6-2).

Process variable	Transmitter	Controller			Measuring		
		Indicating	Recording	Blind	Indicating	Recording	Switches
Flow	FT	FIC	FRC	FC	FI	FR	FS
Pressure	PT	PIC	PRC	PC	PI	PR	PS
Temperature	TT	TIC	TRC	TC	TI	TR	TS
Level	LT	LIC	LRC	LC	LI	LR	LS
Gas analysis	GAT	GAIC	GARC	—	GAI	GAR	—
Differential pressure	DPT	DPIC	DPRC	DPC	DPI	DPR	DPS

Blanks indicate improbable useage.

Water Purification Subsystem Control

A water storage tank makes it possible to operate the water purification subsystem without control interdependence between it and other subsystems. The water storage permits intermittent operation of the water purification subsystem when the fuel production system is operating at less than full power. The storage also provides water to the hydrogen production subsystem during plant start-up.



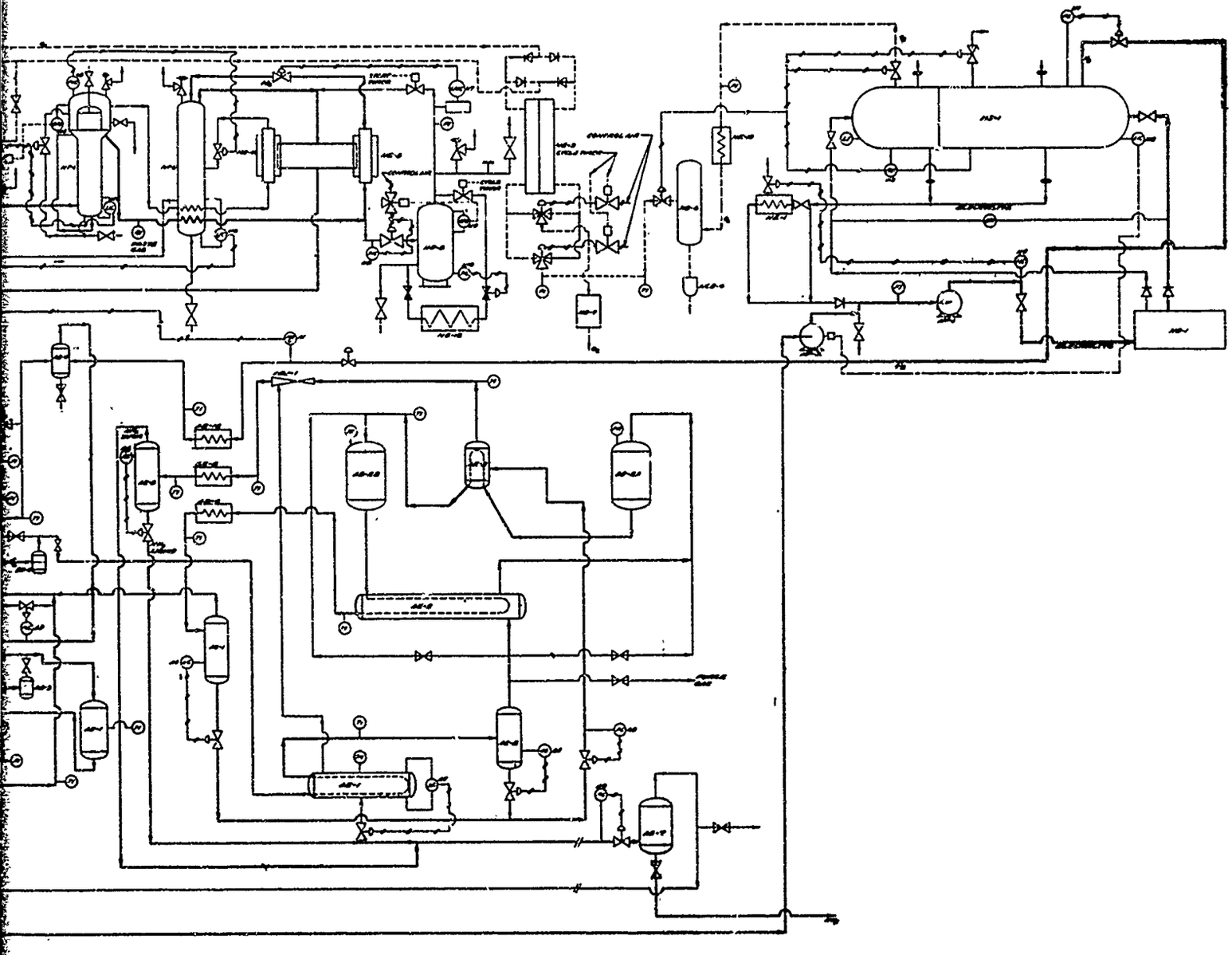


Figure 3.6-2. Fuel production system control schematic.

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The water purification subsystem is a commercially available unit which contains the controls required for stable operation. The principal instrumentation and control points are indicated on the control system schematic. The major control points are numbered on the schematic.

The following functions are performed by the water purification control system.

- Evaporation temperature controller (W1)—The evaporation temperature is controlled by intermittent operation of the electric heaters.
- Level controller (W2)—The immersion heater tank water level is controlled by the water flow rate out of the tank.
- Feedwater pressure controller (W3)—The feedwater pressure is controlled by the pressure drop across a valve in the feed line.
- Cooling water temperature controller (W4)—The cooling water outlet temperature is controlled by the blowdown flow to the vent gas condenser.
- Preheater water temperature (W5)—The water temperature at the outlet of the preheater is controlled by bypassing part of the blowdown.
- The water storage tank level is controlled by switching the unit off when the tank is full (control not shown on schematic).

Hydrogen Production Subsystem Control

The overall functions of the hydrogen subsystem control are to provide a flow of hydrogen to the ammonia synthesis subsystem at a prescribed temperature and pressure, and to provide oxygen for an expansion engine for nitrogen generation subsystem refrigeration. These are accomplished by means of pressure, liquid level, and temperature controls, both within the subsystem and on the gas discharge lines. The following valves and controllers are required for operation and control of the hydrogen subsystem.

- Minimum hydrogen pressure controller (H1)—This controller monitors the hydrogen pressure in the separator-reservoir tank and actuates a valve in the hydrogen gas discharge line to keep the hydrogen pressure above ~250 psi. This control will be used during system start to build up pressure in the separator-reservoir and will protect the subsystem against pressure loss from broken lines between skids.

- Level controller (H2)—A level controller is connected to the hydrogen side of the separator-reservoir tank. The liquid level is maintained by adding feedwater to replace that which is consumed by electrolysis. The level controller has high and low level limit switches which govern feedwater addition by actuating the feedwater pump motor. The level to which the electrolyte is controlled is biased by the power input.

- Differential level control valve (H3)—Differential in liquid levels between the hydrogen and oxygen separator-reservoirs is determined and controlled by a pneumatic differential level controller actuating a valve on the oxygen efflux.

- Temperature control valve (H4)—Temperature of electrolyte to the electrolyzer is determined and controlled by a pneumatic temperature controller. A pneumatic temperature control valve is supplied for regulation of coolant flow through heat exchangers which dissipate heat generated by electrolysis. Electrolyte temperature is measured in the process stream near the module inlet. For the air-cooled systems, the controller actuates a louver system which regulates airflow across the heat exchanger. A magnetic switch, which operates in conjunction with the louvers, turns off one of the fan motors when the louvers reach a preset, partially shut position. (The louvers then automatically return to open position.) This prevents the air pressure drop across the heat exchanger from becoming excessively high, a condition which could cause fan motor damage. Thus, both discrete and variable temperature control is available. Systems 1, 3, and 4 have one air-cooled heat exchanger. For the subsystems with more than one heat exchanger (systems 5, 7, 8, 9, 10, and 11), a single temperature sensor and controller is required, but actuators and switches are needed for each set of louvers. For the water-cooled systems (systems 2 and 6), the water flow to the heat exchanger is regulated by a valve on the water inlet.

- System pressure controller (A3)—A pressure controller senses the pressure in the hydrogen-gas discharge line and controls the pressure by a valve located on the ammonia skid. This valve bleeds high pressure gas from the compressor efflux back into the compressor feed. (This controller is located on the ammonia package No. 3.)

  Systems 5 through 11 require more than one separator-reservoir in their designs. Systems 6 and 9 have two separator-reservoirs, one each for hydrogen and oxygen separation. These tanks have been positioned so that during operation their liquid levels are the same. Also, the electrolyte flow from each tank is combined. Systems 6 and 9 thus act as if there were one separator-reservoir and require no control modifications. Systems 5, 7, 8, 10, and 11 each require three separator-reservoirs, one for oxygen, and two for hydrogen separation. As before, these tanks are positioned for the same liquid levels. The hydrogen separation tanks are joined by large diameter pipes to effectively function as a single tank. Therefore, multiple controls are not needed. The same number of control components suffices for all eleven cases. Only actuator and valve sizes will change.

- Miscellaneous hydrogen production subsystem controls—In addition to the discussed controls, several safety devices are provided. A subsystem overpressure switch is included which will automatically shut the plant down if excessive subsystem pressure occurs. Pressure sensitive rupture disks are provided on the hydrogen and oxygen separator tanks. These disks are set to rupture at a pressure greater than that which will cause the overpressure switch previously mentioned to actuate. Therefore, the rupture disks provide protection in case all other pressure controls fail.

Various system parameters are monitored on the control panel by appropriate instrumentation.

- Pressure indicator—panel-mounted gage to indicate pressure in the separator tanks.
- Flow indicator—panel-mounted gage which indicates electrolyte flow rate into the electrolysis modules.
- Temperature indicator—panel-mounted gage to indicate temperature at electrolysis module inlet.
- Differential pressure indicator—panel-mounted gage to indicate the differential in pressure between the hydrogen and oxygen liquid-gas separator-reservoirs.

A central control cabinet serves all of the subsystems; the location of the cabinet is discussed in the Integration and Packaging Subsection.

Power Conditioning Subsystem Control

The major function of the power conditioning subsystem control is the matching of the power used by the fuel production system to the power available. The power being used by the fuel production system is measured using current transformers on the incoming power supply. This measurement is compared with the power available signal. When they differ, an error signal is sent to the pulse timer on the silicon control rectifiers which control the power absorbed by the electrolysis unit. The other principal control on the power conditioning unit limits transformer and rectifier operating temperatures. Automatic system shutdown results from an over-temperature condition.

Nitrogen Generation Subsystem Control

A major requirement for the control of the nitrogen generation subsystem is to regulate the product nitrogen flow rate to provide the flow rate required to coordinate with the hydrogen flow rate. This controller (A1) is discussed under Ammonia Synthesis Subsystem control. The nitrogen generation subsystem control must also be able to compensate for any variation in flow of oxygen from the hydrogen production subsystem. As the flow of oxygen varies, the flow through the expansion engine or the pressure drop across the engine, or both, must change to provide the refrigeration required in this subsystem.

The nitrogen generation subsystem requires a number of controls which do not directly involve the overall system, but are required to ensure proper operation of the nitrogen subsystem. Following is a list of these controllers.

- Pressure controller (N1)—Compressor discharge pressure is controlled by recirculating part of the compressor air to the inlet of the compressor.
- Temperature controller (N2)—Refrigeration evaporation pressure, and hence temperature, is controlled by bypassing some of the refrigerant around the evaporator.

- Pressure controller (. 3)—Pressure of the high pressure column condenser is controlled by regulating the flow to the lower feed on the low pressure column.
- Level control switch (N4)—A differential pressure switch ensures that the condenser on the high pressure column is covered with liquid by over-riding the reboiler control (N5) and allowing full flow from the high pressure column reboiler.
- Level control (N5)—The high pressure column reboiler level is controlled by the rate of liquid flow from the reboiler to the high pressure column reboiler.
- Level control (N6)—Low pressure column reboiler liquid level is controlled by bypassing part of the oxygen from the hydrogen production subsystem around the expansion engine.
- Purity control (N7)—Product purity is controlled by the feed rate at the top of the low pressure column.
- Flow control (N8)—The flow rate control to liquid nitrogen storage is controlled by a valve in the feed line.
- Level control switch (N9)—Level control on the liquid nitrogen storage tank shuts off the flow when the storage tank is full.
- Pressure control (N10)—Liquid nitrogen storage pressure is controlled by passing some nitrogen through a heat exchanger at start-up.
- Temperature control (N11)—Compressor discharge temperature is controlled by airflow across the compressor discharge heat exchanger to prevent freezing of the moisture in the air.

In addition to the major controls shown on the control schematic, the following switches are required to ensure proper plant operation and to provide sequencing during start-up and shutdown.

- Gas analyzer switch—this switch prevents gas from the nitrogen subsystem from being delivered to the ammonia subsystem until desired purity is reached.
- Reactivation airflow switch—this switch prevents the reactivation air heater from being turned on unless there is reaction airflow.
- Temperature switches—one of these switches controls the temperature of the reactivation to recharge the air dryers by switching the reactivation heater off or on. Another switch keeps the process air temperature above freezing by switching the refrigeration compressor off or on.

Several temperature and pressure indicators are located throughout the nitrogen generation subsystem to provide the operator with information on plant operation. These indicators are shown on the control system schematic.

Ammonia Synthesis Subsystem Control

The ammonia synthesis subsystem control regulates pressures, liquid levels, and temperatures to synthesize hydrogen and nitrogen gases to produce ammonia product. The following valves and controllers are required.

- H_2/N_2 flow ratio controllers (A1 and A2)—The incoming nitrogen flow is controlled to maintain the proper hydrogen to nitrogen ratio in the synthesis gas flow to the ammonia synthesis subsystem. A flow controller (A1) monitors the hydrogen flow to the ammonia subsystem and controls the nitrogen flow. In conjunction with the flow controllers is a bias (A2) which maximizes the ammonia production rate by maintaining the optimum synthesis gas ratio in the ammonia converters. The temperature difference across the ammonia converter is monitored. The nitrogen flow into the system is throttled to attain the greatest temperature difference, which implies the greatest production rate. This allows full hydrogen input to the system to be used, which is desirable because of the relatively high power requirements for hydrogen production.

- Level controllers (A4, A5, A6, and A7)—Four liquid level controllers and valves are required for system operation. The primary liquid-gas separator level is controlled by a pneumatic actuated valve (A4) in the liquid outlet line. Similar controllers are required for the secondary (A5) and product (A6) liquid-gas separators. The secondary ammonia condenser uses low pressure and temperature ammonia from the storage stream to condense ammonia in the recycle stream. The cooling ammonia is brought to low pressure by an expansion valve (A7). The liquid ammonia level in the condenser is controlled by the flow through the expansion valve.

- Pressure controllers (A8 and A9)—There is an expansion valve on the ammonia line which feeds the ejector. This valve (A8) throttles the ejector feed from the 6000-psia loop pressure to 3000 psia. There is also an expansion valve (A9) on the gas stream from the product separator to expand the gaseous ammonia to storage pressure.

Several monitors and gages are provided on the control cabinet for visual observation of system operation.

- Temperature indicators—Nineteen temperature indicators are provided, including the inlet and outlet of each system heat exchanger and compressor inter- and aftercoolers and the deoxo unit. Several of these indicators may be panel-mounted lights which indicate in-limit or out-of-limit temperatures.
- Temperature recording indicator—Each ammonia converter will have thermocouples embedded in the catalyst to continuously record and monitor the converter temperature profile.
- Pressure indicators—Ten pressure indicators are used. Each compressor stage and each pressure controller have indicators. An indicator is also included on the secondary ammonia condenser.

PREDICTED PERFORMANCE

The performance of the system was predicted by means of analog computer analysis. The results are reported in the Control Analysis Subsection of this report.

DISCUSSION

Selection of the control modes for operation of each major subsystem was based on simplicity, safety considerations, cost, and reliability of operation. The design of the nitrogen generation and ammonia synthesis subsystems control modes is based on experience obtained in field units used for many years in the chemical processing industry. The hydrogen production subsystem control has been successfully applied to the Allison laboratory model ten-cell electrolysis system developed under Army Contract DA-44-009-AMC-1021(X). This system ran 1500 hr with no control problems and served to prove the control concept. At the time of the preparation of this report a full-scale single electrolysis module test unit is being readied for further testing using the same control mode.

Integration and Packaging

DESCRIPTION

The fuel production system is composed of the following subsystems which have been described previously:

- Ammonia Synthesis
- Nitrogen Generation
- Hydrogen Production
- Water Purification
- Power Conditioning
- Control

The integration and packaging studies combined these subsystems into optimized ammonia production systems which produce ammonia from water and air using electric power. Eleven system designs were developed which satisfy the mobility and power level restraints specified in Figure 3.7-1.

INVESTIGATION

Parametric data developed in the subsystem design studies were used in the overall system integration and packaging studies. The variation in system production capacity was determined for a variation in system weight and cost. In the systems analysis, consideration was also given to packaging restraints.

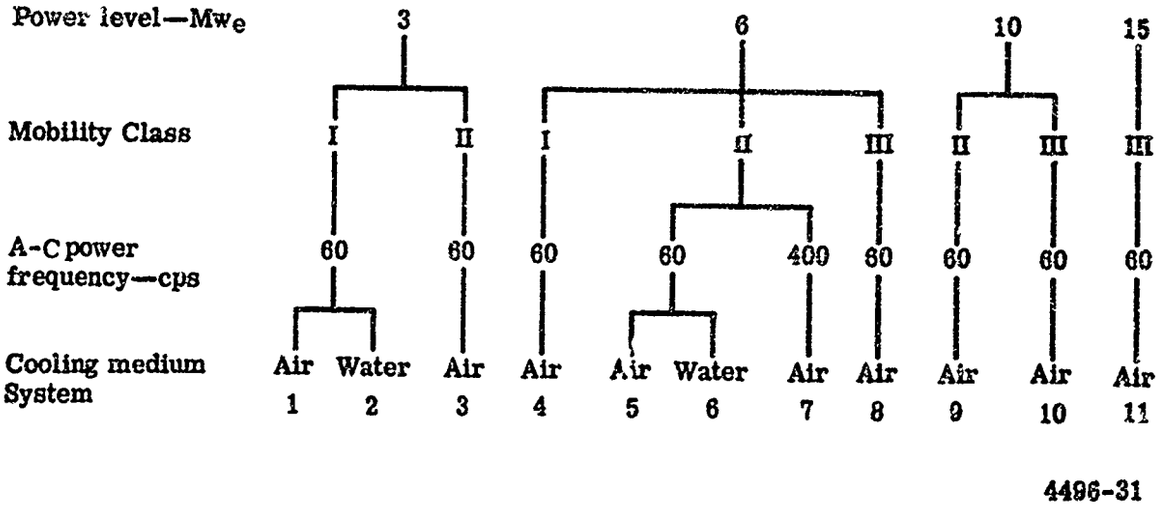


Figure 3.7-1. Ammonia production systems definition

System Analysis

An important limitation imposed on each system was the electrical power input. To increase system production capacity the system efficiency must be increased. Since the power conditioning subsystem transforms and distributes all input electrical power and the hydrogen production subsystem consumes a major portion (approximately 90%) of this power, these subsystems are of primary significance in any system optimization study.

The computer program discussed in the Hydrogen Production Subsystem studies and described in Appendix A was used to optimize the overall system. The power available to the hydrogen production subsystem was system power input less the power conditioning subsystem losses and the power requirements of the other subsystems. The losses in the power conditioning subsystem resulted from transformer and rectifier losses. The rectifier losses were 0.5% of power input to the electrolyzers. The variation in transformer efficiency with weight was determined in the Power Conditioning Subsystem studies and is shown in Figure 3.7-2.

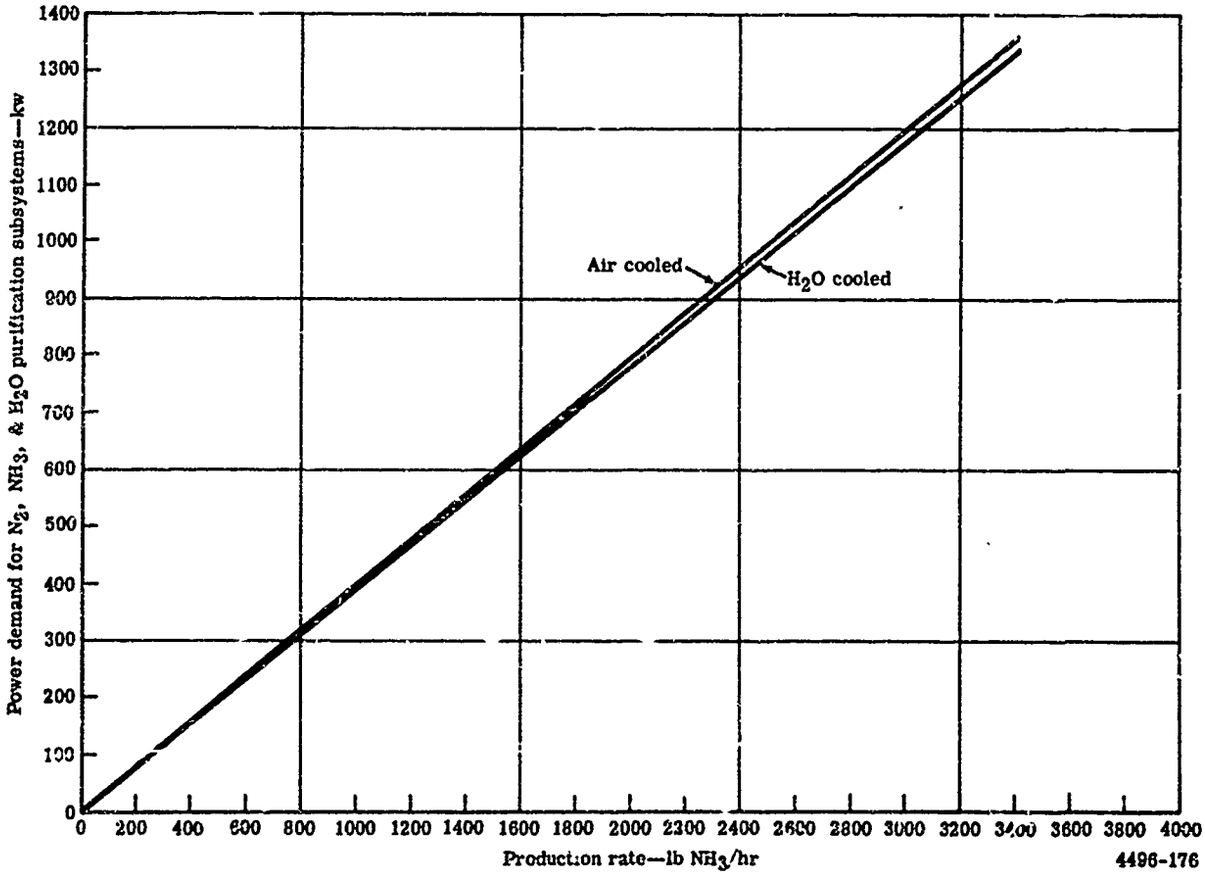


Figure 3.7-2. Power required by N_2, NH_3, and H_2O purification subsystems as a function of NH_3 production rate (based on vendor data).

A preliminary analysis of the cost and weight of the power transformer, when combined with the cost and weight of the nuclear powerplant, indicated that the optimum transformer efficiency ranges from 98.5 to 99% for the 3-Mw_e systems and 98.7 to 99.4% for the 15-Mw_e systems. The efficiency, weight, and dimensions of the transformers used in the final system designs are given in Table 3.7-I. With the exception of system 4, transformer efficiencies are in agreement with those from the optimization study. The transformer for system 4 was affected by packaging considerations and is discussed in more detail later in this subsection.

Table 3.7-I.
Transformer packaging data.

System	Power (Mw_e)	Frequency (cps)	Efficiency (%)	Weight (lb)	Dimensions (in.)
1	3	60	98.9	13,500	101 × 88 × 40
2	3	60	98.9	13,500	101 × 88 × 40
3	3	60	98.9	13,500	101 × 88 × 40
4	3*	60	97.0	7,400*	71 × 70 × 32
5	6	60	99.0	22,200	117 × 90 × 56
6	6	60	99.0	22,200	117 × 90 × 56
7	6	400	99.0	3,200	68 × 60 × 26
8	6	60	99.0	22,200	117 × 90 × 56
9	5*	60	99.0	20,200*	116 × 87 × 56
10	10	60	99.1	35,200	135 × 117 × 54
11	15	60	99.1	49,600	150 × 130 × 61

*One of two transformers in system.

The power consumption of the ammonia synthesis, nitrogen generation, and water purification subsystems was determined over a range of ammonia production rates and was based on the processes recommended in the subsystem design studies. The power requirements are shown in Figure 3.7-2. In general, the water-cooled subsystem requires less electrical power than air-cooled subsystems because the cooling water pump power requirements are less than fan power requirements.

The overall fuel production system was optimized by integrating the several subsystems through the use of the computer program discussed in Appendix A. The output of the integrated system was varied by varying the efficiency of the hydrogen production subsystem (Figure 3.3-6). The output, weight, and power requirements of the other subsystems were varied in a manner compatible with the hydrogen production rate. The total power input was specified at the 3-, 6-,

10-, or 15-Mw_e level for each fuel production system. Many interations were made to determine and evaluate the effect of fuel production system efficiency and output. Other considerations such as packaging arrangements and total system and NPED (including nuclear powerplant) weight and cost were significant in the design point selection for each system.

The final system cost optimization results for the eleven systems are presented in the Supplement to this report. The system design points which were based on preliminary analysis are indicated. With the exception of system 3 the specific cost of the NPED, nuclear powerplant, and fuel production system is within one percent of the theoretical optimum design point. System 3, which is the air-cooled, Mobility Class II system with a power input of 3 Mw_e, appears to be oversized. No advantage to the relaxation of mobility restraints is apparent in the 3-Mw_e plants. System 1, therefore, would be recommended for a 3-Mw_e NPED even if a package weight greater than 35,000 lb were permitted.

The weight and cost optimization results presented in the Supplement indicate the significant effect of the nuclear powerplant on fuel production system optimization. With the weight and cost of the powerplant included the optimum system design is considerably larger and more efficient than system designs optimized solely on fuel production system weight or cost.

Packaging Arrangements

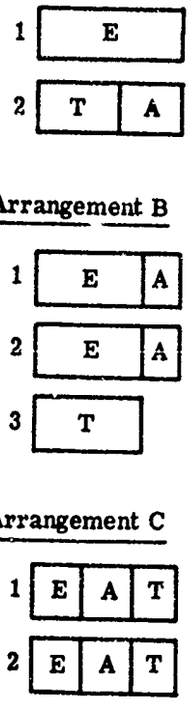
In the investigation of packaging arrangements, consideration was given to the following general criteria.

- The weight and size of individual packages shall not exceed the limitations specified for the mobility class.
- The number of separate packages shall be minimized by combining two or more subsystems on the same skid.
- Plant assembly and disassembly times shall be minimized by limiting the number of large process and electrical interconnections between packages.

When a conflict arose between the last two criteria and/or optimum size and rating of system, engineering judgement was exercised in the selection of packaging arrangement.

In general, the hydrogen production subsystem was the heaviest of the several subsystems. The weight of the optimum hydrogen production subsystem approached or exceeded the package weight limitation. Systems 4 and 9 presented a unique packaging arrangement problem. Each of these two systems had a high power input relative to the package weight limitation. In both cases the optimum hydrogen production subsystem weight significantly exceeded the package weight limit. Several alternate packaging arrangements were considered; three arrangements for the hydrogen production subsystem and the power transformer are indicated in Figure 3.7-3 with comparative data for system 4.

	System capacity (lb NH_3/hr)	Transformer efficiency	Relative specific cost	Total No. of packages	Package weight (lb) 1	2	3
Arrangement A	1,360	98.9%	0.98	5	35,000	35,000	---
Arrangement B	1,415	99.0%	0.985	6	28,500	28,500	22,200
Arrangement C	1,380	97.0%	1.0	5	35,000	35,000	---



E = Electrolyzer module banks
A = Hydrogen production subsystem auxiliaries
T = Power transformer

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Figure 3.7-3. Alternate packaging arrangements – system 4.

In arrangement A of Figure 3.7-3 only two packages are required (five packages for the entire system) and a transformer of high efficiency, 98.9%, can be used. However, the ammonia output is lowest with arrangement A because the electrolyzer weight is limited by package weight limitations. In addition, relatively large electrolyte piping and electrical wiring interconnections must be made between packages.

In packaging arrangement B, two 50% capacity hydrogen production subsystems are used with a separate 6-Mw transformer package. This arrangement produces the most efficient system, but three packages are required (six packages for the fuel production system).

In packaging arrangement C, two 3-Mw transformers and two 50% capacity hydrogen production subsystems are combined to form two identical packages. Some sacrifice in output is required as a result of the reduced transformer and electrolyzer weights. However, major interconnections between packages are minimized.

Packaging arrangement C was selected as a result of an engineering evaluation. The tactical advantage of one less package with fewer interconnections was judged more advantageous than the slightly increased output (2.4%) available with arrangement B. Packaging arrangement A was eliminated because of numerous large package interconnections.

A comparable analysis was made for system 9 and a similar packaging arrangement was selected.

Since the hydrogen production subsystem consumes approximately 90% of the electrical power input and requires relatively large interconnections with the transformer, the hydrogen subsystem and the transformer were packaged together when weight limitations permitted. In addition to systems 4 and 9, discussed previously, system 7 also combines the hydrogen subsystem and the transformers on the same skid.

The nitrogen generation subsystem is a relatively lightweight subsystem and never requires more than one package. Other subsystems, the water purification subsystem and/or the power transformer, are packaged with the nitrogen generation subsystem in several systems. Some revision of the package layouts was necessary when the subsystems were packaged together.

The ammonia synthesis subsystem is somewhat heavier than the nitrogen generation subsystem. In systems 4, 9, 10, and 11, the ammonia subsystem size or weight exceeded the package limitations. Two alternate arrangements were considered for these four systems. First, two 50% capacity subsystems were packages separately. Second, the subsystem was split into two series-connected packages with high pressure equipment on one skid and low pressure equipment on a separate skid. The second arrangement was selected because it resulted in less weight.

Components from other subsystems such as the control cabinet and the cooling water filter were located on the same skid with the ammonia synthesis subsystem when weight limitations permitted.

In the ammonia synthesis and nitrogen generation subsystem studies, two alternate methods of compressing the 17-psia nitrogen product from the air fractionation columns to 315 psia for mixing with the hydrogen product of the electrolyzer. One method used a separate nitrogen compressor on the nitrogen generation subsystem skid and a second method used added stages on the multiservice compressor on the ammonia synthesis subsystem skid. These alternates were evaluated by comparing the cost and weight of the separate nitrogen compressor with the incremental increase in cost and weight of the multiservice compressor with nitrogen stages added. See Table 3.7-II.

In the 3 Mw_e systems, the added stages on the multiservice compressor method was selected because this method has lower cost and weight.

Table 3.7-II.
Alternate methods of nitrogen compression.

System	Power level (Mw_e)	Added stages on multiservice compressor versus separate nitrogen compressor	
		Relative weight	Relative cost
1	3	0.25	0.87
2	3	0.23	0.96
3	3	0.25	0.87
4	6	0.38	1.20
5	6	0.38	1.20
6	6	0.38	1.30
7	6	0.38	1.10
8	6	0.38	1.20
9	10	0.50	1.70
10	10	0.50	1.70
11	15	0.38	1.60

In the 6 Mw_e systems the added stages on the multiservice compressor method was again selected because this method has a lower weight although the cost is slightly higher.

In the 10- and 15-Mw_e systems, the separate nitrogen compressor method was selected because this method has significantly lower cost although the weight is higher.

The distribution of components and subsystems between packages is discussed in the design portion of this subsection.

DESIGN PARAMETERS

The eleven ammonia production systems were designed to meet the requirements indicated in Section II. The package size and weight limitations are given in Table 3.7-II.

The following objectives were established for system design.

- The plant costs, including the cost of the nuclear powerplant supplying electrical power, per unit of fuel production rate design capability will be minimized.
- The number of separate packages and package interconnections will be minimized.
- The reliability and maintainability of the systems will be maximized.

DESIGN

The equipment layouts for the separate packages which make up the integrated fuel production systems are shown in Figures 3.7-4 through 3.7-21. The layouts of the packages containing the hydrogen production subsystem, Figures 3.3-11 through 3.3-21, did not require revision and are not presented here.

Figures 3.7-22 through 3.7-32 show the location of the various subsystems on the packages. The package interconnections are listed and identified. The number of package interconnections varies from 17 in systems 1, 3, 7, and 8 to 40 on system 11. As anticipated, an increase in the number of separate packages increases the number of package interconnections; the use of water cooling requires additional water coolant interconnections; an increase in power input requires additional electrical interconnections. The location of the transformers in the same package with the hydrogen production subsystem reduces the number of interconnections between packages.

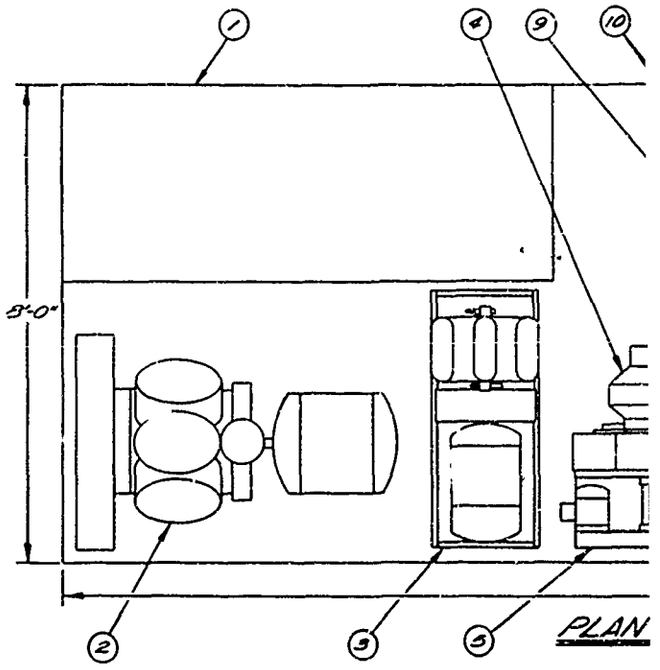
The package interconnections are flexible lines or cables, depending on their function, and are permanently connected at their origin. For example, the power cables will be flexible, multi-conductor, neoprene jacketed cables permanently attached to the transformer secondary. By permanently securing one end of the cables, start-up and shutdown procedures can be simplified. Pneumatic control lines will be bundled with several tubes contained in each bundle. In bundles where water vapor may condense and freeze, provisions will be made for internal heating wires. On the terminus of each line, quick-disconnect fittings will be used. The design of these fittings will include seals to prevent complete loss of pressure when disconnected. By virtue of their large currents, the power cables to the electrolyzer will be of heavy-duty construction. Rigid mechanical connection will be made at the rectifier on the hydrogen package to ensure adequate electrical contact and provide necessary structural support.

Pertinent information on the eleven integrated fuel production systems is summarized in Table 3.7-III. All packages meet size and weight limitations for the appropriate mobility class. In seven of the eleven systems, the subsystems have been combined into three packages. No system required more than five packages.

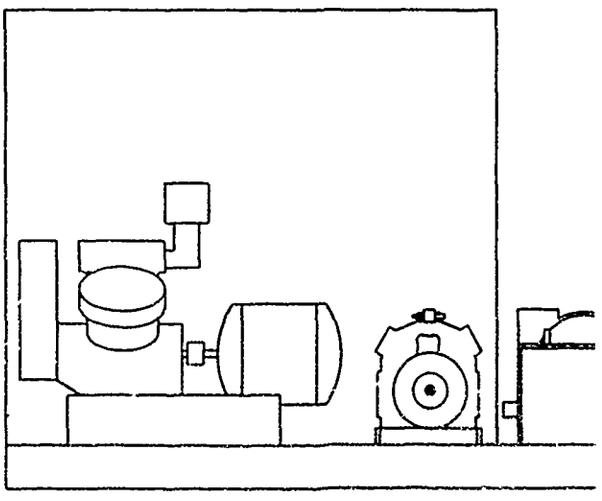
Tables 3.7-IV, 3.7-V, and 3.7-VI provide a weight breakdown for each of the packages in the eleven fuel production systems. The weight of each subsystem for the eleven systems is given in Table 3.7-VII.

LEGEND

1 TRANSFORMER
2 AIR COMPRESSOR
3 REFRIGERATION UNIT
4 AIR COOLED HEAT EXCHANGER
5 FAN & DRIVE UNIT
6 COLD BOX
7 EXPANSION TURBINE
8 EXPANDER-ACCUMULATOR
9 O_2 HEAT EXCHANGER
10 MAIN HEAT EXCHANGER
11 REFRIGERATED HEAT EXCHANGER
12 HIGH PRESSURE COLUMN
13 CRUDE O_2 SUBCOOLER
14 N_2 SUBCOOLER
15 LOW PRESSURE COLUMN
16 WATER PURIFIER
17 AIR PURIFIER
18 LIQUID N_2 STORAGE TANK
19 WATER STORAGE TANK



PLAN



SID

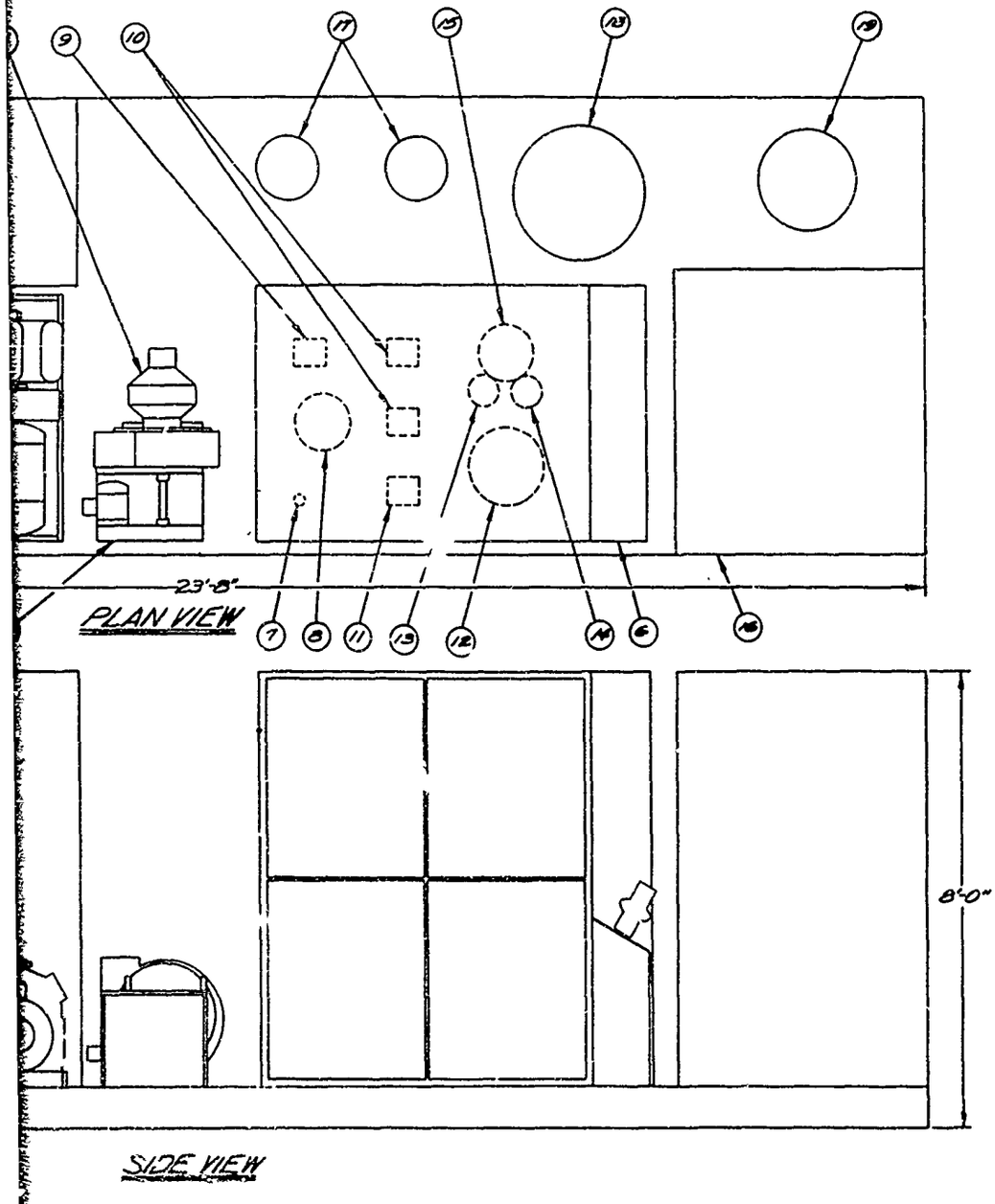
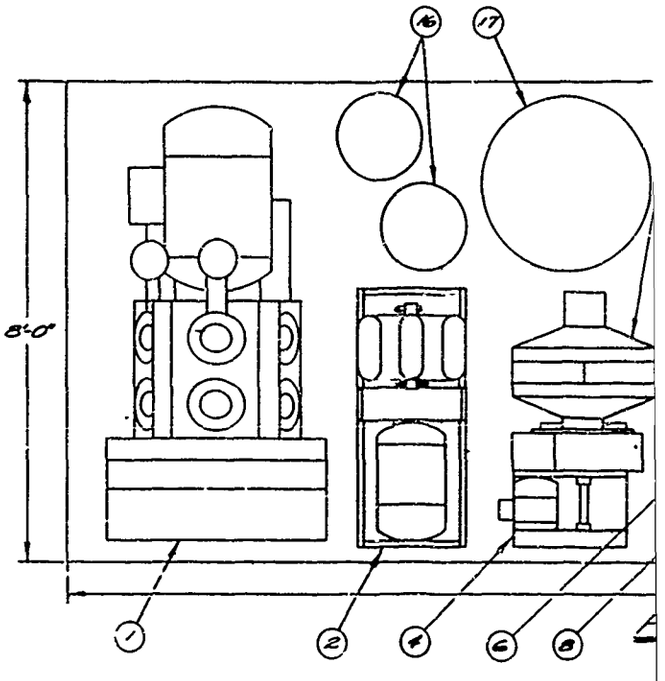


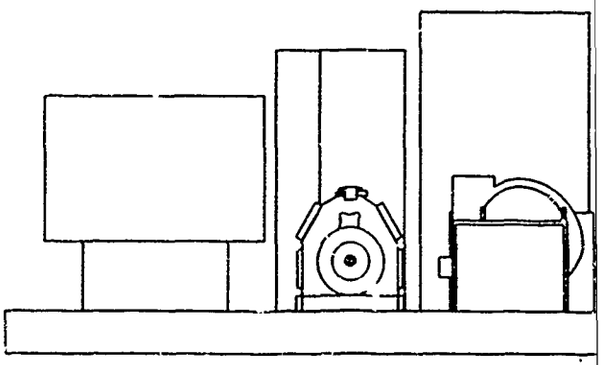
Figure 3.7-4. Nitrogen package layout – systems 1, 2, and 3 – Package 2.

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LEGEND

1 AIR COMPRESSOR
2 REFRIGERATION UNIT
3 AIR COOLED HEAT EXCHANGER
4 FAN & DRIVE UNIT
5 COLD BOX
6 EXPANSION TURBINE
7 EXPANDER-ACCUMULATOR
8 O_2 HEAT EXCHANGER
9 MAIN HEAT EXCHANGER
10 REFRIGERATED HEAT EXCHANGER
11 HIGH PRESSURE COLUMN
12 CRUDE O_2 SUBCOOLER
13 N_2 SUBCOOLER
14 LOW PRESSURE COLUMN
15 WATER PURIFIER
16 AIR PURIFIER
17 LIQUID N_2 STORAGE TANK
18 WATER STORAGE TANK





Figure

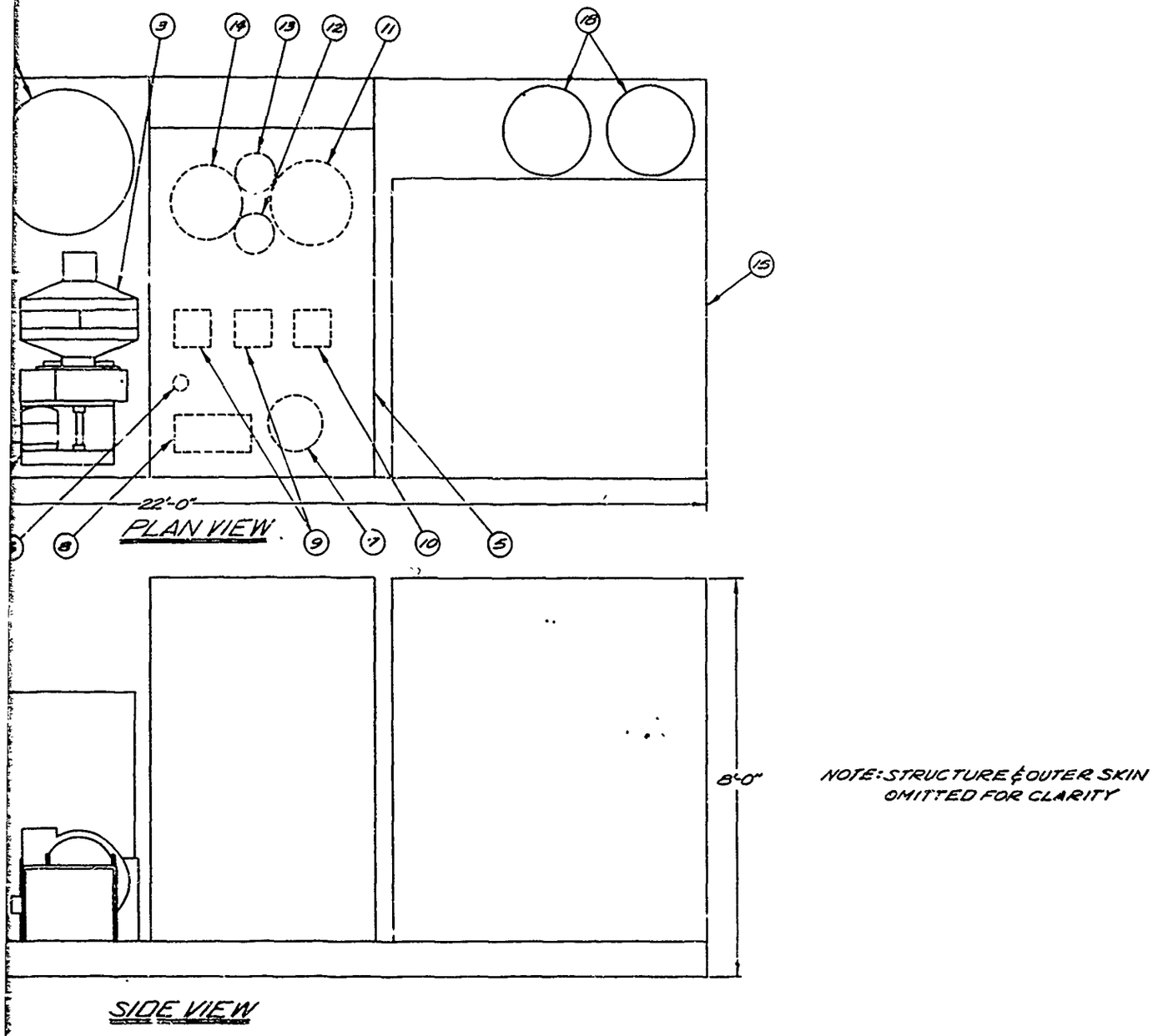
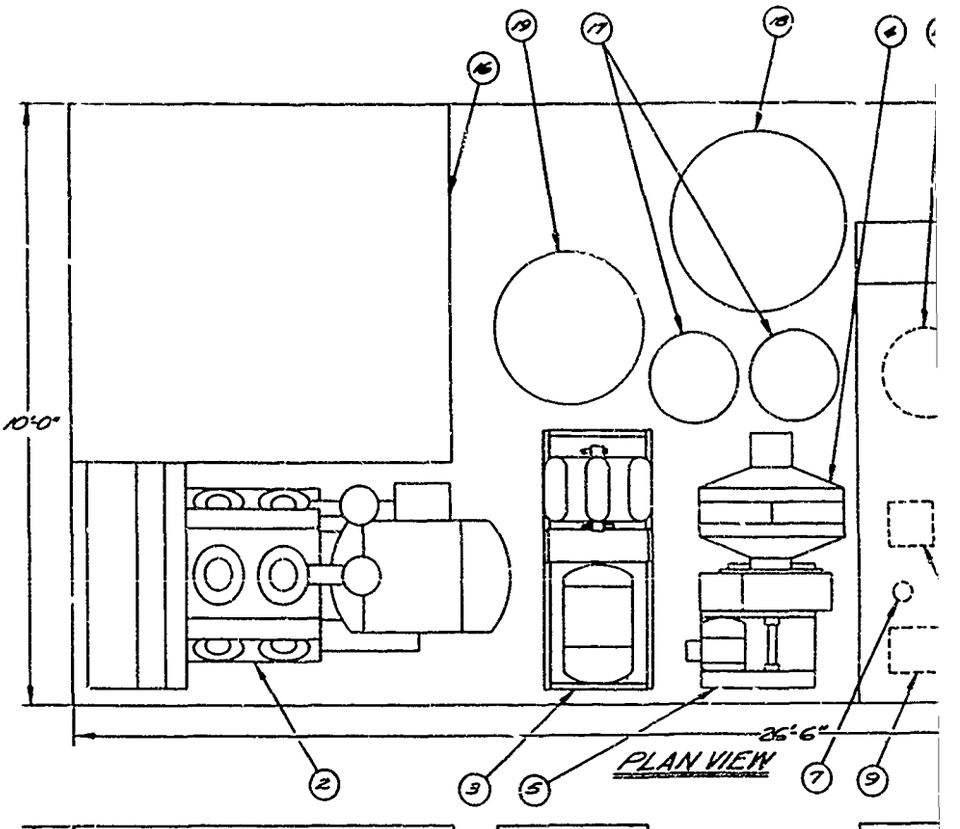


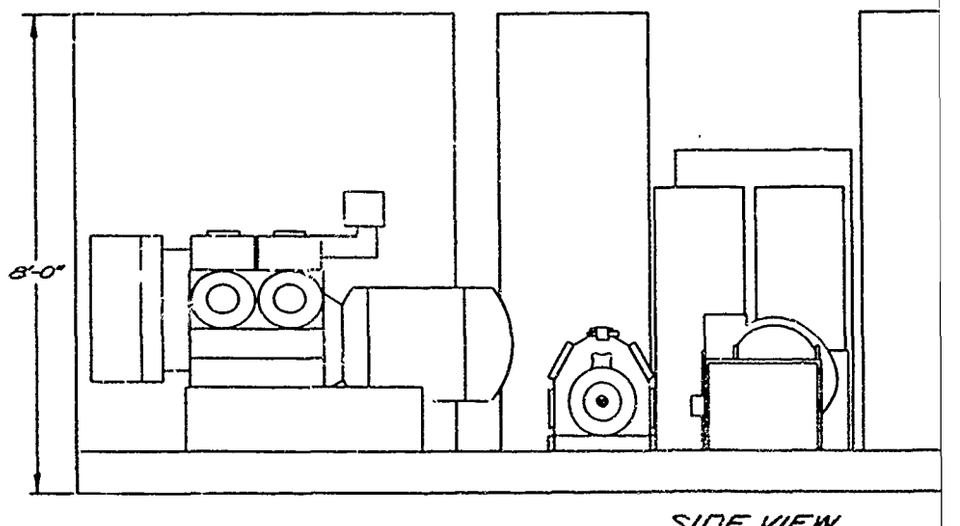
Figure 3.7-5. Nitrogen package layout – systems 4, 7, and 8 – Package 2.

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LEGEND

1 TRANSFORMER
2 AIR COMPRESSOR
3 REFRIGERATION UNIT
4 AIR COOLED HEAT EXCHANGER
5 FAN & DRIVE UNIT
6 COLD BOX
7 EXPANSION TURBINE
8 EXPANDER-ACCUMULATOR
9 O_2 HEAT EXCHANGER
10 MAIN HEAT EXCHANGER
11 REFRIGERATED HEAT EXCHANGER
12 HIGH PRESSURE COLUMN
13 CRUDE O_2 SUBCOOLER
14 N_2 SUBCOOLER
15 LOW PRESSURE COLUMN
16 WATER PURIFIER
17 AIR PURIFIER
18 LIQUID N_2 STORAGE TANK
19 WATER STORAGE TANK





SIDE VIEW

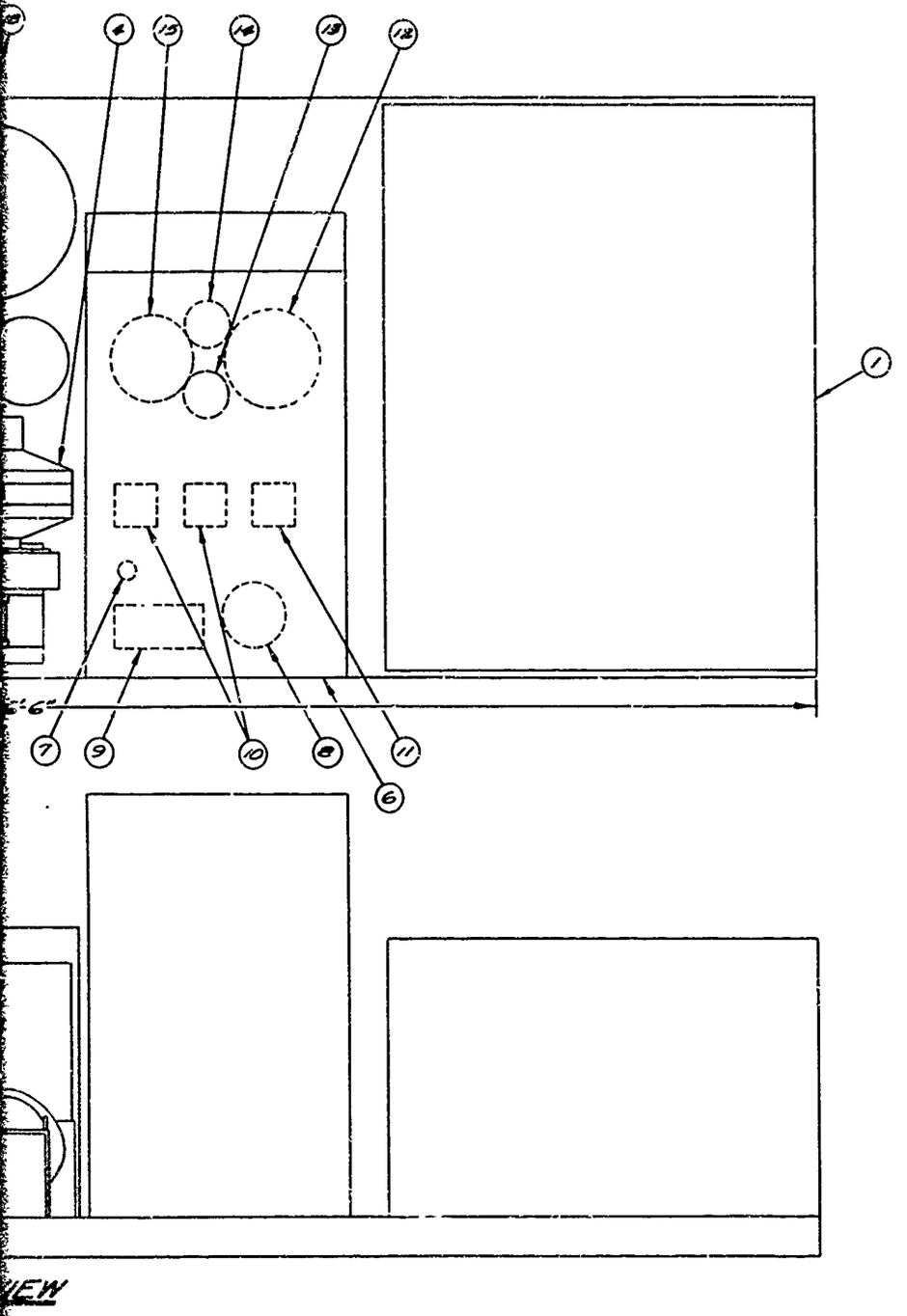
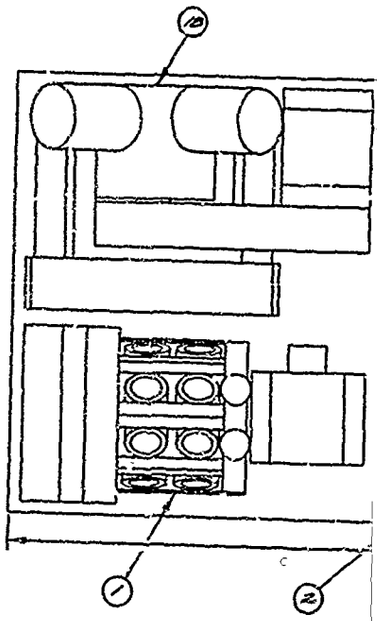


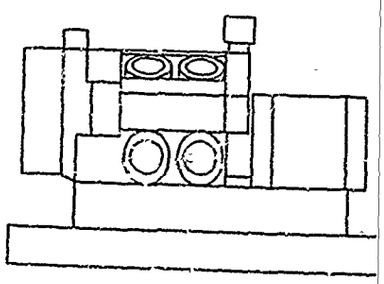
Figure 3.7-6. Nitrogen package layout – systems 5 and 6 – Package 2.

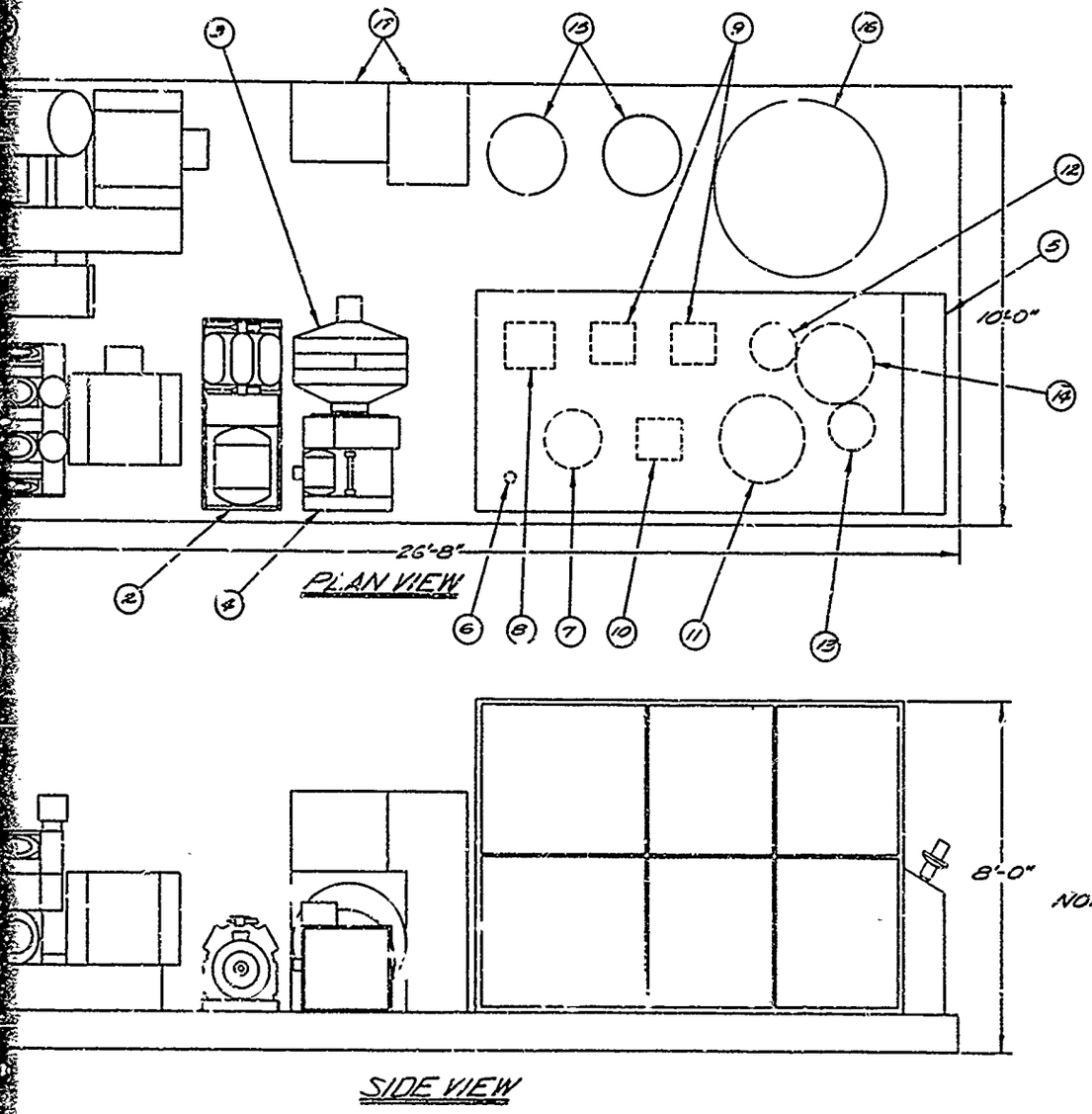
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LEGEND

1 AIR COMPRESSOR
2 REFRIGERATION UNIT
3 AIR COOLER HEAT EXCHANGER
4 FAN & DRIVE UNIT
5 COLD BOX
6 EXPANSION TURBINE
7 EXPANDER-ACCUMULATOR
8 O_2 HEAT EXCHANGER
9 MAIN HEAT EXCHANGER
10 REFRIGERATED HEAT EXCHANGER
11 HIGH PRESSURE COLUMN
12 CRUDE O_2 SUBCOOLER
13 N_2 SUBCOOLER
14 LOW PRESSURE COLUMN
15 AIR PURIFIER
16 LIQUID N_2 STORAGE TANK
17 MOTOR STARTERS
18 NITROGEN COMPRESSOR







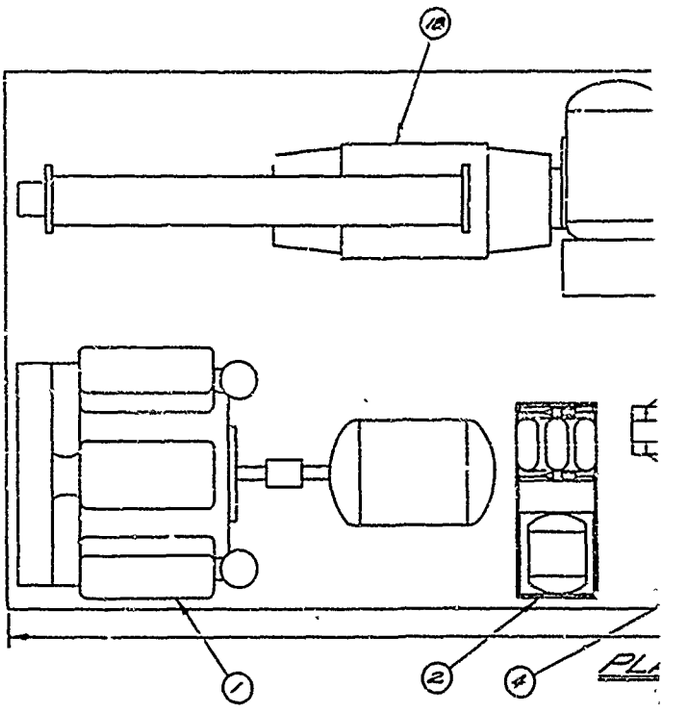
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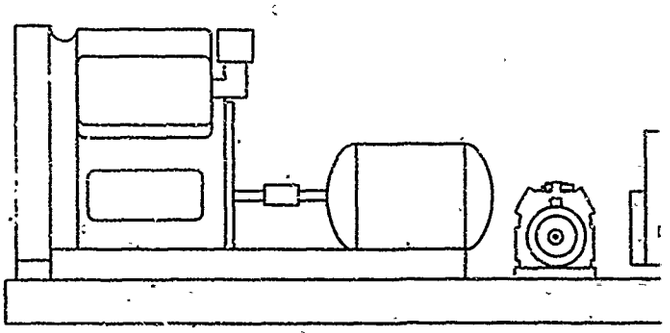
Figure 3.7-7. Nitrogen package layout – systems 9 and 10 – Package 2.

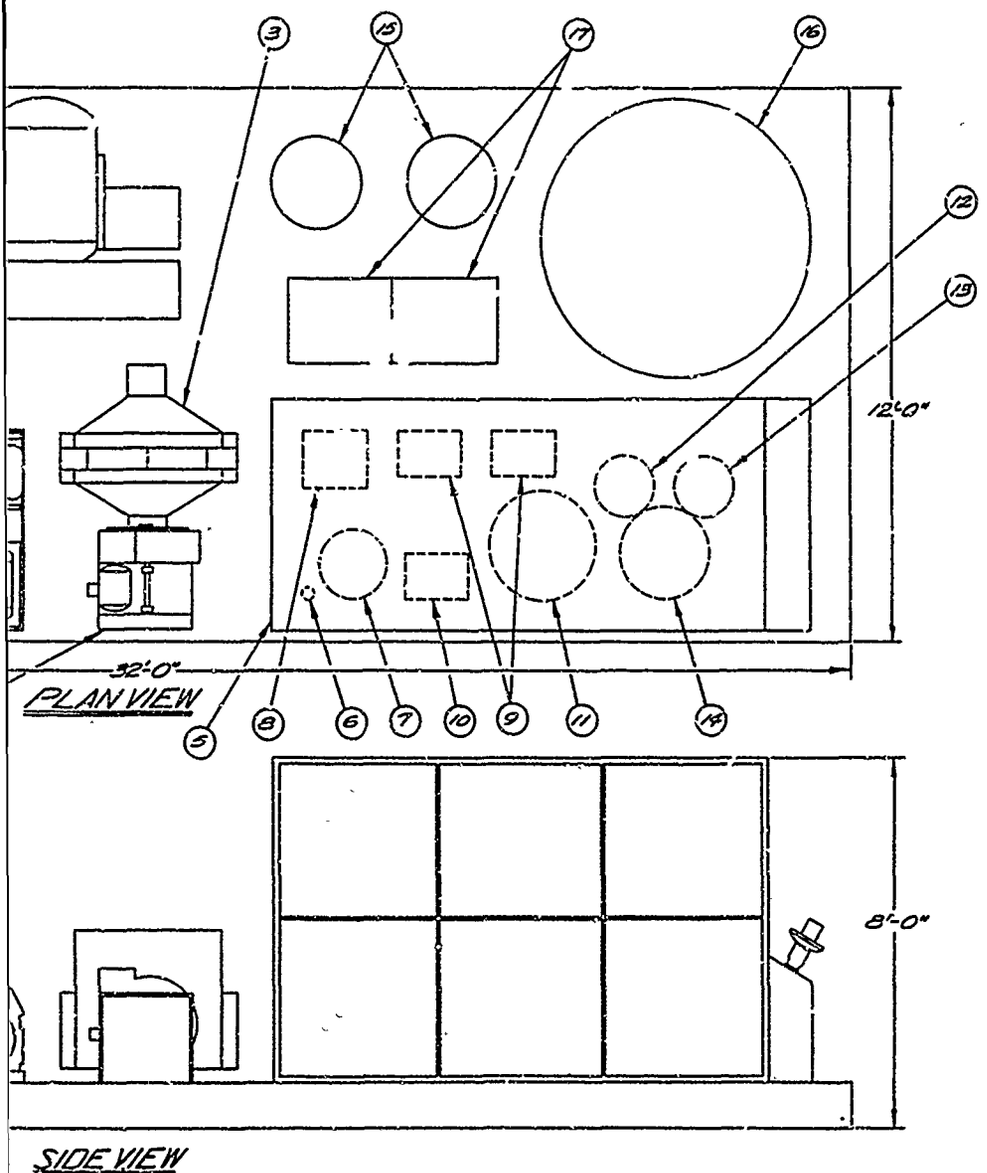
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LEGEND

1 AIR COMPRESSOR
2 REFRIGERATION UNIT
3 AIR COOLED HEAT EXCHANGER
4 FAN & DRIVE UNIT
5 COLD BOX
6 EXPANSION TURBINE
7 EXPANDER-ACCUMULATOR
8 O_2 HEAT EXCHANGER
9 MAIN HEAT EXCHANGER
10 REFRIGERATED HEAT EXCHANGER
11 HIGH PRESSURE COLUMN
12 CRUDE O_2 SUBCOOLER
13 N_2 SUBCOOLER
14 LOW PRESSURE COLUMN
15 AIR PURIFIER
16 LIQUID N_2 STORAGE TANK
17 MOTOR STARTERS
18 NITROGEN COMPRESSOR

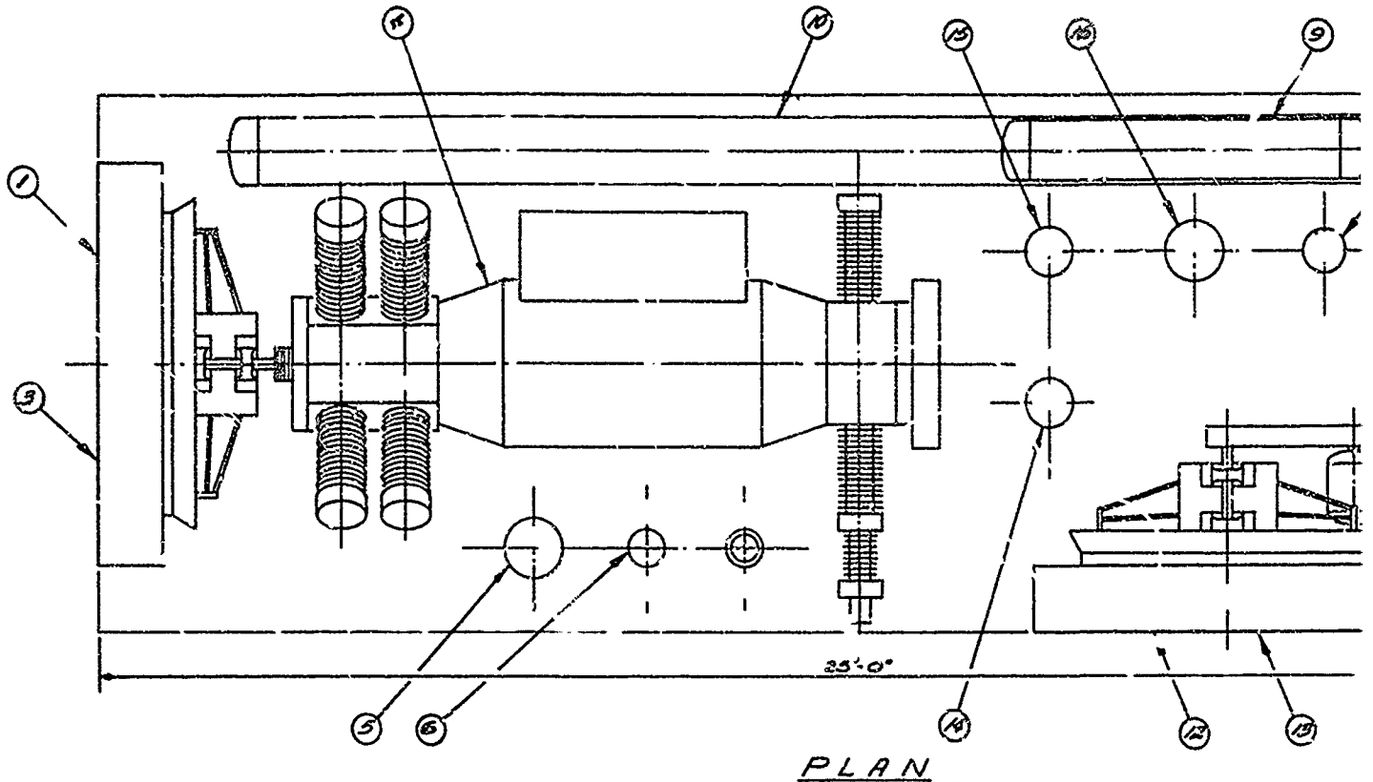




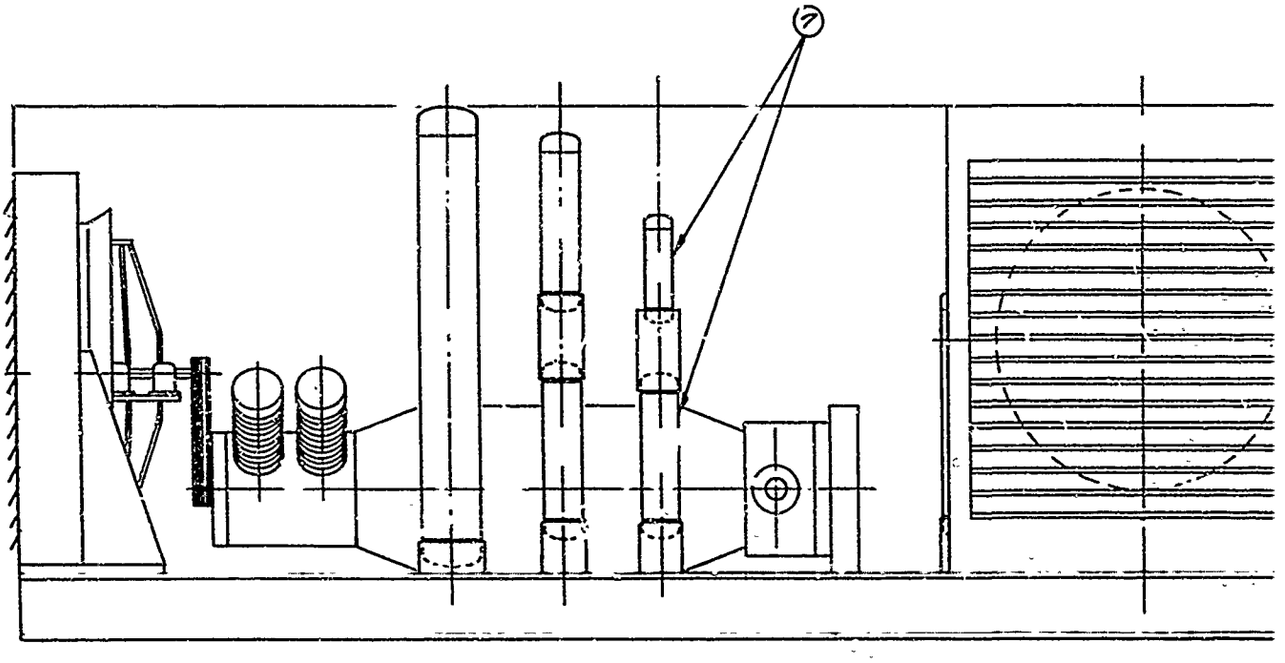


NOTE: STRUCTURE & OUTER SKIN OMITTED FOR CLARITY

Figure 3.7-8. Nitrogen package layout – system 11 – Package 2.



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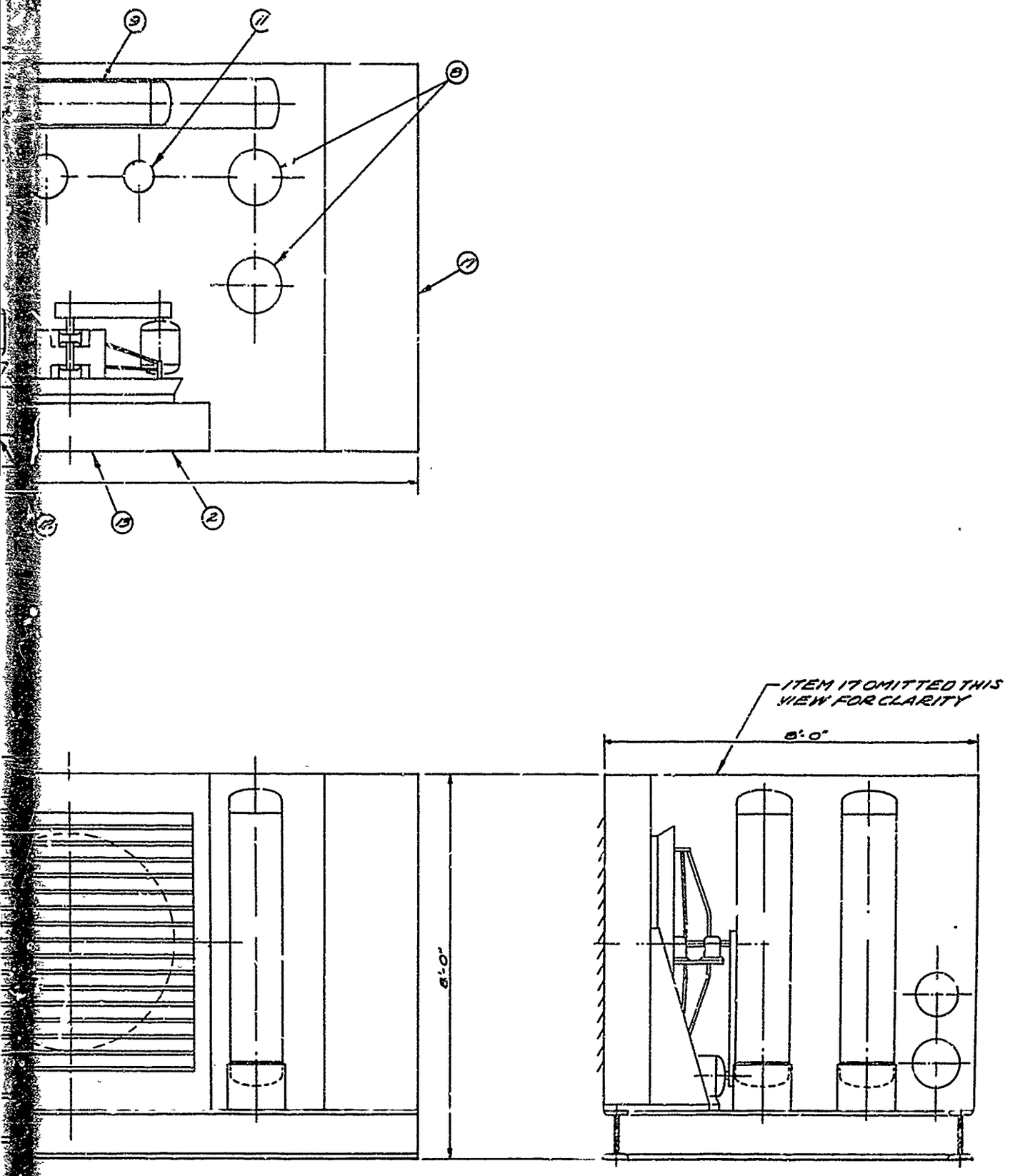
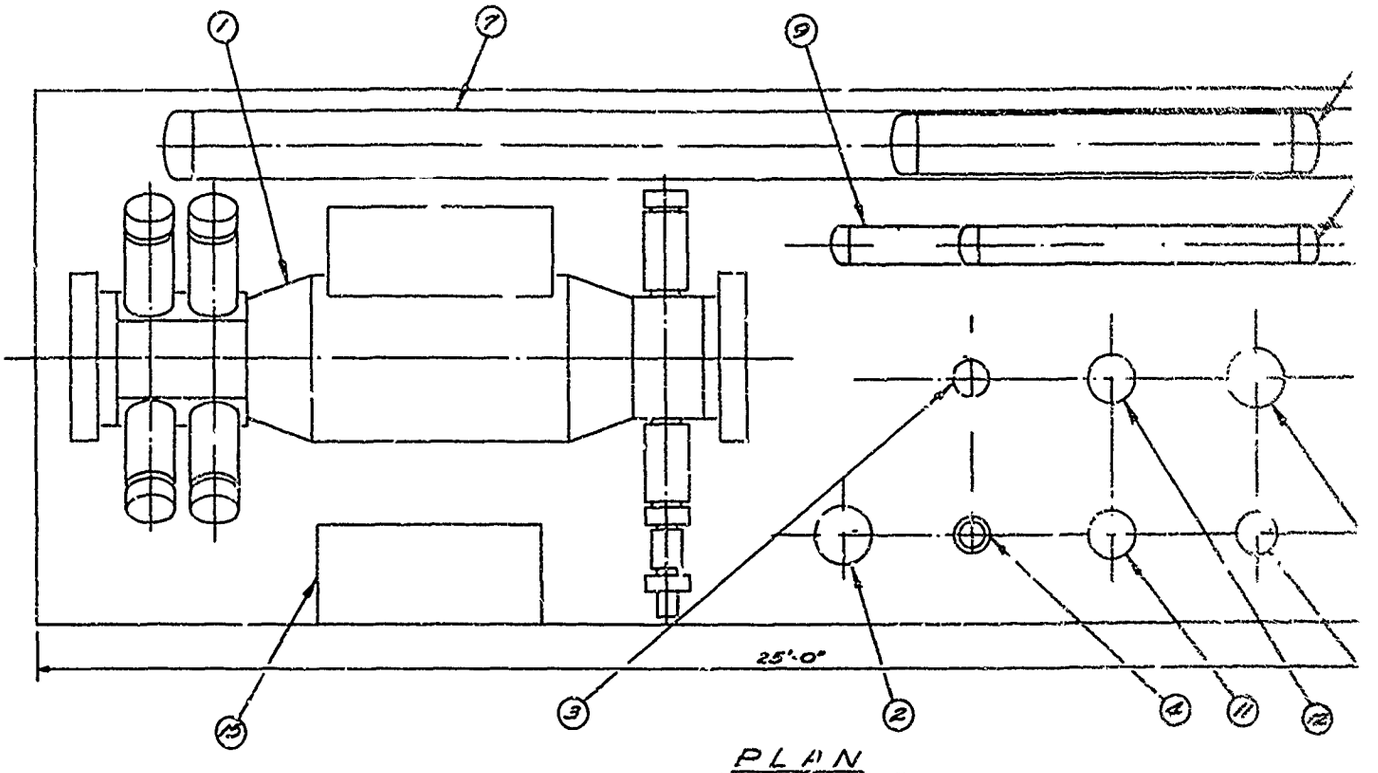
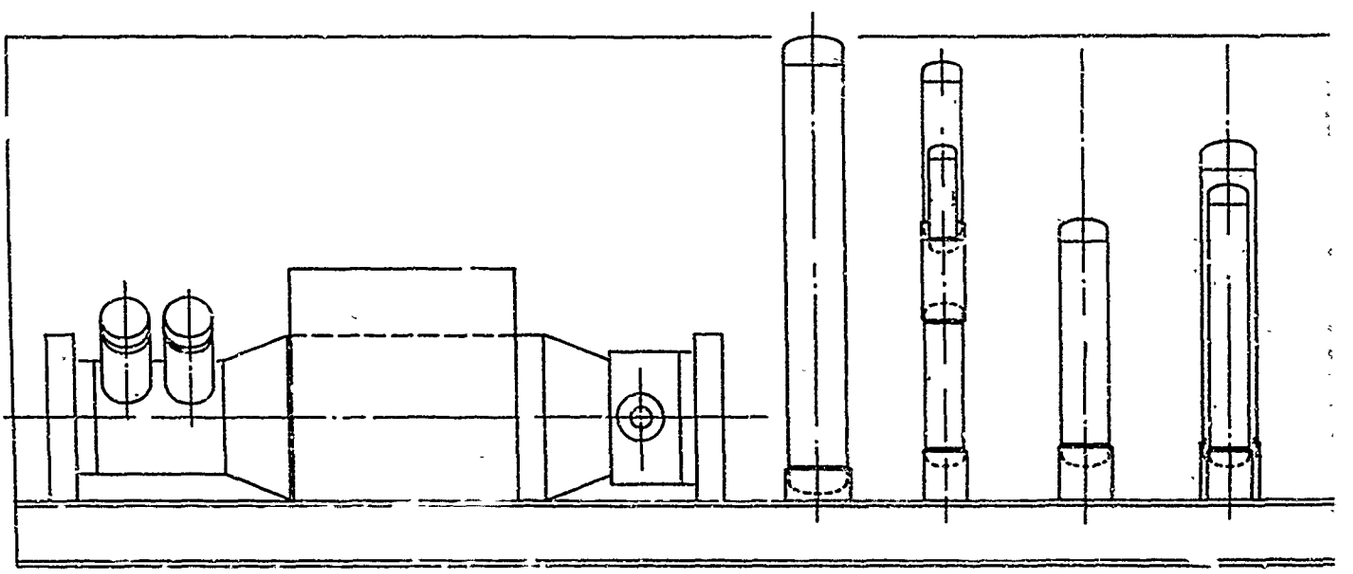


Figure 3.7-9. Ammonia package layout – systems 1 and 3 – Package 3.

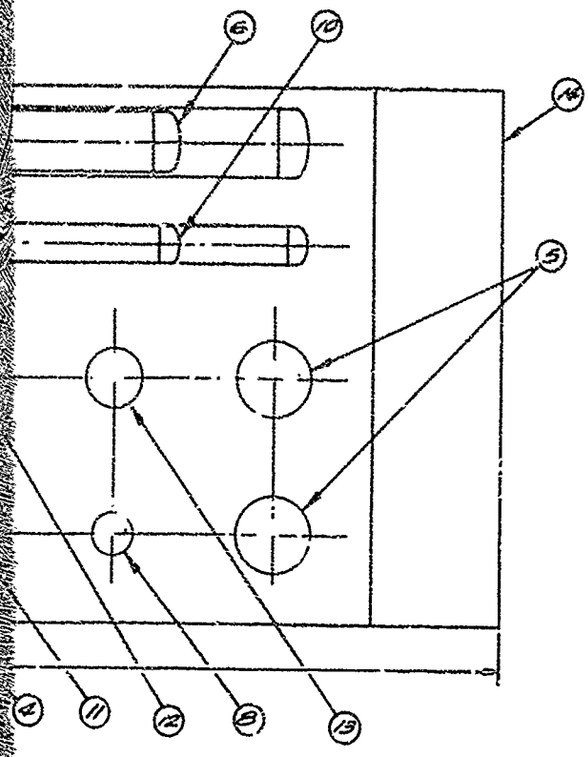


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SIDE ELEVATION

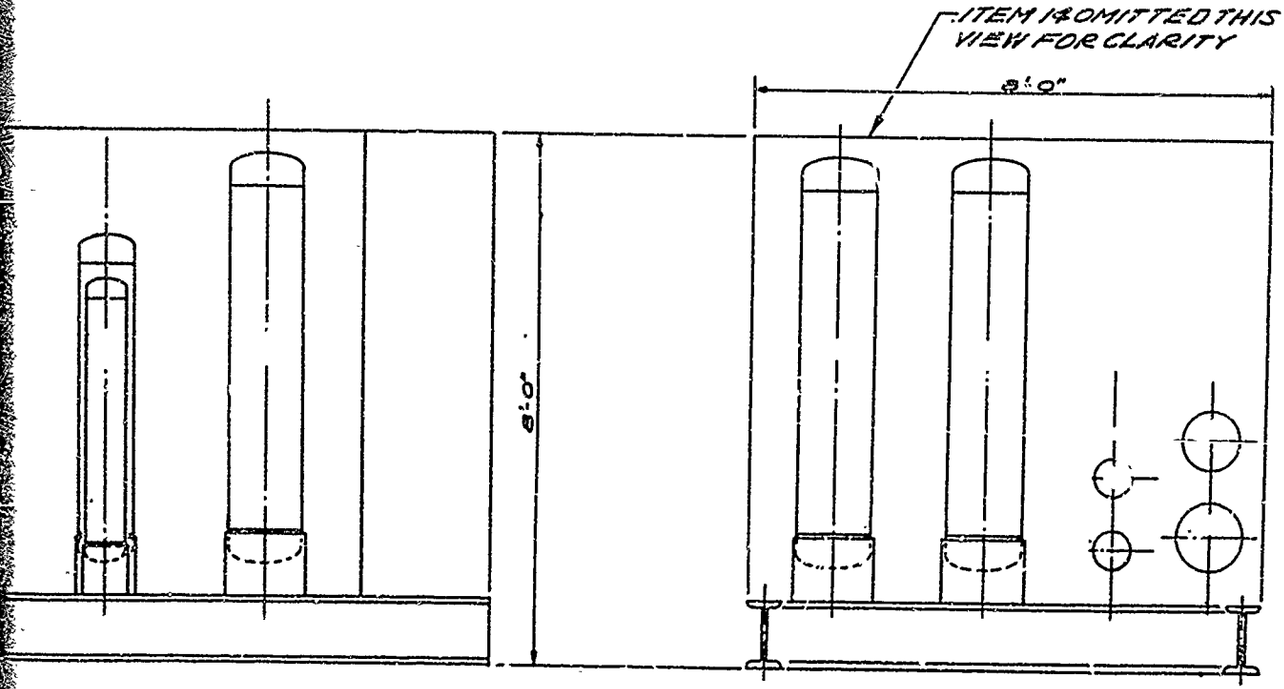
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LEGEND

1 SYNTHESIS GAS COMPRESSOR
2 DEOXO UNIT
3 KNOCK-OUT DRUM
4 GAS FILTER
5 AMMONIA CONVERTER
6 SECONDARY CONDENSER
7 CONVERTER HEAT EXCHANGER
8 AMMONIA VAPORIZER
9 PRIMARY CONDENSER
10 PRODUCT CONDENSER
11 PRIMARY SEPARATOR
12 SECONDARY SEPARATOR
13 PRODUCT SEPARATOR
14 CONTROL CABINET
15 WATER FILTER

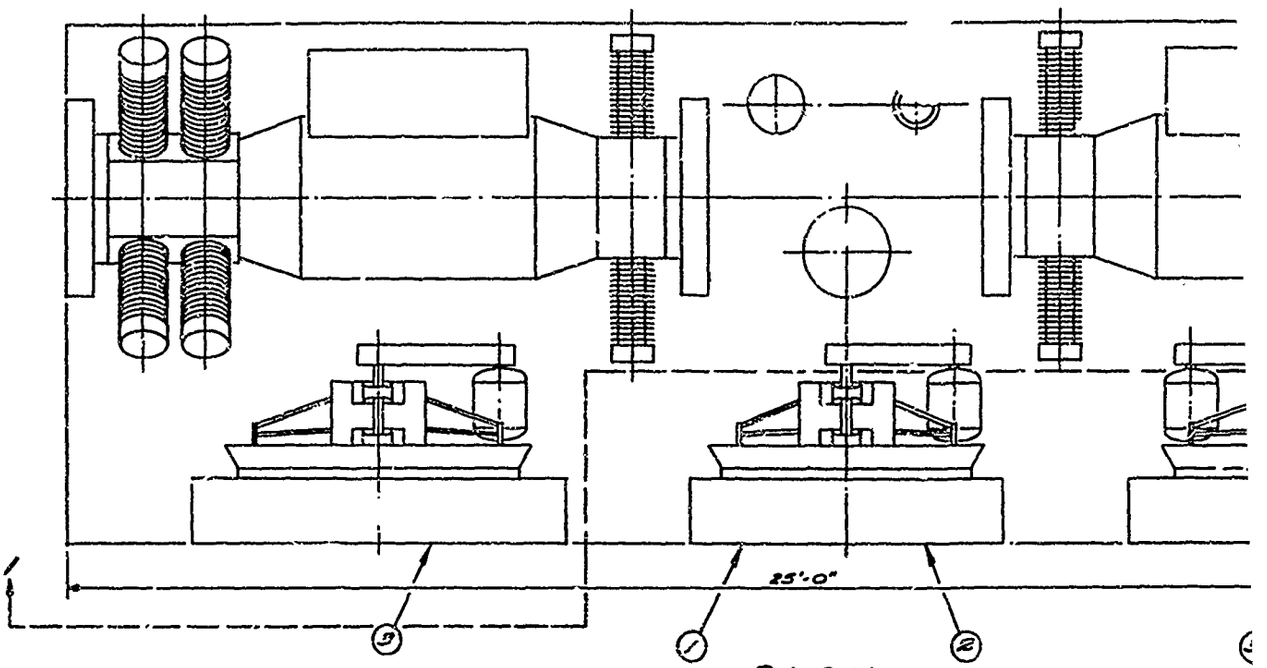
NOTE: STRUCTURE & OUTER SKIN OMITTED FOR CLARITY



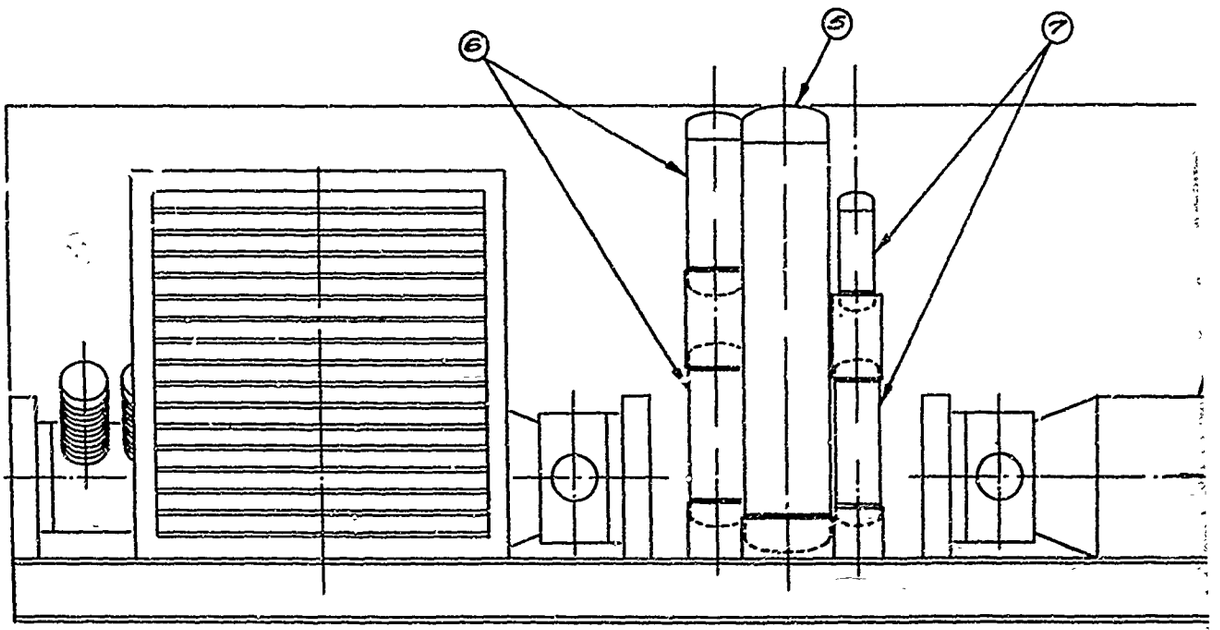
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Figure 3.7-10. Ammonia package layout – system 2 – Package 3.

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PLAN



SECTION "1-1"

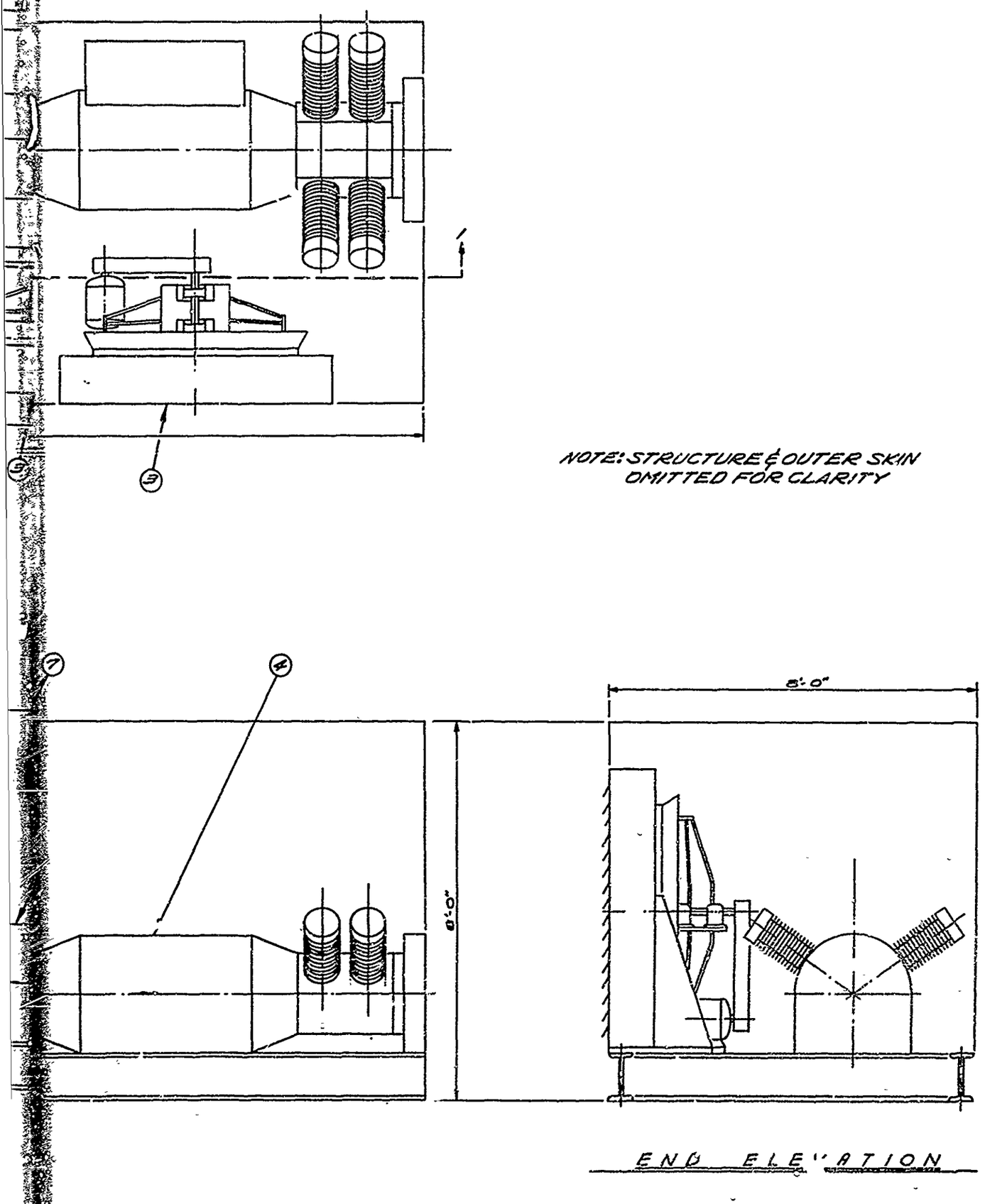
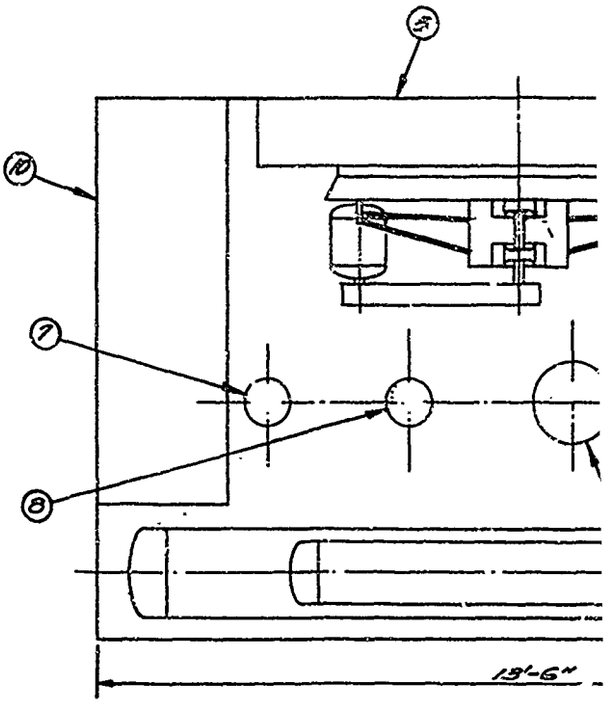
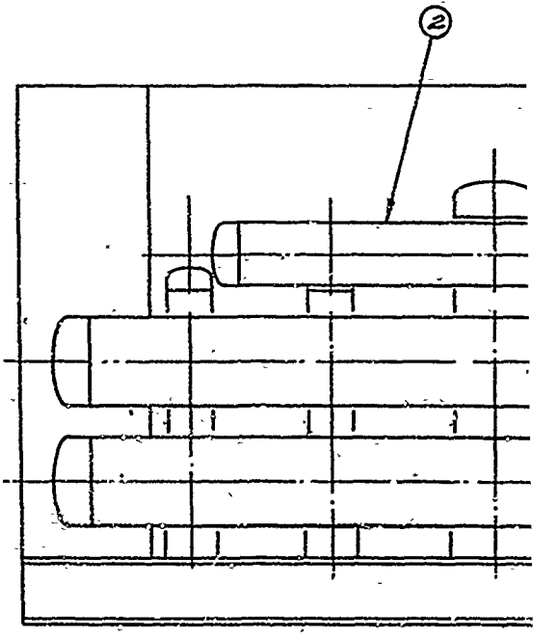


Figure 3.7-11. Ammonia package layout – system 4 – Package 3.

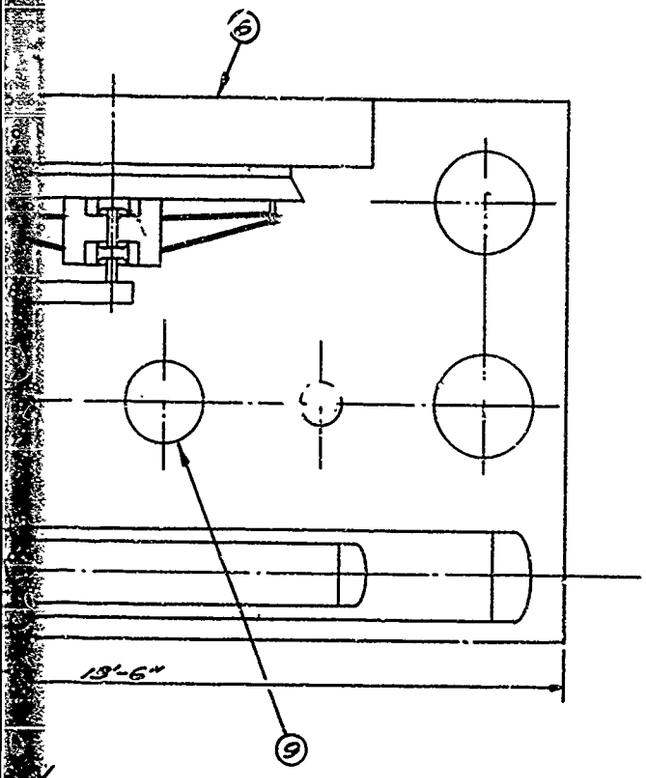
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PLAN



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LEGEND

1 AMMONIA CONVERTER
2 SECONDARY CONDENSER
3 CONVERTER HEAT EXCHANGER
4 AMMONIA VAPORIZER
5 PRIMARY CONDENSER
6 PRODUCT CONDENSER
7 PRIMARY SEPARATOR
8 SECONDARY SEPARATOR
9 PRODUCT SEPARATOR
10 CONTROL CABINET

NOTE: STRUCTURE & OUTER SKIN OMITTED FOR CLARITY

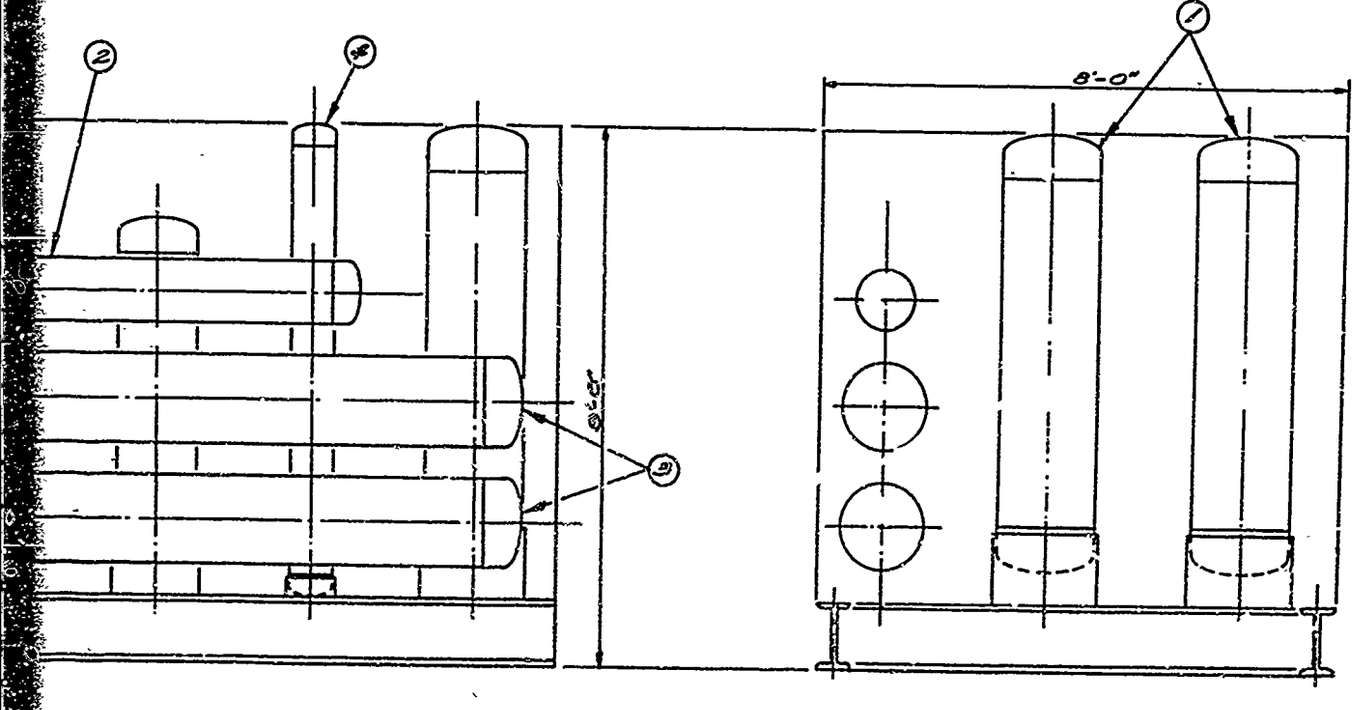
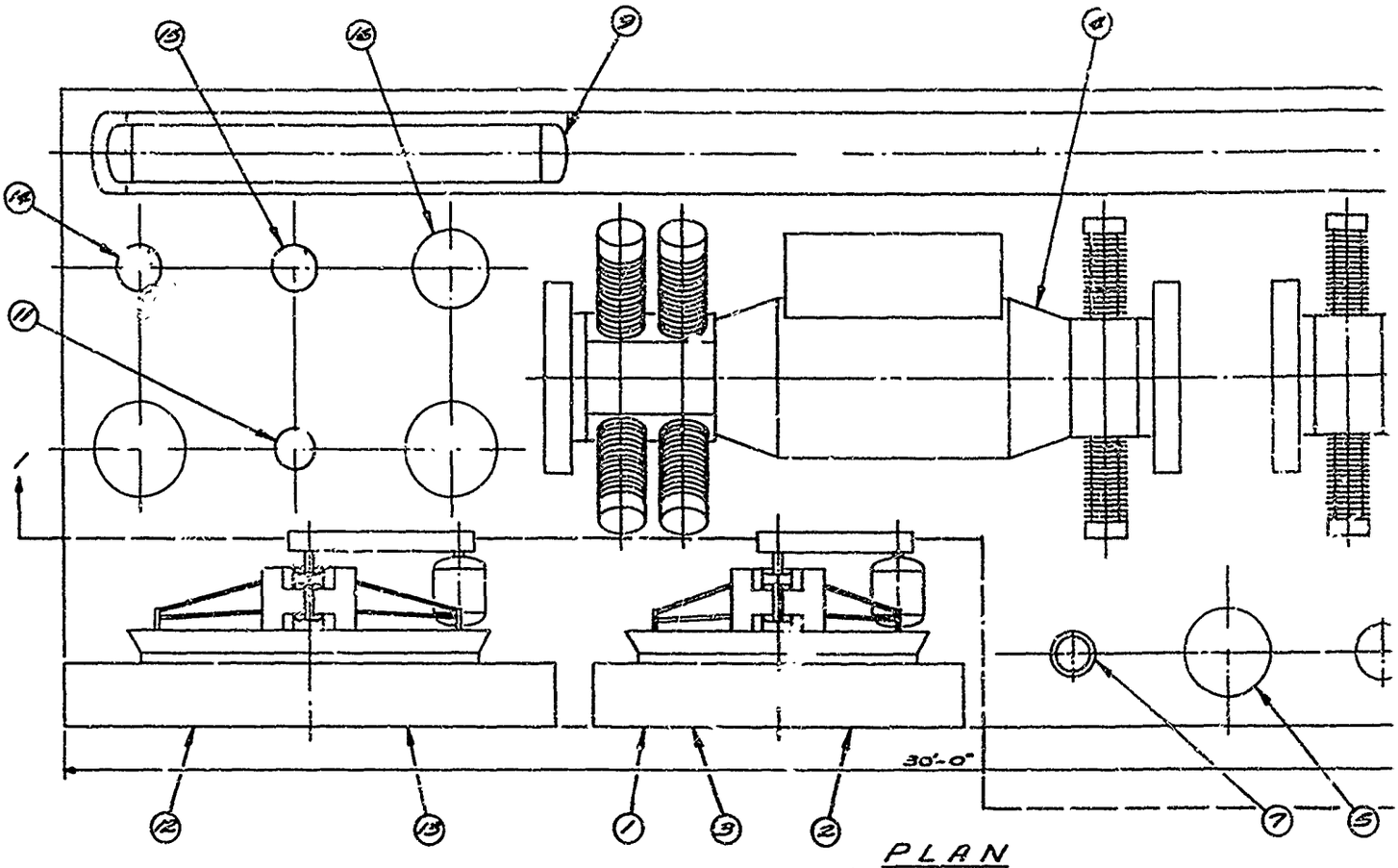
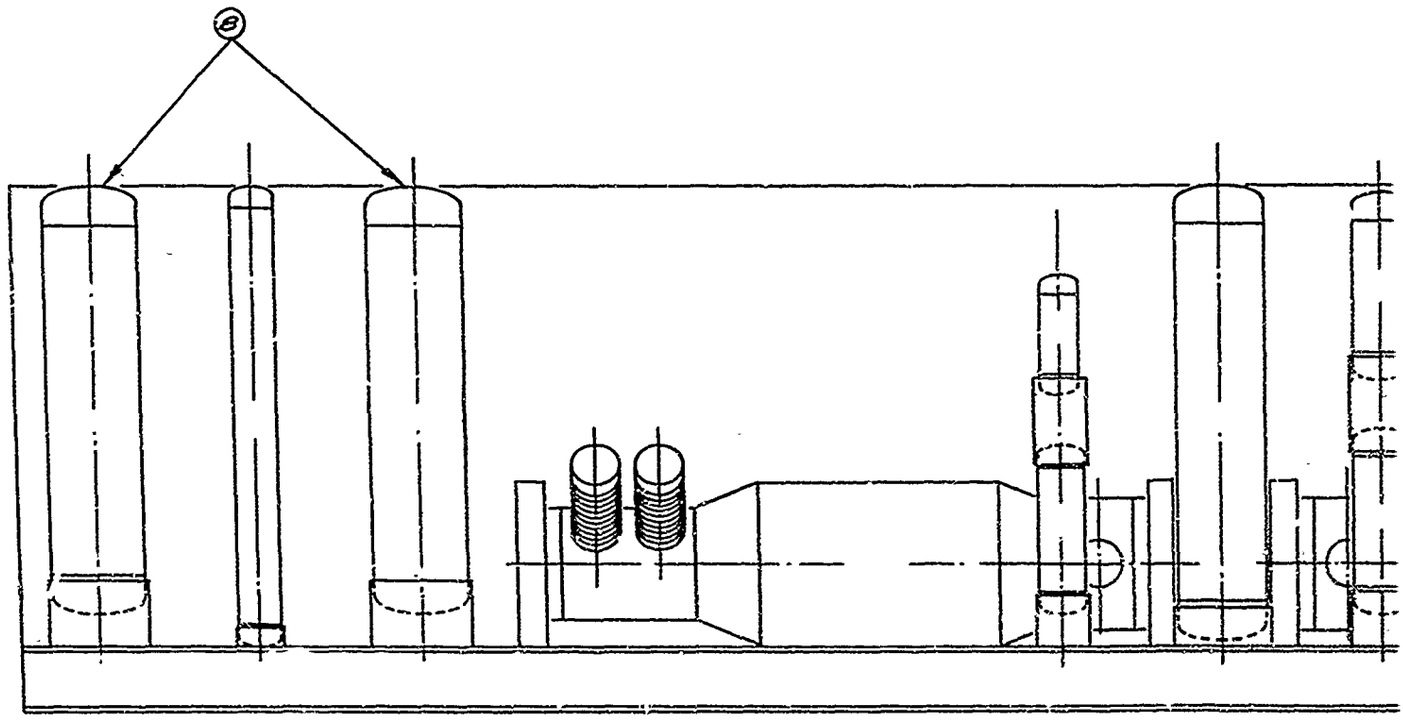


Figure 3.7-12. Ammonia package layout – system 4 – Package 3a.



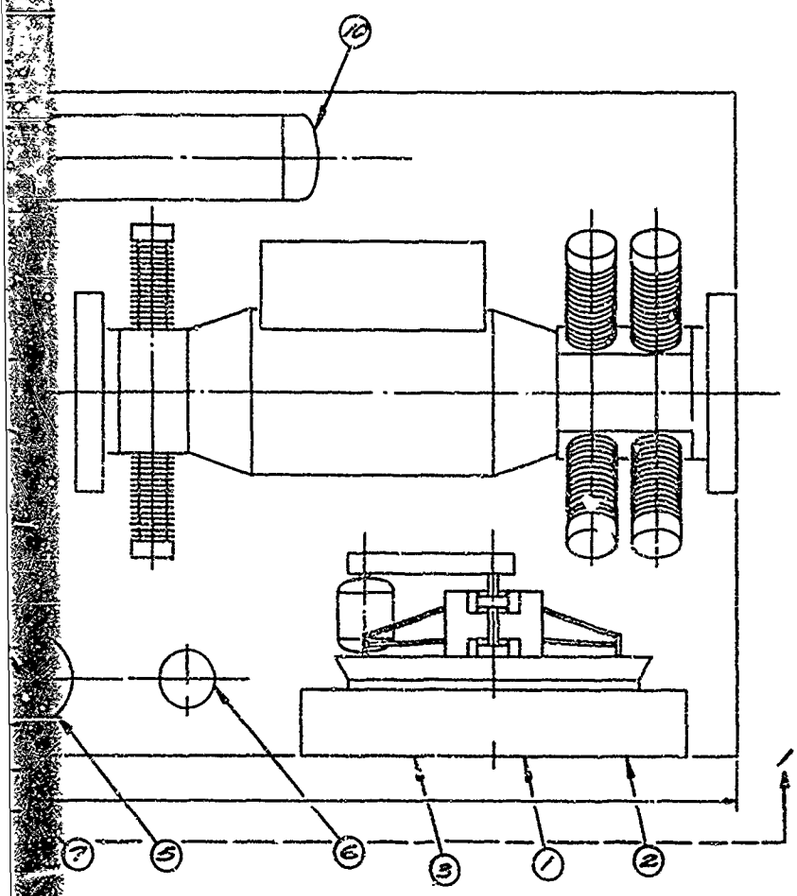
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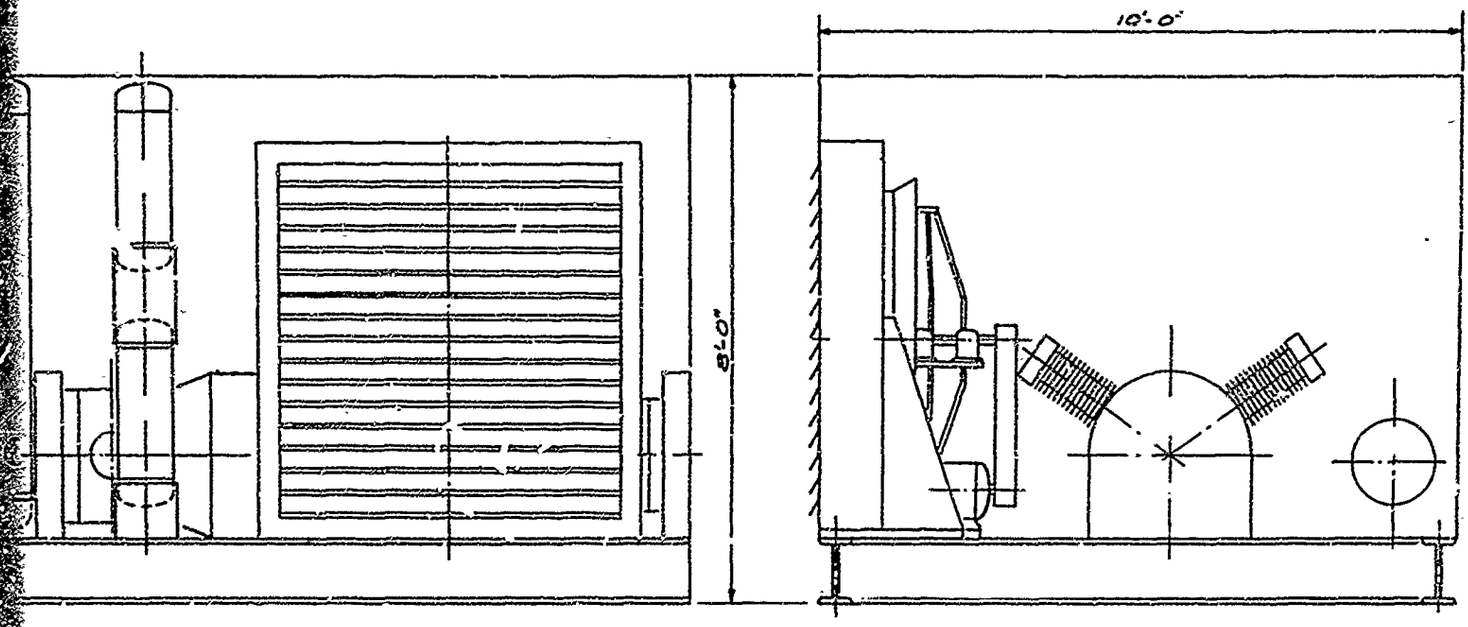
Figure



LEGEND

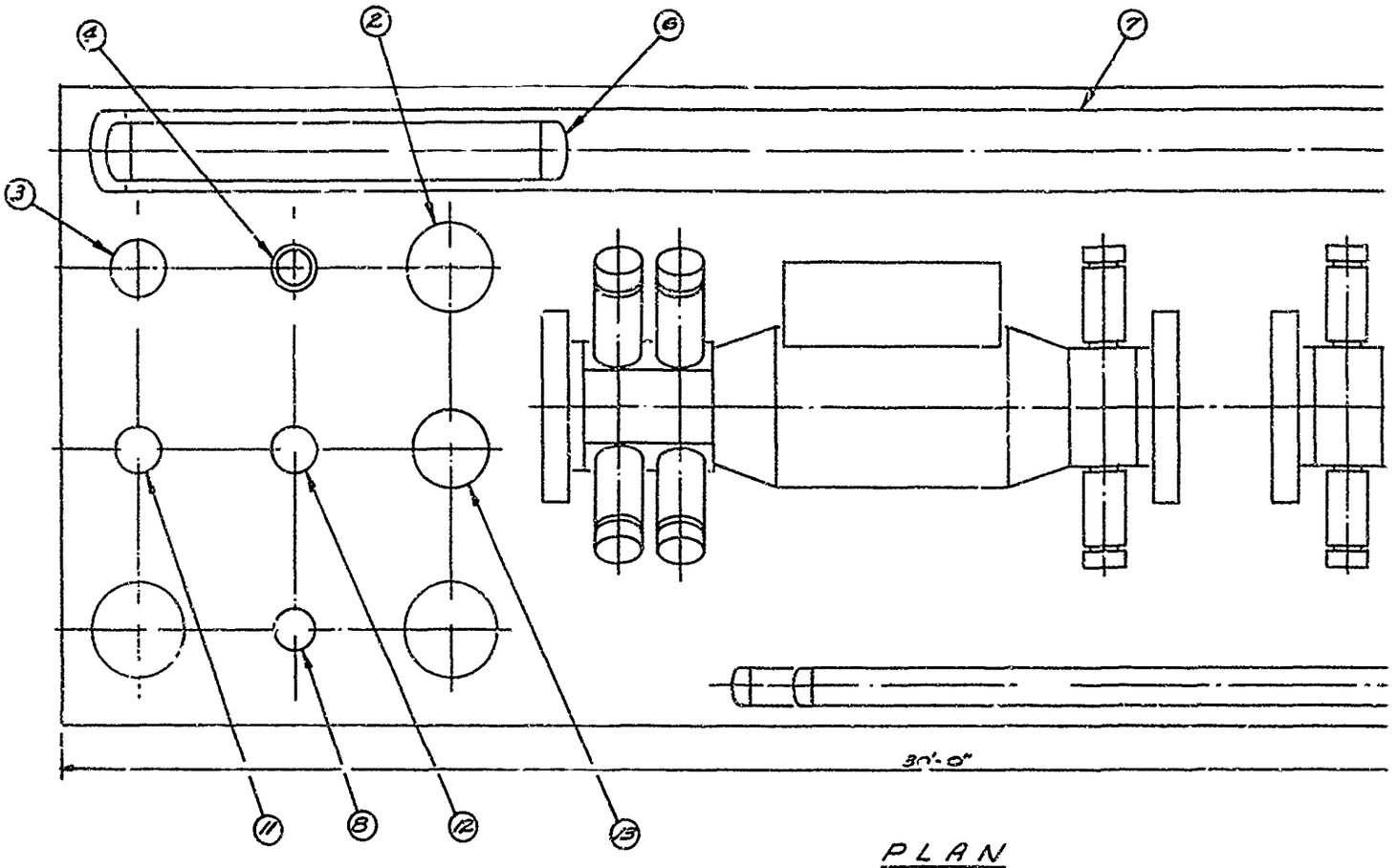
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2 H_2 PRECOOLER
3 SYNTHESIS GAS COMPRESSOR INTER & AFTER COOLERS
4 SYNTHESIS GAS COMPRESSOR
5 DEOXO UNIT
6 1ST KNOCK-OUT DRUM
7 GAS FILTER
8 AMMONIA CONVERTER
9 SECONDARY CONDENSER
10 CONVERTER HEAT EXCHANGER
11 AMMONIA VAPORIZER
12 PRIMARY CONDENSER
13 PRODUCT CONDENSER
14 PRIMARY SEPARATOR
15 SECONDARY SEPARATOR
16 PRODUCT SEPARATOR

NOTE: STRUCTURE & OUTER SKIN OMITTED FOR CLARITY

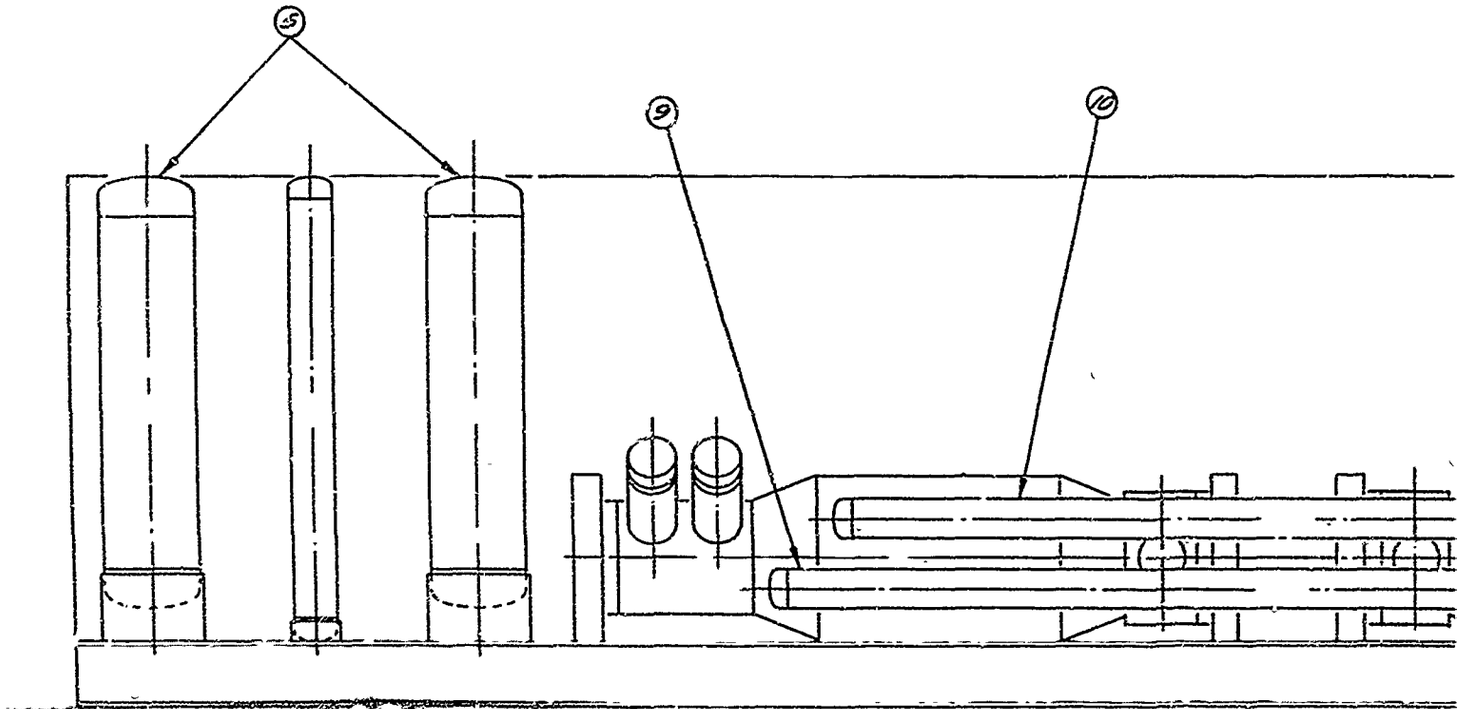


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Figure 3.7-13. Ammonia package layout – systems 5, 7, and 8 – Package 3.

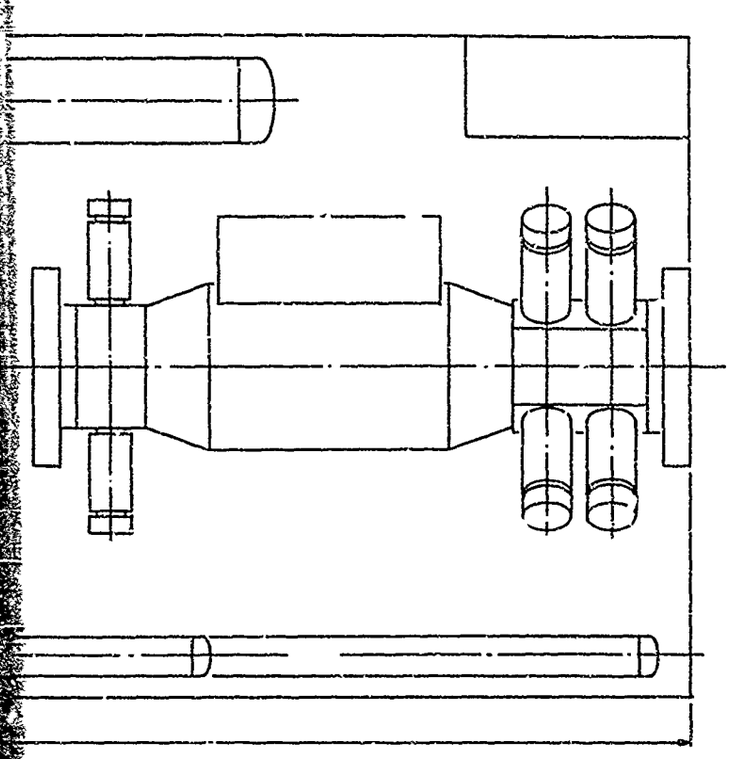


PLAN

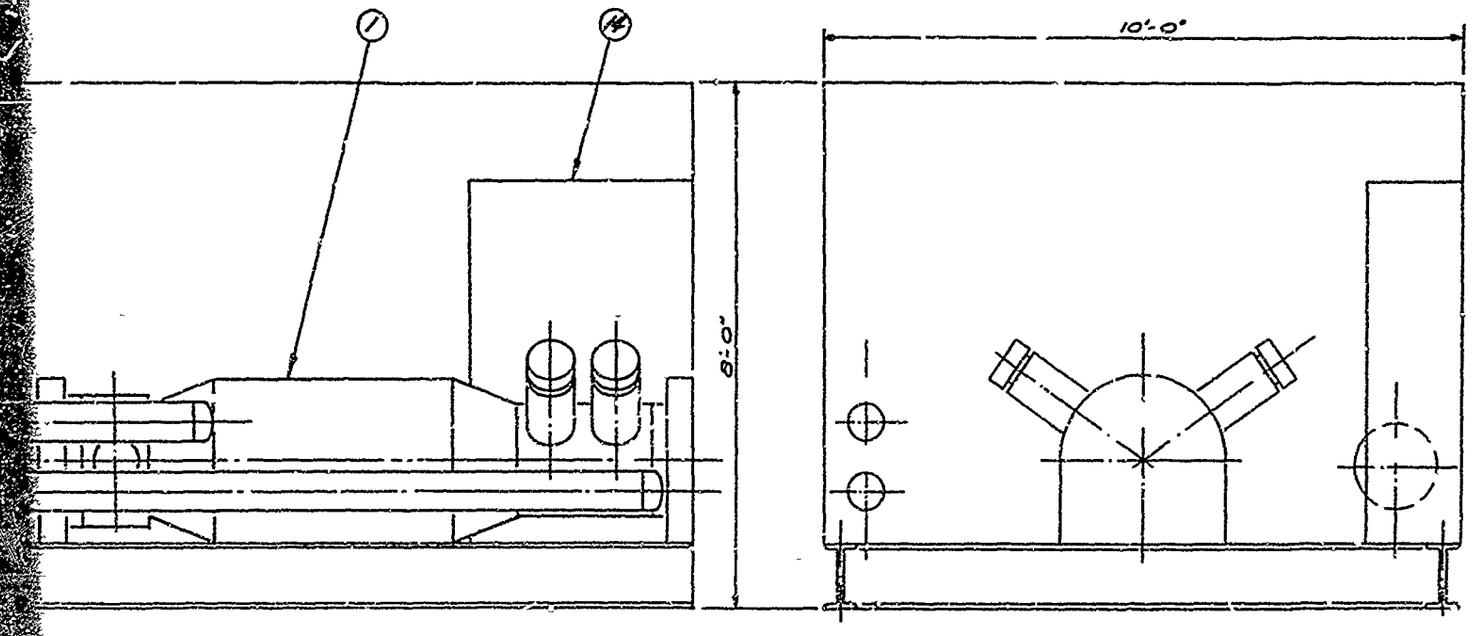


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SIDE ELEVATION



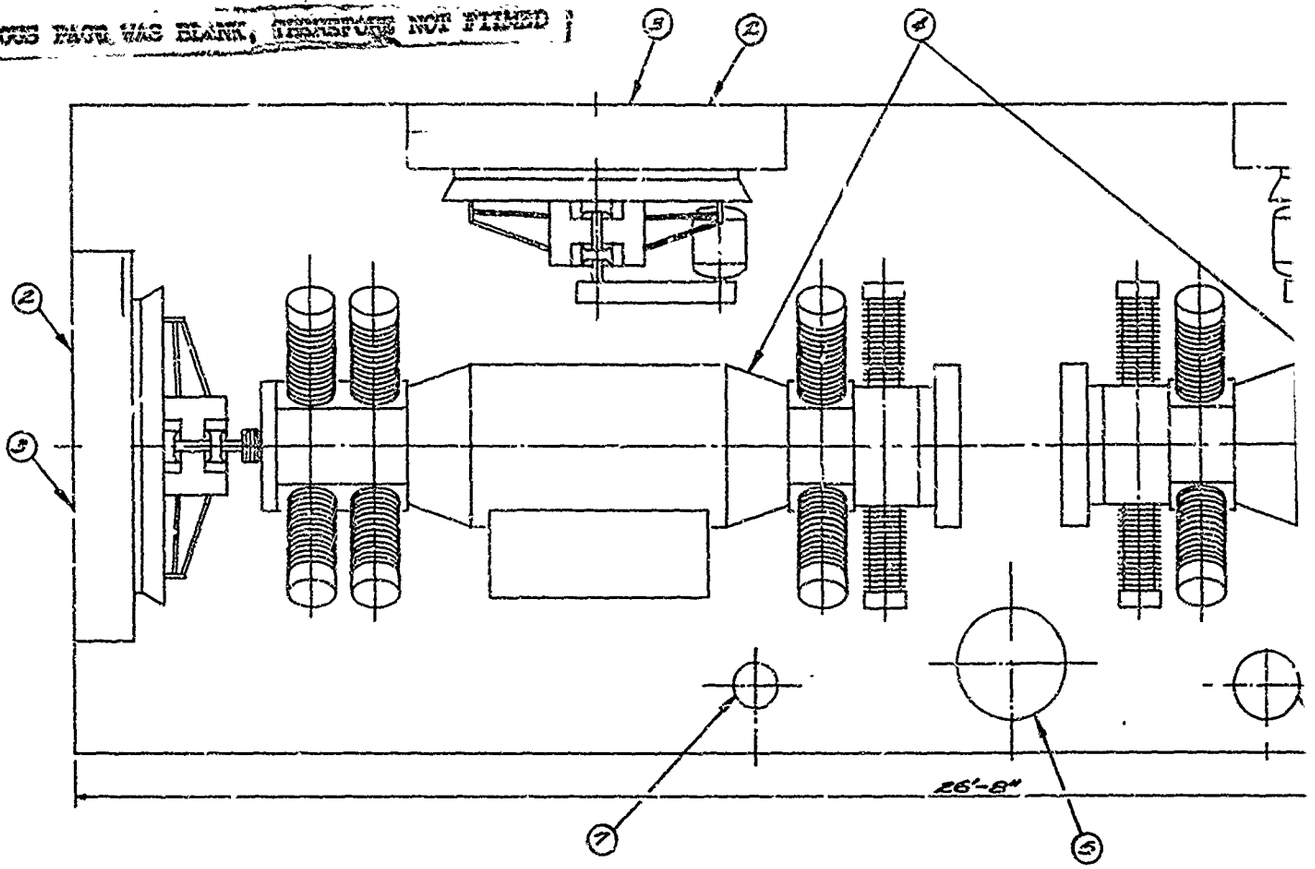
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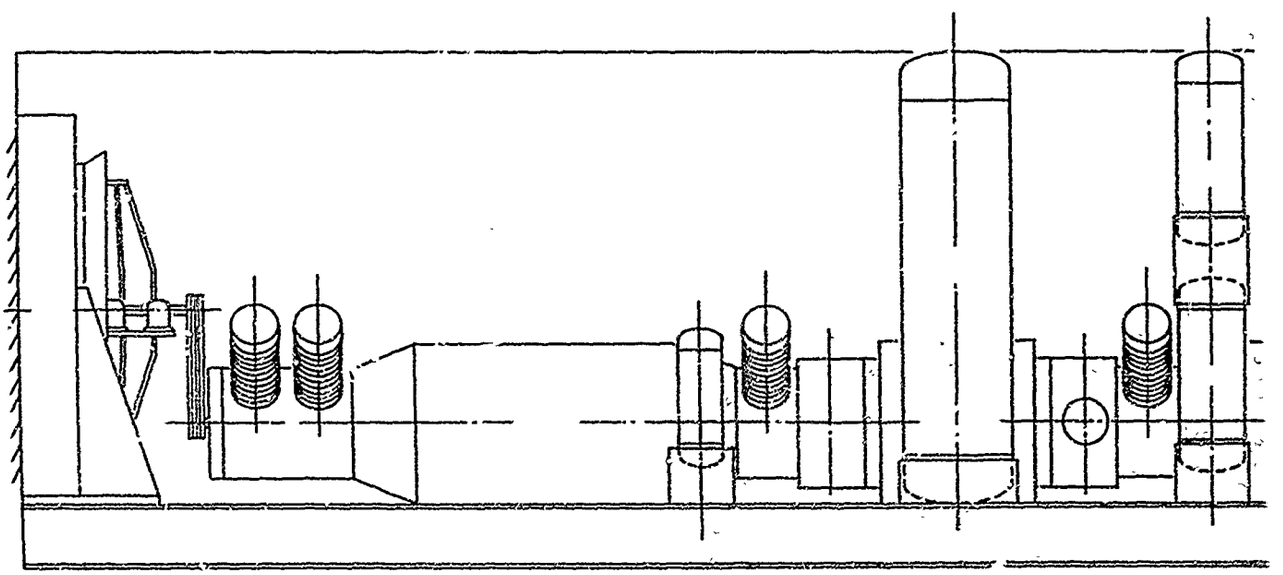
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Figure 3.7-14. Ammonia package layout – systems 6 – Package 3.

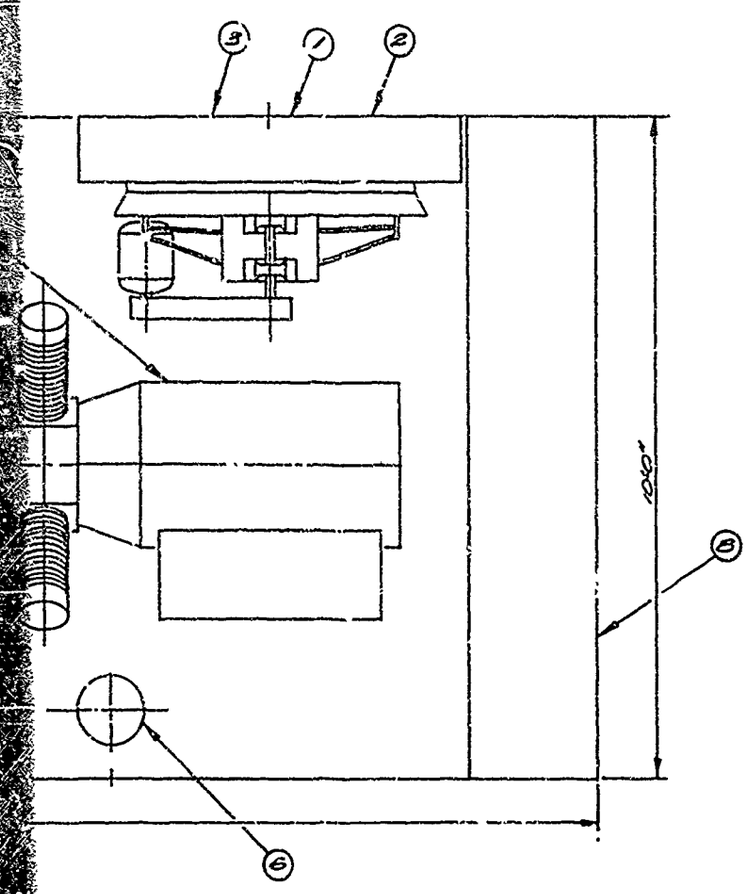
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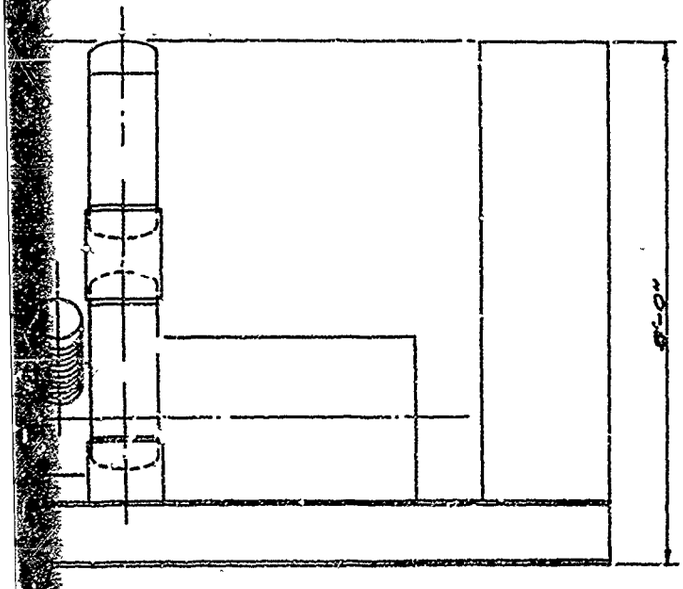


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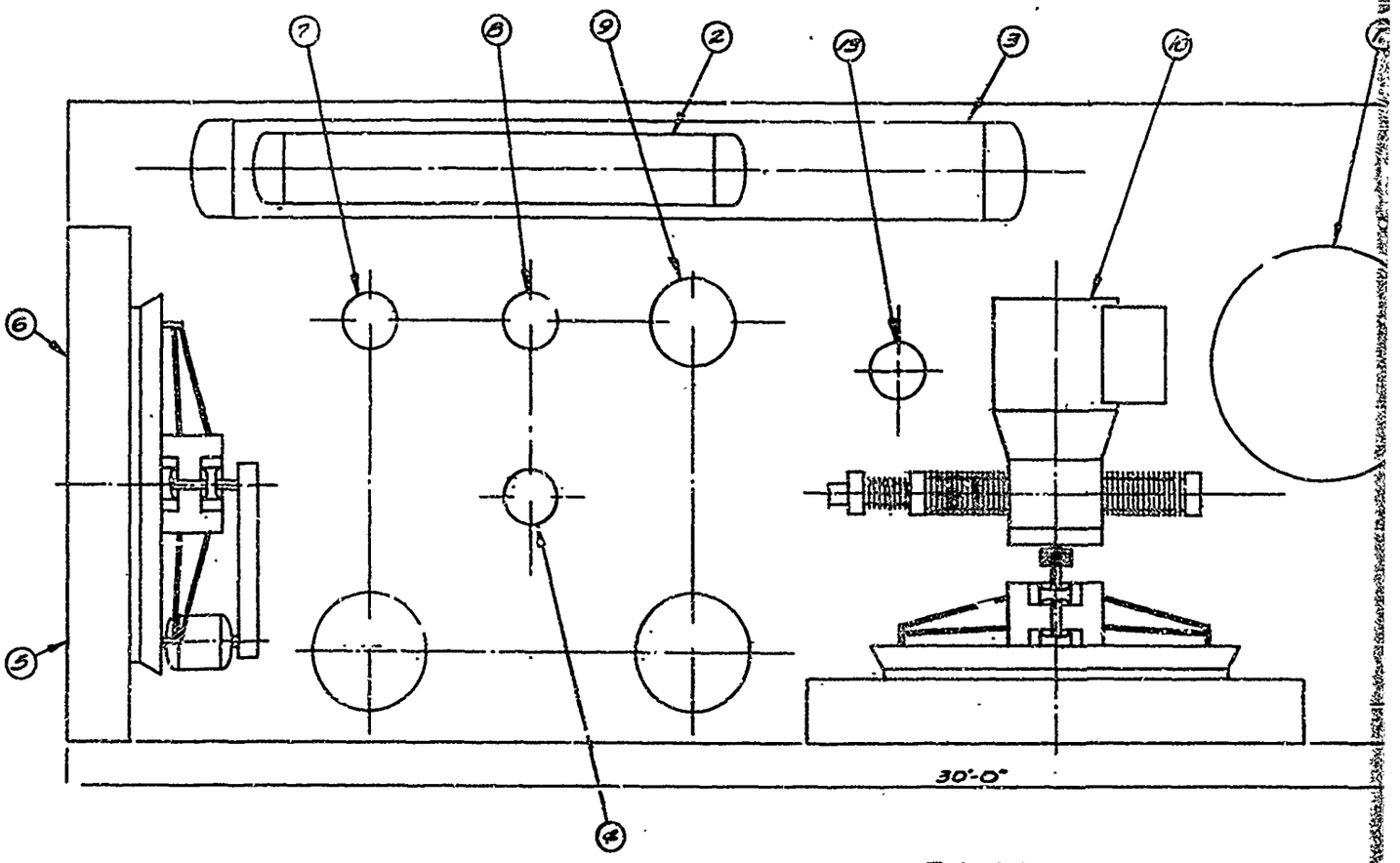


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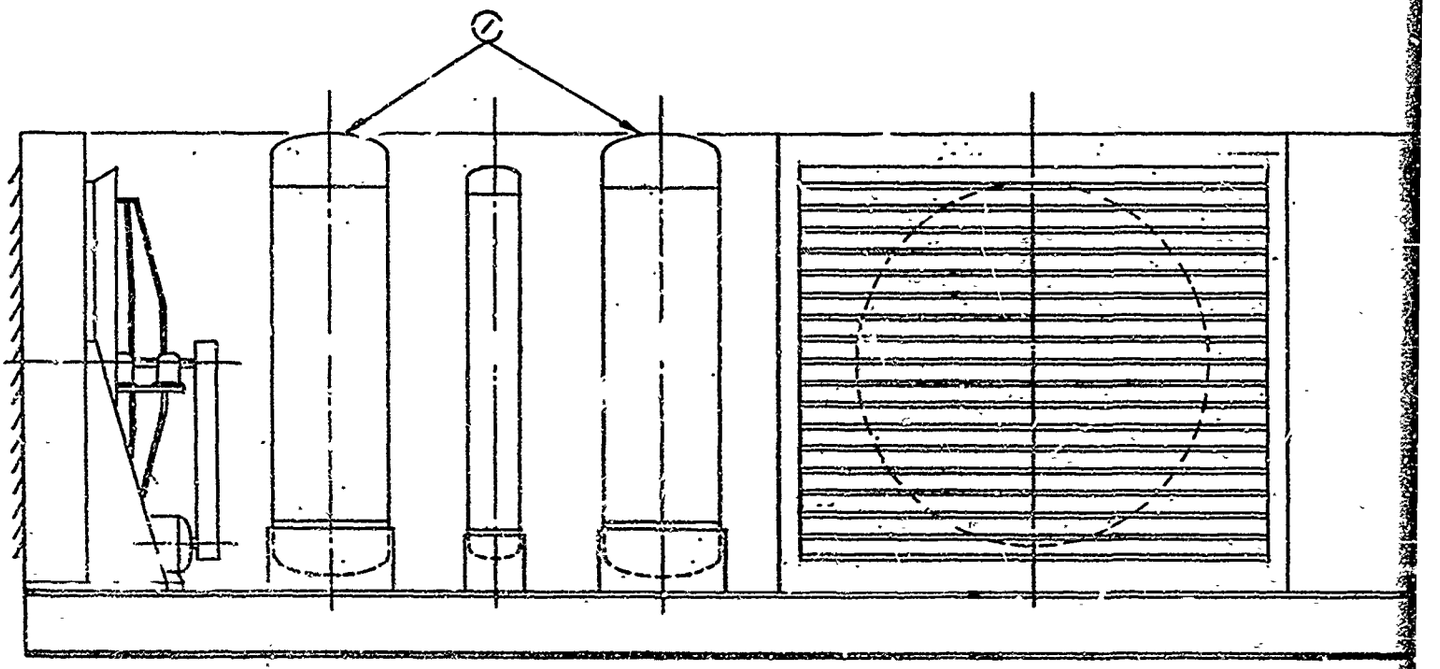
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2 H_2 PRECOOLER
3 SYNTHESIS GAS COMPRESSOR INTER & AFTER COOLERS
4 SYNTHESIS GAS COMPRESSOR
5 DEOXO UNIT
6 KNOCK-OUT DRUM
7 GAS FILTER
8 CONTROL CABINET

NOTE: STRUCTURE & OUTER SKIN OMITTED FOR CLARITY

Figure 3.7-15. Ammonia package layout – Skid 1 – system 9 – Package 3.



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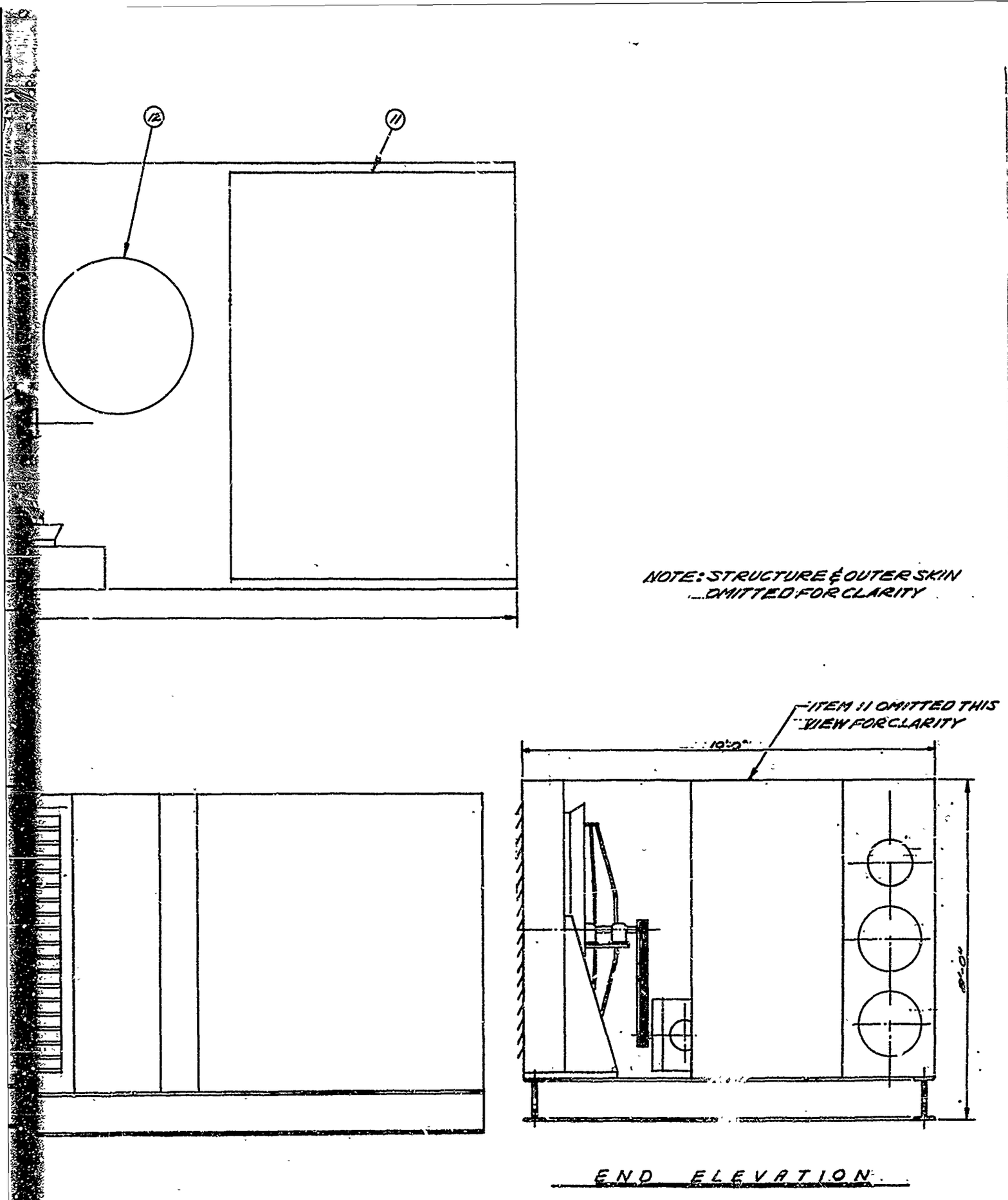
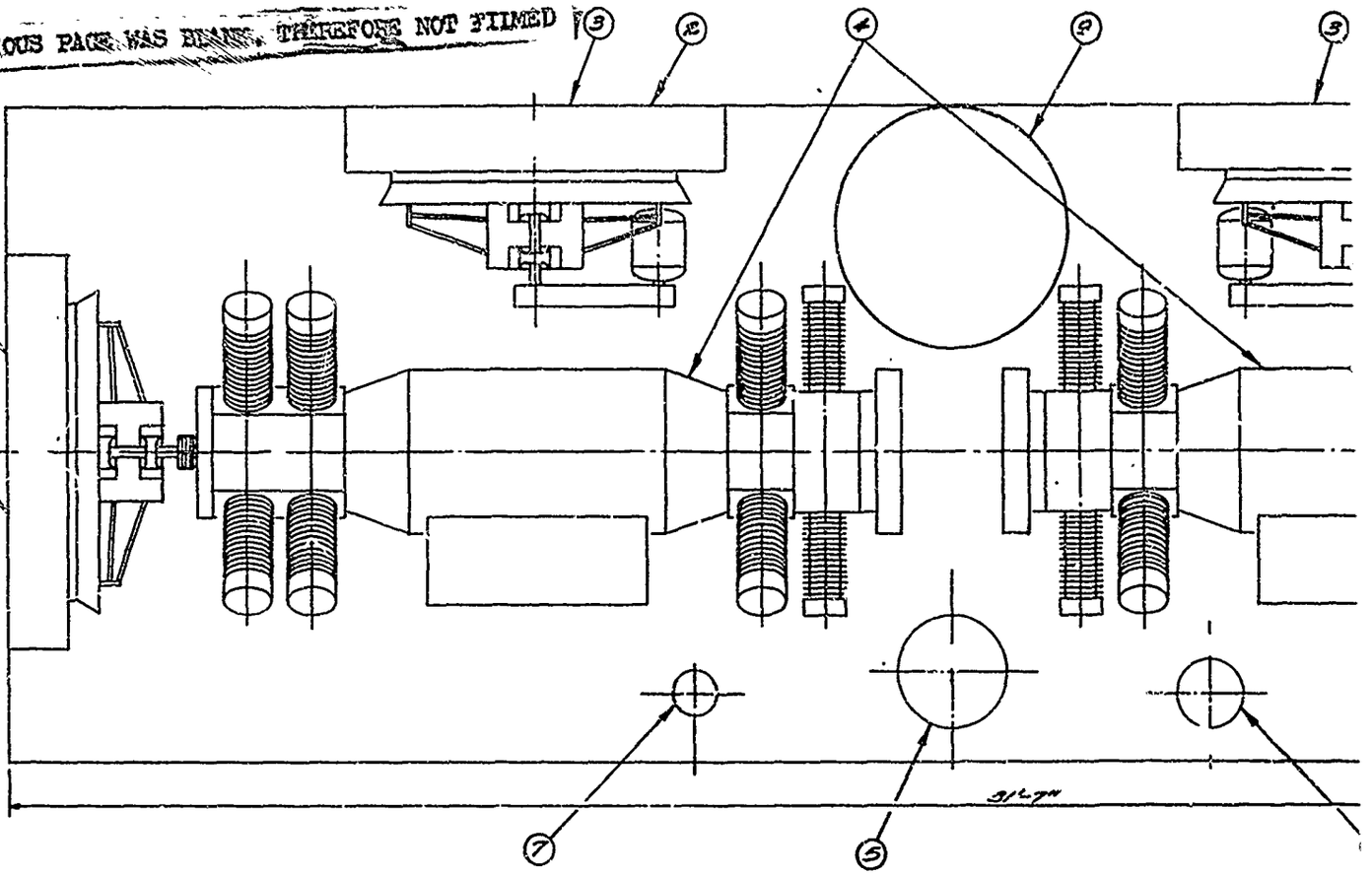
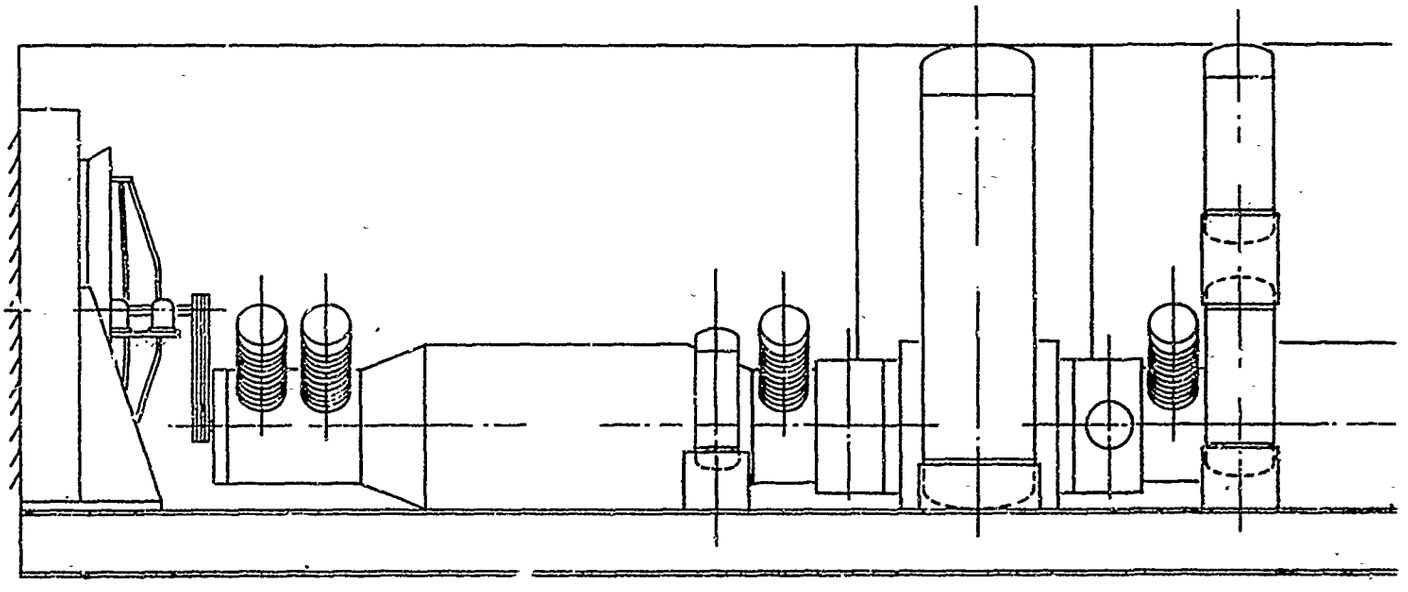


Figure 3.7-16. Ammonia package layout – Skid 2 – System 3 – Package 3a.

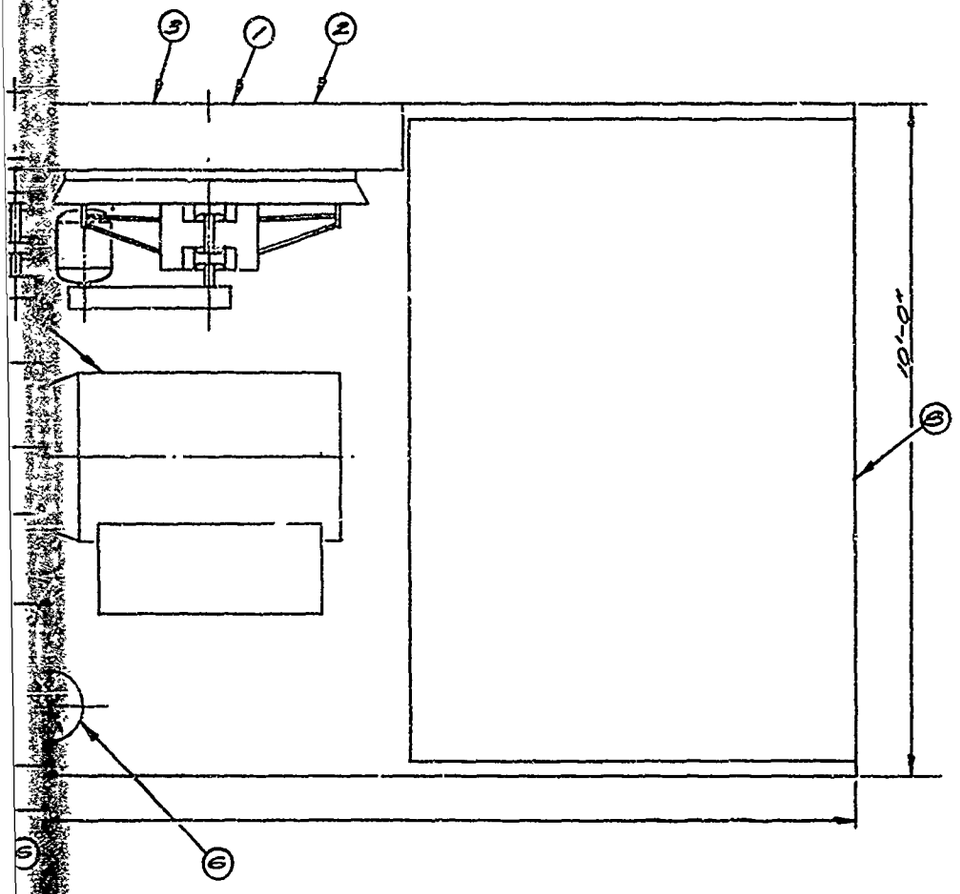
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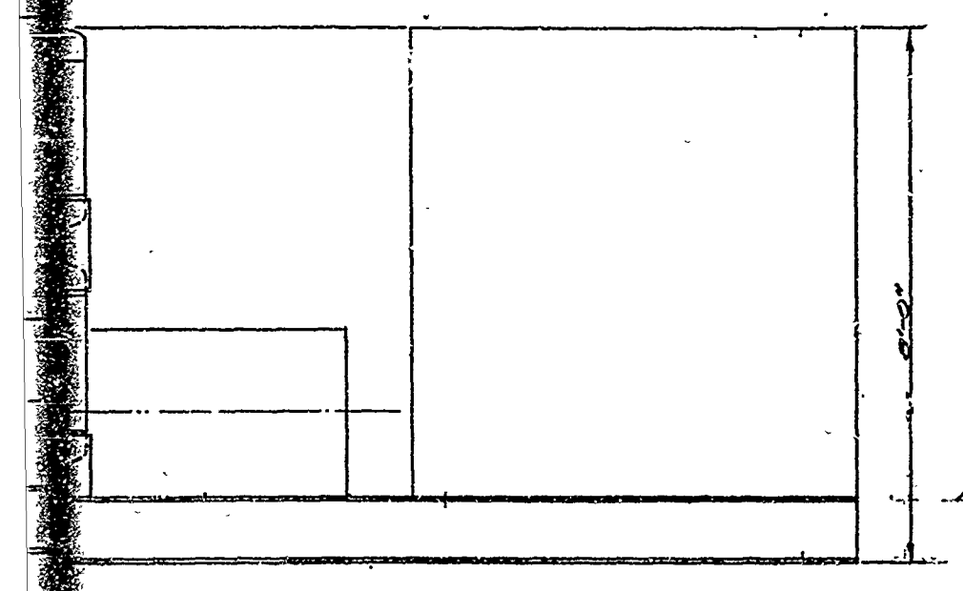


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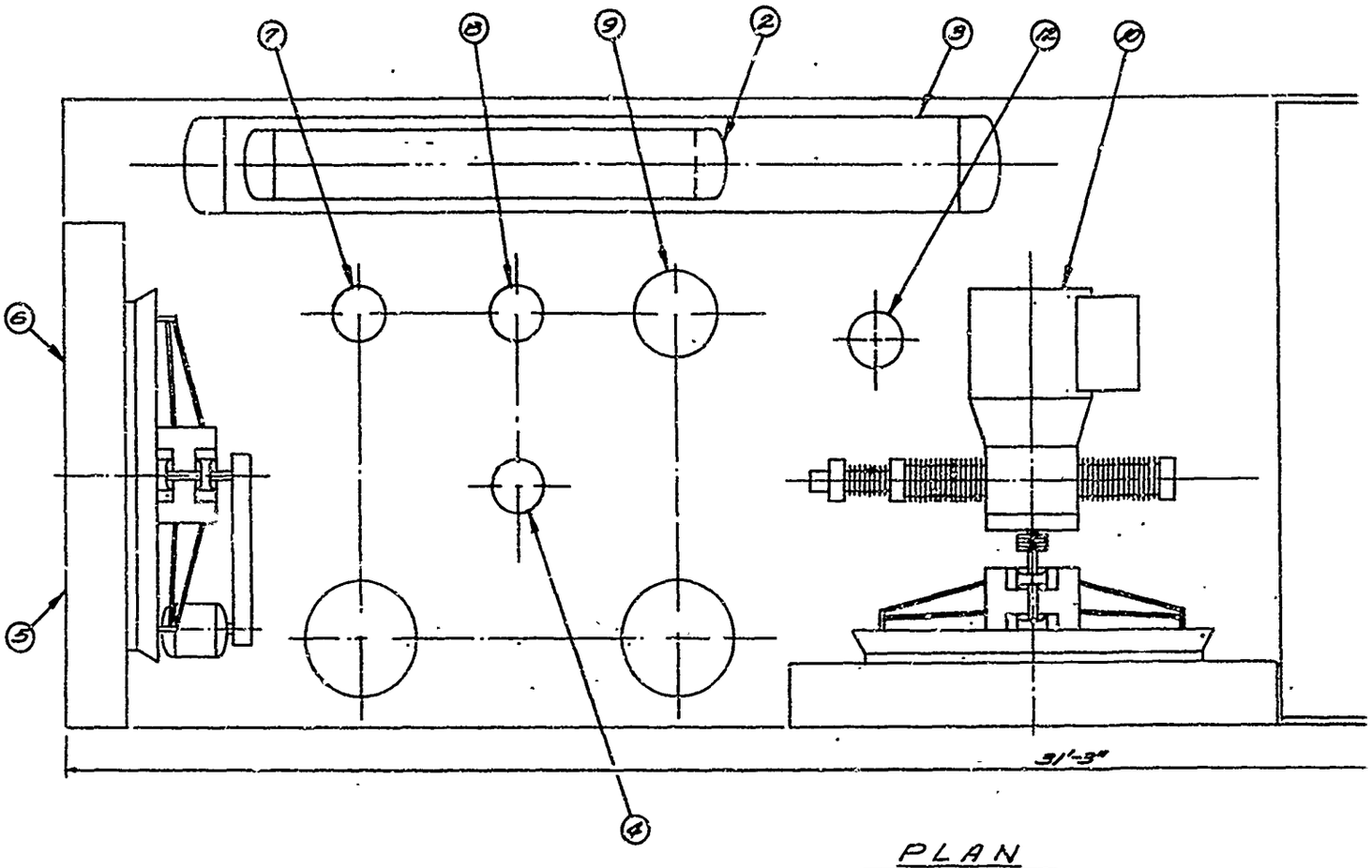


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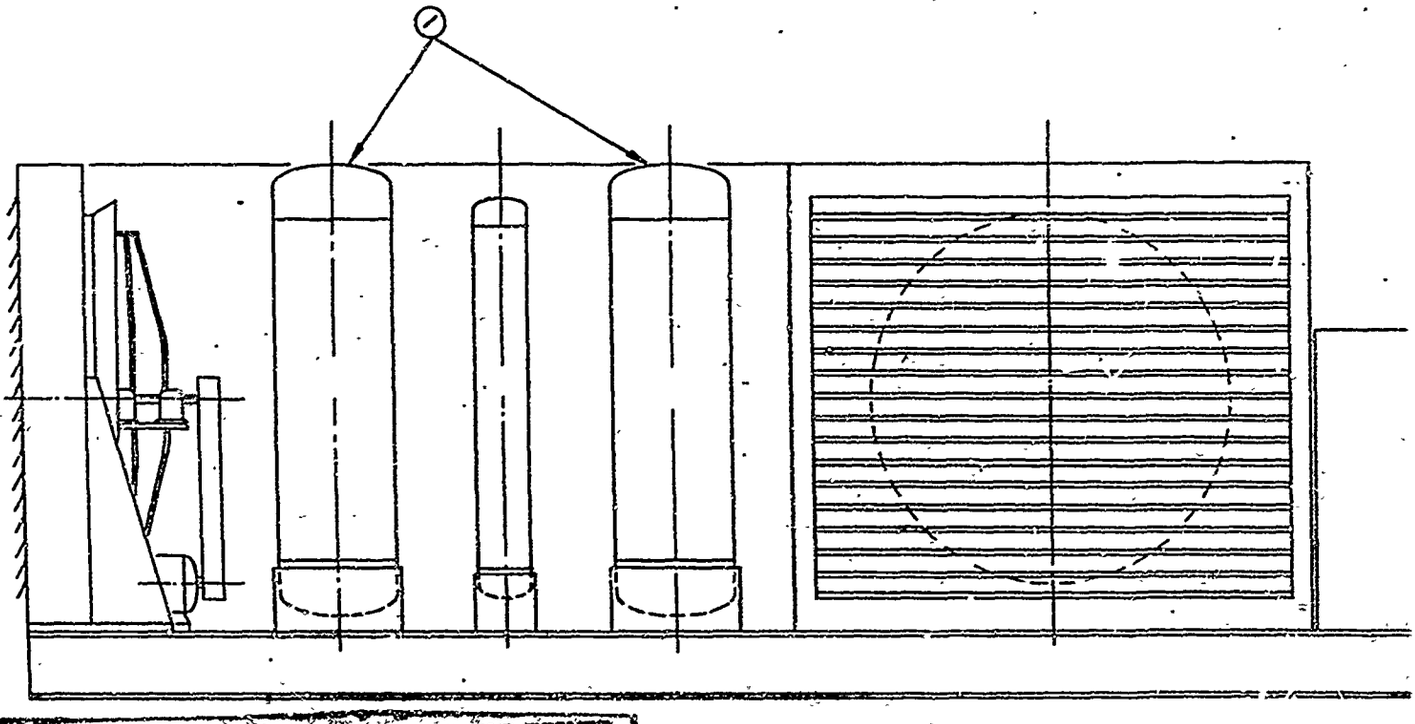
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2 H_2 PRECOOLER
3 SYNTHESIS GAS COMPRESSOR INTER & AFTER COOLERS
4 SYNTHESIS GAS COMPRESSOR
5 DEOXO UNIT
6 KNOCK-OUT DRUM
7 GAS FILTER
8 WATER PURIFIER
9 WATER STORAGE TANK

NOTE: STRUCTURE & OUTER SKIN OMITTED FOR CLARITY

Figure 3.7-17. Ammonia package layout – Skid 1 – system 10 – Package 3.



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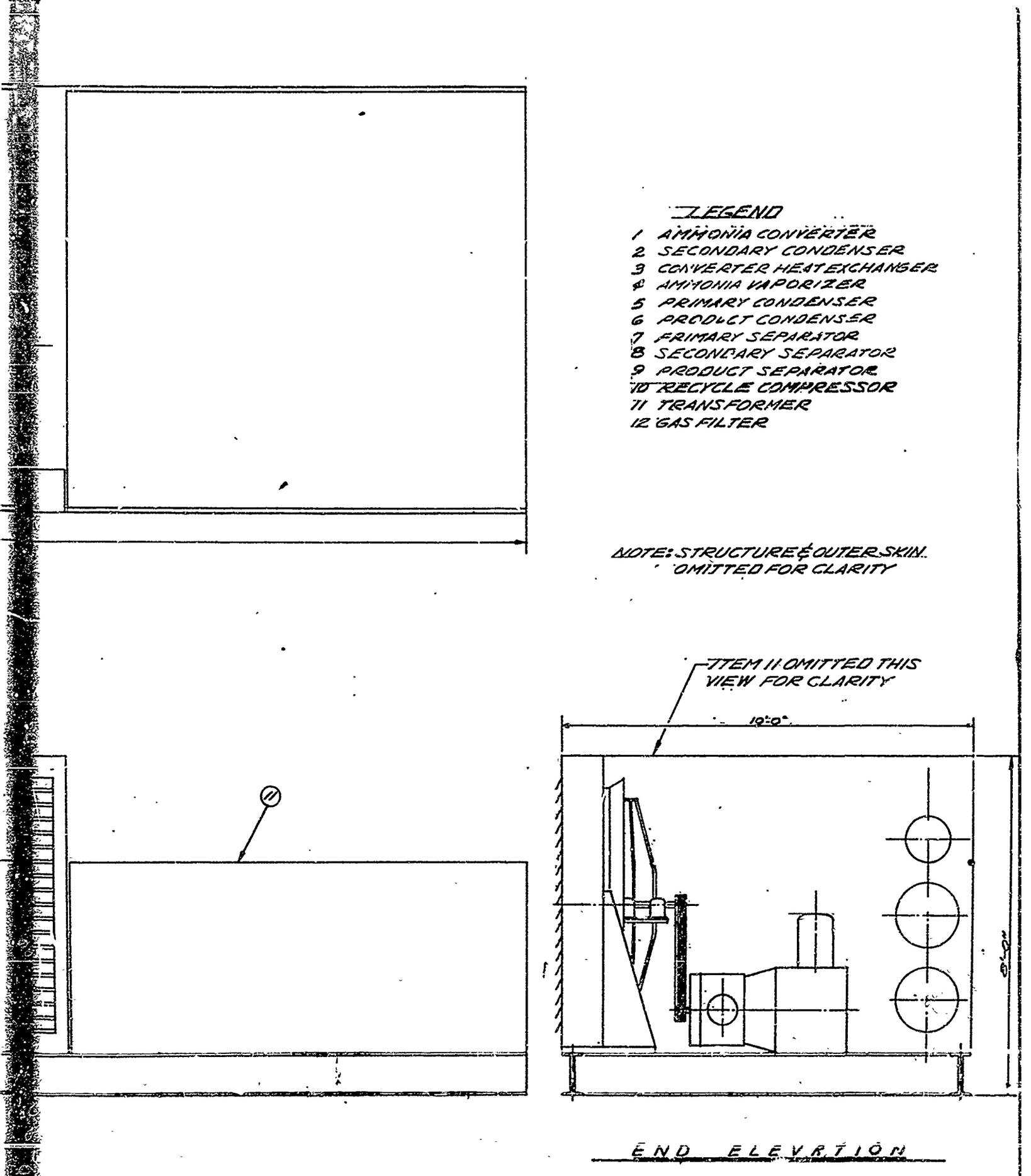
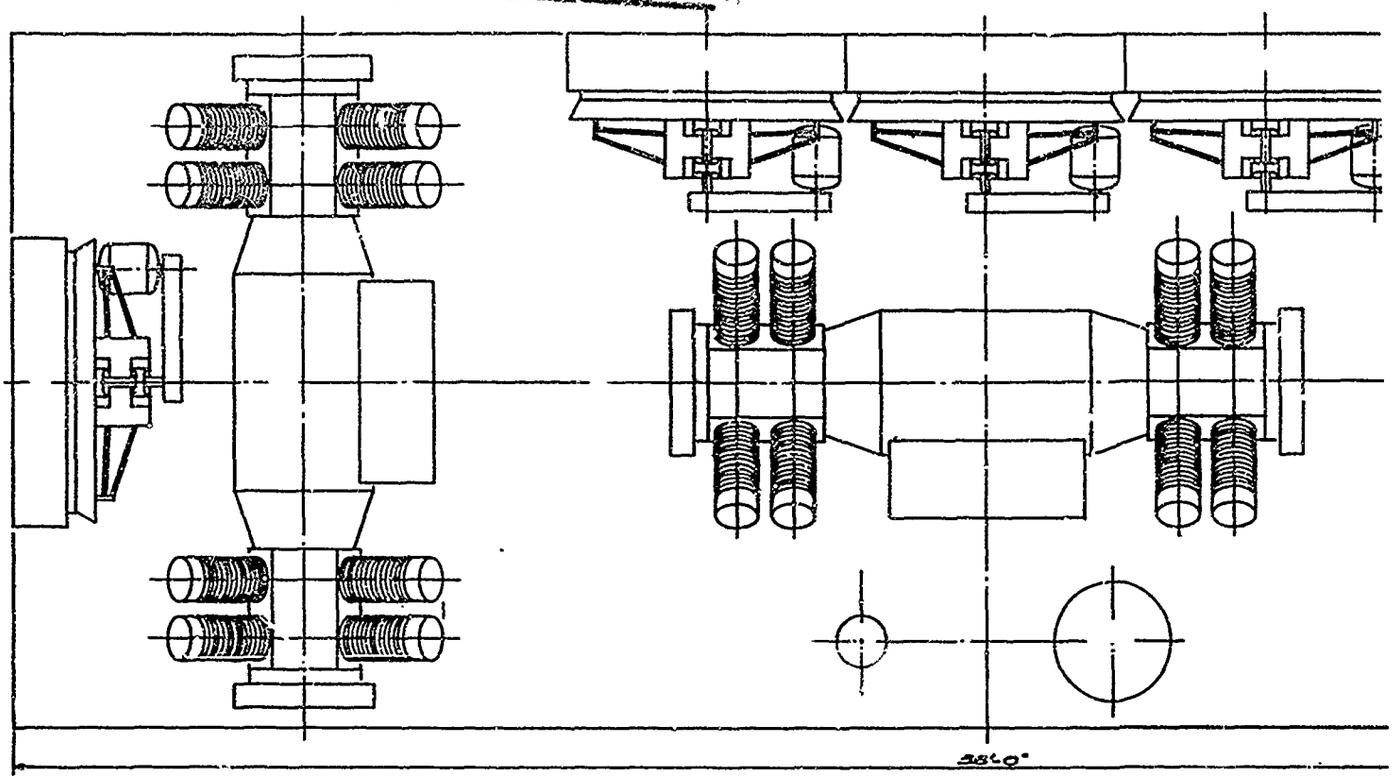
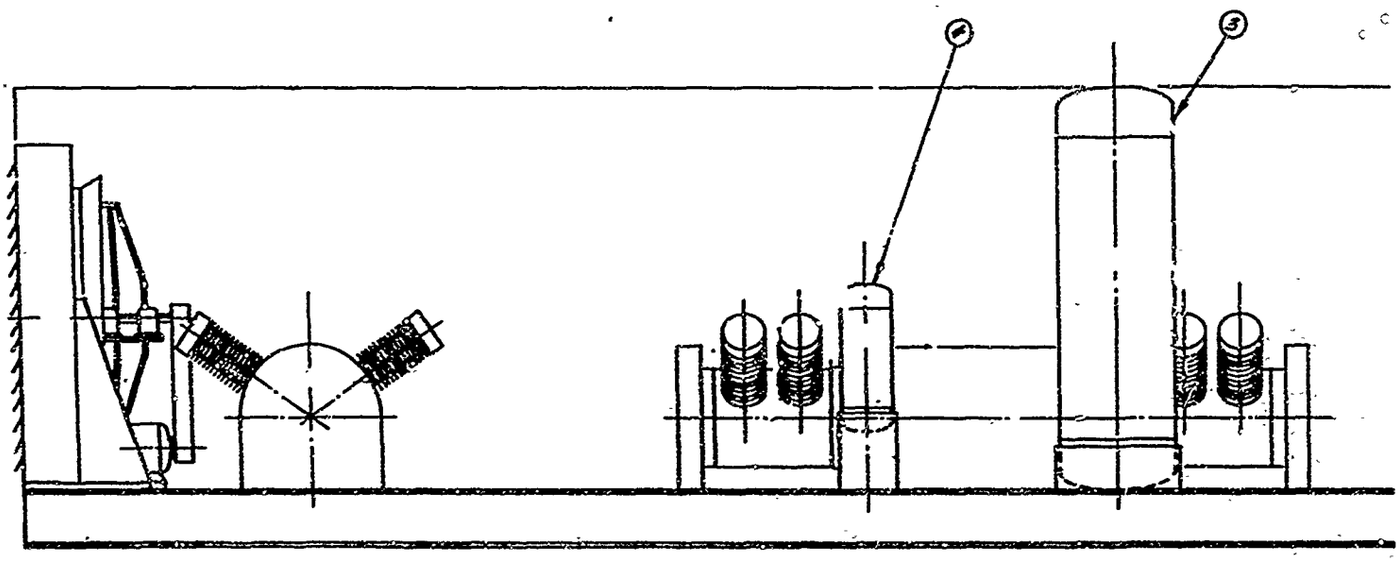


Figure 3.7-18. Ammonia package layout – Skid 2 – system 10 – Package 3a.

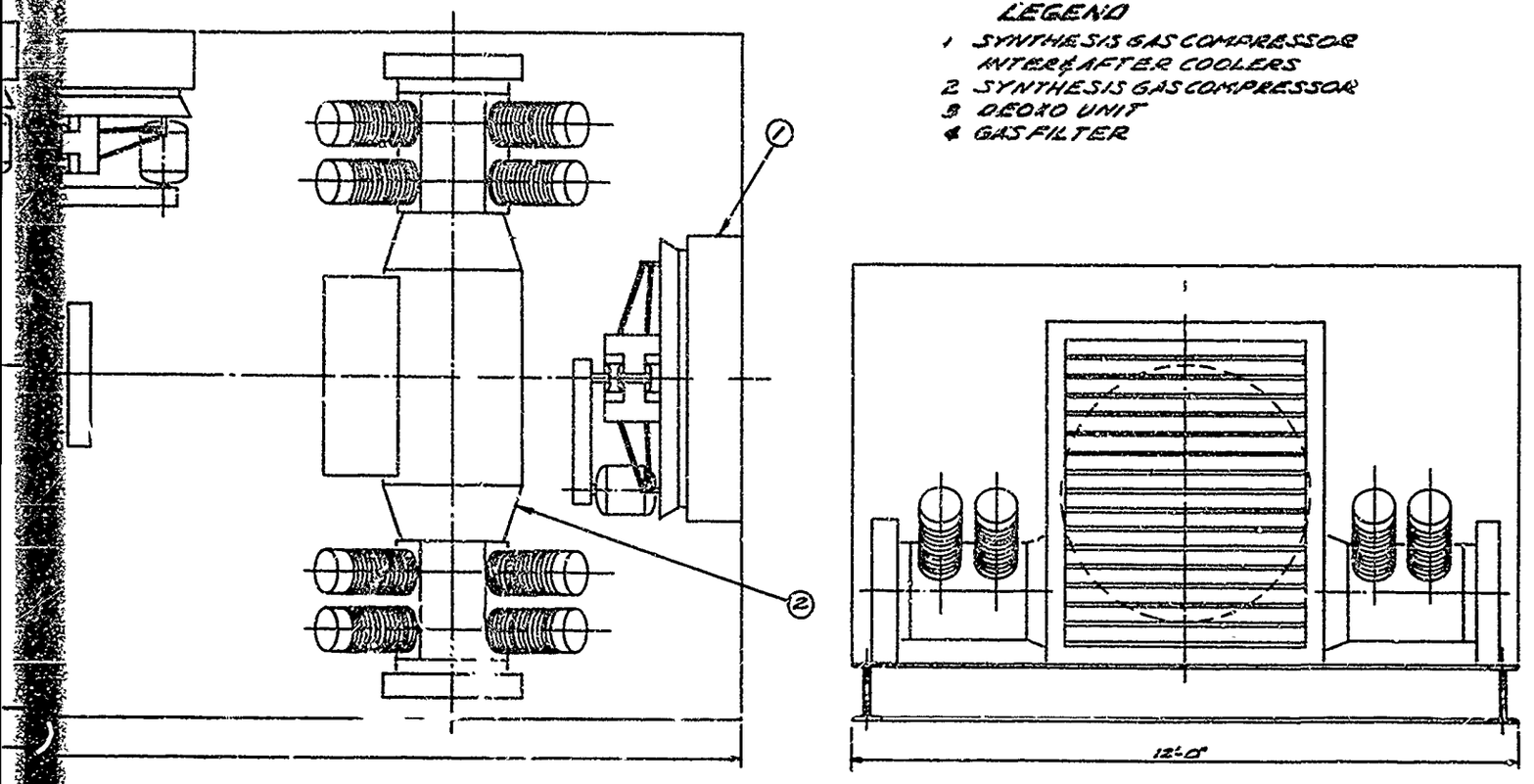
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PLAN



SIDE ELEVATION



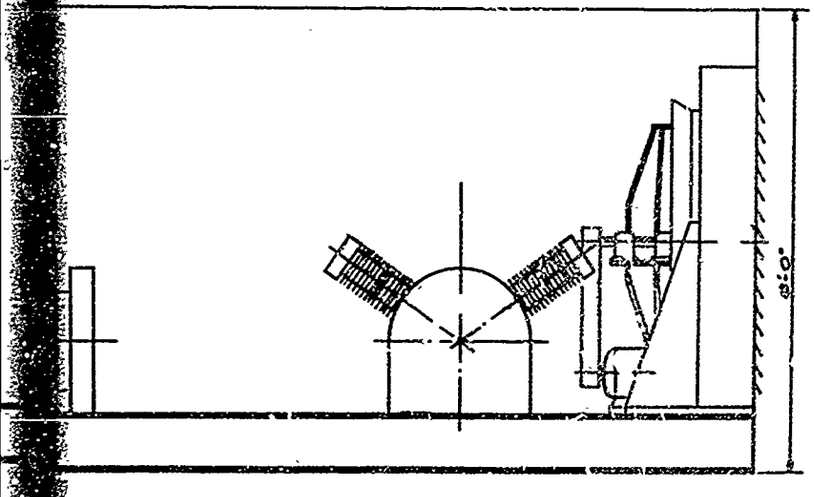
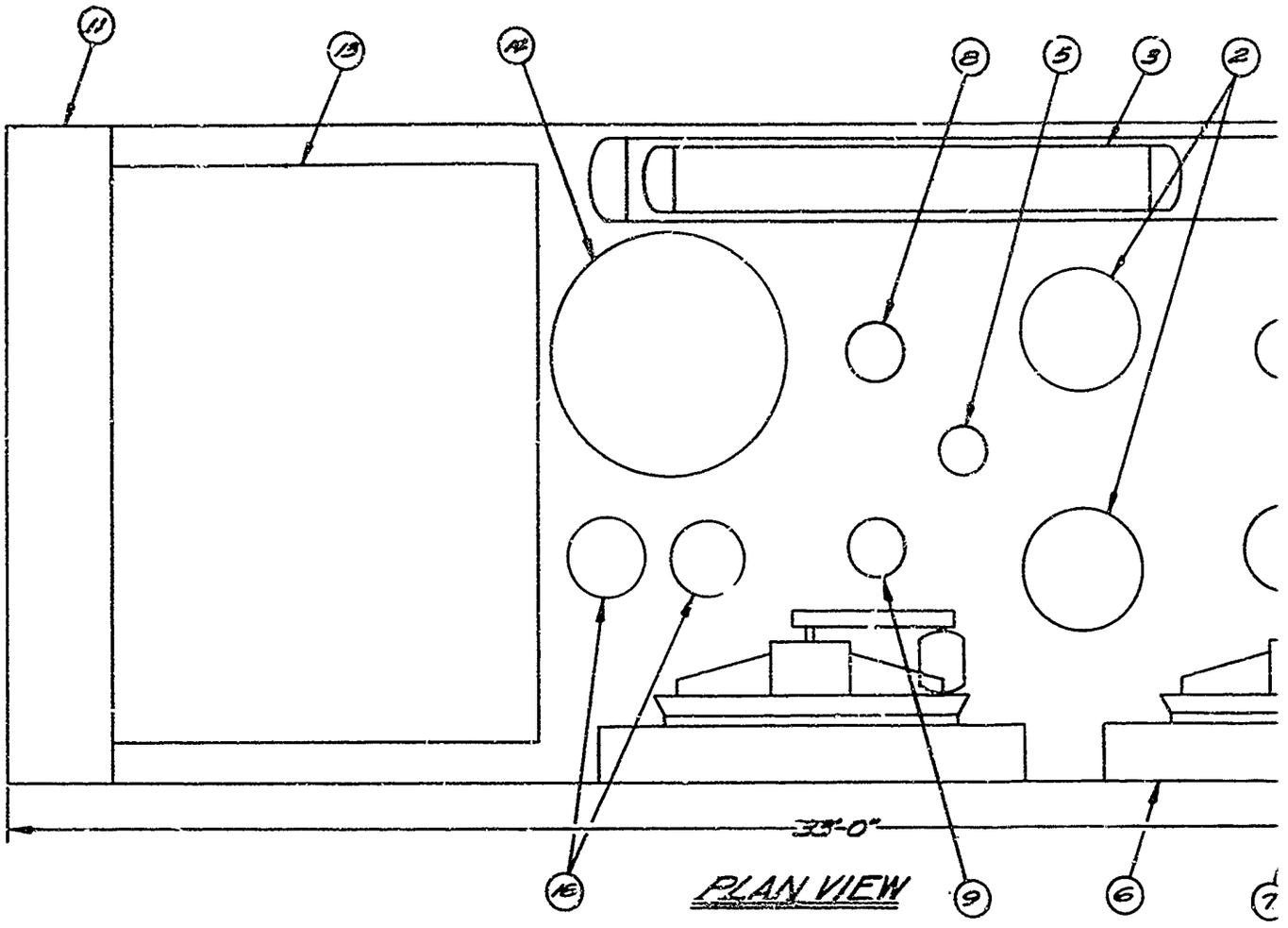
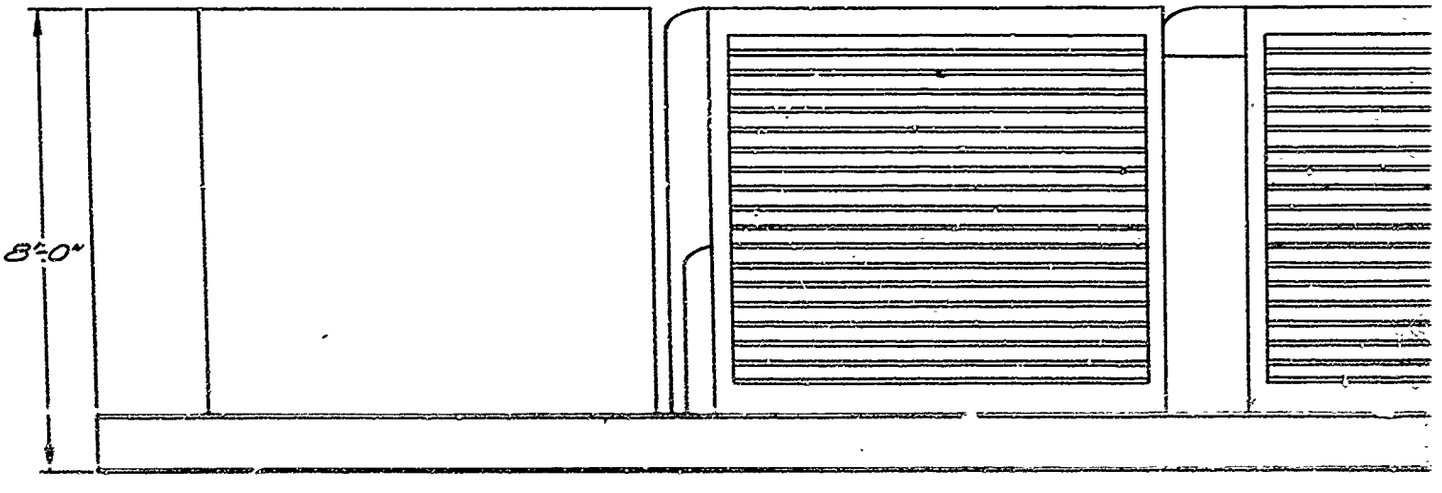


Figure 3.7-19. Ammonia package layout – system 11 – Package 3.

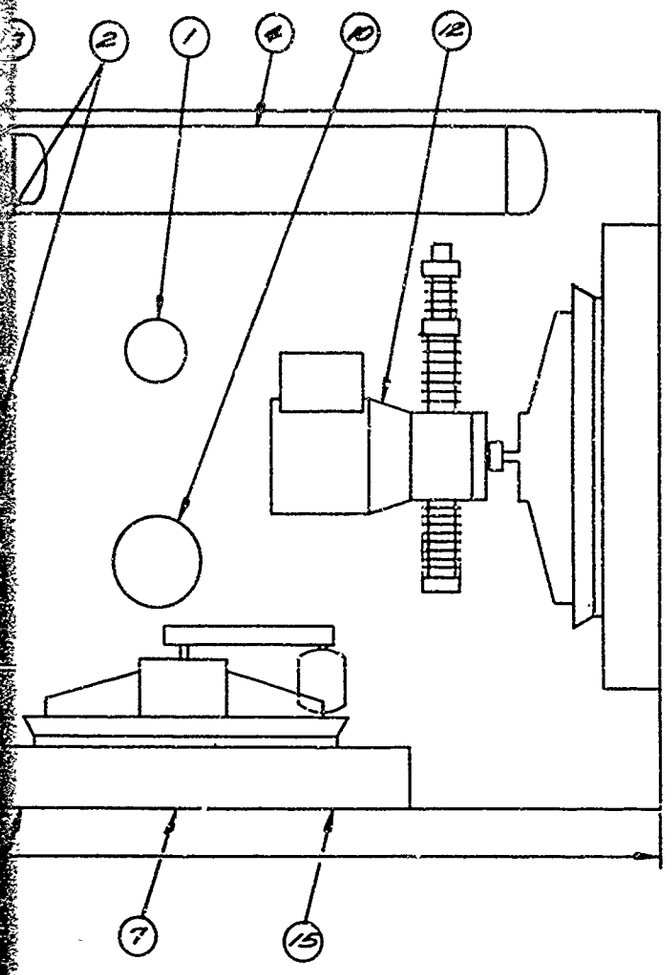


PLAN VIEW

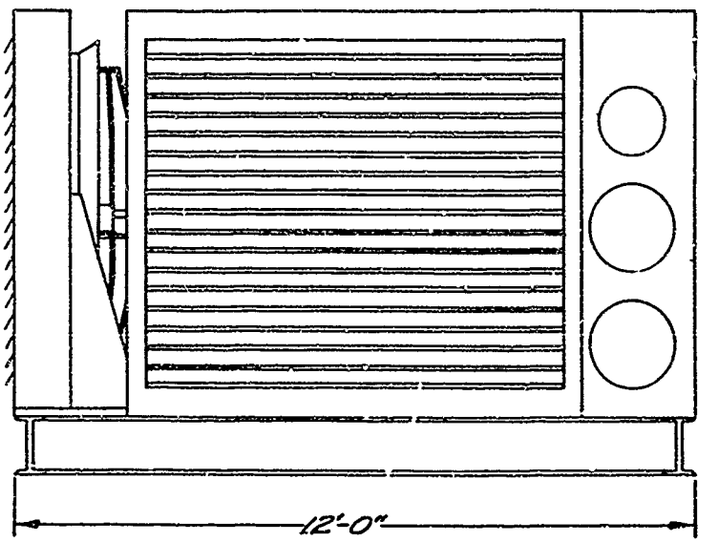


SIDE VIEW

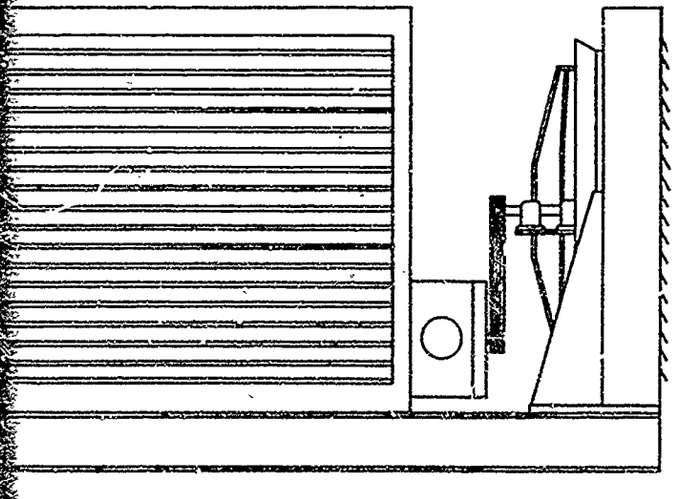
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NOTE: STRUCTURE & OUTER SKIN OMITTED FOR CLARITY



END VIEW



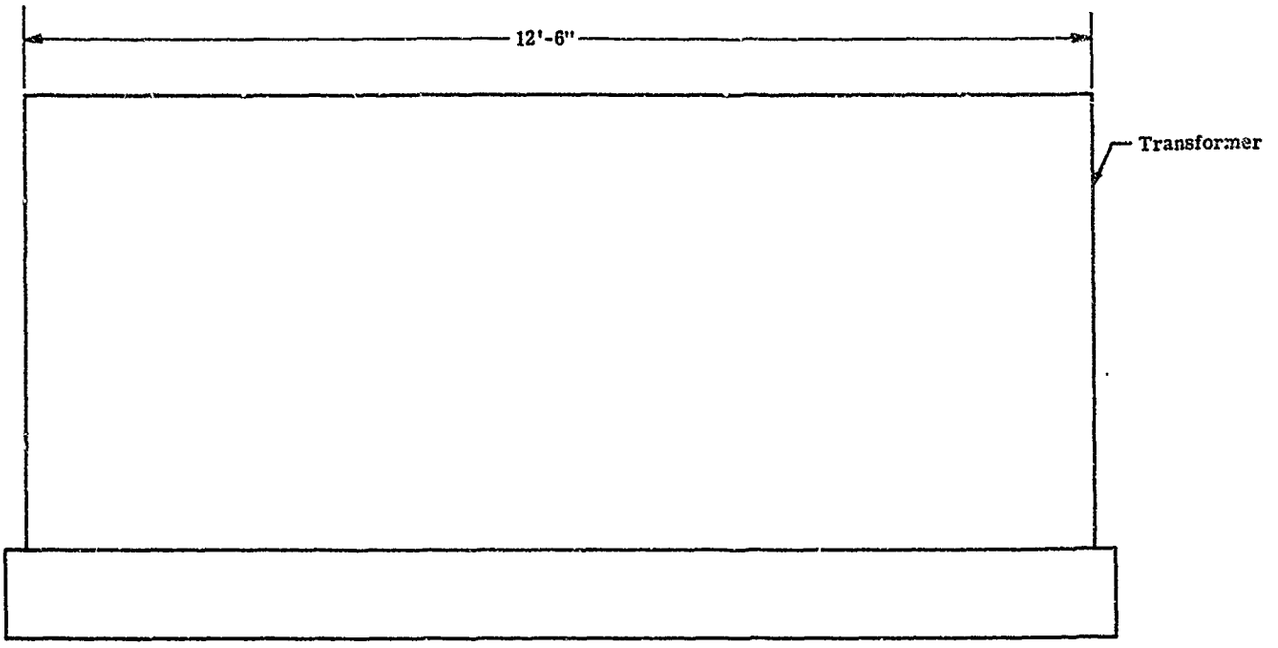
LEGEND

1 GAS FILTER
2 AMMONIA CONVERTER
3 SECONDARY CONDENSER
4 CONVERTER HEAT EXCHANGER
5 AMMONIA VAPORIZER
6 PRIMARY CONDENSER
7 PRODUCT CONDENSER
8 PRIMARY SEPARATOR
9 SECONDARY SEPARATOR
10 PRODUCT SEPARATOR
11 CONTROL CABINET
12 RECYCLE COMPRESSOR
13 WATER PURIFIER
14 WATER STORAGE TANK
15 H_2 PRECOOLER
16 KNOCK-OUT DRUM

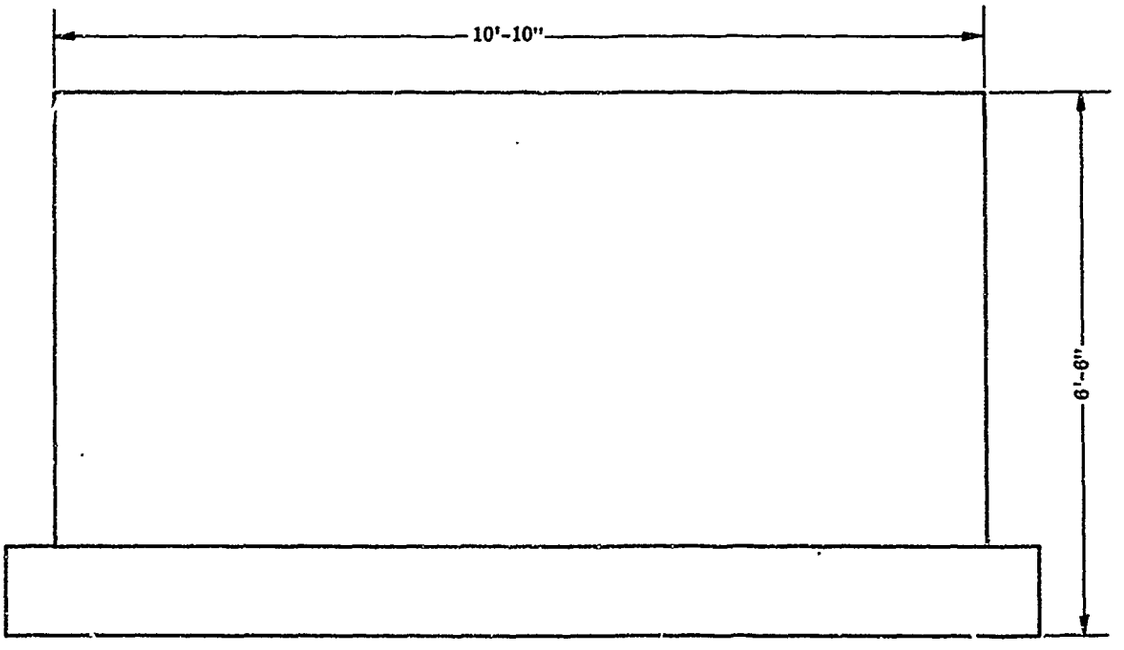
Figure 3.7-20. Ammonia package layout -- system 11 -- Package 3a.

3.7.41

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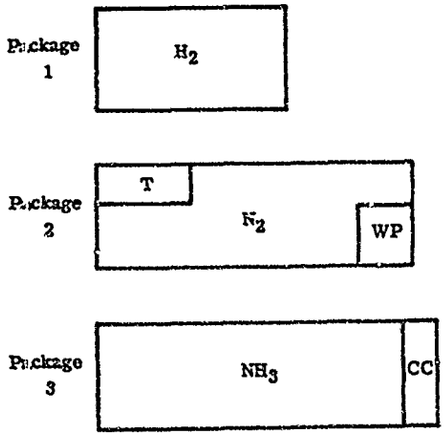
Side elevation



End elevation

4496-195

Figure 3.7-21. Transformer package – system 11 – Package 4.



Nomenclature

H_2	= Hydrogen Production Subsystem	Figure No. 3.3-11
N_2	= Nitrogen Generation Subsystem	Figure No. 3.7-4
NH_3	= Ammonia Synthesis Subsystem	Figure No. 3.7-9
CC	= Control cabinet	Figure No. 3.7-9
T	= Transformer	Figure No. 3.7-4
WP	= Water purifier	Figure No. 3.4-11

System data

Power: 3 Mw
Mobility Class: I
A-C power
Frequency: 60 cps
Cooling medium: air

Reference drawing

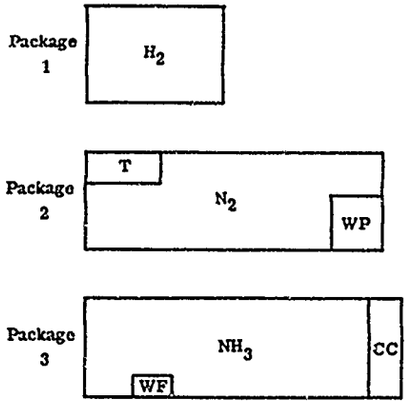
Interconnections*

No.	Purpose of interconnection	Origin (Package No. or name)	Terminus (Package No. or name)
1	Raw water intake	water supply	2
2	Exhaust water from purifier	2	water sump
3	Control—electrical	3	1
4	Control—electrical	3	2
5	Control—air	3	1
6	Control—air	3	2
7	Power	2	1
8	Power	2	1
9	Power	2	1
10	Power	2	3
11	Power	2	3
12	Power	2	3
13	Feedwater	2	1
14	Nitrogen	2	3
15	Oxygen	1	2
16	Hydrogen	1	3
17	Ammonia	3	ammonia storage

*Connections to the powerplant are not included.

4496-196

Figure 3.7-22. Packaging and interconnection data – system 1.



System data

Power: 3 Mw
Mobility Class: I
A-C power
Frequency: 60 cps
Cooling medium: water

Nomenclature

H_2	=	Hydrogen Production Subsystem
N_2	=	Nitrogen Generation Subsystem
NH_3	=	Ammonia Synthesis Subsystem
CC	=	Control cabinet
T	=	Transformer
WP	=	Water purifier
WF	=	Water filter

Reference

Figure No. 3.3-12
Figure No. 3.7-4
Figure No. 3.7-10
Figure No. 3.7-10
Figure No. 3.7-4
Figure No. 3.4-11
Figure No. 3.7-10

Interconnections*

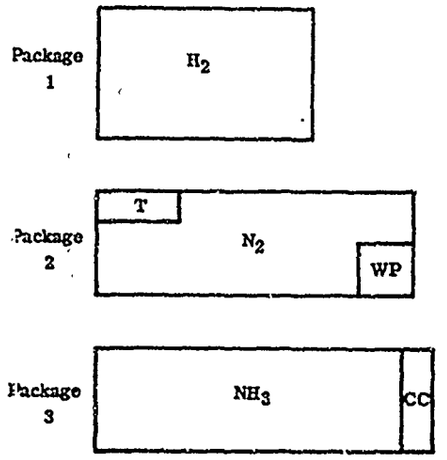
No.	Purpose of interconnection	Origin (Package No. or name)	Terminus (Package No. or name)
1	Raw water intake	water supply	2
2	Exhaust water from purifier	2	3
3	Purifier makeup water	3	2
4	Exhaust water from all plants	3	water sump
5	Control—electrical	3	1
6	Control—electrical	3	2
7	Control—air	3	2
8	Control—air	3	1
9	Power	2	1
10	Power	2	1
11	Power	2	1
12	Power	2	3
13	Power	2	3
14	Power	2	3
15	H_2 Cooling water intake	3	1
16	N_2 Cooling water intake	3	2
17	N_2 Cooling water exhaust	2	3
18.	H_2 Cooling water exhaust	1	3
19	Feedwater	2	1
20	Nitrogen	2	3
21	Oxygen	1	2
22	Hydrogen	1	3
23	Ammonia	3	ammonia storage

*Connections to the powerplant are not included.

4496-197

Figure 3.7-23. Packaging and interconnection data – system 2.

Allison



System data

Power: 3 Mw
Mobility Class: II
A-C power
Frequency: 60 cps
Cooling medium: air

Nomenclature

		Reference drawing
H_2	= Hydrogen Production Subsystem	Figure No. 3.3-13
N_2	= Nitrogen Generation Subsystem	Figure No. 3.7-4
NH_3	= Ammonia Synthesis Subsystem	Figure No. 3.7-9
CC	= Control cabinet	Figure No. 3.7-9
T	= Transformer	Figure No. 3.7-4
WP	= Water purifier	Figure No. 3.4-11

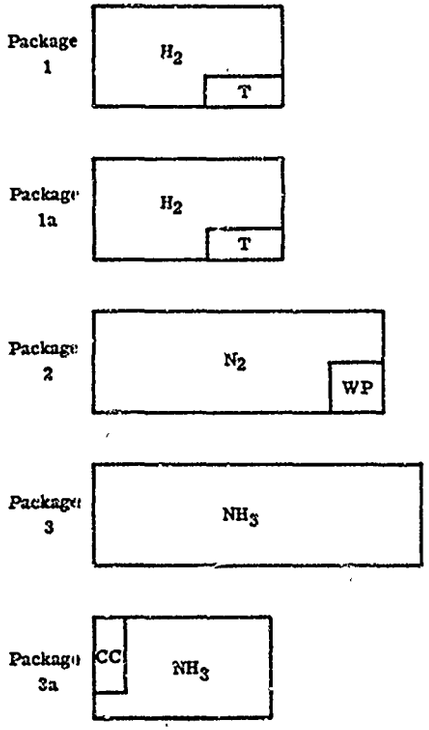
Interconnections*

No.	Purpose of interconnection	Origin (Package No. or name)	Terminus (Package No. or name)
1	Raw water intake	water supply	2
2	Exhaust water from purifier	2	water sump
3	Control--electrical	3	1
4	Control--electrical	3	2
5	Control--air	3	1
6	Control--air	3	2
7	Power	2	1
8	Power	2	1
9	Power	2	1
10	Power	2	3
11	Power	2	3
12	Power	2	3
13	Feedwater	2	1
14	Nitrogen	2	3
15	Oxygen	1	2
16	Hydrogen	1	3
17	Ammonia	3	ammonia storage

*Connections to the powerplant are not included.

4496-198

Figure 3.7-24. Packaging and interconnection data – system 3.



System data

Power: 6 Mw
Mobility Class: I
A-C power
Frequency: 60 cps
Cooling medium: air

Nomenclature

		Reference drawing
H_2	= Hydrogen Production Subsystem	Figure No. 3.3-14
N_2	= Nitrogen Generation Subsystem	Figure No. 3.7-5
NH_3	= Ammonia Synthesis Subsystem	Figure No. 3.7-11 & 3.7-12
CC	= Control cabinet	Figure No. 3.7-12
T	= Transformer	Figure No. 3.3-14
WP	= Water purifier	Figure No. 3.4-12

Interconnections*

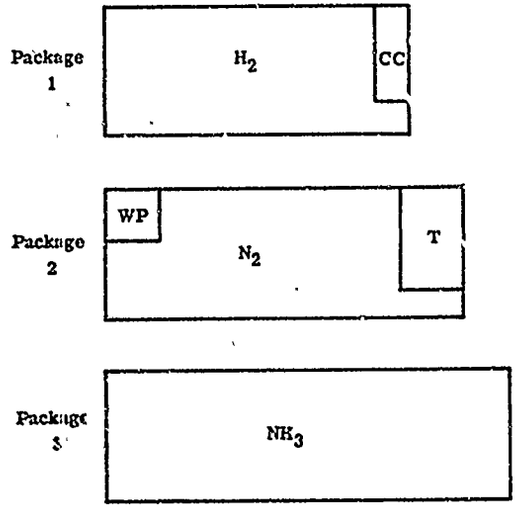
No.	Purpose of interconnection	Origin (Package No. or name)	Terminus (Package No. or name)
1	Raw water intake	water supply	2
2	Exhaust water from purifier	2	water sump
3	Control—electrical	3a	2
4	Control—electrical	3a	1a
5	Control—electrical	3a	1
6	Control—electrical	3a	3
7	Control—air	3a	3
8	Control—air	3a	1
9	Control—air	3a	1a
10	Control—air	3a	2
11	Power	1	1a
12	Power	1	1a
13	Power	1	1a
14	Power	1a	3
15	Power	1a	3
16	Power	1a	3
17	Power	1a	3a
18	Power	1a	3a
19	Power	1a	3a
20	Power	1a	2
21	Power	1a	2
22	Power	1a	2
23	Feedwater	2	1a
24	Feedwater	1a	1
25	Nitrogen	2	3
26	Oxygen	1	1a
27	Oxygen	1a	2
28	Hydrogen	1	1a
29	Hydrogen	1a	3
30	Recycle gas intake	3a	3
31	Recycle gas discharge	3	3a
32	Ammonia	3a	ammonia storage

*Connections to the powerplant are not included.

4496-199

Figure 3.7-25. Packaging and interconnection data – system 4.

Allison



System data

Power: 6 Mw
Mobility Class: II
A-C power
Frequency: 60 cps
Cooling medium: air

Nomenclature		Reference drawing
H_2	= Hydrogen Production Subsystem	Figure No. 3.3-15
N_2	= Nitrogen Generation Subsystem	Figure No. 3.7-6
NH_3	= Ammonia Synthesis Subsystem	Figure No. 3.7-13
T	= Transformer	Figure No. 3.7-6
CC	= Control cabinet	Figure No. 3.3-15
WP	= Water purifier	Figure No. 3.4-12

Interconnections*

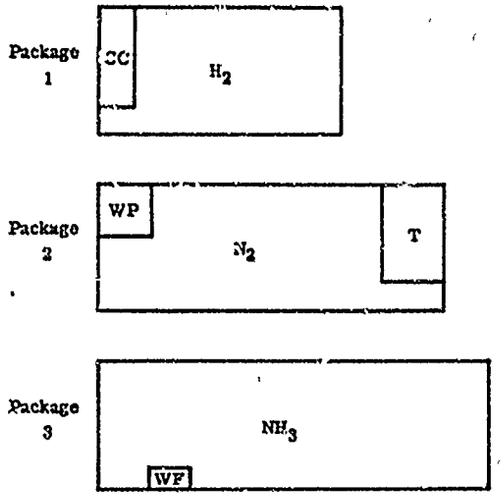
No.	Purpose of interconnection	Origin (Package No. or name)	Terminus (Package No. or name)
1	Raw water intake	water supply	2
2	Exhaust water from purifier	2	water sump
3	Control—electrical	1	2
4	Control—electrical	1	3
5	Control—air	1	2
6	Control—air	1	3
7	Power	2	1
8	Power	2	1
9	Power	2	1
10	Power	2	1
11	Power	2	1
12	Power	2	1
13	Power	2	3
14	Power	2	3
15	Power	2	3
16	Feedwater	2	2
17	Nitrogen	2	3
18	Oxygen	1	2
19	Hydrogen	1	3
20	Ammonia	3	ammonia storage

*Connections to the powerplant are not included.

4496-200

Figure 3.7-26. Packaging and interconnection data – system 5.

Allison



System data

Power: 6 Mw
Mobility Class: II
A-C power
Frequency: 60 cps
Cooling medium: water

Nomenclature		Reference drawing
H_2	= Hydrogen Production Subsystem	Figure No. 3.3-16
N_2	= Nitrogen Generation Subsystem	Figure No. 3.7-6
NH_3	= Ammonia Synthesis Subsystem	Figure No. 3.7-14
CC	= Control cabinet	Figure No. 3.3-16
T	= Transformer	Figure No. 3.7-6
WF	= Water filter	Figure No. 3.7-14
WP	= Water purifier	Figure No. 3.4-12

Interconnections*

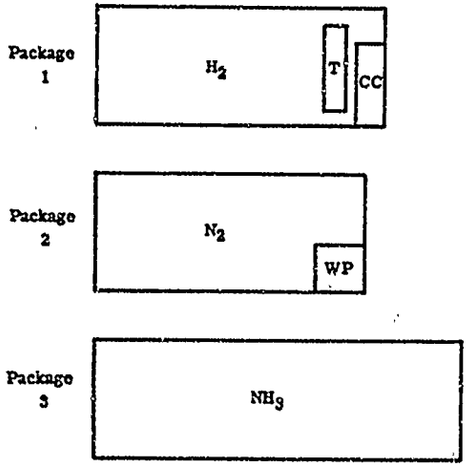
No.	Purpose of interconnection	Origin (Package No. or name)	Terminus (Package No. or name)
1	Raw water intake	water supply	3
2	Exhaust water from all plants	3	water sump
3	Purifier makeup water	3	2
4	Exhaust water from purifier	2	3
5	N_2 Cooling water intake	3	2
6	H_2 Cooling water intake	3	1
7	H_2 Cooling water exhaust	1	3
8	N_2 Cooling water exhaust	2	3
9	Control—electrical	1	2
10	Control—electrical	1	3
11	Control—air	1	2
12	Control—air	1	3
13	Power	2	3
14	Power	2	3
15	Power	2	3
16	Power	3	1
17	Power	2	1
18	Power	2	1
19	Power	2	1
20	Power	2	1
21	Power	2	1
22	Feedwater	2	1
23	Nitrogen	2	3
24	Oxygen	1	2
25	Hydrogen	1	3
26	Ammonia	3	ammonia storage

*Connections to the powerplant are not included.

4496-201

Figure 3.7-27. Packaging and interconnection data – system 6.

Allison



System data

Power: 6 Mw
Mobility Class: II
A-C power
Frequency: 400 cps
Cooling medium: air

Nomenclature

		Reference drawing
H_2	= Hydrogen Production Subsystem	Figure No. 3.3-17
N_2	= Nitrogen Generation Subsystem	Figure No. 3.7-5
NH_3	= Ammonia Synthesis Subsystem	Figure No. 3.7-13
CC	= Control cabinet	Figure No. 3.3-17
T	= Transformer	Figure No. 3.3-17
WP	= Water purifier	Figure No. 3.4-12

Interconnections*

No.	Purpose of interconnection	Origin (Package No. or name)	Terminus (Package No. or name)
1	Raw water intake	water supply	2
2	Exhaust water from purifier	2	water sump
3	Feedwater	2	1
4	Control—electrical	1	2
5	Control—electrical	1	3
6	Control—air	1	2
7	Control—air	1	3
8	Power	1	2
9	Power	1	2
10	Power	1	2
11	Power	1	3
12	Power	1	3
13	Power	1	3
14	Nitrogen	2	3
15	Oxygen	1	2
16	Hydrogen	1	3
17	Ammonia	3	ammonia storage

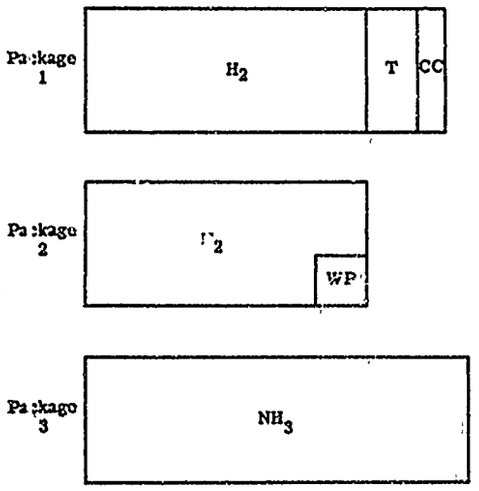
*Connections to the powerplant are not included.

4496-202

3.7-50

Allison

Figure 3.7-28. Packaging and interconnection data -- system 7.



System data

Power: 6 Mw
Mobility Class: III
A-C power
Frequency: 60 cps
Cooling medium: air

Nomenclature

H_2	= Hydrogen Production Subsystem	Figure No. 3.3-18
N_2	= Nitrogen Generation Subsystem	Figure No. 3.7-5
NH_3	= Ammonia Synthesis Subsystem	Figure No. 3.7-13
CC	= Control cabinet	Figure No. 3.3-18
T	= Transformer	Figure No. 3.3-18
WP	= Water purifier	Figure No. 3.4-12

Interconnections*

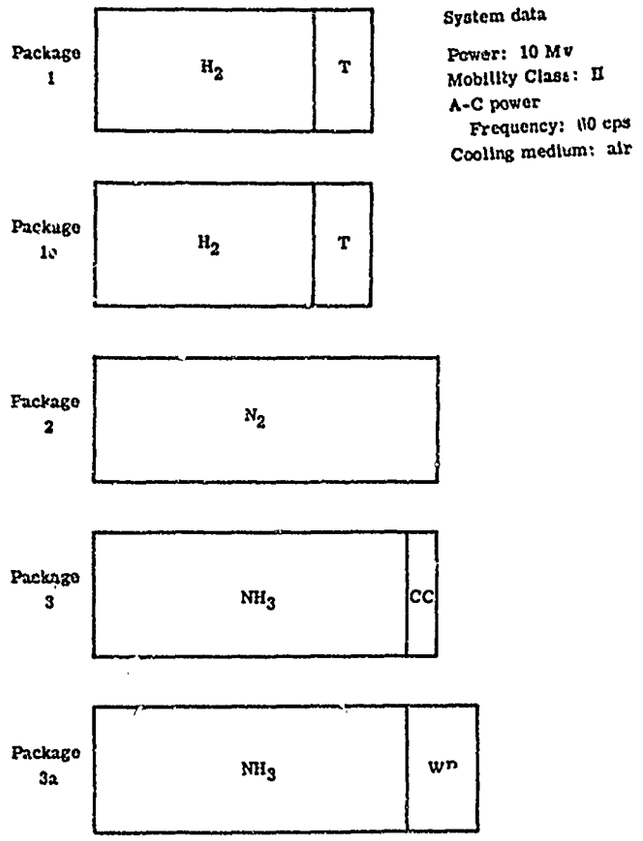
No.	Purpose of interconnection	Origin (Package No. or name)	Terminus (Package No. or name)
1	Raw water intake	water supply	2
2	Exhaust water from purifier	2	water sump
3	Feedwater	2	1
4	Control—electrical	1	2
5	Control—electrical	1	3
6	Control—air	1	2
7	Control—air	1	3
8	Power	1	2
9	Power	1	2
10	Power	1	2
11	Power	1	3
12	Power	1	3
13	Power	1	3
14	Nitrogen	2	3
15	Oxygen	1	2
16	Hydrogen	1	3
17	Ammonia	3	ammonia storage

*Connections to the powerplant are not included.

4496-203

Figure 3.7-29. Packaging and interconnection data – system 8.

Allison



System data

Power: 10 Mw
Mobility Class: II
A-C power
Frequency: 400 cps
Cooling medium: air

Nomenclature		Reference drawing
H_2	= Hydrogen Production Subsystem	Figure No. 3.3-19
N_2	= Nitrogen Generation Subsystem	Figure No. 3.7-7
NH_3	= Ammonia Synthesis Subsystem	Figure No. 3.7-15 & 3.7-16
CC	= Control cabinet	Figure No. 3.7-15
T	= Transformer	Figure No. 3.3-19
WP	= Water purifier	Figure No. 3.4-12

Interconnections*

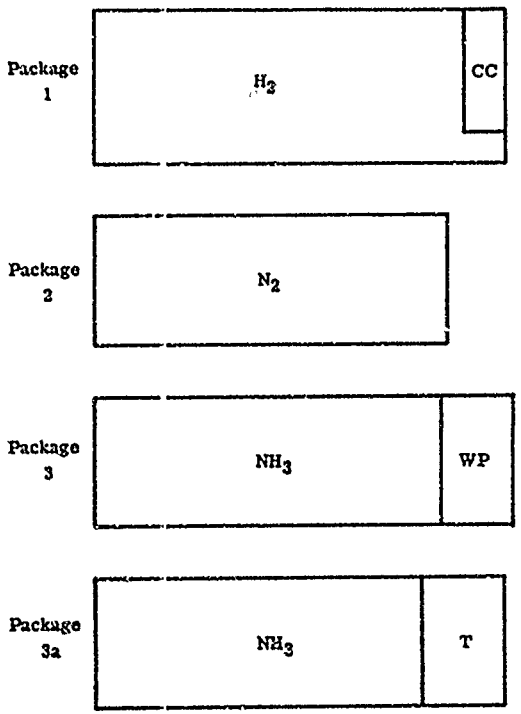
No.	Purpose of interconnection	Origin (Package No. or name)	Terminus (Package No. or name)
1	Raw water intake	water supply	3a
2	Exhaust water from purifier	3a	water sump
3	Feedwater	3a	1a
4	Feedwater	1a	1
5	Control—electrical	3	1
6	Control—electrical	3	1a
7	Control—electrical	3	2
8	Control—electrical	3	3a
9	Control—air	3	1
10	Control—air	3	1a
11	Control—air	3	2
12	Control—air	3	3a
13	Power	1	1a
14	Power	1	1a
15	Power	1	1a
16	Power	1a	3a
17	Power	1a	3a
18	Power	1a	3a
19	Power	1a	3
20	Power	1a	3
21	Power	1a	3
22	Power	1a	2
23	Power	1a	2
24	Power	1a	2
25	Nitrogen	2	3
26	Oxygen	1	1a
27	Oxygen	1a	2
28	Hydrogen	1	1a
29	Hydrogen	1a	3
30	Synthesis gas	3	3a
31	Compressor blow by	3	2
32	Ammonia	3a	ammonia storage

*Connections to the powerplant are not included.

4496-204

Figure 3.7-30. Packaging and interconnection data – system 9.

Allison



System data

Power: 10 Mw
Mobility Class: III
A-C power
Frequency: 60 cps
Cooling medium: air

Nomenclature		Reference drawing
H_2	= Hydrogen Production Subsystem	Figure No. 3.3-20
N_2	= Nitrogen Generation Subsystem	Figure No. 3.7-7
NH_3	= Ammonia Synthesis Subsystem	Figure No. 3.7-17 & 3.
CC	= Control cabinet	Figure No. 3.3-20
T	= Transformer	Figure No. 3.7-18
WP	= Water purifier	Figure No. 3.4-12

Interconnections*

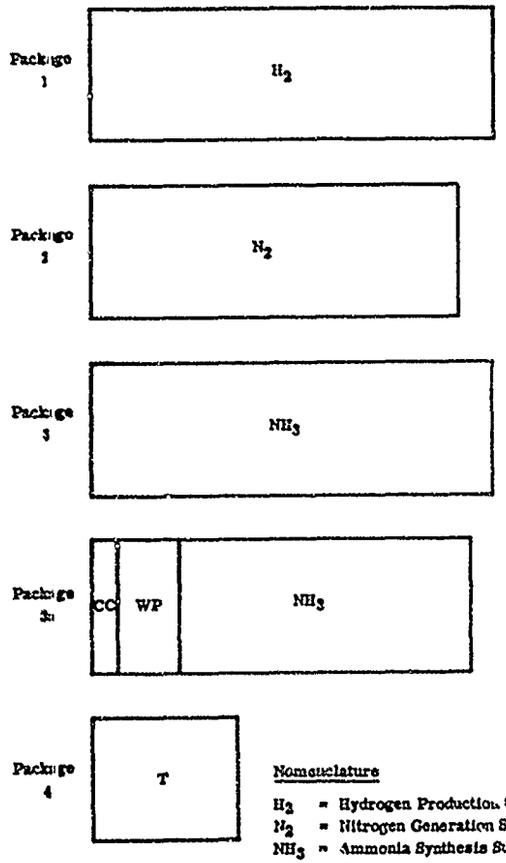
No.	Purpose of interconnection	Origin (Package No. or name)	Terminus (Package No. or name)
1	Raw water intake	water supply	3
2	Exhaust water from purifier	3	water sump
3	Feedwater	3	1
4	Control—electrical	1	2
5	Control—electrical	1	3
6	Control—electrical	1	3a
7	Control—air	1	2
8	Control—air	1	3
9	Control—air	1	3a
10	Power	3a	1
11	Power	3a	1
12	Power	3a	1
13	Power	3a	1
14	Power	3a	1
15	Power	3a	1
16	Power	3a	1
17	Power	3a	1
18	Power	3a	1
19	Power	3a	1
20	Power	3a	1
21	Power	3a	1
22	Power	3a	3
23	Power	3a	3
24	Power	3a	3
25	Power	3a	2
26	Power	3a	2
27	Power	3a	2
28	Nitrogen	2	3
29	Oxygen	1	2
30	Hydrogen	1	3
31	Synthesis gas	3	3a
32	Compressor blow by	3	2
33	Ammonia	3a	ammonia storage

*Connections to the powerplant are not included.

4496-205

Figure 3.7-31. Packaging and interconnection data – system 10.

Allison



System data

Power: 15 Mw
Mobility Class: III
A-C power
Frequency: 60 cps
Cooling medium: air

Nomenclature

H_2	= Hydrogen Production Subsystem	Figure No. 3.3-21
N_2	= Nitrogen Generation Subsystem	Figure No. 3.7-8
NH_3	= Ammonia Synthesis Subsystem	Figure No. 3.7-12 & 3.7-20
CC	= Control cabinet	Figure No. 3.7-20
T	= Transformer	Figure No. 3.7-21
WP	= Water purifier	Figure No. 3.4-12

Interconnections*

No.	Purpose of interconnection	Origin (Package No. or name)	Termination (Package No. or name)
1	Raw water intake	water supply	3a
2	Exhaust water from purifier	3a	water sump
3	Feedwater	3a	1
4	Control—electrical	3a	2
5	Control—electrical	3a	4
6	Control—electrical	3a	1
7	Control—electrical	3a	3
8	Control—air	3a	3
9	Control—air	3a	1
10	Control—air	3a	2
11	Power	4	3a
12	Power	4	3a
13	Power	4	3a
14	Power	4	2
15	Power	4	2
16	Power	4	2
17	Power	4	3
18	Power	4	3
19	Power	4	3
20	Power	4	1
21	Power	4	1
22	Power	4	1
23	Power	4	1
24	Power	4	1
25	Power	4	1
26	Power	4	1
27	Power	4	1
28	Power	4	1
29	Power	4	1
30	Power	4	1
31	Power	4	1
32	Power	4	1
33	Power	4	1
34	Power	4	1
35	Nitrogen	2	3
36	Oxygen	1	2
37	Hydrogen	1	3
38	Synthesis gas	3	3a
49	Compressor blow by	3	3
40	Ammonia	3a	ammonia storage

*Connections to the powerplant are not included.

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Figure 3.7-32. Packaging and interconnection data – system 11.

Table 3.7-III.
System packaging data summary.

	System								
	1	2	3	4	5	6	7	8	9
System electrical power input—Mw	3	3	3	6	6	6	6	6	10
Mobility Class	I	I	II	I	II	II	II	III	II
Cooling medium	air	water	air	air	air	water	air	air	air
A-C frequency	60	60	60	60	60	60	400	60	6
Production rate—lb NH_3/hr	729	730	758	1,380	1,410	1,420	1,410	1,410	2,2
Total number of packages	3	3	3	5	3	3	3	3	5
Package weights (lb)									
H_2 package, package 1	35,000	33,700	54,350	35,000	61,510	58,320	65,000	86,410	63,
package 1a	—	—	—	35,000	—	—	—	—	63,
N_2 package, package 2	31,013	31,013	31,269	26,041	50,291	50,391	26,441	26,441	41,
NH_3 package, package 3	32,295	30,690	32,760	26,990	54,205	51,085	54,205	54,205	31,
package 3a	—	—	—	31,955	—	—	—	—	64,
Transformer package, package 4	—	—	—	—	—	—	—	—	—
Total system weight	98,300	95,400	118,000	155,000	166,000	160,000	146,000	167,000	264,
Package Dimension, length × width × height									
H_2 package, package 1	14' × 8' × 8'	11'10" × 8' × 8'		16'1" × 1' × 8'		14'2" × 8' × 8'		22'7" × 1	
package 1a	—	—		—		14'2" × 8' × 8'		—	
N_2 package, package 2	23'8" × 8' × 8'	23'8" × 8' × 8'		23'8" × 8' × 8'		22' × 8'		26'6" × 1	
NH_3 package, package 3	25' × 8' × 8'	25' × 8' × 8'		25' × 8' × 8'		25' × 3' × 8'		30' × 10'	
package 3a	—	—		—		13'6" × 8' × 8'		—	
Transformer package, package 4	—	—		—		—		—	

Table 3.7-III.

Packaging data summary.

8	9	10	11
6	10	10	15
III	II	III	III
air	air	air	air
60	60	60	60
1,410	2,290	2,290	3,330
3	5	4	5
86,410	63,210	86,130	100,000
--	63,210	—	—
26,441	41,465	41,465	69,607
54,205	31,210	40,920	48,390
--	64,940	90,760	92,575
—	—	—	49,600
167,000	264,000	259,000	360,000

× 8'	22'7" × 10' × 8'	18'5" × 10' × 8'	23'6" × 10' × 8'	28' × 10' × 8'	21'6" × 10' × 8'	30'11" × 12' × 8'	35' × 12' × 8'
× 8'	—	—	—	—	21'6" × 10' × 8'	—	—
	26'6" × 10' × 8'	26'6" × 10' × 8'	22' × 10' × 8'	22' × 10' × 8'	26'8" × 10' × 8'	26'8" × 10' × 8'	32' × 12' × 8'
	30' × 10' × 8'	30' × 10' × 8'	30' × 10' × 8'	30' × 10' × 8'	26'8" × 10' × 8'	31'7" × 10' × 8'	35' × 12' × 8'
× 8'	—	—	—	—	30' × 10' × 8'	31'3" × 10' × 8'	33' × 12' × 8'
	—	—	—	—	—	—	13' × 12' × 6'1"

Table 3.7-IV.

Hydrogen package (package 1) weights* (pounds).

Component	System 1	2	3	**4	5	6	7	8	**9	10	11
Modules	26,210	26,210	41,417	18,765	41,160	41,160	41,160	41,160	27,442	55,012	64,094
Heat exchangers (includes fan and motors)	1,555	396	1,290	1,767	3,443	694	3,443	3,443	2,945	5,899	7,611
Separator-reservoir	870	870	1,575	633	1,594	1,621	1,594	1,594	1,098	2,269	2,834
Electrolyte pump	615	615	580	595	2,132	2,106	2,132	2,132	1,797	2,990	3,863
Rectifier	300	300	449	592	1,576	1,581	1,576	1,576	1,128	2,304	3,168
Bus bars	270	270	423	193	565	567	565	565	340	1,125	1,679
Piping and valves	1,315	1,245	2,189	1,228	2,926	2,791	2,926	2,926	1,506	4,856	5,981
Switch gear	157	75	192	157	299	135	299	299	264	610	680
Feedwater pump	50	50	50	50	60	60	60	50	60	60	75
Miscellaneous (cables, controls, and safety equipment)	443	549	635	430	595	595	595	595	540	710	765
Transformer	—	—	—	7,400	—	—	3,200	22,200	20,200	—	—
Control cabinet	—	—	—	—	2,160	2,160	2,160	2,160	—	2,170	—
Skid and structure	3,215	3,120	5,550	3,190	5,000	4,850	5,290	7,700	5,890	8,125	9,250
Total	35,000	33,700	54,350	35,000	61,510	58,320	65,000	86,410	63,210	86,130	100,000

*Weights are given for the subsystem in its normal transportation mode.

**Two identical packages required.

Note: The electrolyte inventory typically weighs 30 to 35% of the subsystem weight.

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Table 3.7-V.

Nitrogen package (package 2) weights (pounds).

Component	System 1	2	3	4	5	6	7	8	9	10	11
Electric motors	1,460	1,460	1,460	2,320	2,320	2,320	2,320	2,320	5,684	5,684	7,636
Compressors and blowers	2,135	2,135	2,135	3,060	3,060	3,060	3,060	3,060	10,779	10,779	21,774
Continuous drainers	32	32	32	32	32	32	32	32	32	32	32
Liquid N_2 tank	540	540	550	835	860	870	860	860	1,950	1,950	4,450
Other tanks	860	860	880	1,485	1,540	1,540	1,540	1,540	2,350	2,350	2,670
Heat exchangers	1,060	1,060	1,100	2,030	2,090	2,100	2,090	2,090	3,180	3,180	4,300
Filters	66	66	67	92	92	92	92	92	110	110	160
Reactivation heater	75	75	80	140	140	140	140	140	55	55	80
Air purifiers (full)	530	530	560	1,090	1,100	1,110	1,100	1,100	1,720	1,720	2,400
Fractionation columns	800	800	830	1,460	1,500	1,510	1,500	1,500	2,290	2,290	3,100
Turboexpander	10	10	10	12	12	12	12	12	15	15	20
Valves	710	710	720	1,000	1,020	1,030	1,020	1,020	1,680	1,680	2,610
Piping and fittings	890	890	900	1,220	1,250	1,260	1,250	1,250	1,930	1,930	2,930
Electrical controls	145	145	150	205	210	210	210	210	2,070	2,070	2,180
Electric wiring and conduit	150	150	155	200	205	205	205	205	320	320	475
Cold box	460	460	465	555	560	560	560	560	620	620	775
Insulation	800	800	805	955	960	960	960	960	1,180	1,180	1,530
Miscellaneous	100	100	100	130	130	130	130	130	150	150	175
Transformer	13,500	13,500	13,500	—	22,200	22,200	—	—	—	—	—
Water purifier	2,460	2,460	2,500	4,060	4,160	4,180	4,160	4,160	—	—	—
Water storage tanks (wet)	1,300	1,300	1,300	2,600	2,600	2,600	2,600	2,600	—	—	—
Liquid N_2 stored	340	340	350	790	810	820	810	810	2,460	2,460	7,600
Skid and structure	2,590	2,590	2,620	1,770	3,440	3,450	1,790	1,790	2,890	2,890	4,710
Total	31,013	31,013	31,269	26,041	50,291	50,391	26,441	26,441	41,465	41,465	69,607

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Table 3.7-VI.

Ammonia package (packages 3 and 3a) weights (pounds).

Component	System 1	2	3	4	5	6	7	8	9	10	11
Compressor: Package 3	7,900	9,450	7 900	15,300	15,300	18,900	15,300	15,300	16,100	16,100	28,950
Package 3a									2,740	2,740	2,740
Deoxo, D-2: Package 3	450	450	470	790	805	810	805	805	1,250	1,250	1,780
Converter, D-2: Package 3	3,350	3,350	3,450		5,450	5,500	5,450	5,450			
Package 3a				5,350					8,250	8,250	11,400
Filters: Package 3	280	280	285	460	475	480	475	475	110	110	160
Package 3a									725	725	1,290
Knockout drums, D-4 and 5: Package 3	140	140	145	240	245	245	245	245	480	480	
Package 3a											535
Secondary condenser: Package 3	530	530	540		880	885	880	880			
Package 3a				860					1,330	1,330	1,850
Converter heat exchanger: Package 3	5,600	5,600	5,800		9,500	9,550	9,500	9,500			
Package 3a				9,800					15,000	15,000	20,100
Ammonia vaporizer: Package 3	335	335	340		525	530	525	525			
Package 3a				520					770	770	1,055
Coolers E-6 through 12: Package 3	4,250	295	4,250	4,050	8,750	570	8,750	8,750	4,100	4,100	7,000
Package 3a				5,050					9,850	9,850	14,900
Ejector: Package 3	20	20	20		35	35	35	35			
Package 3a				35					55	55	55
Separators: Package 3	1,040	1,040	1,060		1,720	1,730	1,720	1,720			
Package 3a				1,680					2,740	2,740	4,300
Instruments: Package 3	500	500	500	300	500	500	500	500	300	300	200
Package 3a				200					200	200	300
Piping: Package 3	3,350	3,350	3,450	3,350	6,020	6,050	6,020	6,020	3,700	3,700	6,100
Package 3a				3,800					6,900	6,900	9,300
Control cabinet: Package 3	2,050	2,050	2,050						2,170		
Package 3a				2,160							2,120
Water filtration: Package 3		800				1,300					
Water purifier: Package 3										6,180	
Package 3a									6,180		8,260
Water storage tank: Package 3										5,200	
(wet) Package 3a									5,200		7,300
Transformer: Package 3a										35,200	
Skid and structure: Package 3	2,500	2,500	2,500	2,500	4,000	4,000	4,000	4,000	3,000	3,500	4,200
Package 3a				2,500					5,000	7,000	7,000
Total: Package 3	32,295	30,090	32,760	26,990	54,205	51,085	54,205	54,205	31,210	40,920	48,390
Package 3a				31,955					64,940	90,760	92,575

Table 3.7-VII.
Subsystem weights (pounds).

	System										
	1	2	3	4	5	6	7	8	9	10	11
Hydrogen production subsystem	31,785	30,580	48,800	48,820	54,350	51,310	54,350	54,350	74,240	75,835	90,750
Nitrogen generation subsystem*	11,163	11,163	11,349	17,611	17,891	17,961	17,891	17,891	38,575	38,575	64,897
Ammonia synthesis subsystem	27,745	25,340	28,210	51,785	50,205	45,785	50,205	50,205	74,600	74,600	112,015
Water purification and filtration subsystem**	3,760	4,560	3,800	6,660	6,760	8,080	6,760	6,760	11,380	11,380	15,560
Power conditioning subsystem	13,500	13,500	13,500	14,800	22,200	22,200	3,200	22,200	40,400	35,200	45,880
Control subsystem	2,050	2,050	2,050	2,160	2,160	2,160	2,160	2,160	2,170	2,170	2,190
Total skid and structure	8,305	8,210	10,670	13,150	12,440	12,300	11,080	13,490	22,670	21,515	28,880
Total	98,308	95,403	118,379	154,986	166,006	159,796	145,646	167,056	264,035	259,275	360,172

*Includes liquid N_2 stored.
**Includes H_2O stored.

PREDICTED PERFORMANCE

The integration and packaging data for the eleven systems is summarized in Table 3.7-VIII. From the data presented, it can be seen that the number of packages range from a minimum of three to a maximum of five. The number of package interconnections ranges between 17 and 40. The total system weights vary between 98,300 lb and 360,000 lb.

To ensure maximum production under field operating conditions, steps have been taken to allow the overall system to continue operation by bypassing defective equipment. In the case of the water purifier, sufficient feedwater storage capacity has been provided to ensure normal operation for 2 hr. This provision also makes hydrogen production subsystem feedwater available during plant start-up.

In the nitrogen generation subsystem, two air purifiers are used. Normally one purifier is operated while the other is rejuvenated by a refrigeration unit which is operated using an automatic switching cycle. If the refrigeration system or one of the purifiers fails, the remaining purifier can continue to operate. The length of time that such a failure can be tolerated depends on how soon the surviving purifier becomes depleted. The depletion rate depends on atmospheric conditions. The nitrogen generation subsystem also contains a 0.5-hr supply of liquid nitrogen in a storage tank. If the plant is shut down, nitrogen can be drawn from the tank while temporary repairs are made.

In systems 4 through 11, the ammonia synthesis subsystem uses more than one compressor. If one fails, the remaining compressors can continue to run and the subsystem will be able to operate at reduced capacity.

Table 3.7-VIII.
Integration and packaging data.

System	Power level (Mw)	Mobility Class	Cooling medium	A-C frequency	No. of packages required H_2	N_2	NH_3	Total No. of packages	Location of transformer	Location of water purifier	Location of control cabinet	Package interconnections	Total system weight (lb × 10⁻³)
1	3	I	air	60	1	1	1	3	N_2	N_2	NH_3	17	98.3
2	3	I	water	60	1	1	1	3	N_2	N_2	NH_3	23	95.4
3	3	II	air	60	1	1	1	3	N_2	N_2	NH_3	17	118
4	6	I	air	60	2	1	2	5	H_2*	N_2	NH_3**	32	155
5	6	II	air	60	1	1	1	3	N_2	N_2	H_2	20	166
6	6	II	water	60	1	1	1	3	N_2	N_2	H_2	26	160
7	6	II	air	400	1	1	1	3	H_2	N_2	H_2	17	146
8	6	III	air	60	1	1	1	3	H_2	N_2	H_2	17	167
9	10	II	air	60	2	1	2	5	H_2*	NH_3**	NH_3†	32	264
10	10	III	air	60	1	1	2	4	NH_3**	NH_3†	H_2	33	259
11	15	III	air	60	1	1	2	††	††	NH_3**	NH_3**	40	360

*One transformer on each skid.
**Located ir Package 3a.
†Located in Package 3.
††Separate transformer skid.

The hydrogen production subsystem can be operated with 25% of the module banks shut down. When the hydrogen production subsystem is operated with the normal input power but with less than the normal number of banks, the current of each remaining bank increases and the efficiency decreases. The heat load increases simultaneously. Because of the added current and heat loads, the capability of operating with less than the normal number of modules requires the electrical distribution system and the heat exchanger to be oversized. The extra capacity of these components also allows the subsystem to operate at approximately a 15% overload condition. In the event that the feedwater pump fails, the liquid in the separator-reservoir could allow the subsystem to operate for a short period. In system 10, one of the two parallel-connected electrolyte pumps could continue to operate with the other pump shut down.

DISCUSSION

The final packaging data for the eleven systems was summarized in Table 3.7-VIII. From the table it can be seen that only two of the hydrogen production subsystems require more than one package, i.e., systems 4 and 9. In both cases the relatively high power level and low package weight limit (6 Mw, 35,000 lb and 10 Mw, 65,000 lb, respectively) require more than one package to satisfactorily accommodate the subsystem equipment.

The nitrogen generation subsystem is the lightest of the three major subsystems, and in each of the eleven designs, requires only one package.

Most of the ammonia synthesis subsystems require only one package. However, systems 4, 9, 10, and 11 each require two packages with the high pressure equipment located on one and the compressor and purification equipment on the other.

Seven of the eleven systems have three skids each. The largest number of packages, five, is required by systems 4, 9, and 11. Unlike systems 4 and 9, previously discussed in the Investigation subsection, the fifth package in system 11 is a separate transformer package. For system 11, the transformer weighs 49,600 lb and is 150 in. × 130 in. × 61 in. It was too large to be accommodated with existing equipment on any of the other packages; hence, a separate package was needed.

The locations of the secondary subsystems are given in Table 3.7-VIII. As indicated for transformer location, in nine of the systems the transformer is located in either the hydrogen or nitrogen package. It is preferable to place the transformer in the hydrogen package to eliminate several large interconnecting power cables. However, since the hydrogen package is frequently weight limited, it was necessary, in six instances, to place the transformer elsewhere. In the majority of these six cases, the transformer was place in the nitrogen package.

Since the hydrogen production subsystem is the only subsystem that consumes feedwater, it first appears advisable to package the water purification and hydrogen production subsystems together. However, when considering the problem of package interconnections, it can be seen that removal of the water purifier from the hydrogen production subsystem results in only one additional connection. For this reason it is better to place the transformer (with a minimum of three heavy power connections) on the hydrogen production subsystem provided the weight limit is not exceeded.

Packaging of the control cabinet posed no serious problems; because of its light weight and small size, it was readily accommodated.

The number of package interconnections given in Table 3.7-VIII indicates that the water-cooled systems, 2 and 6, require more interconnections than air-cooled systems because of the additional cooling water lines. As stated previously, packaging the transformer with the hydrogen production subsystem reduces the number of interconnections. This can be seen in systems 7 and 8 where the least number of interconnections are required. In systems 4 and 9, which also have the transformer or the hydrogen package, a direct comparison cannot be made since two transformers are used in each subsystem.

REFERENCE

Final Report Energy Depot Electrolysis Systems Study. Allison Division of General Motors, EDR 3714, 1 June 1964.

Control Analysis

DESCRIPTION

The control subsystem, described in subsection 3.6 of this report, coordinates and regulates the various subsystems of the fuel production system to establish and maintain a stable operating condition which maximizes the ammonia fuel production rate consistent with input power and design limitations.

An analog computer program was developed which simulates the dynamic characteristics of the fuel production system with the control subsystem influences. The block diagram for this computer program is shown in Figure 3.8-1. The program emphasizes the interrelationships between the control, ammonia synthesis, nitrogen generation and hydrogen production subsystems. The water purification subsystem is isolated dynamically from the other subsystems by a water storage tank. The response time of the power conditioning subsystem is relatively short, a few cycles, and therefore can be represented by step change inputs to the computer.

INVESTIGATION

The complexity of the fuel production system dictated that the dynamic analysis be done with the aid of a computer program. Both analog and digital computers can be used to perform dynamic analysis and several approaches were studied.

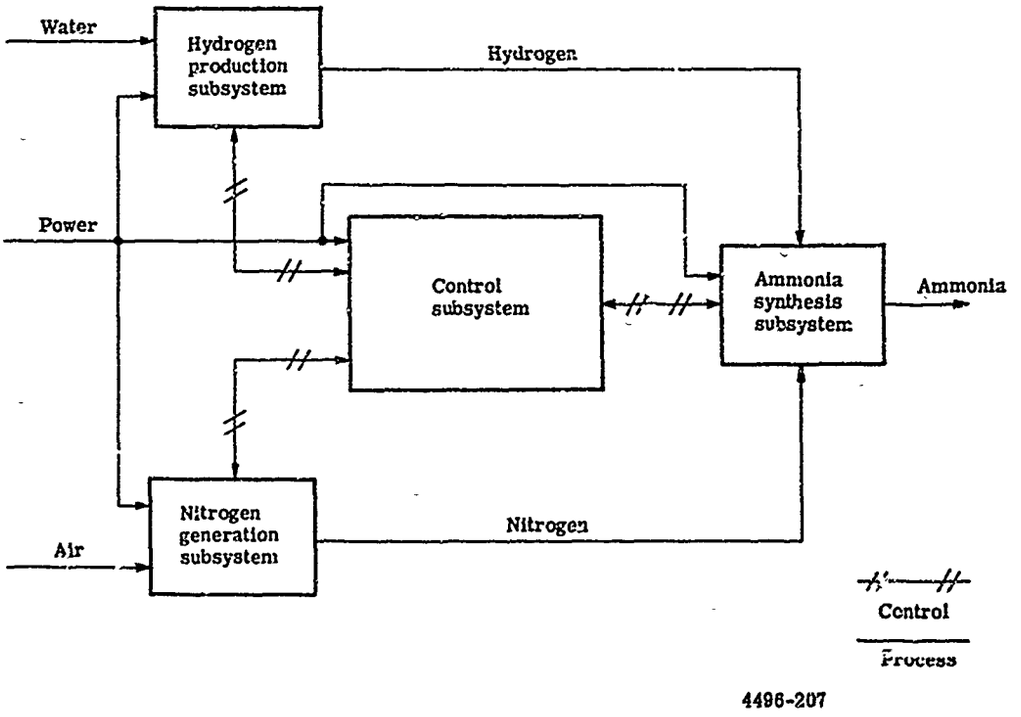


Figure 3.8-1. Fuel production system computer program block diagram.

- Digital—direct programming of the mathematical equations.
- Digital—program which converts the digital computer to analog terminology.
- Analog—wide range simulation of the dynamics equations.
- Analog—simulation of perturbation equations around the design point, with separate programs for start-up.

The last method was chosen for the following reasons.

- Excellent description of the system dynamics for control stability analysis is obtained with a minimum of program complexity.
- The steady-state operating point is established from the system performance calculations This minimizes the steady-state error in dynamic analysis.
- The mathematical models are scaled so that small differences, such as the pressure difference between the hydrogen subsystem liquid-gas separator-reservoirs, yield good resolution.
- Simplified start-up models provide sufficient accuracy to provide start-up times since a portion of the process and some of the controllers are not operative during system start.

The mathematical description for each component of each subsystem was developed in a manner to be consistent with the analog computer. Considerations were made in the selection of each mathematical model to be sure that the detail was sufficient to give an accurate representation of the dynamics of the system without causing undue program complexity. When the time constants of a particular component were small in comparison with the overall system response time, they were ignored or lumped with other time constants. Care was taken not to omit time constants which might be coupled together to give a system instability within a small section of the system. An example of the inclusion of small time constants is the pressure balance loop in the hydrogen production subsystem.

In the development of the mathematical models, a literature survey was made to find what techniques have been used for the solution of similar problems. A method of approximation of fluid inertia developed by Burgreen[1] was used in developing the inputs for hydrogen production subsystem simulation.

The method used to simulate the air fractionation columns on the nitrogen generation subsystem was developed by Williams.[2] Equations from this work were rewritten in perturbative form and changes were made to account for the lumping of several distillation column trays into a single simulated tray.

The representation of the system dynamic equations was in the form of a superposition of the plant dynamics upon the steady-state performance equations or empirical data. This approach to the writing of the dynamic computer program requires as inputs the steady-state operating conditions for the fuel manufacturing plant. In instances where information for off-design conditions was available from the subsystem analysis, these functions were incorporated as part of the analog program. Specifically, in the ammonia synthesis subsystem the operation of the ammonia converter was simulated using the synthesis rate as a function of temperature, space velocity (SV), pressure, ammonia input concentration, and hydrogen-nitrogen ratio within the system as presented in Figures 3.1-5 to 3.1-10.

In the case of the nitrogen generation subsystem, the concentrations of nitrogen in the liquid and gas phase in each fractionation column tray were those developed by the computer analysis presented in summary Table 3.2-III. For the hydrogen production subsystem the steady-state performance was that given in EDR 3714.[3]

When similar components were used throughout the fuel production system, the basic mathematical models for each component were derived by writing the dynamic equations relating temperature, flow, pressure, volume, concentrations, and empirical or calculated performance data. Examples of these components are heat exchangers and process fluid collection tanks (liquid-gas separator-reservoirs in the hydrogen production and ammonia synthesis subsystems).

The equations which represent the dynamic operating characteristics of ammonia synthesis converters, air fractionation columns, and heat exchangers are presented in the following agraphs.

Ammonia Synthesis Converters

The amount of ammonia produced in the converter depends primarily on the following input conditions:

- Synthesis temperature
- Synthesis pressure
- Space velocity
- Ammonia concentration at the inputs to the converter
- Hydrogen-nitrogen ratio
- Argon concentration

The variation of production around the system design point is given in the ammonia subsystem section of this report—Figures 3.1-5 to 3.1-10. The variation of ammonia production with each of the above inputs is linear except for the temperature and hydrogen-nitrogen ratio, as shown in the following equation.

$$\%\,NH_3 = F(T_{conv}) \left[\frac{\partial \%\,NH_3}{\partial \%\,A} \Delta \%\,A + \frac{\partial \%\,NH_3}{\partial \%\,NH_{3\,in}} \Delta \%\,NH_{3\,in} + \frac{\partial \%\,NH_3}{\partial P_{conv}} \Delta P_{conv} + \frac{\partial \%\,NH_3}{\partial SV} \Delta SV + F\left(\frac{H_2}{N_2}\right) \right] + (\%\,NH_3)_o \quad (1)$$

where:

$\%\,NH_3$ = concentration of ammonia in exit stream of second converter

$(\%\,NH_3)_o$ = steady-state ammonia concentration at second converter exit = 19.68% NH_3

$\%\,NH_{3\,in}$ = concentration of ammonia in inlet stream of first converter

T_{conv} = weighted average temperature in both ammonia converters

$\%\,A$ = concentration of inerts (principally argon) in input stream of ammonia converter

P_{conv} = synthesis gas pressure in converters

SV = space velocity of synthesis gas through converters

H_2/N_2 = hydrogen-nitrogen molar ratio in the input stream of the first converter

When the change in the ammonia concentration at the exit of the converter is linear, a partial derivative notation is used to represent the change. Function generators were used for variation in ammonia production rate with converter temperature and hydrogen-nitrogen ratio.

Air Fractionation Columns

The dynamics of the air fractionation column are represented by the concentration of the more volatile product, nitrogen, both in the liquid on each tray and the gas above each tray. The mathematical representation is:

$$H_N \frac{\partial X_N}{\partial t} = L_{N+1} X_{N+1} - L_N X_N + V_{N-1} Y_{N-1} - V_N Y_N \quad (2)$$

$$Y_N = a^* X_N \quad (3)$$

$$\frac{\partial L_N}{\partial t} = \frac{L_{N+1} - L_N}{T_{HL}} \quad (4)$$

where:

- H = the amount of liquid contained in a distillation tray
- X = concentration of volatile product in the liquid
- Y = concentration of volatile product in the gas
- L = liquid flow rate down the column
- V = gas flow rate up the column
- T_{HL} = hydraulic lag time constant associated with the increase in level on a distillation tray with increased flow rate
- α^* = the linearized relation between liquid concentration and gas concentration of nitrogen

Subscript N refers to the distillation tray represented by the equation

This equation is a dynamic material balance for each tray in the air fractionation column. A complete development of the equations is given by Williams.[2] Variations of these equations were reduced for the column reboilers and condensers and feed trays.

Heat Exchangers

The most common component in the fuel production system is the counter-flow heat exchanger.

The fundamental equations used to represent the heat exchangers when no condensation or vaporization takes place are:

$$Q_n = U_n A_n (T_{n\,avg} - T_S) = \dot{\omega}_n C_{p_n} (T_{n\,in} - T_{n\,out}) \tag{5}$$

$$T_{n\,avg} = \frac{T_{n\,in} + B\,T_{n\,out}}{1 + B} \tag{6}$$

$$T_S = \frac{1}{\omega C_p} \int \left(\sum_{n=1}^{K} Q_n \right) \partial t \tag{7}$$

Where one of the fluids is vaporized or condensed, the equations are

$$Q_n = U_n A_n (T_{n\,avg} - T_S) = \dot{\omega}_n C_{p_n} (T_{n\,in} - T_{n\,out}) + q\,\dot{\omega}_n\,\Delta H_n \tag{8}$$

for those streams where condensation takes place. The amount condensed is given by:

$$q = K (T_{n\,out} - T_{n\,cond}) \tag{9}$$

The weighting factor B is adjusted to provide the proper average temperature if it is required.

Where:

- Q = heat transferred to one fluid in the heat exchanger
- UA = overall heat transfer coefficient-area product for the heat exchanger
- T = temperature of fluid designated by subscript
- $\dot{\omega}$ = flow rate of fluid designated by subscript
- C_p = specific heat of fluid designated by subscript
- B = weighting factor for average fluid temperature in heat exchanger
- ωC_p = thermal inertia of the heat exchanger
- ΔH = heat of vaporization of the fluid
- q = ratio of fluid condensed to total flow
- K = relation of condensation ratio and temperature in a mixed fluid
- S = heat exchanger structure

Subscript n refers to fluid stream in heat exchanger

DESIGN PARAMETERS

The computer runs for the fuel production system were made for the design point ambient condition of 77°F temperature and sea-level pressure. For the following reasons no significant difference in plant controllability is anticipated for off-design ambient conditions.

- The operating temperatures and pressures in the parts of the system germain to fuel manufacturing are controlled to the same levels.
- The rejection of heat to the air does vary with ambient temperature; however, these temperature controls are involved in systems which have large thermal inertias and may be set slow enough to show good controllability at the most critical temperature.

DESIGN

The diagram shown in Figure 3.8-2 was used as a basis for the computer program for the dynamic analysis of the fuel production system. As this figure indicates, detailed analog programs were built for the nitrogen generation, hydrogen production, and ammonia synthesis subsystems. The control subsystem, which coordinates all the other subsystems and regulates the internal operation of each subsystem, was simulated as required with these subsystems. The coupling between the nitrogen generation subsystem, hydrogen production subsystem, and the ammonia synthesis subsystem was accomplished by generating lumped models of the hydrogen and nitrogen subsystems and coupling these models at the ammonia synthesis subsystem simulation.

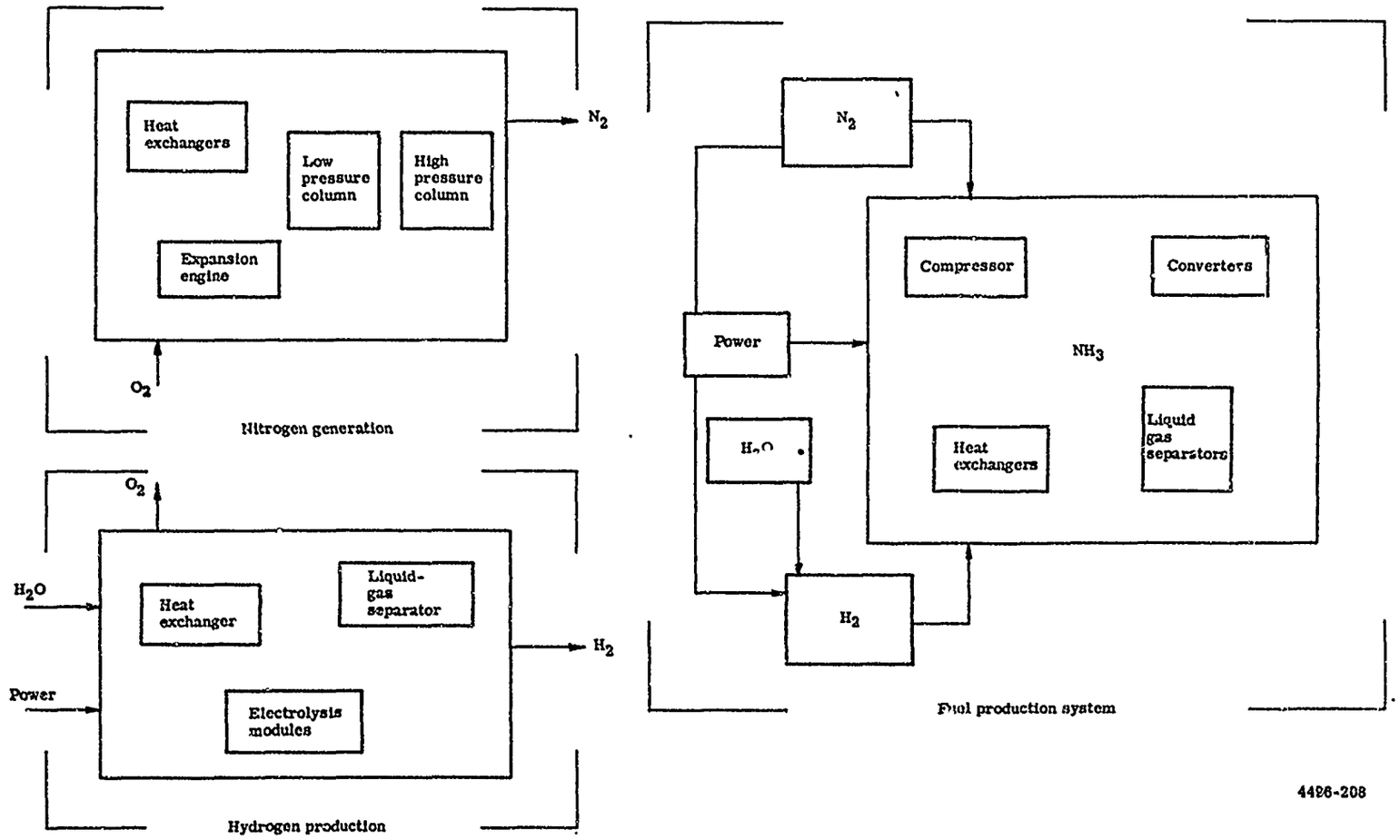


Figure 3.8-2. Basis for computer analysis of the ammonia fuel production system.

Allison

The power conditioning subsystem was not programmed because the voltage regulation method recommended has been used for other applications and is known to be stable. The transients in the power conditioning subsystem are so short that this subsystem may be represented by a pure step or ramp input to the hydrogen production subsystem and other subsystems. This approach to overall dynamic performance and stability analysis was used in conjunction with the mathematical models to form analog computer programs. A detailed description of these programs is presented in Appendix B.

PREDICTED PERFORMANCE

Four different fuel production systems, one for each of the different input power levels, 3 Mw_e, 6 Mw_e, 10 Mw_e and 15 Mw_e, were programmed on the analog computer. The differences between systems with the same power level were not significant in the dynamic representation, thus the four models were sufficient to represent all eleven fuel production systems. This program was used to show that the control subsystem would regulate and stabilize each case both for steady-state operation and for change in operating point. Control settings were selected for each case to produce rapid response and good stability.

The system response to change in system operating condition is described later in this section. The response of the hydrogen and nitrogen subsystems to several perturbations is presented in Appendix B.

Analysis of plant start-up was made on the analog computer to provide the information needed to determine the overall response time for each of the eleven systems. In addition to these runs, computer analysis of start-up and shutdown was made to determine the liquid nitrogen storage required to allow ammonia synthesis until the nitrogen generation subsystem is cooled to operating temperature. Settling times for the hydrogen production subsystems were calculated to ensure that their times and production rates are consistent with other operations during start-up.

The sequence of events and the times for the start-up operations are tabulated in the Performance Analysis subsection (3.9) of this report. The time between the start of electrolysis and the completion of start-up is 30 min for system one. The time calculated to come to 100% output is 32 min. The 100% output, therefore, occurs at approximately the same time the start-up sequence is completed. System 3, however, takes 48 min to reach full hydrogen production, but reaches 95% of full production by the time the start-up sequence is complete. The stabilization time of the hydrogen production subsystem can be decreased by placing electric heating elements in the electrolyte flow stream. Since the initial output is approximately 60% of full output when the electrolyzer is first energized at an ambient temperature of 70°F, no requirement for this time reduction is envisioned. However, electric heating elements may be desirable at ambient temperatures substantially below 70°F. Hydrogen production subsystems

have shorter settling times for the 6, 10 and 15 Mw systems than for the 3 Mw systems. In these systems the hydrogen production subsystem is at full output before the start-up sequence is completed.

During start-up of the nitrogen generation subsystem, liquid nitrogen from storage flows through the cold box components to evaporate and cool them, the gaseous nitrogen is then compressed for mixing with hydrogen or vented. The time required for the nitrogen generation subsystem to begin generation is dependent on configuration and the temperature to which the cold box components have warmed during shutdown. In the anticipated operational mode system 1 will start generation within 30 min and system 11 in less than one hour. Nitrogen storage is designed to supply nitrogen to the ammonia synthesis subsystem until generation begins.

Since both nitrogen and hydrogen are immediately available in substantial quantities, ≥ 60% of design production rate, the time required for the ammonia synthesis loop to reach synthesis temperature and pressure will determine time required to begin ammonia fuel production.

For each of the eleven systems, the time required for the catalyst beds to achieve synthesis temperature and pressure was determined using an analog program. Results of this determination (expressed in minutes) follow:

Systems 1, 2, 3	7.7
System 4	10.0
Systems 5, 6, 7, 8	9.8
Systems 9, 10	13.5
System 11	17.1

The dynamic response of the system during warm-up is presented in Appendix B.

After the synthesis process begins, the synthesis gas flow is increased as rapidly as allowed by temperature and pressure limitations in the ammonia converters. Design synthesis gas flow can be achieved within 5-10 min of the beginning of synthesis.

Response of the fuel production system to a change in operating conditions is illustrated in Figures 3.8-3 through 3.8-10. System response to a step decrease in hydrogen production rate from 100% to 75% is shown in Figures 3.8-3 through 3.8-6. These figures indicate the change in significant parameters during the transient and the level when a new steady-state operating point is reached. Since the hydrogen production rate is controlled to maintain the electrical power input at a preset level, the effect of an equivalent step decrease in electrical power available to the system is identical.

The nitrogen flow ra to the ammonia synthesis subsystem is controlled to be in the proper ratio to hydrogen flow rate. Following the 25% decrease in hydrogen flow rate, the nitrogen flow rate also drops to 75% within one minute.

The ammonia synthesis subsystem is self-regulating in a stable manner. With the decreased inflow of nitrogen and hydrogen synthesis gases, the ammonia synthesis loop pressure begins to decrease from 6000 psia to a new operating point of 4240 psia; the gas recycle flow rate decreases; the ammonia synthesis rate decreases as the flow rate through the converter decreases; the average converter temperature increases slightly as the heat generation rate and heat rejection rate stabilize at the new operating point.

The small fluctuations in converter temperature are caused by the variations in system pressure, flow rates, and space velocity which occur during the transient.

The slight, damped oscillations in ammonia delivery are caused by the level controller on the product liquid-gas separator tank.

All parameters reach new steady-state operating conditions within thirty minutes after initiation of the transient. Figure 3.8-3 illustrates the variation in these parameters for systems 1, 2, and 3.

In the larger systems, the effects of a decrease in hydrogen production rate is similar to the effect on systems 1, 2 and 3. Figures 3.8-4, 3.8-5, and 3.8-6 indicate the variation in system parameters with an instantaneous decrease of hydrogen production rate from 100% to 75% for the larger systems.

Response of the fuel production system to an instantaneous increase in hydrogen production rate from 75% to 100% of rating is shown in Figures 3.8-7 through 3.8-10. The transition from steady-state operation at 75% to 100% hydrogen flow rate is smooth. Operation at full rate output is stable and all parameters are well controlled.

DISCUSSION

The control analysis program for the fuel production system provided data which indicate that no significant change in the control subsystem is required. However, several areas of control sensitivity were found.

In the ammonia synthesis subsystem, degenerative system shutdown can occur if a decrease in synthesis rate decreases the average synthesis temperature significantly. The reduction in temperature of synthesis gas leaving the second converter reduces the heat transferred in the

0
10 min
30 min
Nitrogen flow rate to ammonia subsystem (percent)
100
80
60
40
20
0
(a)
Flow rate from ammonia subsystem recycle compressor (percent)
0
20
40
60
80
100
(b)
Change in average converter temperature (°F)
+160
+80
T_o = 845°0
-80
-160
(c)
0
500
1000
1500
2000
Transient initiation
Time—seconds

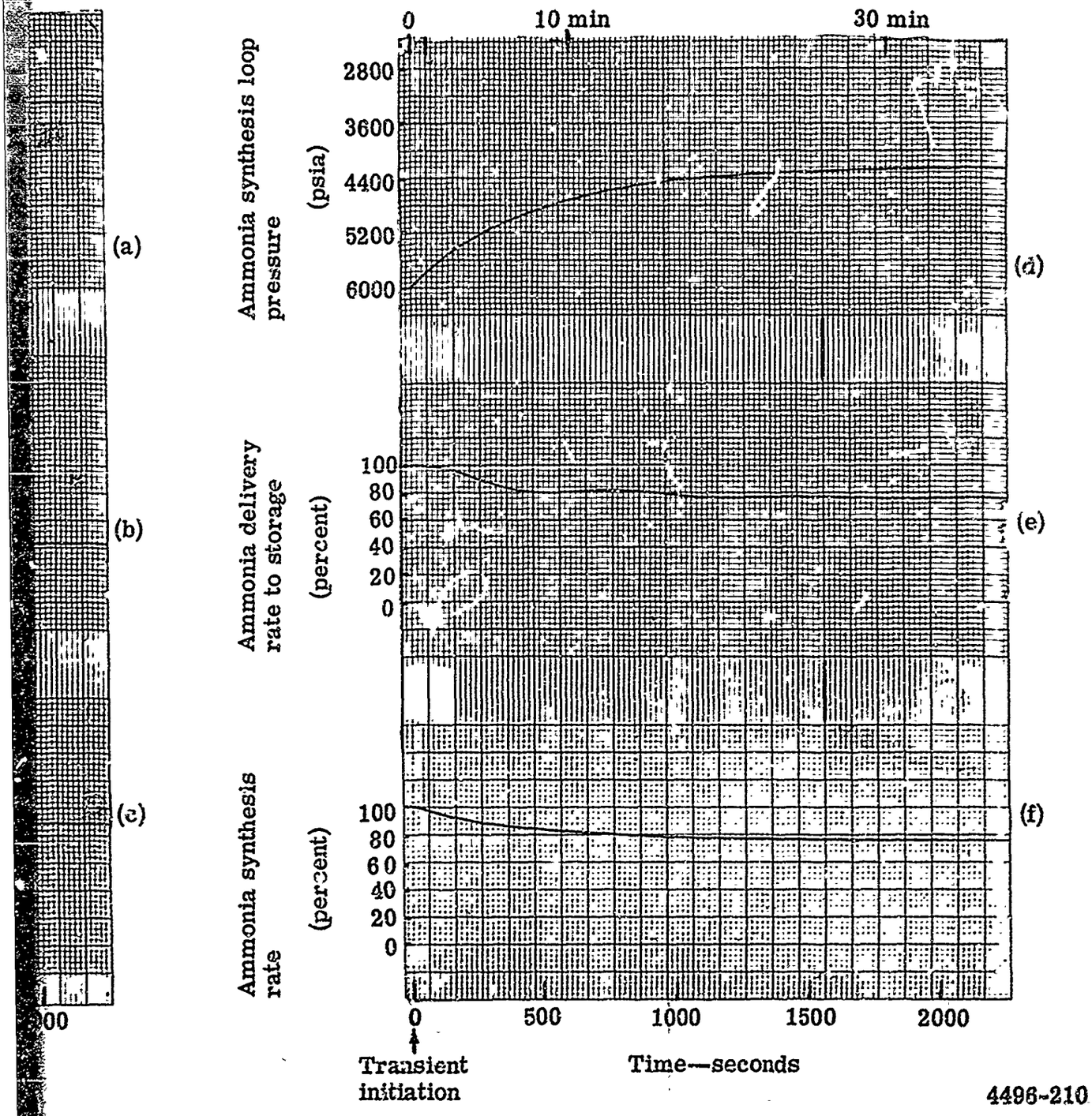


Figure 3.8-3. Fuel production system response to an instantaneous decrease in hydrogen production rate from 100% to 75% of design output (systems 1, 2, and 3).

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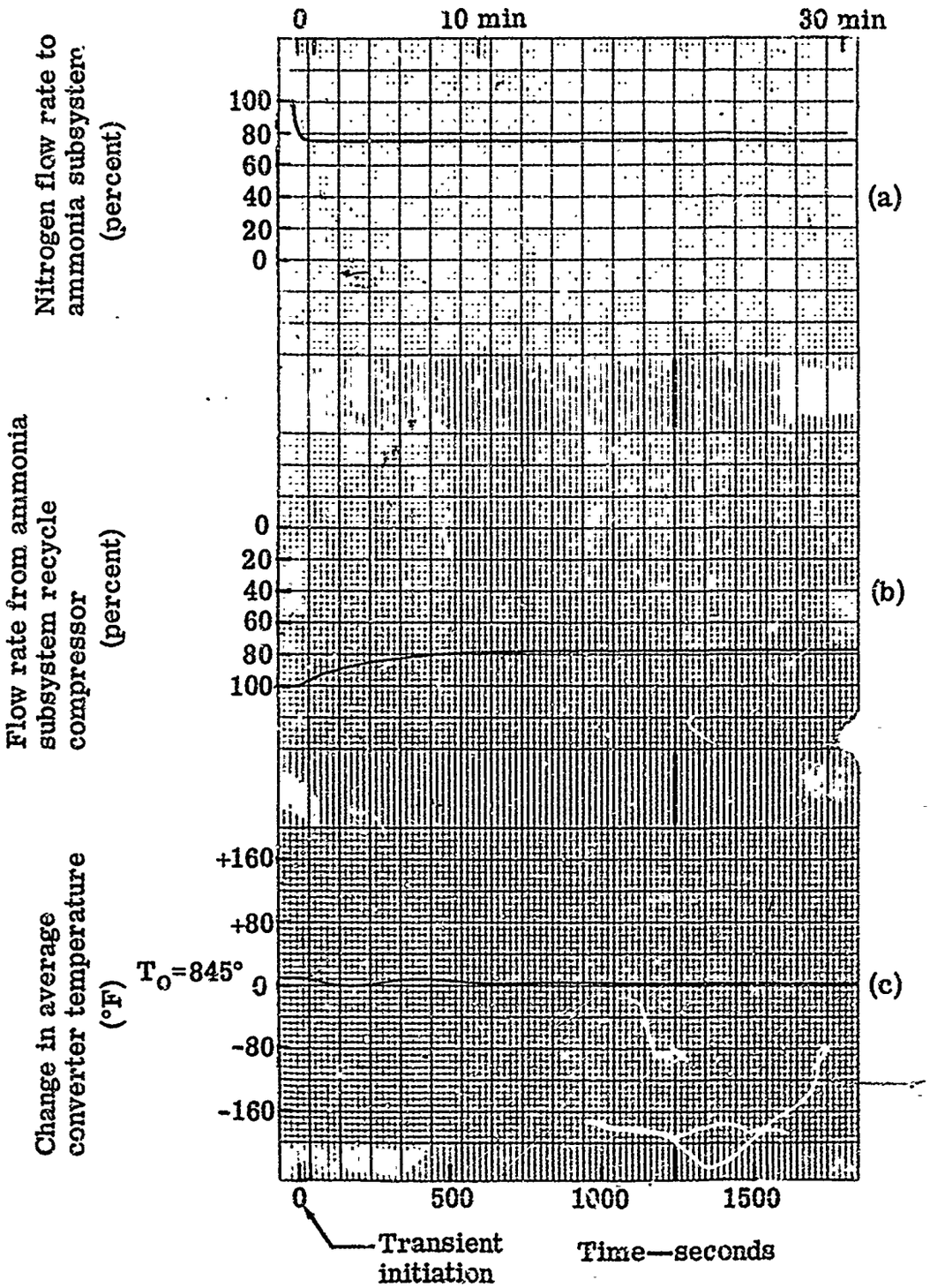


Figure 3.6-
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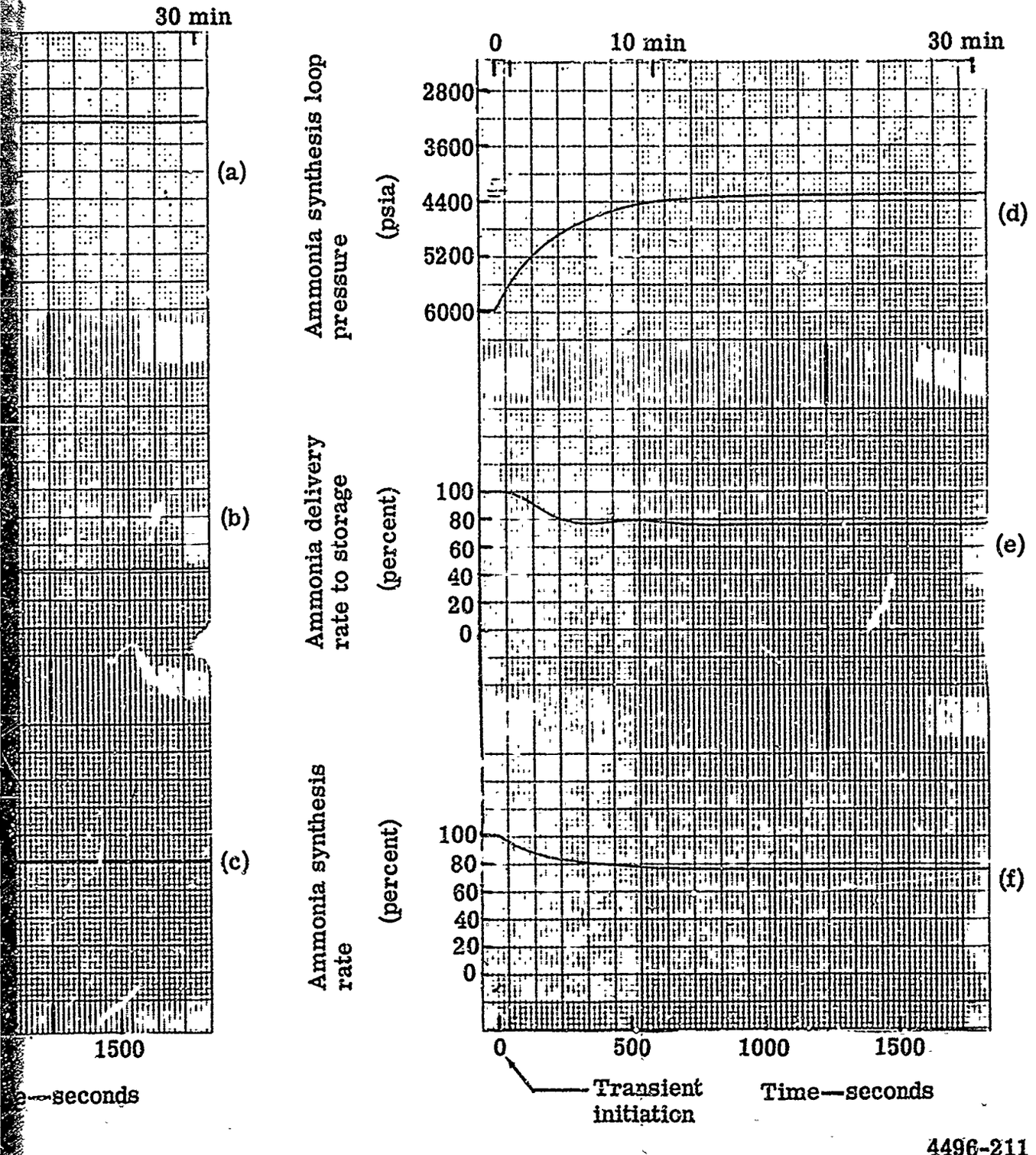


Figure 3.8-4. Fuel production system response to an instantaneous decrease in hydrogen production rate from 100% to 75% of design output (systems 4, 5, 6, 7, and 8).

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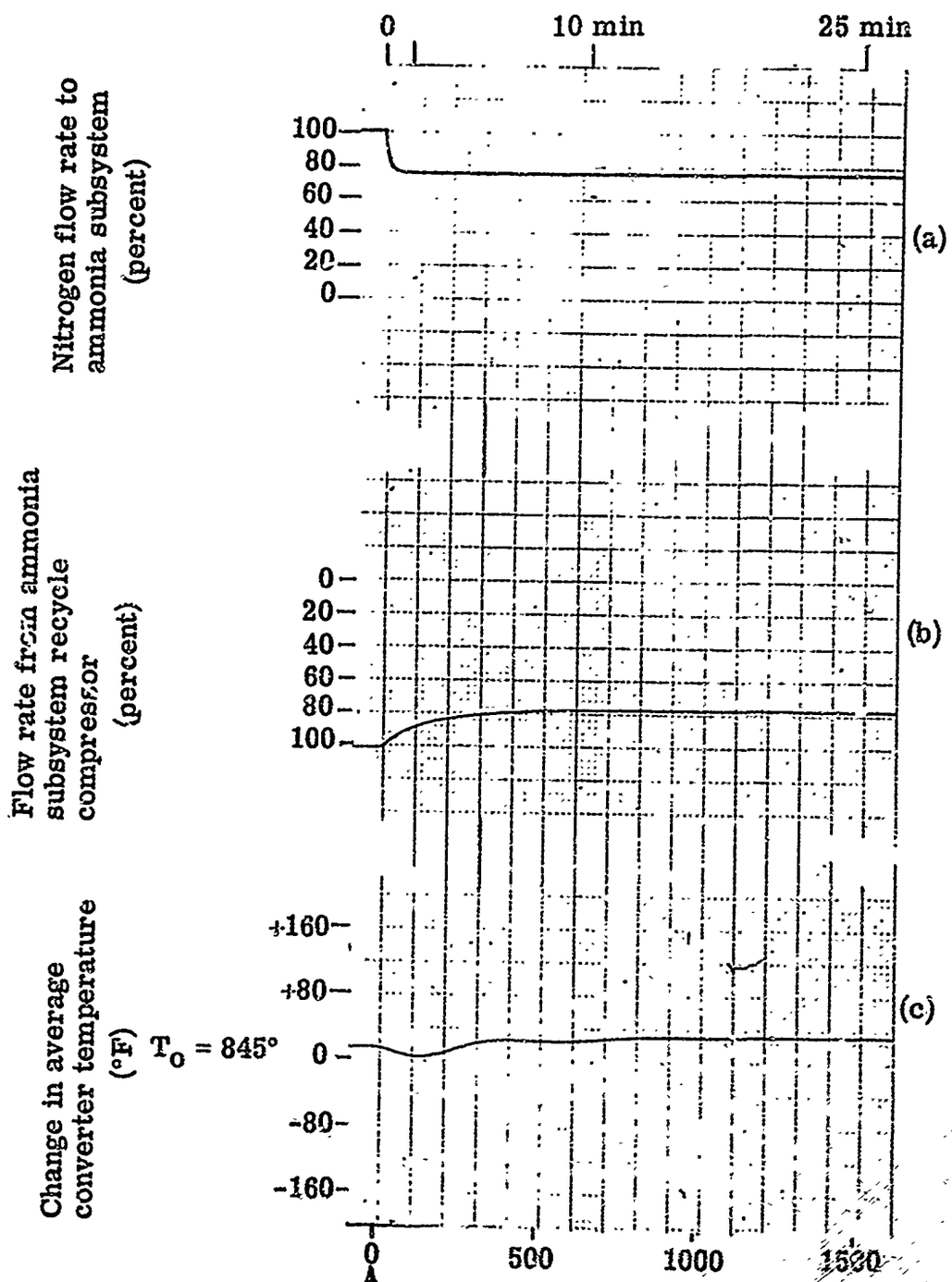


Figure 3. 8-5
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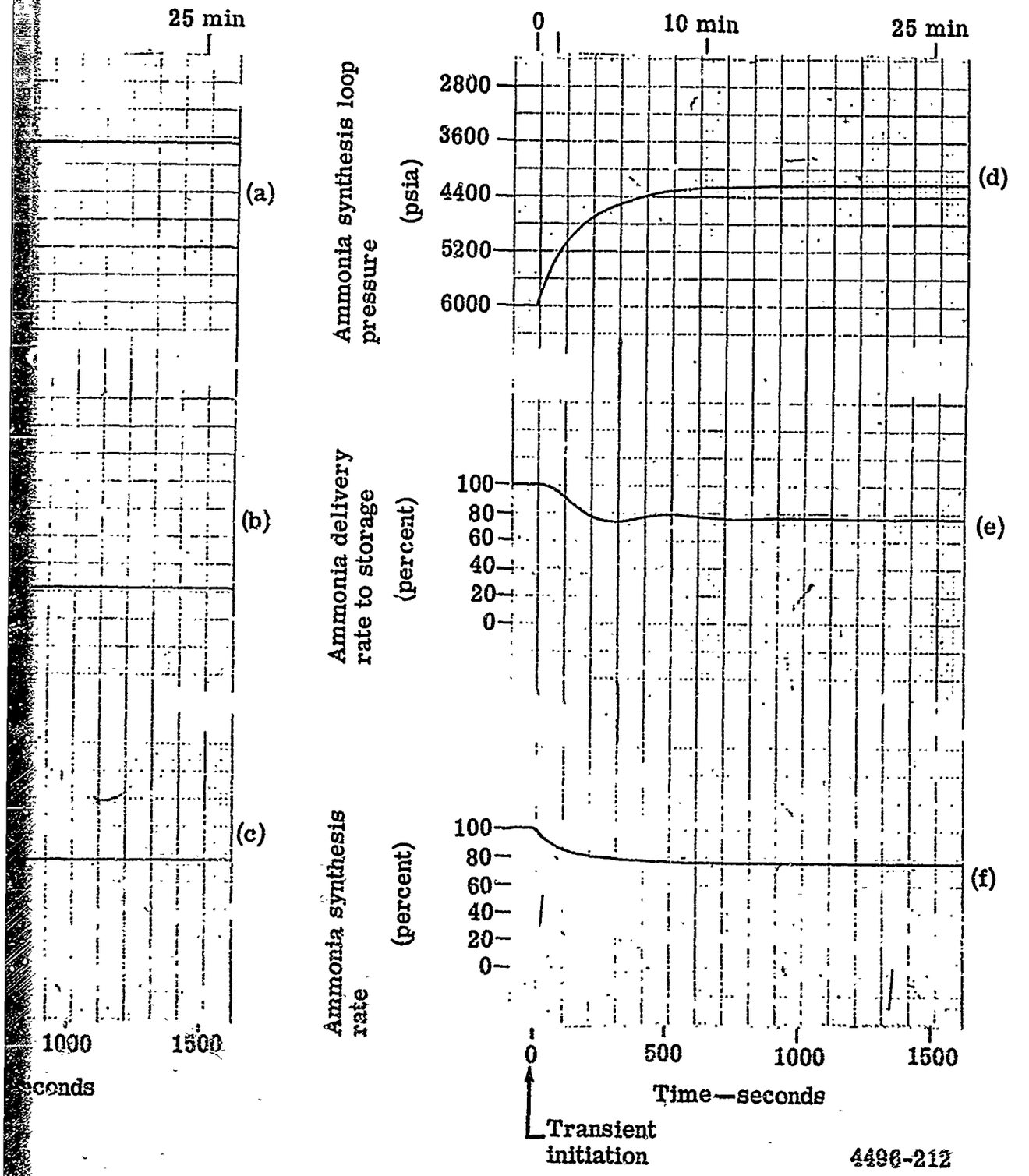
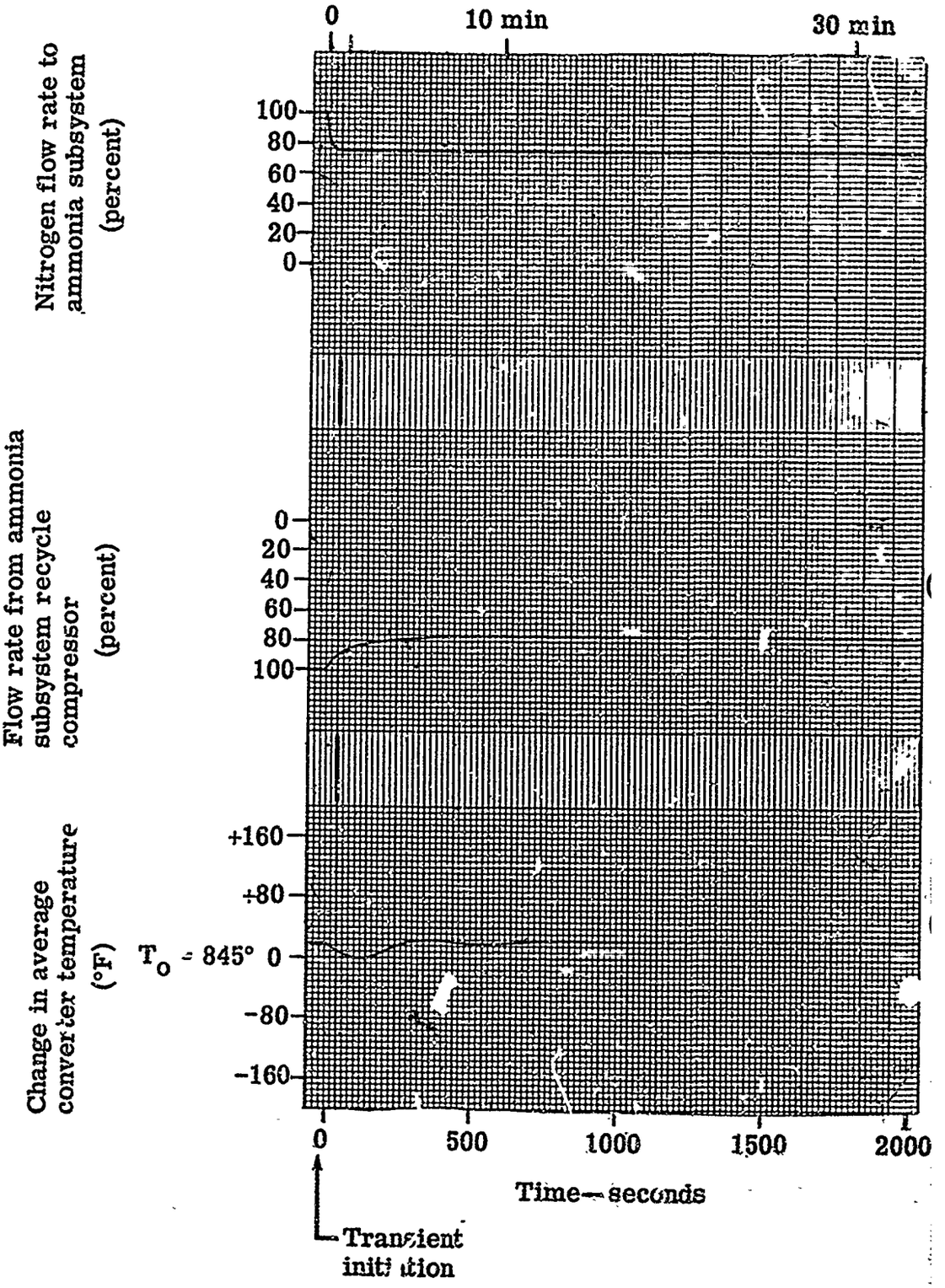


Figure 3.8-5. Fuel production system response to an instantaneous decrease in hydrogen production rate from 100% to 75% of design output (systems 9 and 10).

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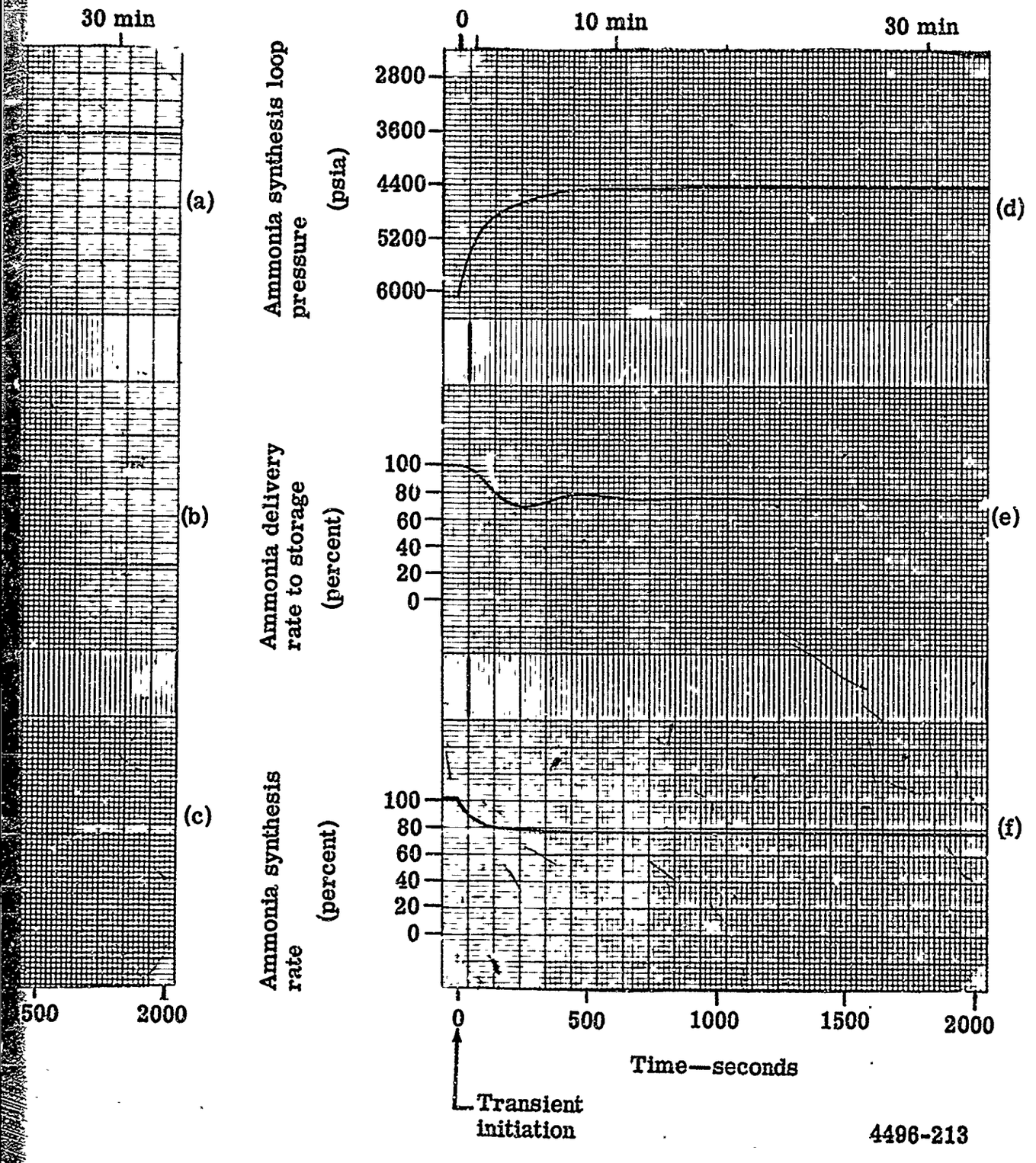
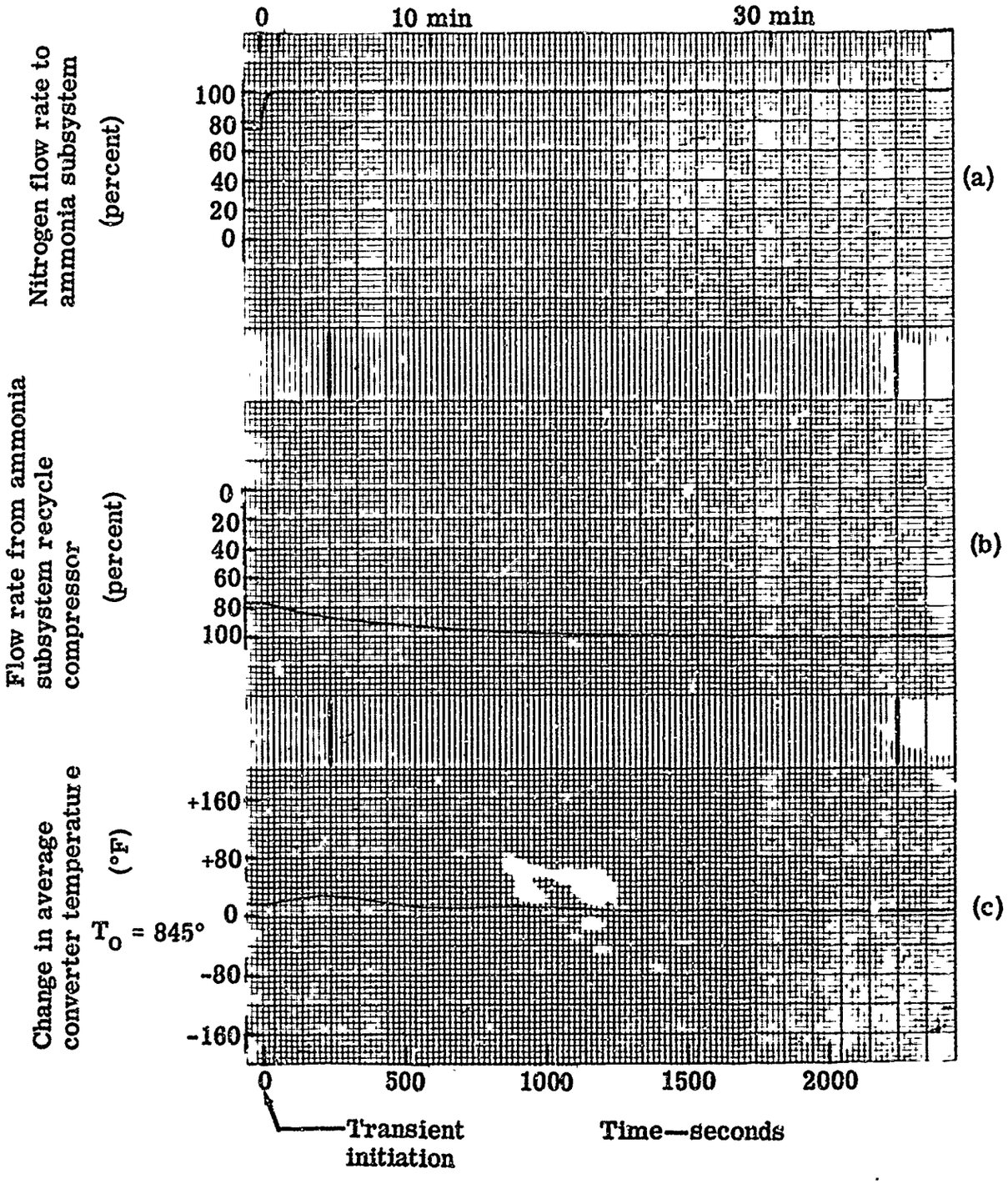


Figure 3.8-6. Fuel production system response to an instantaneous decrease in hydrogen production rate from 100% to 75% of design output (system 11).

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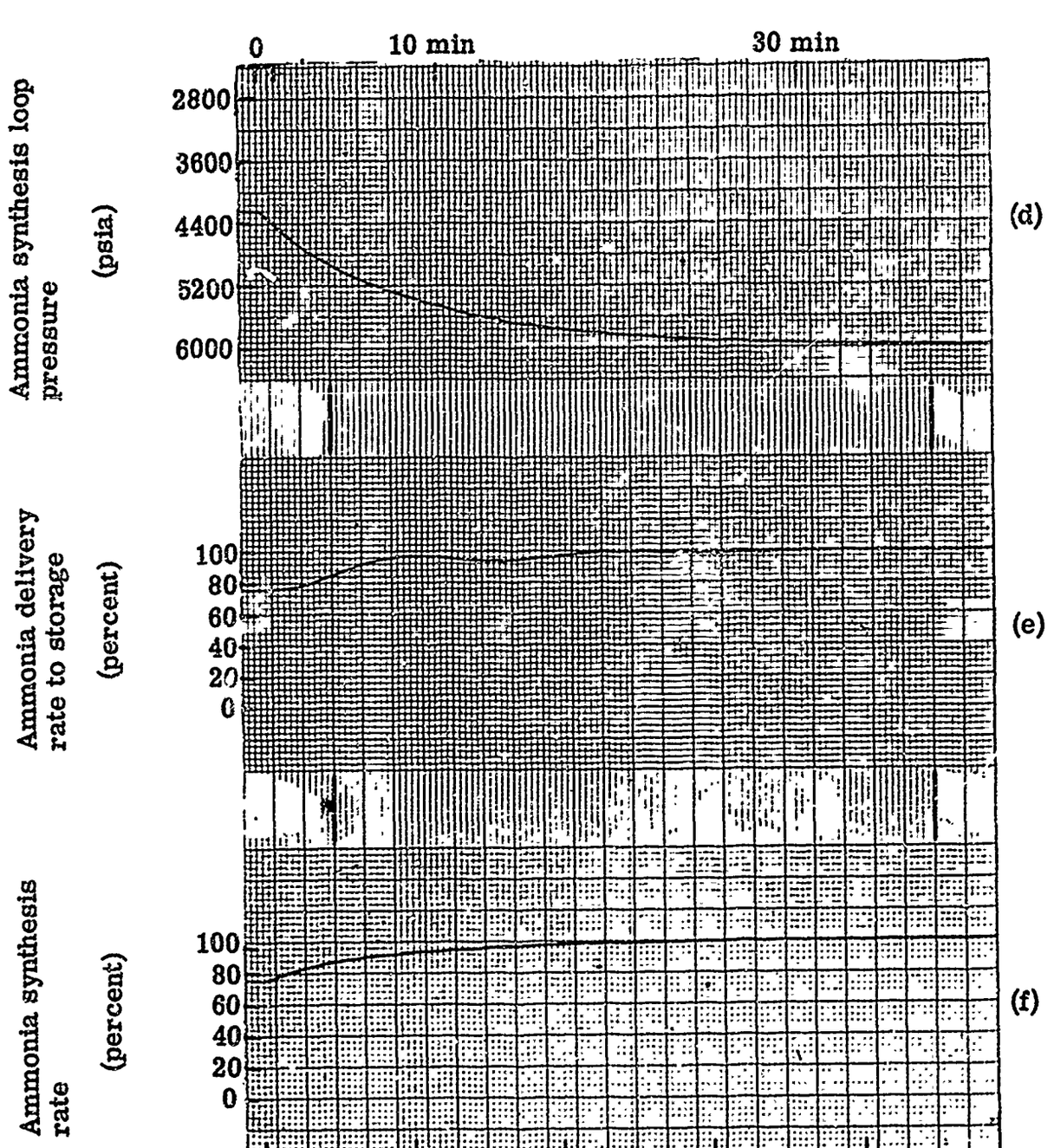


Figure 3.8-7. Fuel production system response to an instantaneous increase in hydrogen production rate from 75% to 100% of design output (systems 1, 2, and 3).

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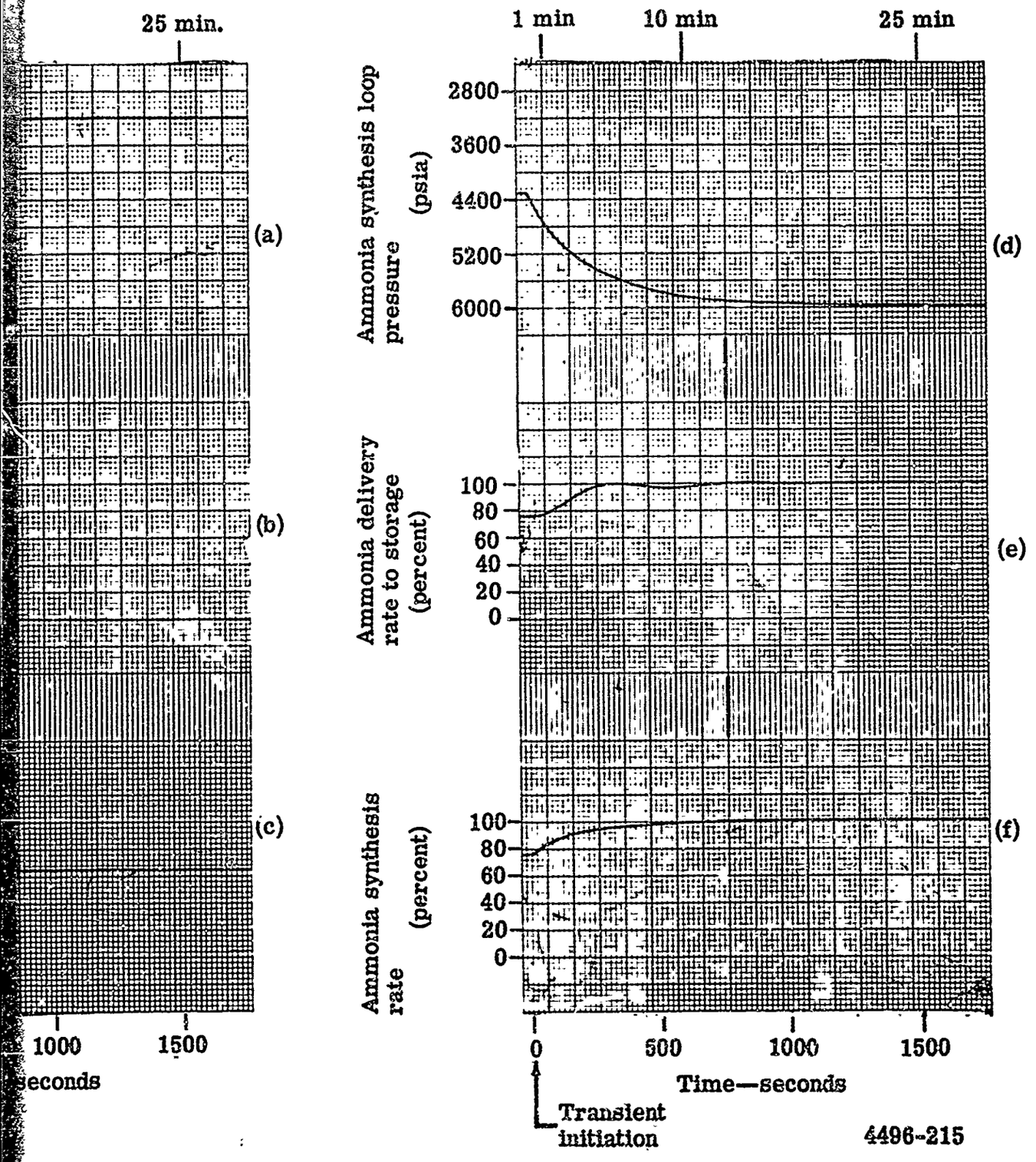
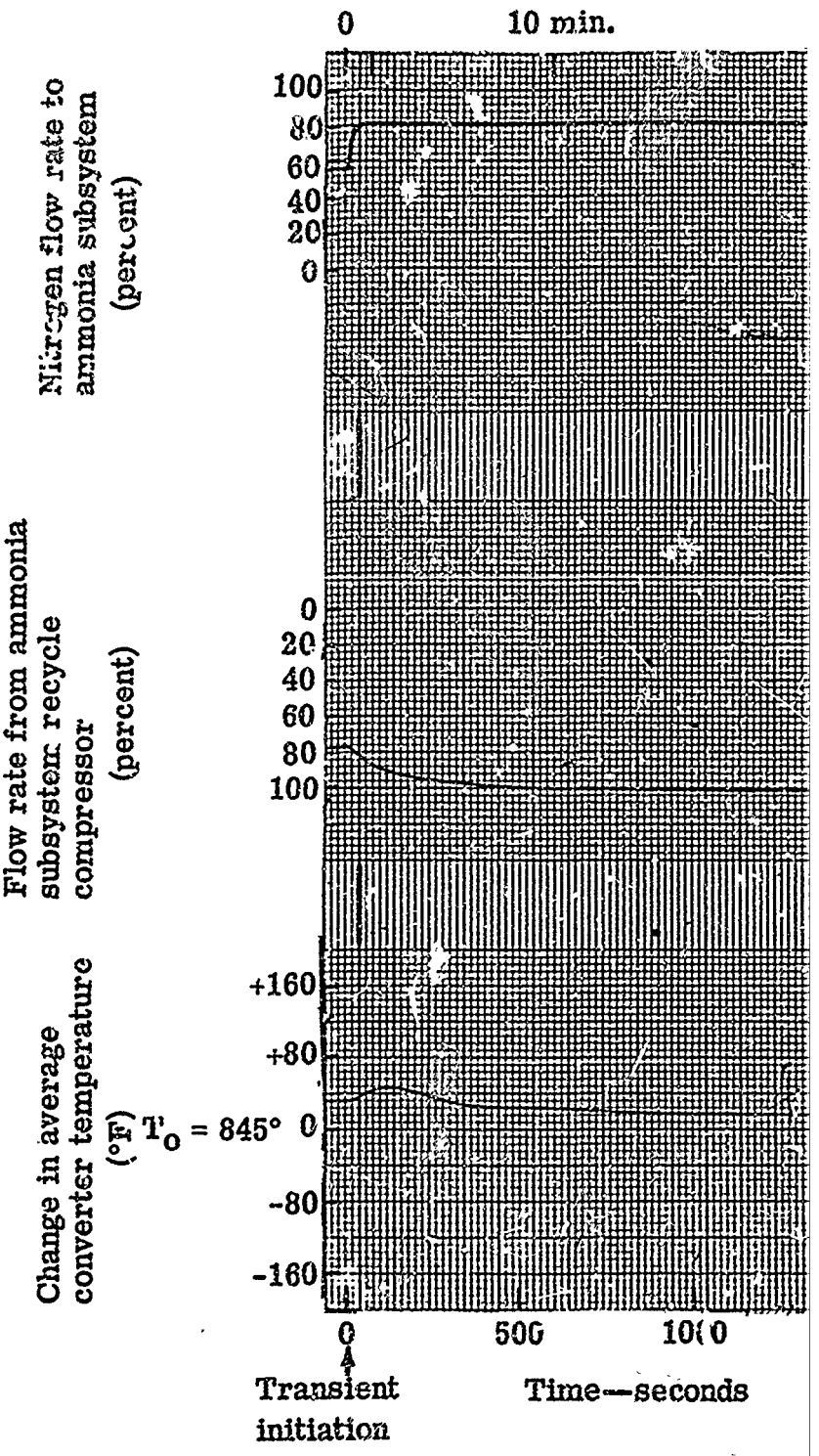


Figure 3.8-8. Fuel production system response to an instantaneous increase in hydrogen production rate from 75% to 100% of design output (systems 4, 5, 6, 7, and 8).

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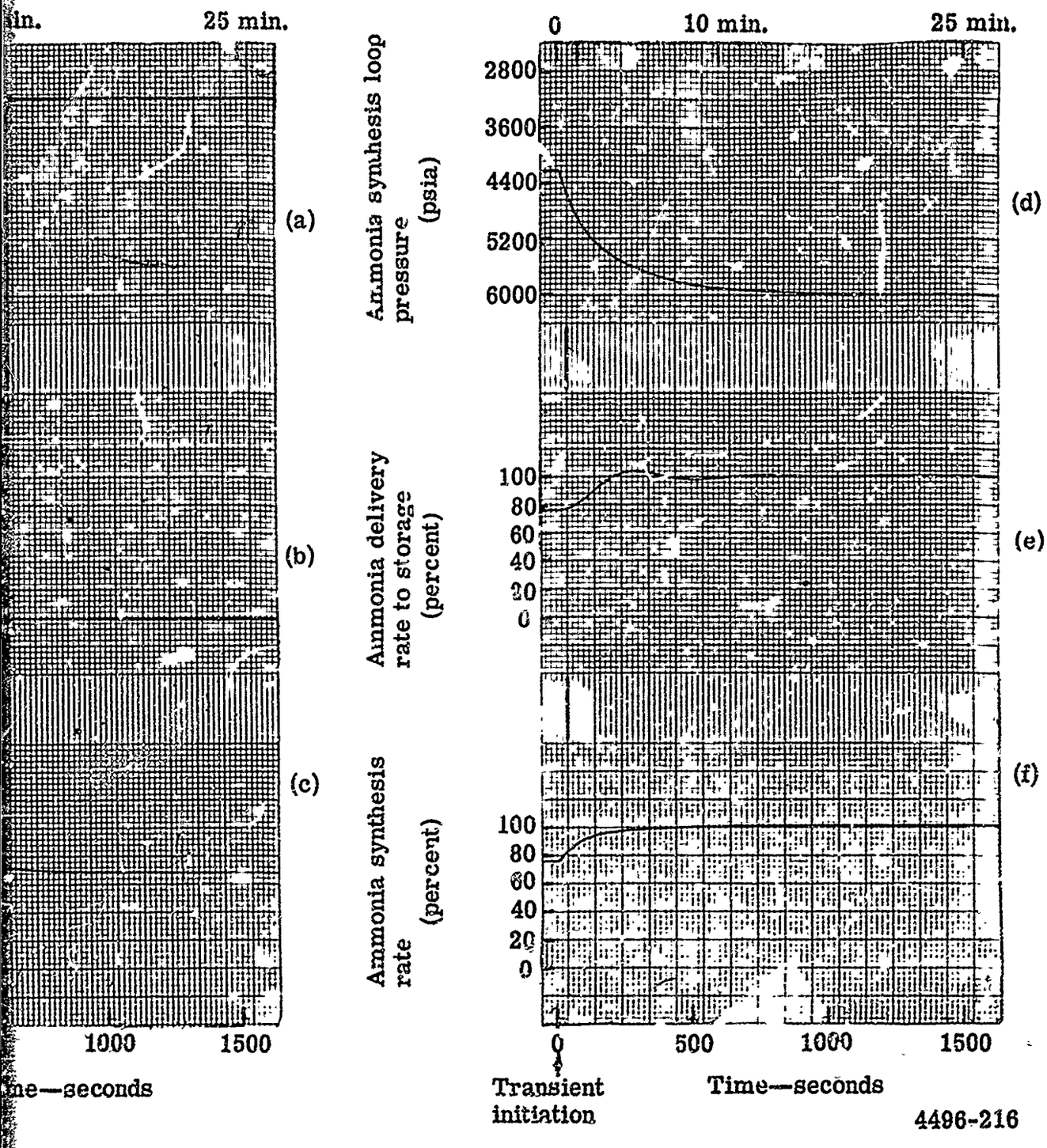
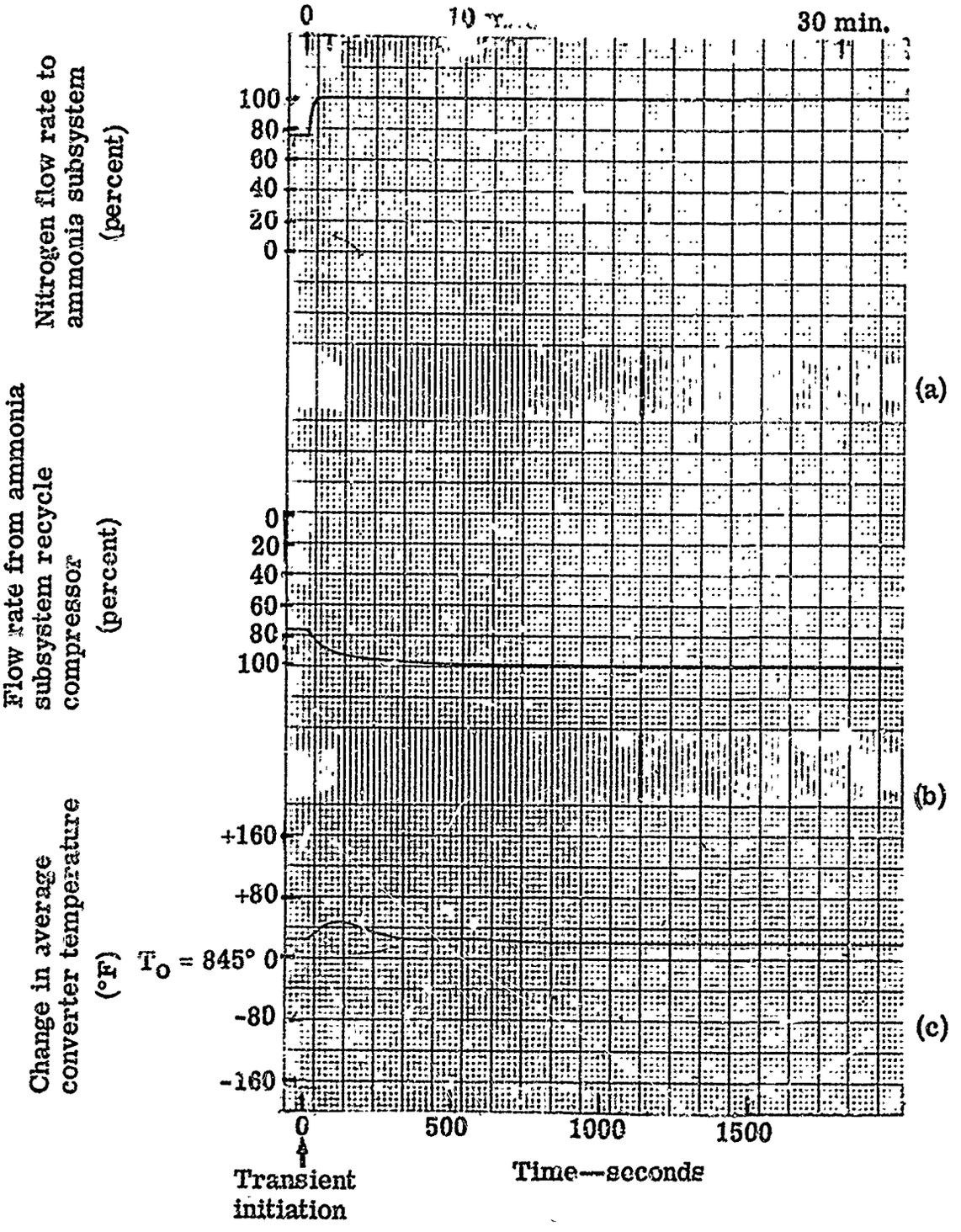


Figure 3.8-9. Fuel production system response to an instantaneous increase in hydrogen production rate from 75% to 100% of design output (systems 9 and 10).

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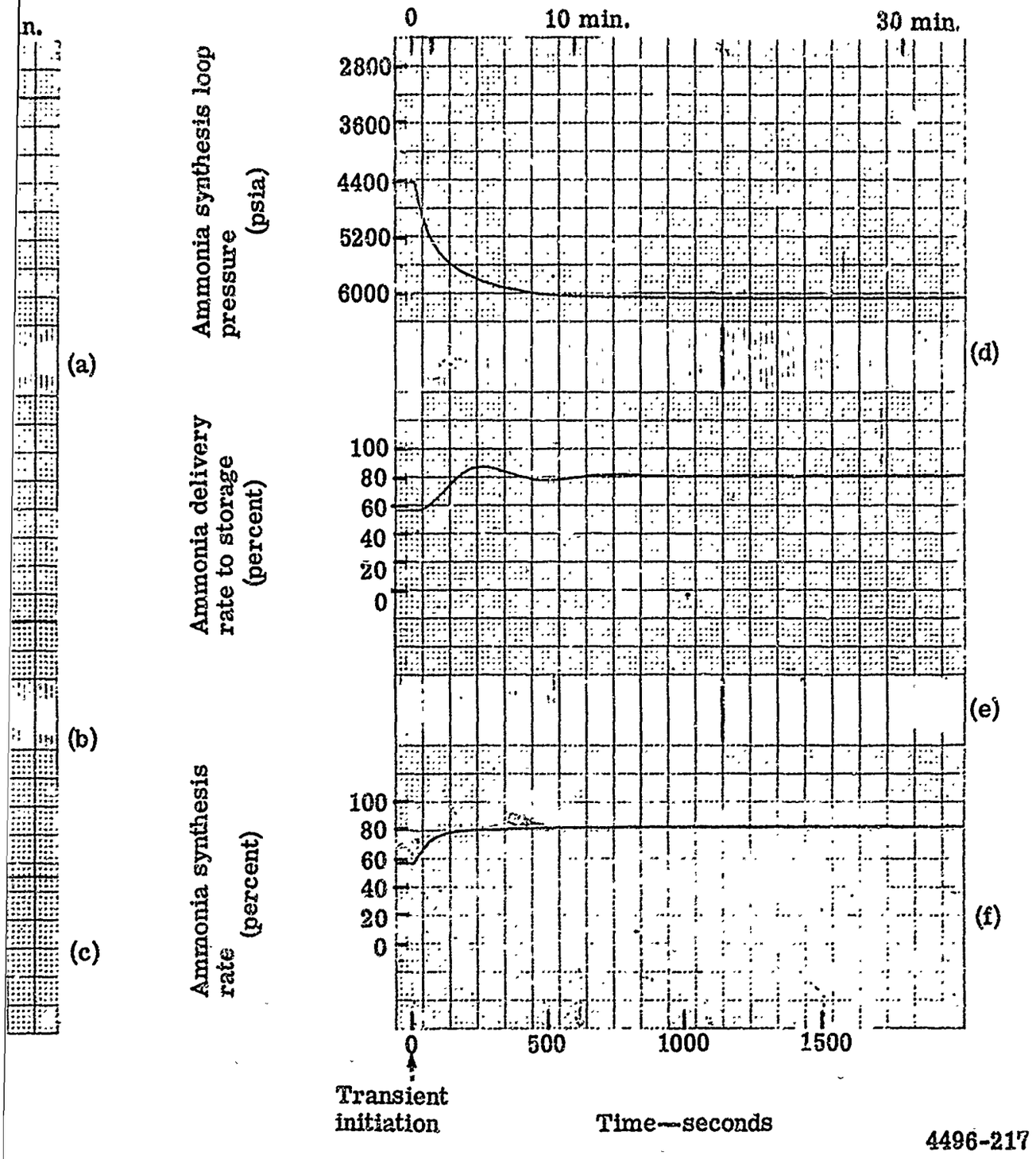


Figure 3.8-10. Fuel production system response to an instantaneous increase in hydrogen production rate from 75% to 100% of design output (system 11).

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converter heat exchanger to the synthesis gas entering the first converter (see Figure 3.1-14). The reduction in temperature of the incoming synthesis gas reduces ammonia synthesis rate; this reduces heat generation rate, thus further reducing the average synthesis temperature and shutting the system down. At design point operating temperature and pressure, the variation in synthesis loop pressure, if the synthesis rate is perturbed, serves to restore operating conditions or establish a new stable operating point. However, if the average synthesis temperature is below approximately 700°F, such as during start-up, manual control of loop pressure and temperature as well as supply of supplementary heat is required.

In the hydrogen production subsystem, unstable operation can occur if the control gains for hydrogen subsystem pressure controller (A-3, lower center section of Figure 3.6-2) and differential level controller (H-3, upper right hand corner of Figure 3.6-2) are not properly adjusted. The differential level controller (H-3) influences subsystem pressure by varying oxygen flow. Improper coordination of controller action is shown in Figure 3.8-11 (b). The period of the oscillations in the illustration with unstable control is approximately six seconds. The frequency of the oscillations is dependent on the control adjustments. Operation with correct control gain adjustment is shown in Figure 3.8-11 (a).

As indicated in the control subsystem section (3.6) of this report, the ammonia synthesis rate is a function of the hydrogen-nitrogen molar ratio in the synthesis gas with the optimum ratio at approximately 2.5 to 1. The control of the synthesis gas hydrogen-nitrogen ratio was simulated on the analog computer. The ammonia synthesis rate was sampled every six minutes by measurement of synthesis gas flow rate and temperature rise through the converters. The nitrogen feed rate was adjusted in accordance with whether the production rate increased or decreased during the six minute interval. Starting at a hydrogen-nitrogen molar ratio of 3 to 1 in the synthesis gas the controller biased the nitrogen flow to bring the hydrogen-nitrogen ratio to the optimum operating point in 50 min (see Figure 3.8-12). A description of the program is presented in Appendix B.

CALCULATIONS

The basic mathematical equations which represent the dynamics characteristics of the major components are presented in the Investigation portion of this subsection (3.8). A detailed description of the analog computer program which was used to find the solution to these equations is presented in Appendix B.

REFERENCES

1. Burgreen, O., "Flow Coastdown in a Loop after Pumping Power Cutoff." Nuclear Science and Engineering. (Dec 1959) p. 306-42.
2. Williams, T. J., Ph.D. Thesis, The Pennsylvania State University, 1955.
3. Final Report Energy Depot Electrolysis Systems Study. Allison Division of General Motors, EDR 3714, June 1964.

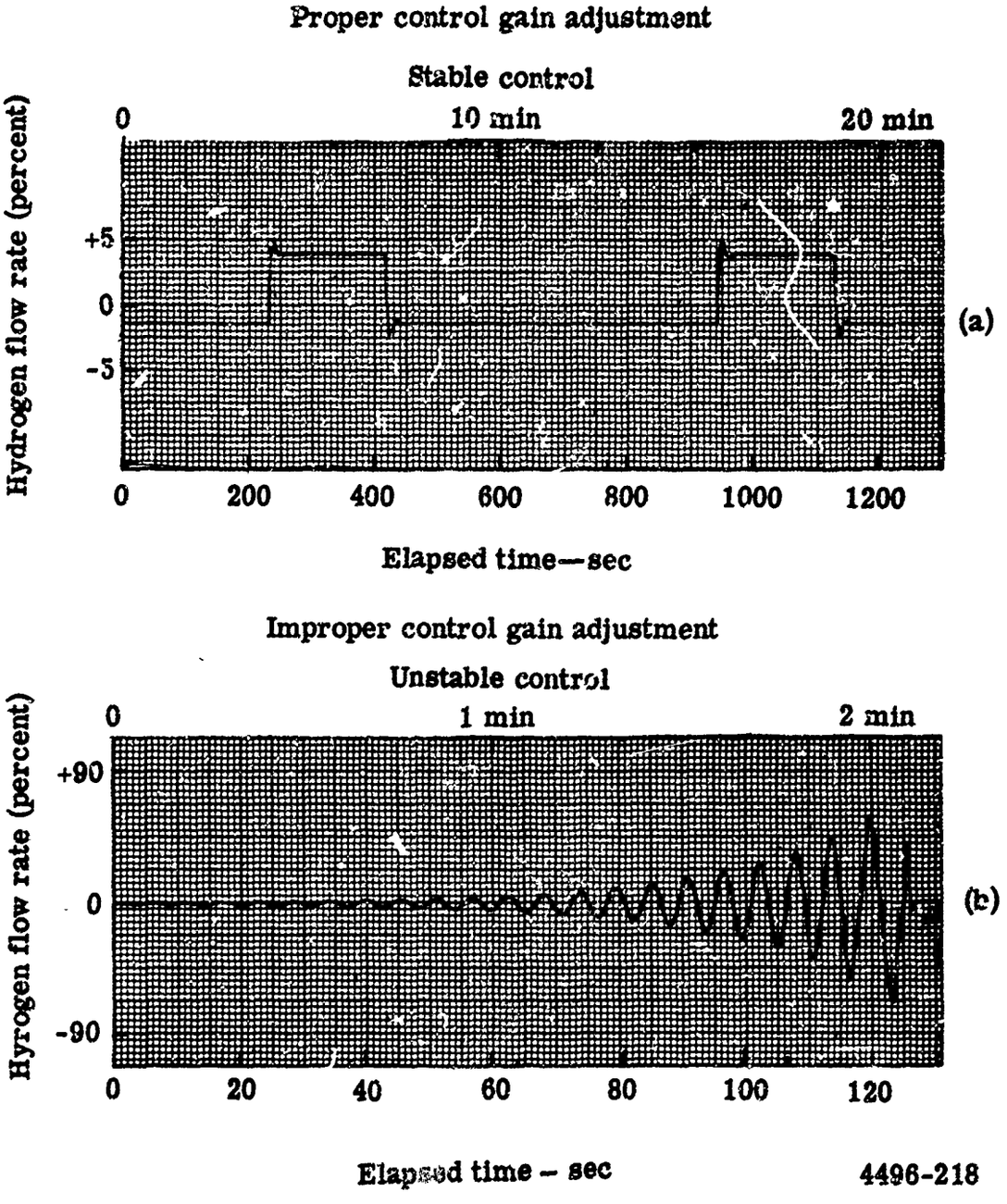


Figure 3.8-11. Effect of control gain adjustment on electrolyzer stability.

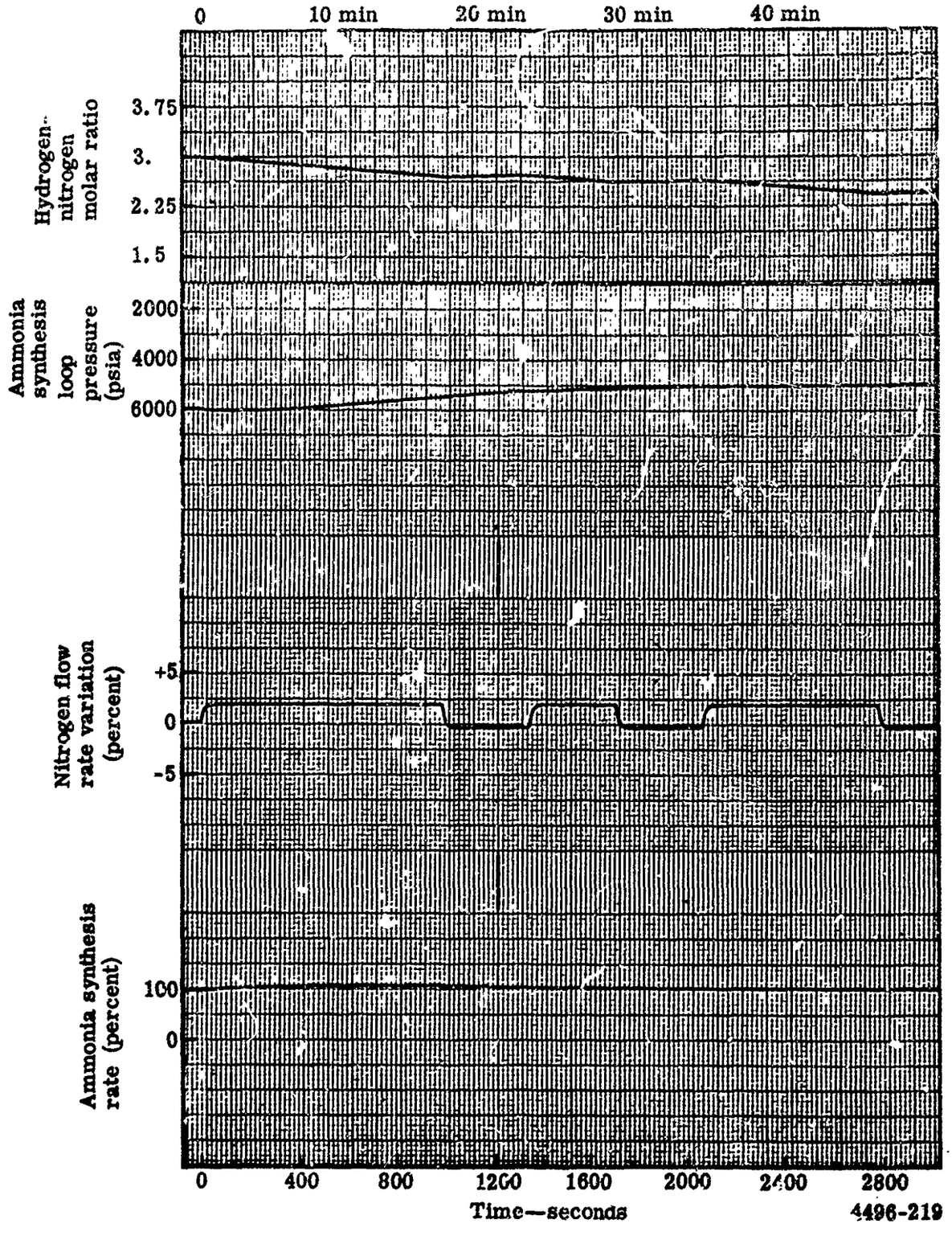


Figure 3.8-12. Control subsystem optimization of hydrogen-nitrogen molar ratio in ammonia synthesis loop.

Performance Analysis

DESCRIPTION

The eleven integrated and optimized fuel production systems presented in the Integration and Packaging subsection (3.7) are further defined by presenting the following regarding each system:

- Performance characteristics under design point and selected nondesign point ambient conditions
- The system modifications required for operation at nondesign point ambient conditions
- The manpower requirements for operation, start-up, shutdown, assembly and disassembly
- General considerations of the effect of input power variations

INVESTIGATION

In coordination with the integration and packaging studies, the performance of each of the eleven integrated and optimized fuel production systems was defined based on information developed in the subsystem design studies. Where necessary, the subsystem data were modified to reflect the changes in subsystem operating conditions resulting from the integration and optimization of the systems.

The operational procedures described later in this section were developed using engineering judgment to combine the subsystem information in a logical sequence. Manpower requirements and response times were determined similarly.

In the analysis of nondesign point performance, the ammonia production rate was established by the subsystem production rate most reduced at the nondesign conditions. In this way, the subsystem which exhibited the greatest reduction in production rate was used to establish the production rates from the other subsystems.

DESIGN PARAMETERS

The eleven plant designs were optimized for operation based on the criteria and parameters presented in the Introduction, Section II of this report. No new criteria were required to complete the performance analyses.

DESIGN

The eleven integrated systems were designed for the optimum ammonia production rate of specified ambient conditions consistent with the constraints appropriate to each system. A discussion of the techniques used to optimize the systems was presented in the Integration and Packaging subsection (3.7).

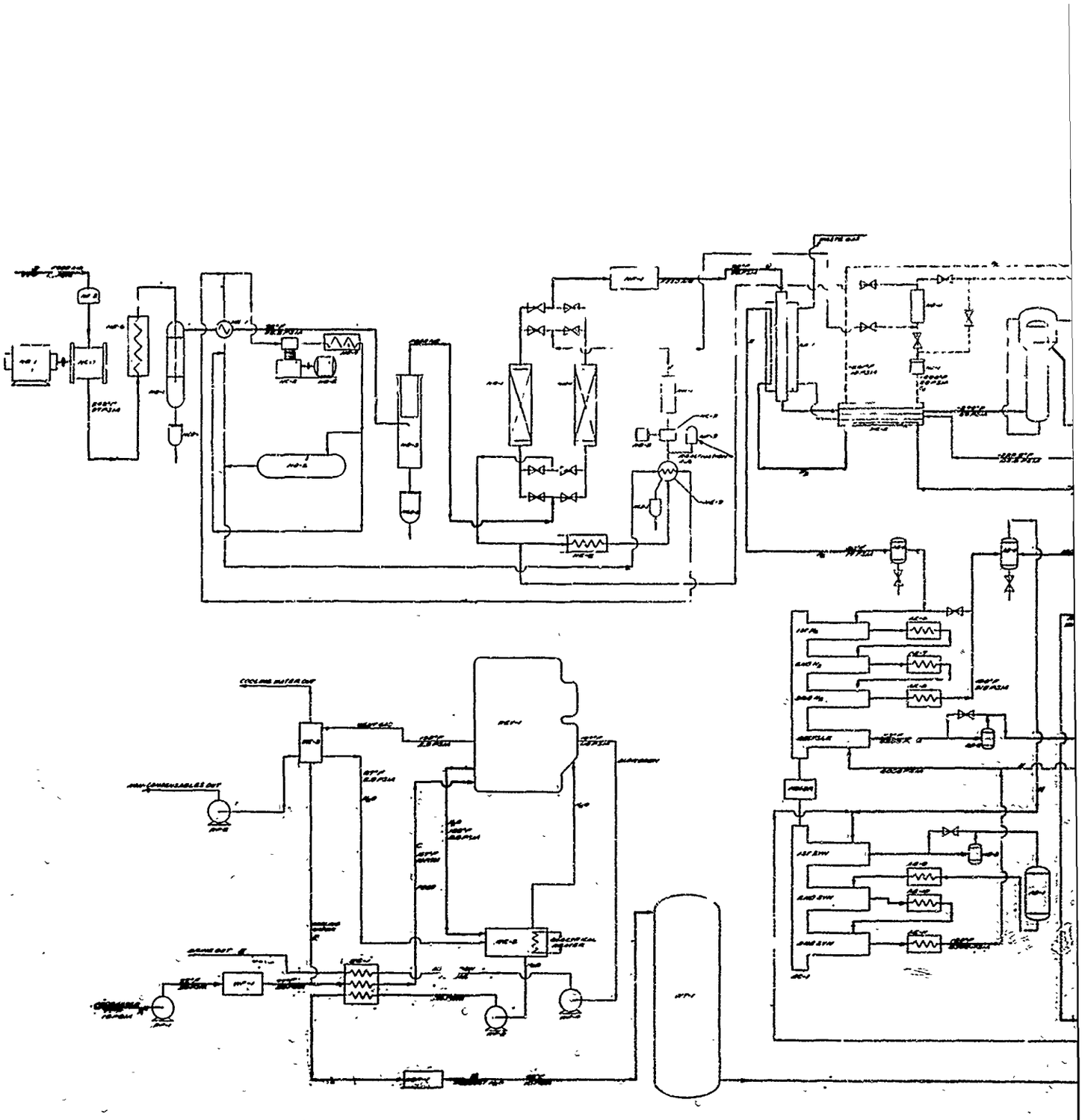
Nondesign point operation evolves principally around variations in ambient conditions. The effects of these variations are manifested in changes in the cooling fluid for the subsystems, and in freezing problems where water collects. No equipment modifications are required to accommodate changes in the conditions of the cooling fluid for the subsystems. Each skid will be totally enclosed by a skin cover that is louvered so that panels can be opened or closed to admit more or less ambient air as required for cooling purposes. Accordingly, more panels would be opened for operation at 125°F ambient temperature than for -65°F operation. Similarly, 10 psia ambient pressure conditions require more open panels than 14.7 psia ambient pressure conditions. To avoid freezing problems, separators, drainers, and process lines which function with streams containing water have to be insulated and heated. The amount of insulation required for system 1 was estimated to weigh 250 lb. No other equipment modifications are envisioned.

The power required by electrical heaters needed to supply heat in addition to the waste heat available was estimated to be insignificant.

PREDICTED PERFORMANCE

An overall system flow schematic was prepared by combining the water purification, nitrogen, hydrogen, and ammonia subsystem flow schematics. The combined overall system flow schematic is presented in Figure 3.9-1. The various pieces of equipment shown in Figure 3.9-1 are defined in the Equipment List given in Table 3.9-I. The subsystem operating temperatures and pressures are shown in Figure 3.9-1 at several selected, key positions. Since each system produces a different amount of ammonia product, stream flow rates are represented in Figure 3.9-1 with a code letter. The code letters were used in preparing Tables 3.9-II and 3.9-III, which show the stream flow rates and compositions for each lettered position in the overall flow schematic. The key stream flow rates and compositions are given in Table 3.9-II for system 1 operating at full power and design point conditions. These data were obtained from the systems integration work described in subsection 3.7 by adjusting each subsystem production rate using the hydrogen production rate as the basis. For example, the production rate of the nitrogen generation subsystem (in moles) was established by multiplying the moles of hydrogen produced by one-third. The moles of ammonia produced was then determined by assuming a 99% conversion efficiency, i.e., 99% of the nitrogen and hydrogen generated ultimately leaves the system as ammonia, and 1% of the nitrogen and hydrogen is lost by venting.

All of the stream flow rates for the remaining ten systems are determined for full power operation at design point conditions by multiplying the rates given in Table 3.9-II by the multiplying factors presented in Table 3.9-III. For example, the system ammonia production rate is represented in Figure 3.9-1 by the code letter L, which has a value of 42.8 lb-mole/hr for plant design 1 as shown in Table 3.9-II. The multiplying factor for plant design 11 is 4.53 for stream L as shown in Table 3.9-III. Therefore, the ammonia production rate for plant design 11 is 42.8 × 4.53 or 194 lb-moles/hr for full power operation at design point conditions. Other stream rates are determined similarly.



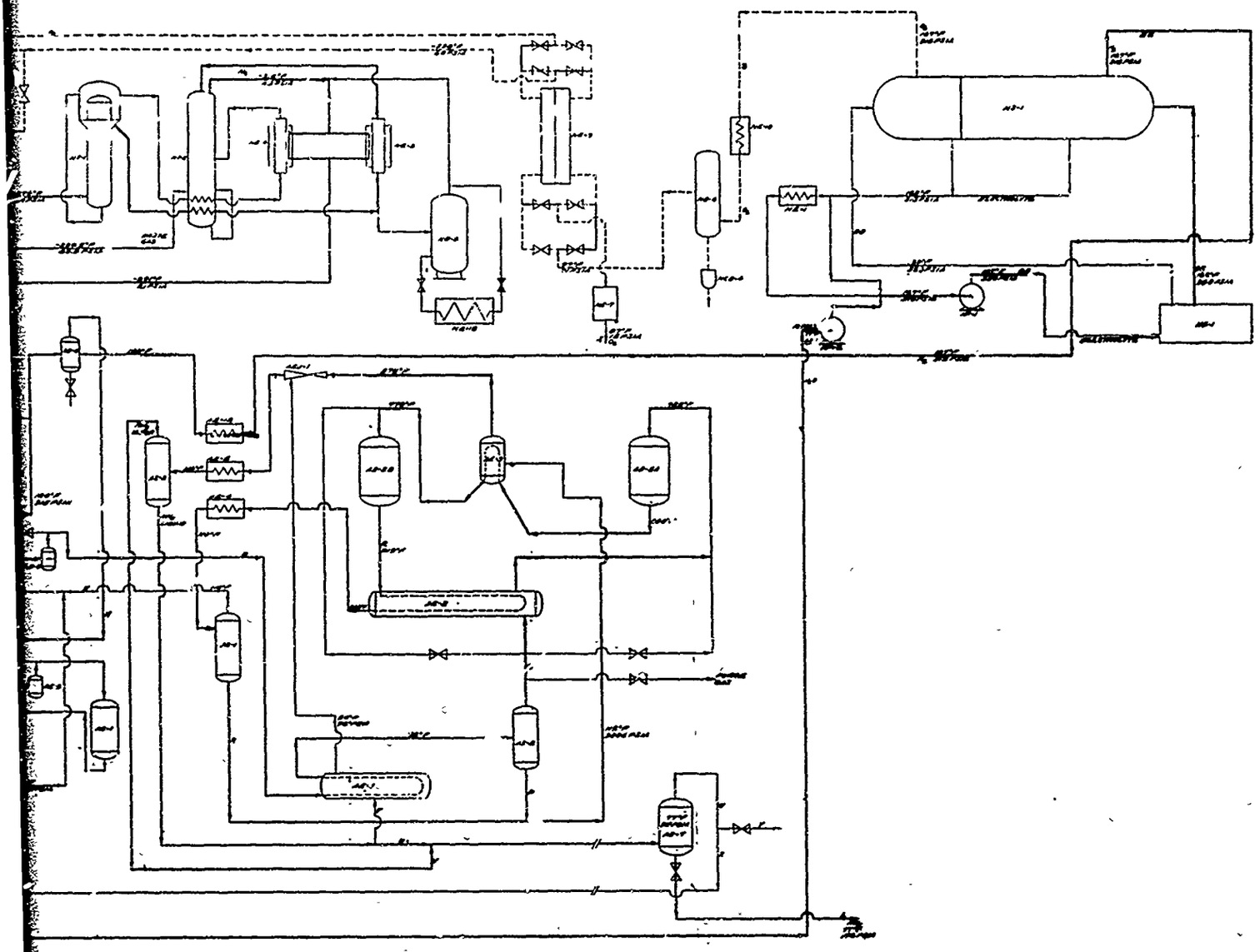


Figure 3.9-1. Fuel production system flow schematic.

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Table 3.9-I.
Equipment list.

Water purification subsystem

WE-1	Feedwater preheater
WE-2	Immersion heater tank
WE-3	Vent condenser
WEV-1	Evaporator vessel
WF-1	Feedwater filter
WP-1	Feedwater pump
WP-2	Vacuum pump
WP-3	Distilled water pump
WP-4	Blowdown water pump
WSP-1	Distilled water salinity probe
WT-1	Water storage tank

Nitrogen generation subsystem

NB-1	Air compressor motor
NB-2	Refrigeration compression motor
NB-3	Reactivation air blower motor
NC-1	Air compressor
NC-2	Refrigerant compressor
NC-3	Reactivation air blower
NDC-1 through 4	Continuous drainers
ND-1	Air compressor separator
ND-2	Refrigerant receiver
ND-3	Precooler separator
ND-4	Expander accumulator
ND-5	Liquid nitrogen storage tank
ND-6	Oxygen separator
ND-7	Silencer
NE-1	Main heat exchanger
NE-2	Refrigerated heat exchanger
NE-3	Oxygen reversing heat exchanger
NE-4	Crude oxygen subcooler
NE-5	Nitrogen subcooler
NE-6	Air compressor aftercooler
NE-7	Refrigerant condenser
NE-8	Reactivation gas cooler
NE-9	Refrigerated reactivation gas cooler
NE-10	Oxygen cooler

Table 3.9-I. (Cont)

Nitrogen generation subsystem (cont)

NE-11	Air precooler
NE-12	Liquid nitrogen heater
NF-1	Dust filter
NF-2	Air intake filter
NF-3	Reactivation air filter
NH-1	Reactivation air heater
NR-1	Air purifiers
NT-1	High pressure air fractionation column
NT-2	Low pressure air fractionation column
NX-1	Expansion turbine

Hydrogen production subsystem

HP-1	Electrolyte circulation pump
HP-2	Feedwater pump
HD-1	Electrolyzer
HE-1	Electrolyte heat exchanger
HS-1	Liquid-gas separator-reservoir

Ammonia synthesis subsystem

AC-1	Multiservice compressor
AD-1	Deoxo unit
AD-2A & 2B	Ammonia converters
AD-3	Gas filter
AD-4 & 5	Knockout drums
AD-6	Gas filter
AD-7	Ammonia storage tank
AE-1	Secondary condenser
AE-2	Converter heat exchanger
AE-3	Ammonia vaporizer
AE-4	Primary condenser
AE-5	Product condenser
AE-6 through 11	Compressor interstage and aftercoolers
AE-12	Hydrogen precooler
AEJ-1	Ejector
AS-1	Primary separator
AS-2	Secondary separator
AS-3	Product separator

Stream flow rates fo

Table 3.9-II.
EDR 4496

	H_2O purification subsystem					N_2 subsystem						Ammonia product				
Stream	A	B	C	D	E	F	G	H	I	J	K	L	M	N	O	P
H_2O	271	27	244	65.7	177	0.46				0.228	0.096					
Air						28.4	28.4			8.79						
N_2								21.55	0.440			0.005	21.55	59.55	81.5	0.06
O_2								0.015	5.46		32.3					
A								0.022	0.241			0.001	0.022	16.3	16.4	0.01
H_2												0.011	64.55	179	244	0.16
NH_3												42.8		27.0	28.2	8.35
KOH (aqueous)																
KOH (aqueous) + H_2																
KOH (aqueous) + O_2																

*Rates expressed in gal/min.

Table 3.9-III.
EDR 4496

System number
Factor for streams A through
Factor for stream BB
Factor for streams CC and DD

Table 3.9-II.

n flow rates for system 1 at full power and design point conditions—lb-moles/hr.

	NH_3 subsystem											H_2 subsystem					
	O	P	Q	R	S	T	U	V	W	X	Y	Z	AA	BB	CC	DD	EE
												0.382	65.697				0.765
55	81.5	0.033	81.5	60.0	0.306	0.010	0.005	0.365	0.365	0.303	0.062						
												32.275					
3	16.4	0.016	16.4	16 4	0.106	0.001	0.001	0.121	0.121	0.100	0.021						
	244	0.166	244	180	0.850	0.005	0.011	1.00	1.00	0.832	0.172						64.55
0	28.2	8.35	19.8	62.7	35.8	18.4	38.7	5.43	1.34	1.11	0.231						
														2450*			
															1139*		
																569*	

Table 3.9-III.

Stream flow rate multiplying factors.

	1	2	3	4	5	6	7	8	9	10	11
ams A through AA and EE	1.00	1.001	1.04	1.89	1.94	1.95	1.94	1.94	3.14	3.14	4.53
am BB	1.00	1.00	0.855	1.93	1.64	1.61	1.64	1.64	2.61	2.50	3.38
ams CC and DD	1.00	1.00	0.884	1.93	1.69	1.66	1.69	1.69	2.68	2.59	3.55

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The overall performance of each system is ultimately measured in terms of the amount of ammonia product that each is capable of producing while operating at full power at design point conditions. These ammonia production rates are shown summarized in Table 3.9-IV for each of the eleven systems along with the subsystem production rates.

Table 3.9-IV.
Design point plant production rates.

Plant design	Power input (Mw)	Subsystem production rates (lb/hr)			
		H_2O	N_2	H_2	NH_3
1	3	1180	603	129	729
*2	3	1190	604	129	730
3	3	1230	627	134	758
4	6	2240	1140	244	1380
5	6	2300	1170	250	1410
*6	6	2310	1180	252	1420
**7	6	2300	1170	250	1410
8	6	2300	1170	250	1410
9	10	3720	1890	406	2290
10	10	3720	1890	406	2290
11	15	5350	2730	585	3300

*H_2O cooled systems. All other systems are air cooled.
**400-cps power. All other systems use 60-cps power.

Half-Power Performance

It was anticipated that there may be rare occasions when the systems will be required to operate at less than full power input. An analysis was made to estimate the performance of each system when only half the design power is available. The subsystems were examined to determine the effects of one half power on each.

Since the water purification subsystem has provision for water storage, the unit was assumed to operate at full power but for only half time. The average power input to the water purification subsystem is then half the design power input.

The water filtration unit (required only on water-cooled systems) is small and was allowed to operate at full power at reduced system power inputs.

The nitrogen generation subsystem would require variable ratio belt drives, two-speed motors, or two half-capacity motors in place of one full-capacity motor to operate at half power. Since

half power operation will be rare and the nitrogen subsystem is a relatively low power consumer, it was considered best to allow the subsystem to operate at full power and simply provide recycle lines around the air compressor and the nitrogen compressor where necessary.

In the case of the ammonia synthesis subsystem, it was also considered impractical to equip motors with variable belt drives or two-speed motors. Instead, the subsystem was permitted to operate at full power except for these systems equipped with multiple, identical units such as compressors. For the latter, the appropriate fraction of the multiple units were simply turned off.

To determine the performance of the hydrogen production subsystem with the entire system at half power, the power requirements for the other subsystems were first summed and then subtracted from the total power input. This power difference was then distributed to the hydrogen production subsystem and the resulting subsystem performance was evaluated. In those cases where a number of half power modes of operation were possible, that mode was selected which made the most product with the available power without overloading any equipment.

A summary of the results of the foregoing analyses are presented in Table 3.9-V. The performance of each system is given as the percent of the full power ammonia production rate. The modes of operation of the subsystems are also given except for the water purification, nitrogen, water filtration, and control subsystems where the mode of operation is invariant as previously outlined. In general, the results presented in Table 3.9-V indicate that the systems are capable of producing approximately half of their full power production rates when half the design power is supplied.

Performance of Off-Design Ambient Conditions

The Energy Depot systems are required to operate in various parts of the world under varying conditions of ambient temperature and pressure. Accordingly, a study was made to establish the performance of the Energy Depot for an appropriate range of ambient conditions. The range of ambient air temperatures studied was from -65°F to 125°F, and the range of ambient air pressures studied was from 10 to 14.7 psia. For the water-cooled systems, the range of cooling water inlet temperatures studied was from 35 to 105°F. The results of these analyses are discussed starting with the air cooled systems.

Ambient Temperature Effects

There are three major effects of ambient temperature on the performance of the nitrogen subsystem. First, the quantity of feed air processed will be altered because the air density is inversely proportional to the ratio of absolute temperature. There is a corresponding change in the amount of nitrogen produced. At 125°F ambient temperature, the nitrogen production rate is 92% of the rate at 77°F. At -65°F ambient temperature, the nitrogen production rate

is 138% of the 77°F rate. Second, the quantity of refrigeration required to precool the air before purification is altered. Below 55°F ambient temperatures no auxiliary refrigeration is required. As the ambient temperature increases above 55°F, the amount of auxiliary refrigeration delivered automatically increases as pointed out in the nitrogen generation subsystem (3.2). Third, the heat leak to the cold box is influenced by the difference between the ambient temperature and the cold box temperature. At ambient temperatures below 77°F, the amount of refrigeration required from the electrolyzer oxygen expansion engine decreases. At ambient temperatures greater than 77°F, the higher heat leak is accommodated by increasing the oxygen expander inlet pressure. Thus, the nitrogen production rate is not affected by these latter two effects.

Table 3.9-V.
System performance at half input power operation.

System	Ammonia production rate (% of full power production rate)	Subsystem mode of operation
1 & 2	47.7	Transformer operates at half power. Ammonia subsystem operates at full power. Hydrogen subsystem uses balance of power with all modules on line.
3	46.5	Same as plant design 1.
4	49.2	One transformer, hydrogen production subsystem, and ammonia compressor shut down. The other multiple units operate at full power.
5, 6, 7, & 8	49.3	Transformer operates at half power. One ammonia compressor shutdown. Hydrogen subsystem uses balance of power with 6 of 13 module banks shutdown.
9	51.2	Both transformers operate at half power. One ammonia compressor shutdown, two operate at full power. Hydrogen subsystem uses balance of power with all modules on line.
10	51.2	Transformer operates at half power. One ammonia compressor shutdown, two operate at full power. Hydrogen subsystem uses balance of power with all modules on line.
11	52.4	Same as plant design 10.

Ambient temperature variations have a pronounced effect on the ammonia synthesis subsystem performance. Temperatures below 77°F increase the capacity of the fin-fan units. Increased

cooling lowers the compressor inlet temperature and raises the capacity. Also, the concentration of ammonia in the gas going to the converter is lowered which increases the converter capacity. At ambient temperatures of -65°F, the subsystem capacity is about 120% of the 77°F capacity. When the ambient temperature rises to 125°F, the temperature achieved in the air-cooled primary condenser rises from the 110°F design value to about 145°F. The temperature of the condensed product in the product separator also increases which raises the pressure. The increased pressure adversely effects the ejector performance. The reduced quantity of product collected in the primary separator reduces the motive gas going to the ejector, thus reducing the capacity of the secondary condenser. The higher gas temperatures of the gas leaving the compressor inter- and aftercoolers reduce the compressor capacity. These changes reduce the subsystem capacity to about 40% of the design rate.

The production capability of the hydrogen production subsystem is limited by the temperature of the electrolyte which is influenced by the ability of the heat exchangers to dissipate the heat generated in the electrolyzer. At lower than design ambient temperatures, more heat can be removed, and, consequently, the hydrogen production rate may be increased. At an ambient temperature of 5°F, for example, the hydrogen production rate limit is about 140% of the design value. At ambient temperatures greater than 77°F, the heat exchangers limit the hydrogen production rate to values less than the design rate. At 125°F ambient temperature, the capacity of the subsystem is reduced to about 65% of the design capacity.

The power conditioning subsystem is limited by the ability of the heat exchanger to dissipate the heat generated in the transformer. In addition, however, the maximum d-c voltage of 619 v limits the output of the subsystem. At an ambient temperature of 125°F, the transformer limits the output to about 75% of the design output. At an ambient temperature of -65°F, the rectifiers limit the output to about 133% of the design output.

Ambient Pressure

The primary effect of ambient pressure variations on the nitrogen generation subsystem is on the air compressor performance, since the air density is proportional to the absolute pressure. At an ambient pressure of 10 psia, the capacity of the compressor is reduced to about 66% of the design point value and the amount of nitrogen produced is decreased to 68% of the design value.

Lowered atmospheric pressure has no effect on the actual ammonia loop operation but it does reduce the capacity of the air cooler fans. At an ambient pressure of 10 psia, the decreased air-cooling drops the ammonia production rate to about 67% of the design rate. The effect of ambient pressure on the hydrogen production subsystem is the same as described for the ammonia subsystem. The effect of ambient pressure on the power conditioning subsystem is similar. At an ambient pressure of 10 psia, the hydrogen production rate is decreased to about 80% of the design point rate, and the transformer output is reduced to about 85% of the design output.

Overall Effects of Ambient Air Variations

All of the preceding effects of ambient air temperature and pressure variations were collected and plotted. The results are shown in Figures 3.9-2 and 3.9-3. The percent of design point production is plotted on the ordinate, and the ambient air temperature is plotted on the abscissa of both figures. Figure 3.9-2 is for 14.7 psia ambient pressure, and Figure 3.9-3 is for 10 psia ambient pressure. The overall system performance is limited by the individual subsystem which exhibits the lowest production rate at any set of ambient conditions. At -65°F and 14.7 psia, Figure 3.9-2 shows that the ammonia synthesis subsystem limits the system production rate to ~120% of the design point production rate provided the required additional electrical power is available. At -65°F and 10 psia, Figure 3.9-3 shows that the system production rate is limited to about 93% of the design rate by the nitrogen generation subsystem. At 125°F and 14.7 psia, the ammonia synthesis subsystem limits production to about 40% of the design rate (Figure 3.9-2). At 125°F and 10 psia, the ammonia subsystem again limits the production rate, and the rate is reduced to about 27% of the design point capacity (Figure 3.9-3). At 77°F and 10 psia, the production is limited by both the nitrogen and the ammonia subsystems to about 68% of the design rate (Figure 3.9-3).

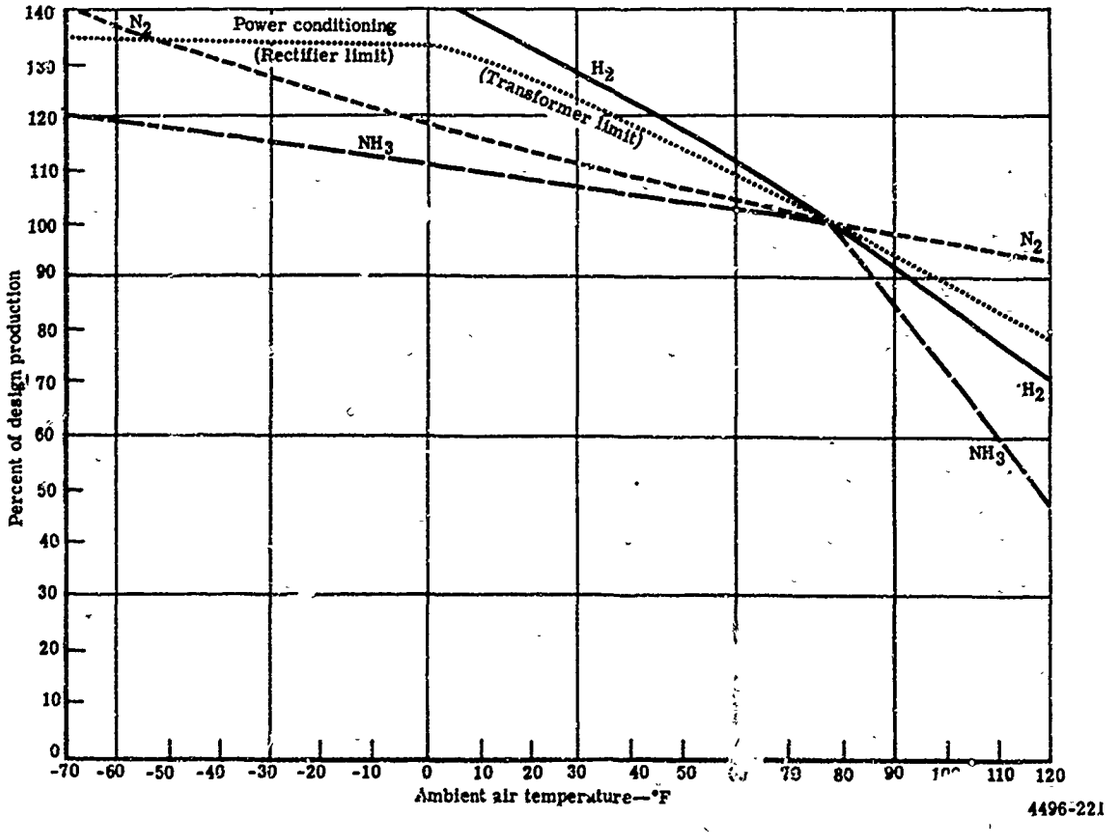


Figure 3.9-2. Subsystem off-design point performance (air-cooled—14.7 psia pressure).

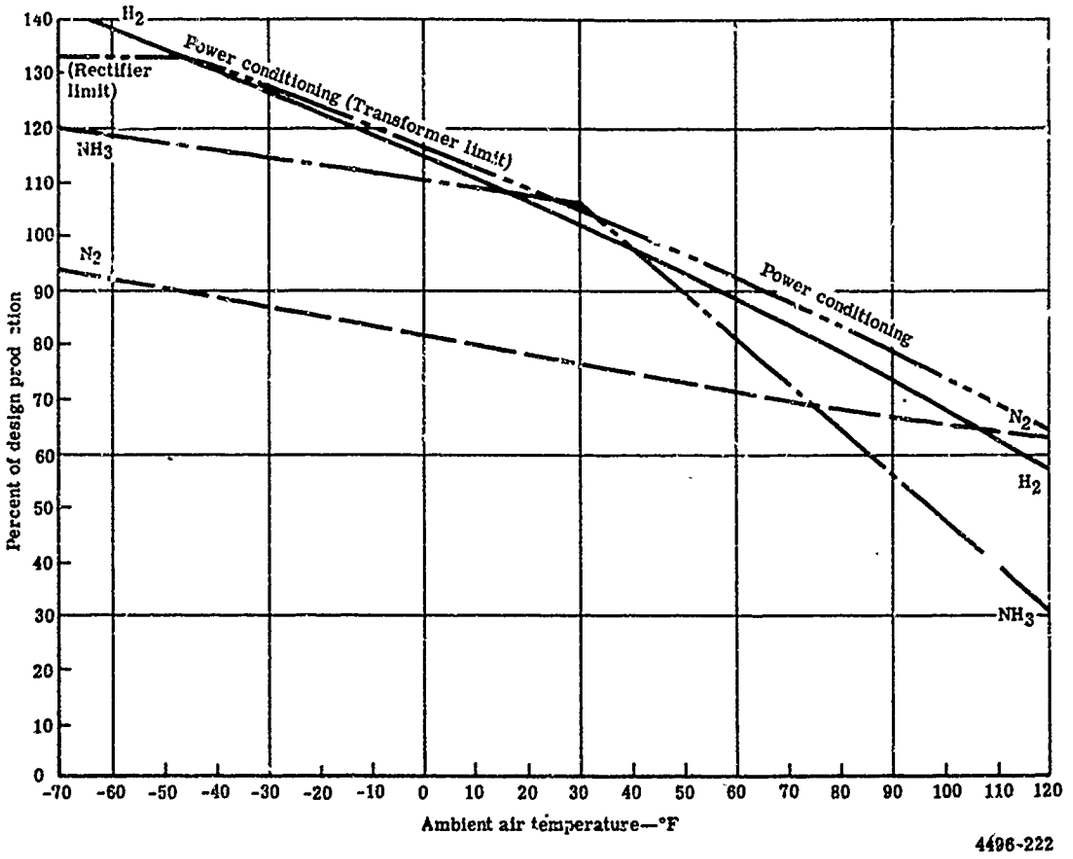


Figure 3.9-? Subsystem off-design point performance (air-cooled—10 psia).

Cooling Water Temperature

Systems 2 and 6 were designed using water as the primary cooling fluid. The amount of cooling water required by each of the subsystems is shown in Table 3.9-VI for design point conditions. To avoid scaling problems in heat exchangers, the maximum temperature of the cooling water leaving any heat exchanger was limited to 110°F for extended periods of operation.

Cooling water temperature variations influence the inter- and aftercooler gas temperatures around all the compressors as described earlier for ambient air temperature variations. Heat exchanger exit temperatures are influenced similarly as also outlined previously. Hence, at a cooling water inlet temperature of 35°F, the nitrogen generation subsystem production rate is increased to about 108% of the design rate, the ammonia synthesis subsystem production rate is increased to about 106% of the design rate, the hydrogen production subsystem production rate capability is increased to about 133% of the design rate, and the power conditioning subsystem capacity is increased to about 120% of the design value. At a cooling water inlet temperature of 105°F, the nitrogen generation subsystem production rate is reduced to about 53%

of the design rate, the ammonia synthesis subsystem production rate is reduced to about 15% of the design rate, the hydrogen production subsystem production rate is reduced to about 35% of the design rate, and the power conditioning subsystem capacity is reduced to about 40% of the design value. All of these data are shown plotted in Figure 3.9-4. The overall system production rate is again limited by the subsystem yielding the minimum amount of product. Therefore, as shown in Figure 3.9-4, the ammonia subsystem is the limiting subsystem over the entire temperature range of interest. For a 35°F cooling water inlet temperature, the ammonia subsystem limits the system production rate to ~106% of the design value provided that the required additional electrical power is available. For a 105°F cooling water inlet temperature, the ammonia subsystem limits the system production rate to about 15% of the design rate.

Table 3.9-VI.
Cooling water flow rate—gal/hr.

System	2	6
Hydrogen production subsystem	3,960	9,900
Nitrogen generation subsystem	880	1,780
Ammonia synthesis subsystem	5,350	10,700
Total	10,190	22,380

These data suggest that heat exchangers should be oversized if extended operation in high temperature environments is anticipated to avoid the marked reduction in production capacity which otherwise results.

Cross-Contamination of Hydrogen and Oxygen

Any oxygen entering the ammonia synthesis subsystem must be removed in the deoxo unit to avoid poisoning the catalyst. As mentioned in the nitrogen generation subsystem section (3.2), a very slight amount of oxygen impurity is present in the nitrogen product stream (0.069%). This impurity oxygen was accounted for when the subsystems were designed, and is reflected in the system performance data. However, cross-contamination of the two hydrogen production subsystem product streams will also affect the system performance. Some hydrogen is expected to be present in the leaving oxygen stream, and some oxygen is expected to be present in the leaving hydrogen stream because the liquid-gas separator cannot separate the dissolved O_2 and H_2 from the electrolyte. These electrolyte streams are then combined for circulation back through the modules. Thus, the electrolyte entering the cathodic chamber will contain some dissolved O_2 which is displaced by hydrogen, thus releasing the oxygen which then appears in the hydrogen gas stream. A similar process occurs in the anodic chamber of the

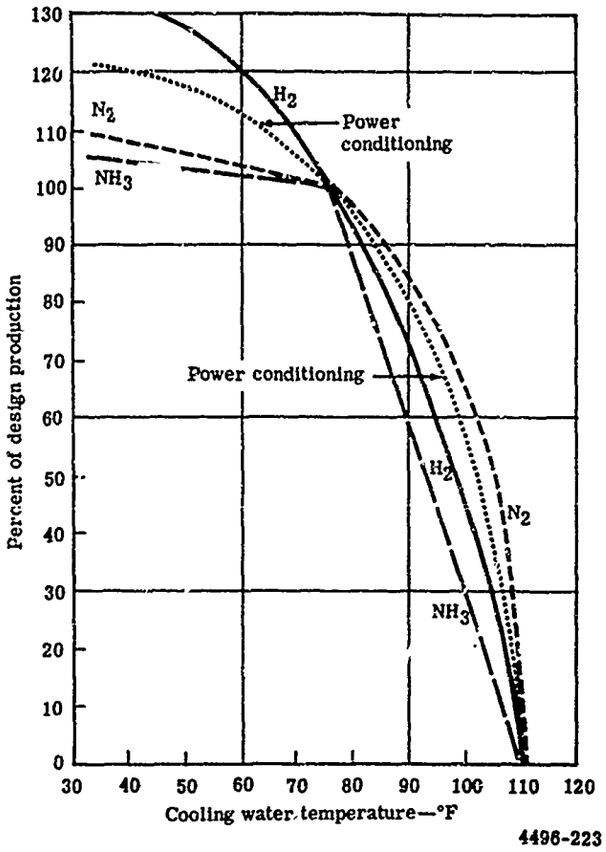


Figure 3.9-4. Subsystem off-design point performance—water cooled.

modules accounting for the hydrogen in the oxygen stream. This cross-contamination is discussed more completely in Reference 1 at the end of this subsection.

Hydrogen contamination in the oxygen stream is lost as far as ammonia production is concerned, and oxygen contamination in the hydrogen stream consumes hydrogen in the deoxo unit when it is removed. This hydrogen is also lost to the ammonia synthesis subsystem since it is converted to water. Accordingly, an analysis was performed to determine the effect of these two hydrogen losses on the system production rate. An equation was derived which expressed this effect parametrically. The derivation of the equation is given in the Calculations portion of this subsection. The results of the analysis are presented in Figure 3.9-5. The percent of full ammonia production is plotted on the ordinate, and the mole percent of oxygen in the hydrogen stream leaving the hydrogen production subsystem liquid-gas separator is plotted on the abscissa of the figure. The parametric curves shown in the figure represent the different mole percents of hydrogen lost in the oxygen stream leaving the liquid-gas separator. To illustrate how these data are used, assume that there is 2% of oxygen in the hydrogen stream and 1% of hydrogen in the oxygen stream. The data in Figure 3.9-5 indicate that at these conditions, the ammonia production rate is about 95.4% of the rate for no hydrogen losses or oxygen contamination.

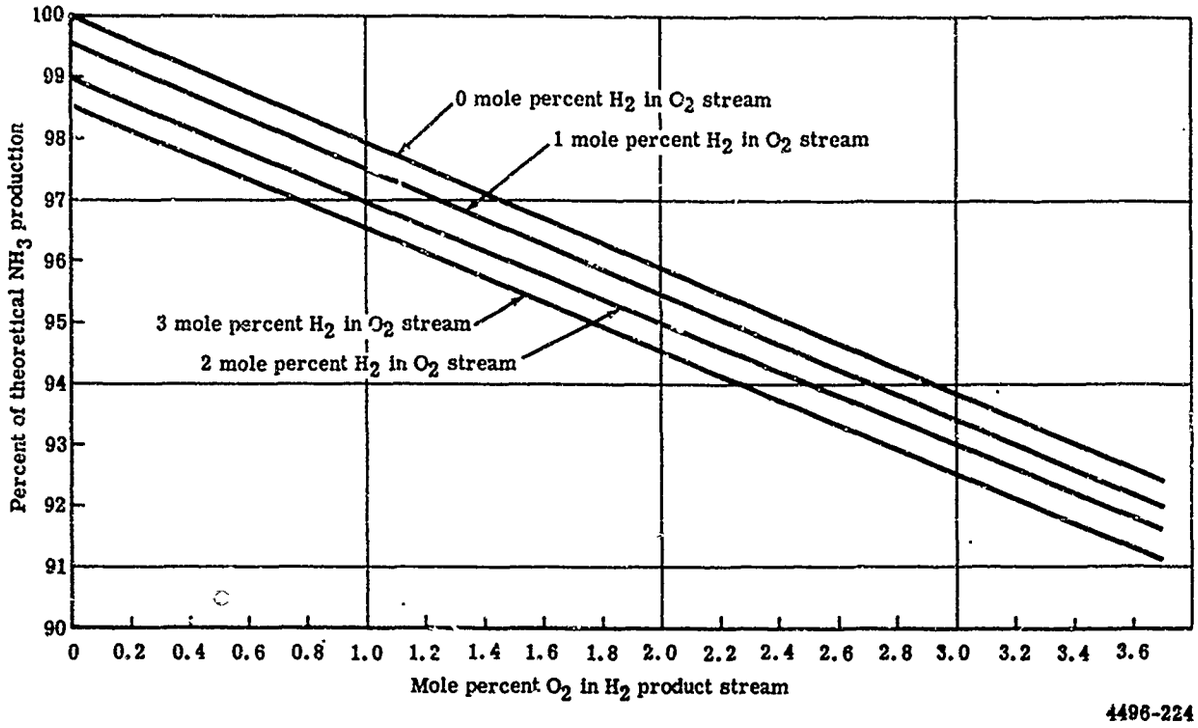


Figure 3.9-5. Effect of O_2 contamination of H_2 product stream and H_2 lost in O_2 stream on NH_3 production.

Manpower Requirements

The fuel production system is designed to be fully automatic during normal operation. The operator function will be primarily that of monitoring the system operating parameters. A single control cab is used to operate the entire fuel production system. Two operators per working shift plus an officer in charge for a total of seven men would be required. On each shift, one operator would monitor the control console while the other would provide relief and visual observation of the fuel production system. During system assembly and disassembly phases, all six operators would assist. It is probable that the reactor powerplant could be monitored by the same operators.

It is not anticipated that operation at off-design conditions will alter the manpower requirements.

Procedures

The procedures that must be followed to assemble, start-up, shutdown, disassemble, and establish standby conditions for the fuel production system were determined by combining the procedures for each subsystem in the proper sequence. The procedures which follow are based on the premise that the operating site has been selected and properly prepared, transport

trailers have been properly positioned, and that 4160 v electrical power is available at the site. The tractors may be disconnected from the system transport trailers if the tractors are required elsewhere.

Assembly Procedure

1. Install leveling jacks and level packages.

2. Open package enclosure panels to ensure proper air flow and personnel access to equipment.

3. Visually inspect plant apparatus for damage incurred during transit. If necessary, make arrangements for repairs to damaged equipment.

4. Make necessary interconnections between packages as discussed in Integration and Packaging subsection (3.7).

5. Unload air, nitrogen, and synthesis gas compressor cylinders.

6. Open manual shutoff valves on hydrogen skid in hydrogen and oxygen collection pipes at inlet to liquid-gas separator.

7. Open manual electrolyte shutoff valves at entrance to each module bank on hydrogen skid.

8. Open manual feedwater valve to water purifier.

Start-up Procedure

1. Energize power conditioning subsystem.

2. Turn on control subsystem air compressor and energize system controls.

3. Conduct prestart check—check instrument panel to verify subsystems in operating mode and no safety problems exist.

4. Unblock nitrogen and ammonia subsystems and open hydrogen and nitrogen vent valves.

5. Close circuit breaker to start electrolyte circulation pump on hydrogen subsystem.

6. Start flow of liquid nitrogen from storage tank into air fractionation columns and heat exchangers. Flow stops automatically when operating conditions reached.

7. For systems 2 and 6, close circuit breaker to start cooling water filtration unit.

8. Apply power to electrolyzer modules to begin generation of hydrogen and oxygen.

9. Close circuit breakers to start nitrogen generation subsystem air compressor, cooling air blower, refrigeration compressor, and reactivation air blower.

10. Open valve admitting electrolyzer oxygen to nitrogen generation subsystem and start oxygen switching heat exchanger sequence.

11. Start water purification subsystem. This subsystem has an automatic startup sequence.

12. Start nitrogen gas compressor (part of ammonia synthesis multiservice gas compressor on 3 and 6 Mw_e systems. (See step 14.)

13. Close nitrogen vent valve.

14. Start synthesis gas and recycle compressor including inter- and aftercooler fans.

15. Close hydrogen vent valve.

16. When synthesis gas purity reaches preestablished level, admit synthesis gas to ammonia loop.

17. Start electric start-up heaters in ammonia converters.

18. Increase ammonia loop pressure to match increasing loop temperature until reaction "lights off."

19. Place primary ammonia condenser into service and adjust exit gas temperature.

20. Increase synthesis gas flow rate through converters until electric heaters not needed to sustain reaction. When reaction becomes self-sustaining, open circuit breakers to turn off ammonia converter electric heaters.

21. When ammonia level in primary separator reaches set point, place ammonia vaporizer into service.

22. Place ammonia product condenser into service and adjust exit gas temperature.

23. When ammonia level in product separator reaches set point, admit product ammonia to storage.

24. Admit sidestream of product ammonia to secondary condenser and place ejector into operation.

25. When ammonia level in secondary seperator reaches set point, combine stream with primary separator effluent.

When the hydrogen production subsystem design temperatures, pressures, and liquid levels are reached, the cooling fans and feedwater pump start automatically. The system now operates in the automatic, on-stream mode of operation.

Shutdown Procedure

1. Open hydrogen and nitrogen gas vents, unloading the synthesis gas compressor and nitrogen compressor cylinders as required.

2. Turn off water purification unit and close manual feedwater valve.

3. Turn off electric power to electrolyzer modules but keep electrolyte circulation pump on to sweep gas out of piping and to separator tank.

4. Lower pressure in ammonia synthesis loop, until reaction ceases, by venting.

5. Open circuit breaker to shut down air, nitrogen, and synthesis gas compressors and blowers and refrigeration compressor in nitrogen subsystem.

6. Stop flow of cooling fluid to all subsystems and open circuit breaker to turn off electrolyte pump.

7. Vent the hydrogen production subsystem to atmospheric pressure and block in positive pressure in nitrogen generation and ammonia synthesis subsystems to prevent the leakage of air.

8. Close manual valves at entrance to hydrogen production subsystem separator tank and modules.

9. Drain water from drainers in nitrogen generation subsystem and deoxo unit and knockout drums in ammonia subsystem.

10. Turn off control subsystem.

11. Deenergize power conditioning subsystem.

Disassembly Procedure

1. Remove skid interconnections and store lines.

2. Close skid enclosure panels.

3. Remove leveling jacks.

Standby Procedure (Standby is defined as being prepared to bring the system on-stream)

1. Complete assembly procedure. (Not required if system has operated previously at location.)

2. Energize power conditioning subsystem.

3. Turn on control subsystem air compressor and energize system controls.

4. Conduct prestart-up check—check instrument panel to verify subsystems in operating mode and no safety problems exist.

5. Unblock nitrogen and ammonia subsystems.

Off-design Procedures

In general, the procedures will be the same for all operating conditions. When operating at high ambient temperature or low pressure, a greater number of enclosure panels must be opened to ensure adequate coolant flow. When operating at subfreezing conditions, the appropriate electrical heaters located on water drainers and line must be turned on.

Time Requirements

The time requirements to carry out the preceding procedures were established using subcontractor data and engineering judgement to determine how much time is needed to perform each step. Where sufficient manpower was available and the steps were not sequential, an effort was made to perform steps concurrently to minimize the overall time requirements. To obtain the time required to connect and disconnect the system packages, it was estimated that two man-minutes would be required to make or break each package interconnection. Also, it was estimated that one man-minute would generally be required to open or close each manually operated valve.

The results of this analysis are shown graphically for system 1 in Figure 3.9-6 for the assembly and start-up phases, and in Figure 3.9-7 for the shutdown, disassembly, and standby phases of operation. The first column presents the specific task being performed. The second column presents the steps required to perform the task. The step numbers are keyed to the numbers of the step descriptions presented previously in the Procedures subsection. The third column presents the number of men required to perform each step. For this analysis, it was assumed that all six operating personnel would be available during the assembly and disassembly phases, and that only the two shift operators would be available during the start-up, shutdown, and standby phases. The last column headed "Time" presents the total elapsed time required to perform each step. Referring to Step 1 in the Assembly phase, for example, the data in Figure 3.9-6 show that it takes 12 min for the six men to perform this step, or a total of 72 man-minutes are required. Other steps are analyzed similarly.

The total elapsed time required to perform each task is obtained by determining how much time has elapsed when the final steps are completed. For example, referring again to the Assembly task in Figure 3.9-6, Steps 5 and 7 are the last steps completed and are finished at the same time, viz, after 26 min have elapsed from beginning the assembly procedure. Therefore, the assembly time for system 1 is 26 min. The time required to perform the other tasks is obtained similarly.

A summary of the time required to perform the assembly, start-up, shutdown, disassembly, and standby tasks for all eleven systems is provided in Table 3.9-VII. In addition, the response time for each system is shown in the table. The response time was obtained by summing the time requirements for the assembly, start-up, shutdown, and disassembly phases. Two standby phase time requirements are given. The first entry shows the time required if the fuel production plant must be assembled at a new site, and therefore, includes the system assembly time. The second entry shows the standby time requirement if the fuel production plant has already been operated at the site, and therefore, is already assembled.

Table 3.9-VII.
System time requirements for assembly, start-up, shutdown, disassembly, and standby phases (minutes).

System	1	2	3	4	5	6	7	8	9	10	11
Assembly & Start-up	26	26	30	45	30	32	29	22	47	43	55
Assembly	26	29	30	45	30	32	29	29	47	43	55
Start-up	39	39	40	51	49	49	49	49	66	65	74
Total	65	68	70	93	79	81	78	78	113	108	129
Shutdown & Disassembly											
Shutdown	23	23	23	27	26	26	26	26	32	35	37
Disassembly	12	14	12	21	13	16	12	12	21	19	27
Total	35	37	35	48	39	42	38	38	53	54	64
Response time	100	105	105	144	118	123	106	116	166	162	193
Standby	36(a) 10(b)	39(a) 10(b)	40(a) 10(b)	57(a) 13(b)	40(a) 10(b)	42(a) 10(b)	38(a) 10(b)	39(a) 10(b)	59(a) 12(b)	55(a) 13(b)	67(c) 12(b)

(a) If assembly required
(b) If no assembly required

Task	Step	Personnel required
Assembly	1	6
	2	3
	3	3
	4	6
	5	3
	6	1
	7	1
	8	1
Startup	1	1
	2	1
	3	2
	4	2
	5	1
	6	1
	7	N.A.
	8	1
	9	1
	10	1
	11	1
	12	N.A.
	13	1
	14	1
	15	1
	16	1
	17	1
	18	1
	19	1
	20	1
	21	1
	22	1
	23	1
	24	1
	25	1

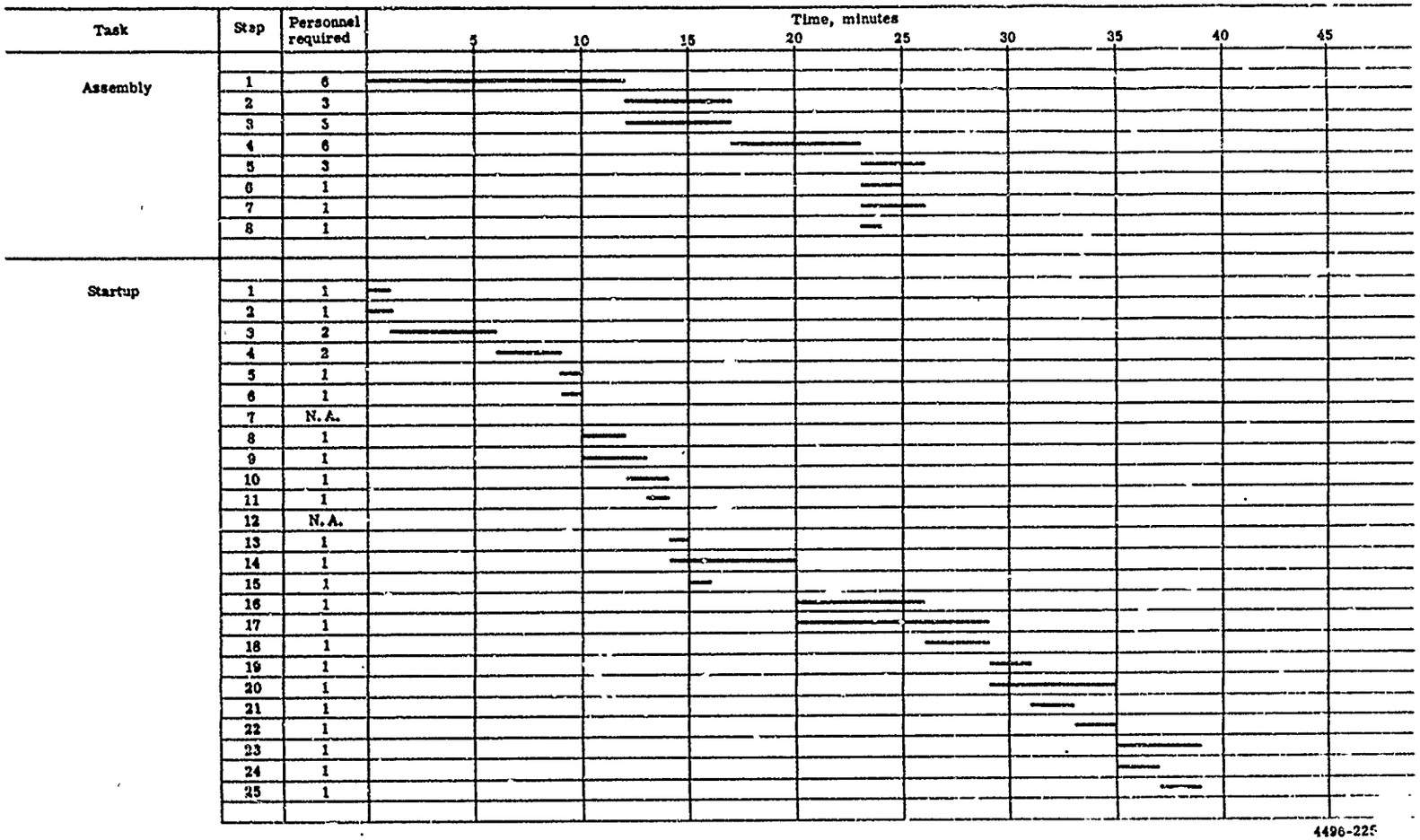


Figure 3.9-6. Assembly and start-up time—system 1.

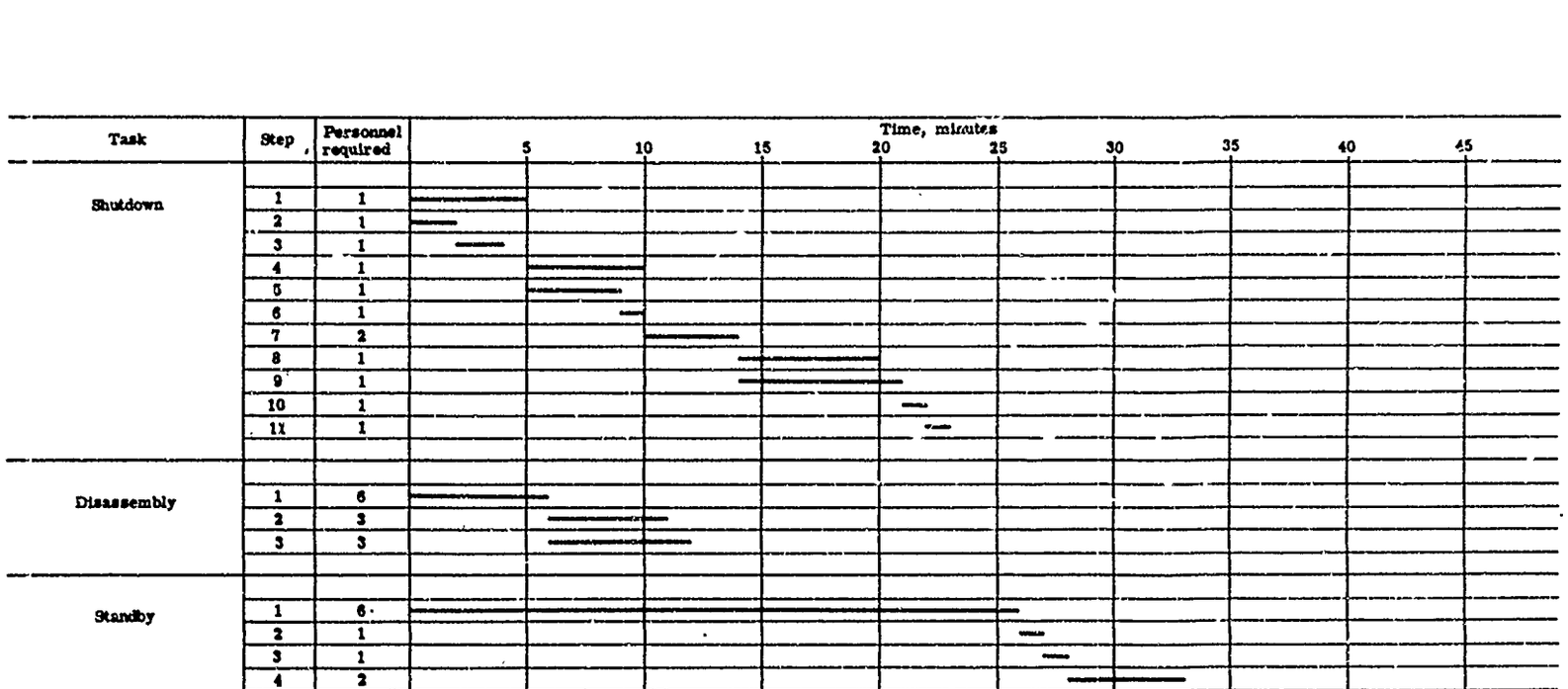


Figure 3.9-7. Shutdown, disassembly, and standby time—system 1.

Allison

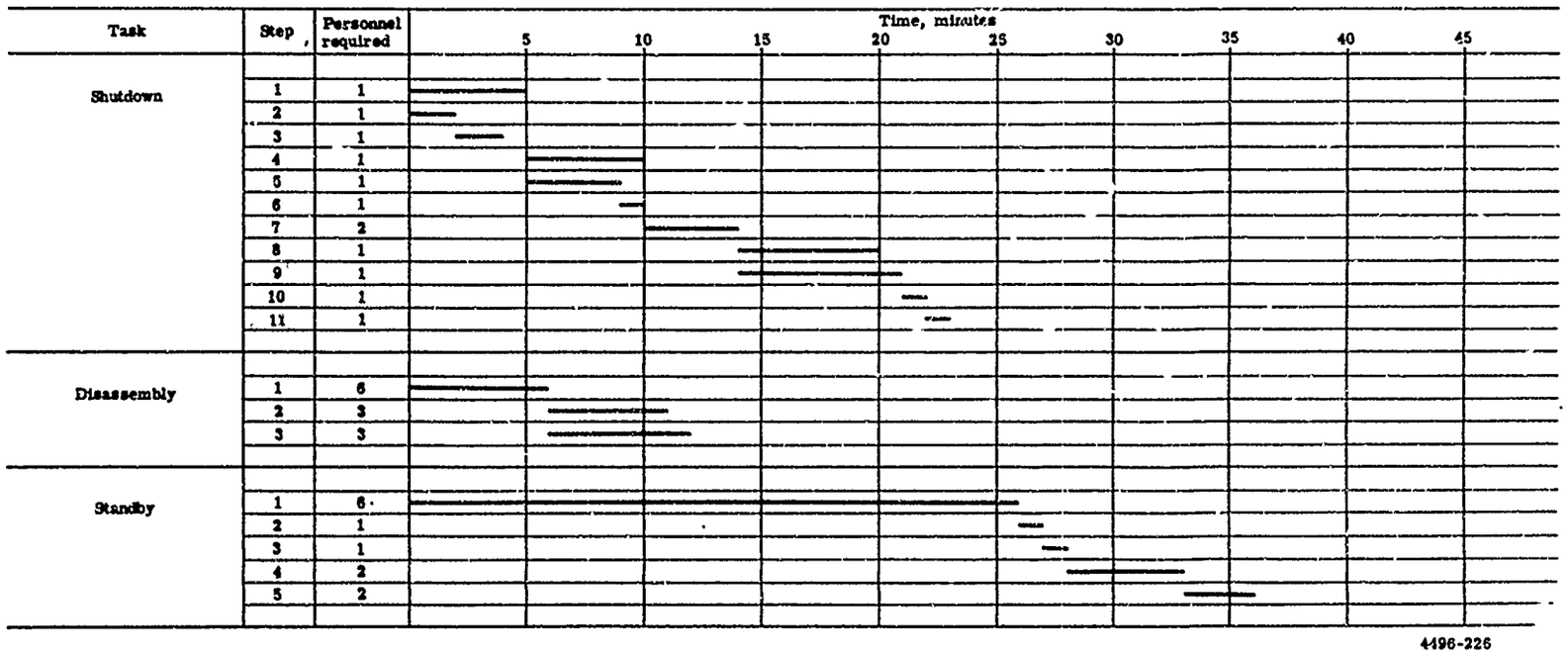


Figure 3.9-7. Shutdown, disassembly, and standby time—system 1.

Allison

Electrical Requirements

The electrical power requirements differ for each of the eleven systems due to the variations in system design parameters and constraints. A circuit diagram for the integrated systems was presented in the Power Conditioning Subsystem section (see Figure 3.5-10). A schematic diagram of the overall system power distribution is shown in Figure 3.9-8. The values of the lettered variables shown in Figure 3.9-8 are presented in Table 3.9-VIII keyed to the letters shown in the schematic for full power and half-power operation. Both the reactive and the significant resistive electrical load characteristics of each of the subsystems are presented. The standby condition was defined as being prepared to start the system on-stream. Consequently, the only power required during standby is that needed to energize the control subsystem, i.e., entry J in Table 3.9-VIII.

The resistive electrical load for the ammonia synthesis subsystem, item E in Figure 3.9-8, is the load required for the start-up heaters, which are not on during operation, but only during start-up periods as shown later. The heater loads are 100 kw for the 3 Mw_e systems, 150 kw for the 6 Mw_e systems, 200 kw for the 10 Mw_e systems, and 250 kw for the 15 Mw_e system.

For half-power operation, it was assumed that half of the design power was available to each of the eleven systems. This amount of power was distributed to the subsystems as described earlier under Predicted Performance.

The start-up and shutdown power requirements were determined by examining the system start-up and shutdown procedures presented earlier. The procedures indicate the sequence followed in bringing power consuming pieces of equipment on line during start-up, and the sequence followed in removing these components from the line during shutdown. Using the procedures as a guide, Figures 3.9-9 and 3.9-10 were developed. In Figure 3.9-9, the relative power delivered to the system is plotted for each step during the start-up sequence. The same information is presented in Figure 3.9-10 for the shutdown steps. The power delivered is shown as the percent of power delivered under full power, full production conditions. The curves shown are typical of all eleven systems; differences between systems are insignificant. It should be noted that the electric start-up heaters in the ammonia synthesis subsystem are started at a point in the start-up sequence such that greater than 100% of system design power would be needed unless power is cut back somewhere else. To accommodate this need, the control subsystem reduces the power to the electrolyzer modules sufficiently to keep the power demand at 100% of full system power. No similar problems are manifested during the shutdown sequence.

System Availability Analysis

An analysis was made to estimate the availability of the fuel production systems. The system availability was defined as the percent of the total time that the system is capable of operating.

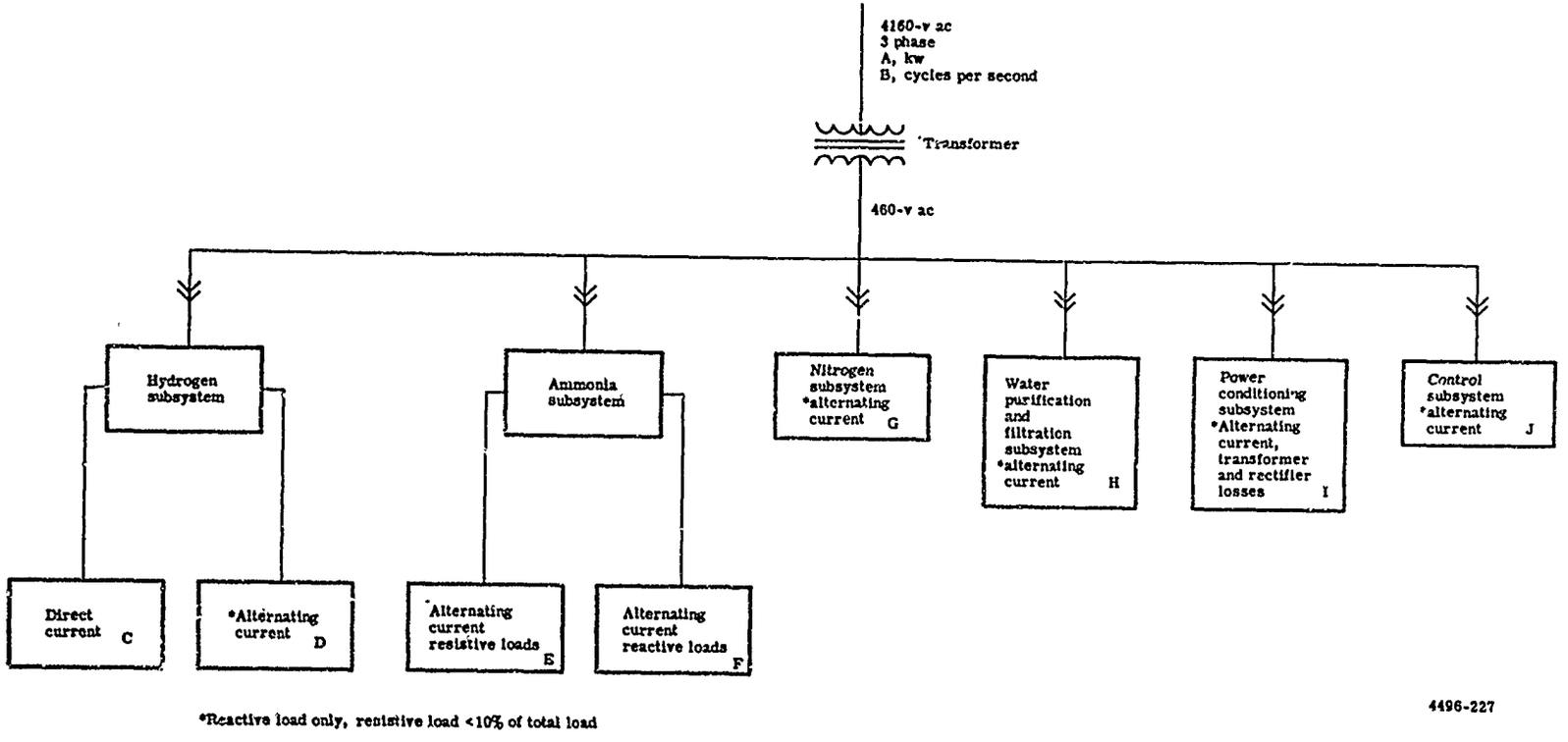


Figure 3.9-8. Schematic of system electrical power distribution.

Allison

Table 3.9-VIII

System electrical power

Plant design	1				2				3		
Operation	Full power		Half power		Full power		Half power		Full power		Ha
A, kw	3000		1500		3000		1500		3000		
B, cps	60		60		60		60		60		
Quantity	kw	amp	kw	amp	kw	amp	kw	amp	kw	amp	k
C, H_2 subsystem, dc	2643	4460	1171.7	2130	2648	4460	1175.1	2130	2630	4360	115
D, H_2 subsystem, ac	19.3	30.3	19.3	30.3	15.2	23.9	15.2	23.9	21.7	34.1	2
E, NH_3 subsystem, ac resistive	0	0	0	0	0	0	0	0	0	0	
F, NH_3 subsystem, ac reactive	233.8	366	233.8	366	229.2	359	229.2	359	243.1	381	24
G, N_2 subsystem, ac	45	70.5	45	70.5	45	70.5	45	70.5	47	74	
H, H_2O subsystem, ac	7.7	12.1	3.8	6.0	13	20.4	9.1	14.3	7.9	12.4	
I, Power conditioning subsystem, ac	47.2	X	22.4	X	45.6	X	22.4	X	46.3	X	2
J, Control subsystem, ac	4	6.3	4	6.3	4	6.3	4	6.3	4	6.3	
Total	3000	X	1500	X	3000	X	1500	X	3000	X	1

Plant design	6				7				8		
Operation	Full power		Half power		Full power		Half power		Full power		Ha
A, kw	6000		3000		6000		3000		6000		
B, cps	60		60		400		400		60		
Quantity	kw	amp	kw	amp	kw	amp	kw	amp	kw	amp	k
C, H_2 subsystem, dc	5329	9060	2598.4	4470	5299	9000	2576.1	4440	5299	9000	257
D, H_2 subsystem, ac	32.4	50.8	32.4	50.8	53.1	83.1	53.1	83.1	53.1	83.1	5
E, NH_3 subsystem, ac resistive	0	0	0	0	0	0	0	0	0	0	
F, NH_3 subsystem, ac reactive	437.4	686	218.7	343	457.6	716	230.6	361	457.6	716	23
G, N_2 subsystem, ac	85	133	85	133	86	135	86	135	86	135	
H, H_2O subsystem, ac	24.9	39.0	18	28.3	13.6	21.3	6.8	10.7	13.6	21.3	
I, Power conditioning subsystem, ac	86.8	X	43.0	X	86.7	X	42.9	X	86.2	X	4
J, Control subsystem, ac	4.5	7.0	4.5	7.0	4.5	7.0	4.5	7.0	4.5	7.0	
Total	6000	X	3000	X	6000	X	3000	X	6000	X	3

ble 3.9-VIII.
ical power distribution.

3			4				5			
power	Half power		Full power		Half power		Full power		Half power	
00	1500		6000		3000		6000		3000	
0	60		60		60		60		60	
amp	kw	amp	kw	amp	kw	amp	kw	amp	kw	amp
4360	1157.9	2030	5202	8920	2556.5	4380	5299	9000	2576.1	4440
34.1	21.7	34.1	44.6	69.8	22.3	34.9	53.1	83.1	53.1	41.5
0	0	0	0	0	0	0	0	0	0	0
381	243.1	381	445.7	698	224.7	353	457.6	716	230.6	361
74	47	74	83	130	83	130	86	135	86	135
12.4	4.0	6.9	13.3	20.8	6.6	10.3	13.6	21.3	6.8	10.7
X	22.3	X	206.9	X	102.4	X	86.2	X	42.9	X
6.3	4	6.3	4.5	7.0	4.5	7.0	4.5	7.0	4.5	7.0
X	1500	X	6000	X	3000	X	6000	X	3000	X

8			9				10				11			
power	Half power		Full power		Half power		Full power		Half power		Full power		Half power	
00	3000		10,000		5000		10,000		5000		15,000		7500	
0	60		60		60		60		60		60		60	
amp	kw	amp	kw	amp	kw	amp	kw	amp	kw	amp	kw	amp	kw	amp
9000	2576.1	4440	8864	15260	4144.9	7810	8849	15190	4125.1	7760	13344	23200	6233.1	1215
83.1	53.1	83.1	84.2	132	84.2	132	108.8	173	108.8	173	152	238	152	238
0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
716	230.6	361	564.6	883	368.6	577	564.6	883	368.6	577	793.2	1240	519.2	814
135	86	135	316	496	316	496	316	496	316	496	476	744	476	744
21.3	6.8	10.7	21.1	33.1	10.6	16.6	21.1	33.1	10.6	16.6	29.8	467	14.9	23.4
X	42.9	X	144.9	X	70.5	X	135.3	X	65.7	X	202	X	98.8	X
7.0	4.5	7.0	5.2	8.1	5.2	8.1	5.2	8.1	5.2	8.1	6.0	9.4	6.0	9.4
X	3000	X	10000	X	5000	X	10000	X	5000	X	15000	X	7500	X

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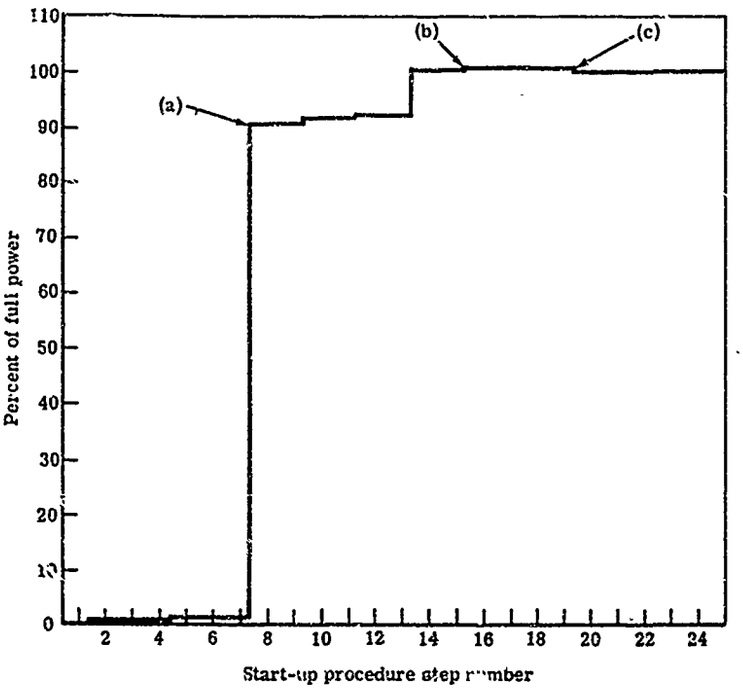
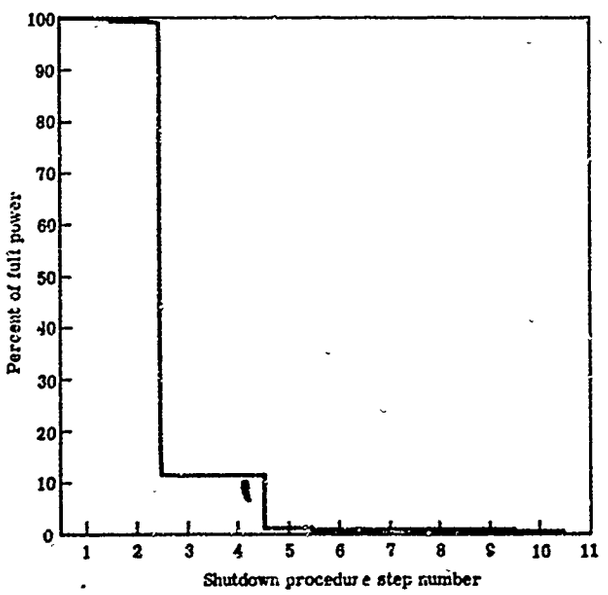


Figure 3.9-9. Typical start-up power demand curve.

Notes: (a) Power delivered to modules not stepped but gradual as discussed in Control Analysis section.
(b) When ammonia start-up heaters started, power decreased to modules to keep demand at 100%.
(c) When ammonia start-up heaters stopped, full power is restored to modules.

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Figure 3.9-10. Typical shutdown power demand curve.

The results which are based on the scheduled system maintenance requirements and the expected component failure rates, indicate that the systems are capable of operating at full production capacity from 95 to 97% of the time. In general, the availability was affected by the number of system component and package interconnections. Since systems 4 and 9 have duplicate hydrogen production and power conditioning subsystems, a failure in one of either of these subsystems does not incapacitate the entire system. Hence, systems 4 and 9 can operate at half-production capacity in the event failures are experienced in these duplicate subsystems.

Input Power Variations

General consideration was given to various system input power arrangements and their influences on the system yield, weight, and control. Analyses were made with the realization that the power generation and transmission equipment would require extensive modification to satisfy some of the conditions chosen. However, the major emphasis was placed upon defining any potential advantages or disadvantages to the fuel production plant. The following criteria were used along with engineering judgement in making the analyses.

- D-C equipment is larger and heavier than a-c equipment.
- 60 cps equipment is larger and heavier than 400 cps equipment.
- Power cable weights increase as current carried increases.
- D-C motors are slightly less efficient than a-c motors.

The reference power distribution design is shown in Figure 3.9-11. Major weight savings can be made in this design by eliminating the transformer, rectifier, and voltage regulator. These components can be eliminated if the power source provides d-c power with voltage regulation for the electrolyzer requirements and a-c power for the control and motor loads. Further weight reductions are possible if the a-c power frequency is increased from 60 cps to 400 cps. Another possible alternate is to operate the entire fuel production plant on d-c power. The various alternates are described in the following.

- Alternate A—one potentially attractive arrangement examined is similar to the reference design except the transformer is removed. The motors would operate at 4160 v ac, and the electrolyzer would be designed to operate at the resulting rectified d-c level (~5000 v dc). A large weight saving would result from the elimination of the transformer, but the electrolyzer design would be complicated considerably and significantly increased electrical insulation to structures would be required. Current leakage through electrolyte piping, connections, and structure would be difficult to prevent, and a cell failure would shut down the entire unit since the cells would be in series. Also, the rectifier peak-inverse-voltage (piv) would be too high for state-of-the-art devices, i.e., the piv would be about 8 kv, and the maximum allowable is presently about 1.6 kv. Consequently, this power approach was abandoned.

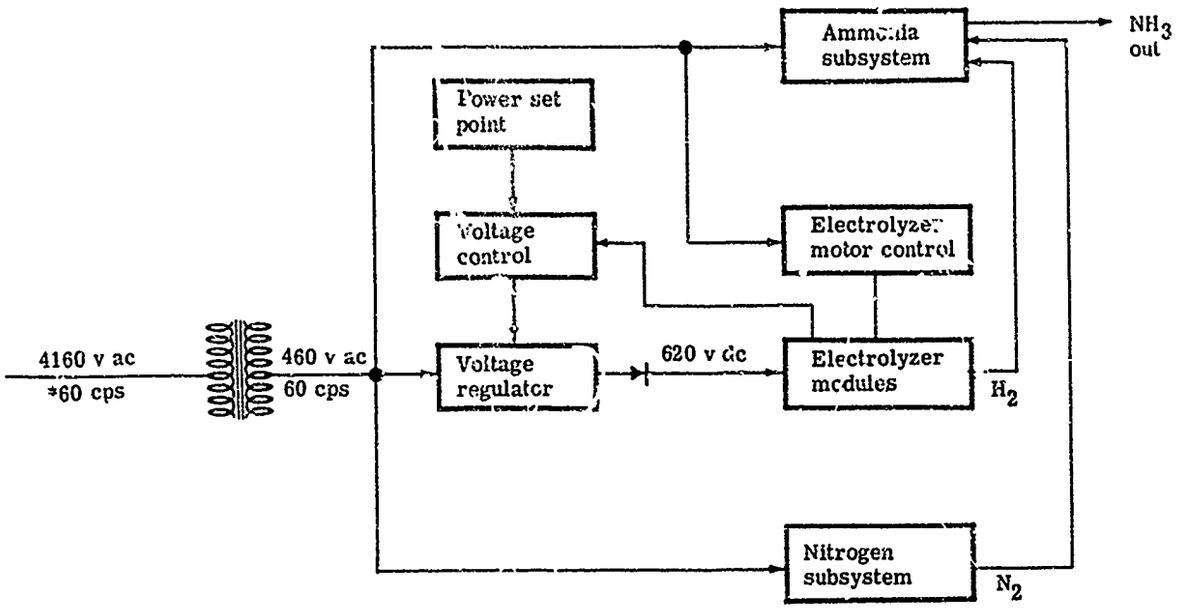


Figure 3.9-11. Simplified reference design system power arrangement.

- Alternate B—the alternate B system arrangement shown in Figure 3.9-12 realizes a weight savings in the fuel production system by eliminating the 4160 v/460 v transformer, and supplying 460 v power to the fuel production plant. Since the transformer is eliminated, transformer losses are eliminated. This results in a slightly more efficient system and, consequently, slightly higher yields (~1%). However, equipment external to the fuel production system, such as electrical power transmission cables and generating equipment, would be significantly affected.

  The method of control would remain the same as in the reference case.

- Alternate C—the system shown in Figure 3.9-13 calls for low voltage d-c power to be delivered to the fuel production system. Voltage regulation is provided by integrating the electrical power generator control and the fuel production system control. The transformer, rectifier, and voltage regulator would be eliminated resulting in a weight saving. Weights of motors will increase slightly. However, the overall effect would be to lower the fuel production system weight.

  Since transformer power losses would be eliminated and d-c motors are only slightly less efficient than a-c motors, the system yield should increase slightly (~1%).

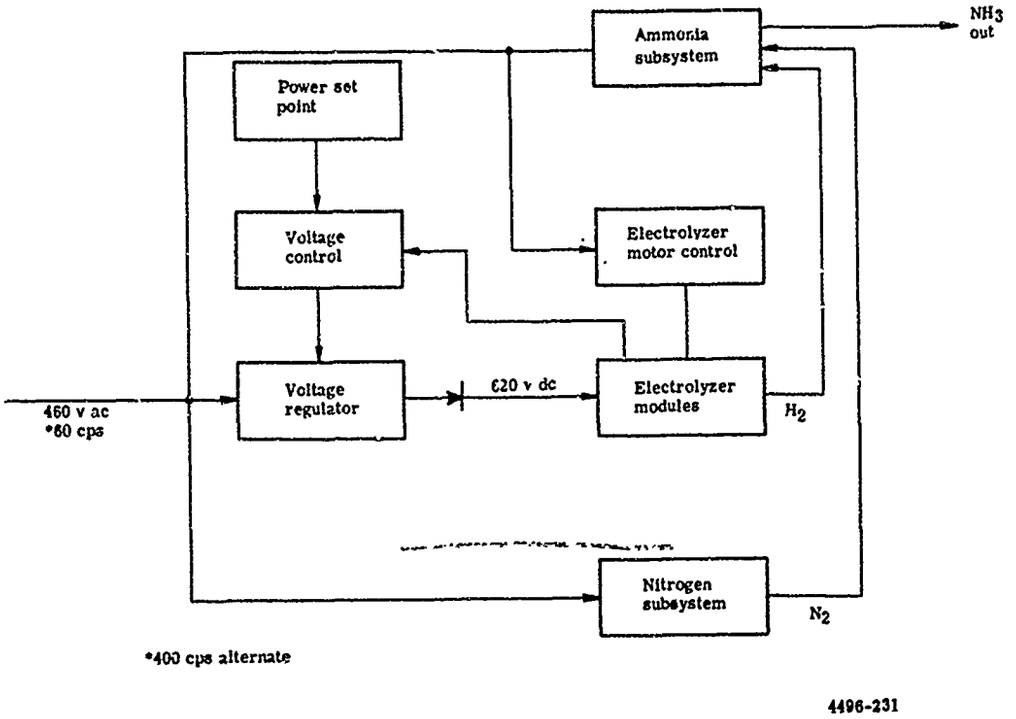


Figure 3.9-12. Simplified alternate B power arrangement.

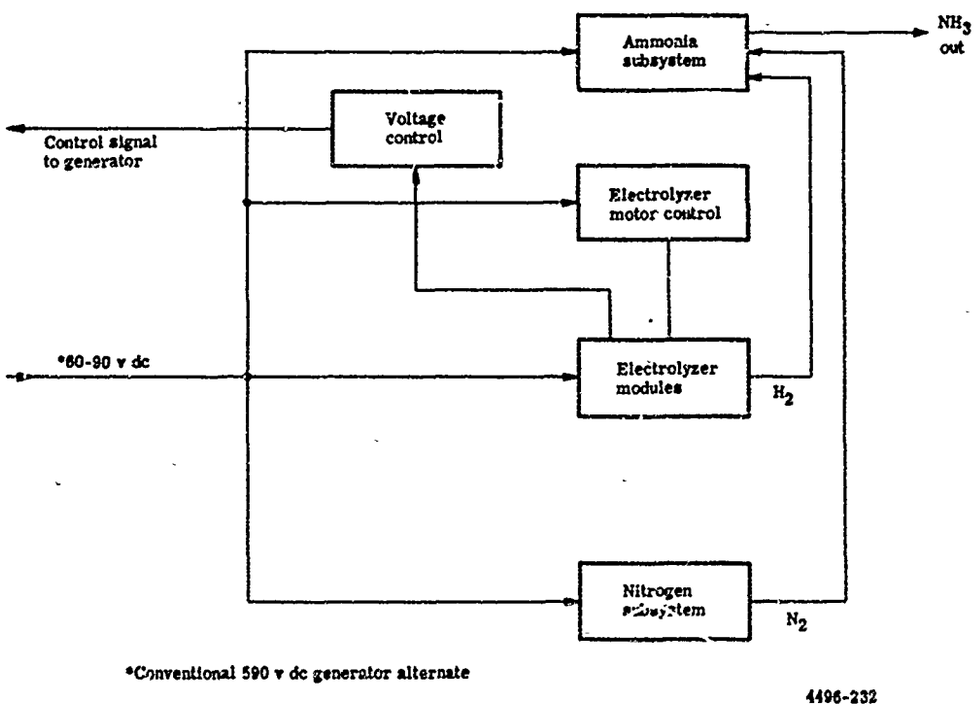


Figure 3.9-13. Simplified alternate C power arrangement.

This arrangement may introduce relatively severe problems into the power generating plant. These effects should be carefully assessed.

A possible alternate system would be to generate 590 v dc by means of conventional d-c generators. This would enable the electrolyzer module weight to remain unchanged and provide a higher level dc to other components, thereby reducing the size and weight of the components.

- Alternate D—this alternate is similar to the preceding alternate C. However, it is based on the power source generating both a-c and d-c power. The electrolyzer is supplied 60 to 90 v dc and motors and other subsystems are supplied with 460 v a-c, 60- or 400-cps power.

The advantages of this system are that transformers, rectifiers, and regulators are eliminated, and motor and component sizes and weights are minimized.

The fuel production system yield should be slightly higher and the weight lower than for the reference design.

It appears clear that generation of both a-c and d-c power is highly desirable from the standpoint of the fuel production system weight. However, it is emphasized that the weight changes in the powerplant generating and transmission equipment have not been included in the analysis. The fuel production system yield, however, is not affected significantly by variations in the power generation quality (~1% losses in transformer are eliminated).

Alternate D, or a similar variation, appears the most attractive arrangement from fuel production system considerations. However, full evaluation of these and other alternates will require an integrated study of the consequences in both the power generation and the fuel production plants of power quality variations.

DISCUSSION

The performance of the eleven ammonia fuel production systems was analyzed for design and nondesign point operation. The ammonia production rates were optimized based primarily upon the input power and the total cost of each system. Other design points could be obtained if other factors, such as weight, were considered more important and additional optimization work were done. For example, the electrodialysis process could be used for the water purification task instead of the vapor compression process selected. The resulting system would be slightly lighter and the ammonia production rate would increase slightly due to the difference in power requirements for the two water purification processes. However, the system cost would increase significantly, and this difference does not justify the slight gain in ammonia production.

The disadvantage of decreased system flexibility caused by coupling the operation of the nitrogen generation subsystem to the oxygen generated in the hydrogen production subsystem was discussed in the nitrogen generation subsystem section (3.2). An alternate nitrogen generation scheme could be substituted without serious difficulty if this disadvantage proves objectionable. Alternate schemes would require slightly more electrical power, and the system ammonia production rate would decrease slightly.

The performance of the systems could be improved at off-design conditions by oversizing the system heat exchangers. In general, for low ambient temperature operation, the ammonia production rate could be increased if more power were available. For high ambient temperature operation, the ammonia production rate could be increased if the heat exchangers were large enough to dissipate the waste heat generated.

The design of the ammonia synthesis subsystem was based on a hydrogen to nitrogen molar ratio of 3:1 in the ammonia synthesis loop. This results in a conservative design and the performance analysis maintains this conservatism. Since the ammonia synthesis subsystem limits system capability at most off-design conditions, it is anticipated that system output at off-design conditions will be somewhat higher than the performance analysis indicates. The effect of an optimization of the hydrogen to nitrogen molar ratio at the design point is illustrated in the control analysis section (see Figure 3.8-12).

CALCULATIONS

Multiplying factors were presented earlier for determining stream flow rates for systems 2 through 11 (Table 3.9-III). These factors were derived based on the hydrogen subsystem production rate achieved through the systems integration work. This was made possible since all the selected stream flow rates are proportional to the subsystem production rate. Thus, the multiplying factor for plant design 2 was obtained by calculating the ratio of the hydrogen production rate for plant design 2 to that for plant design 1, i.e.,

$$M.F._2 = \frac{(HPR)_2}{(HPR)_1}$$

$$= \frac{64.65}{64.55}$$

$$= 1.001$$

where

$M.F._2$ = multiplying factor for plant design 2
$(HPR)_2$ = hydrogen production rate for plant design 2
$(HPR)_1$ = hydrogen production rate for plant design 1

Similar calculations were performed to determine the remaining multiplying factors.

Curves were presented in the preceding Predicted Performance subsection which showed the effects on the production rate of oxygen contamination in the hydrogen product stream and hydrogen losses in the oxygen stream (Figure 3.9-5). These curves were obtained by means of an equation derived from material balances around the hydrogen subsystem and the deoxo unit. The derivation of the equation follows:

Let A = moles of oxygen generated
B = moles of hydrogen generated
C = total moles leaving oxygen side of liquid-gas separator
D = total moles leaving hydrogen side of liquid-gas separator

Then by a material balance,

$$A + B = C + D$$

Or, since 2 moles of hydrogen are generated per mole of oxygen,

$$C + D = 0.5\,B + B = 1.5\,B$$

The fraction of hydrogen in the leaving oxygen stream is defined as f_1, and the fraction of oxygen in the leaving hydrogen stream is defined as f_2. Then the moles of hydrogen in the oxygen stream is f_1C, and the moles of oxygen in the hydrogen stream is f_2D. The moles of water formed in the deoxo unit by combustion is defined as F, and the moles of hydrogen finally delivered to the ammonia subsystem is defined as E. The fraction of full hydrogen production, which is equivalent to the fraction of full ammonia production, is the ratio of the hydrogen leaving the deoxo unit to the hydrogen initially generated, i.e.,

$$\eta = \frac{E}{B}$$

The problem now is to express η in terms of f_1 and f_2, the fractions which describe the contamination of the two streams. A hydrogen balance around the system gives

$$B = C_h + E + F_h$$

where the subscript "h" refers to hydrogen, e.g., C_h means the moles of hydrogen in stream C. Rearranging this equation gives

$$\eta = \frac{E}{B} = 1 - \frac{C_h + F_h}{B}$$

or, since $C_h = f_1C$,

$$\eta = 1 - \frac{f_1C + F_h}{B}$$

Two moles of hydrogen are needed in the deoxo unit per mole of oxygen, so,

$$F_h = 2\, f_2 D$$

Substituting in the above equation gives

$$\eta = 1 = 1.5\, f_1 + \frac{D}{B}\,(f_1 - 2\, f_2)$$

By another material balance it is seen that

$$D = E - C_h + D_o$$

where D_o = moles of oxygen in stream D

Substituting previous identities into this equation yields

$$D = \frac{B\,(1 - 1.5\, f_1)}{1 - (f_1 + f_2)}$$

Inserting this equation into the equation given for η and simplifying gives the final working equation

$$\eta = (1 - 1.5\, f_1)\left[1 + \frac{(f_1 - 2\, f_2)}{1 - (f_1 + f_2)}\right]$$

This equation was used to develop the subject curves by inserting values for f_1 and f_2. Each parameter was varied from zero to 3%, i.e., the amount of hydrogen in the oxygen stream and the amount of oxygen in the hydrogen stream was calculated allowing the contamination of each stream to reach 3 mole %.

REFERENCE

1. Schade, C.W. and Thomson, B.N. Final Technical Report Water Electrolyzer Module Development, EDR 4610 (Unpublished).

Cost Analysis

DESCRIPTION

Cost data were developed for each of the eleven ammonia production systems. The cost data were broken down by subsystem for each of the systems analyzed. In addition, the fractional cost of each of the major components of the nitrogen generation, hydrogen production, and ammonia synthesis subsystems is furnished. Costs are of three types; development costs (excluding existing contracts); production costs in quantities of one, ten, and one-hundred per year; and operating costs. Transportation costs for each system were also developed for alternative means of transportation. The data are presented in a format which complies, as closely as possible, to that desired by Stanford Research Institute, the organization carrying out the Energy Depot cost-effectiveness evaluation.

The cost data are based on the system analysis and layout drawings prepared as a part of this contract and should not be construed to be firm quotations. Firm prices would require equipment specifications and detailed drawings. The cost data, however, were prepared with sufficient attention to details that it should provide an excellent basis for evaluation in the cost-effectiveness studies.

INVESTIGATION

Four distinct and different types of costs were desired for each of the eleven different ammonia production systems. To determine these costs as accurately as possible, each was divided into subunit costs, to the greatest extent possible. For example, manufacturing costs were determined for each subsystem component, for each subsystem assembly, and for overall system assembly. These were then accumulated for determination of the total plant cost. All costs were based, where possible, on firm vendor price estimates. In those cases where vendor estimates were not available, engineering estimates of costs were used.

Production Costs

The systems analysis effort carried out for each of the eleven ammonia production plants defined the operating characteristics, sizes, and weights of the subsystem components. These details of plant design provided the basis for determination of plant costs. Wherever possible, off-the-shelf components were selected for use; qualified vendors were furnished necessary design information and asked to submit cost estimates for subsystem components. Competitive vendors were contacted on many components to provide the best product at the least expense to the government. For those components for which insufficient data were available to allow a vendor to quote prices, engineering estimates were made based upon the best available information.

Estimates were prepared by shop personnel for necessary Allison production effort. These estimates included manpower, materiel, facilities, and tooling estimates, and were prepared in considerable detail as to numbers and type of personnel required. All furnished cost data were developed for each of three production levels: one per year, ten per year, and one-hundred per year. Ten years of production were assumed at each level. Different learning curves were used for different components, depending upon their complexity and existing experience factors.

All cost data were compiled and reviewed by Allison cost estimating personnel.

Development Costs

System development costs were determined for each subsystem of the ammonia production plant. These costs were further projected by fiscal year. For determination of development costs, a five-year development cycle was postulated. The development cycle, consisting of eleven program tasks, was detailed as nearly as possible in accordance with existing Army publications.[1] The development cycle, with time-phased program tasks was used to determine manpower and material estimates, which were developed for each program task. The development program included component development, system design, and fabrication and testing of a prototype plant. Cost data from vendors (e.g., the Girdler Corporation for the ammonia subsystem and Gas Equipment Company for the nitrogen subsystem) were used to the extent possible. In estimating development costs, no attempt was made to evaluate those costs accrued by Army personnel or facilities; for example, in service testing of the equipment. Cost estimates were included, however, for anticipated contractor support of the Army's evaluation and testing program.

The major portion of the development costs is expended for design, fabrication, and testing of a prototype plant. Use of off-the-shelf equipment, wherever possible, has limited the cost of the development program. The existing electrolyzer module design was incorporated into all hydrogen plants, thus limiting the development costs for the electrolyzer. The major subsystem requiring component development is the electrolyzer, and component development costs are included for the following activities:

- Development of lightweight components, including liquid-gas separators and controls
- Testing of these components
- Investigation of means for improvement of electrolyzer performance

Estimates were included in the development costs for necessary manpower to revise designs after prototype testing, and for required repair parts support during testing.

Operating Costs

Operating costs were estimated for each plant. These costs comprise operating and maintenance labor, maintenance materials, and required tools for maintenance personnel. No attempt was made to factor in personnel training costs or operator attrition, as this requires knowledge of Army experience factors not available to Allison. These operating costs are expressed on the basis of cents per pound of ammonia, considering 5000 hr of operation per year.

The basis for estimating operating labor was a requirement for two operators per shift, or a total of six assigned operators. The cost of these operators was based upon $11,000 per man per year. Maintenance labor requirements for an electrolyzer were developed in some detail as a part of Allison's current Water Electrolyzer Module Development Contract.[2] These maintenance data were used to develop maintenance labor requirements for each of the eleven hydrogen production subsystems. Gas Equipment Company furnished—based on their operating experience for nitrogen plants—a factor for maintenance manhours per hour of operation. Similar maintenance labor requirements were projected for other subsystems. Maintenance tools were estimated to cost $25,000 based upon a tool list projected by maintainability personnel. Maintenance materials and repair parts were estimated based upon 3% of the capital cost per year.

Transportation Costs

Allison, as a part of its independent program of analysis of the Energy Depot concept, developed transportation costs for military supplies by alternative means of transportation. These data, developed for transportation of POL, were modified to apply to transportation of the various Energy Depot plants. The following methodology or reference material was used in developing the transportation costs.

Military Truck

A contractual study was completed for Allison by Planning Research Corporation.[3] This study developed operation and maintenance costs of Army supply vehicles, excluding operators. The data were revised for the Energy Depot to utilize the M123C[4] for systems 1, 2, and 4; and the XM523/4-E1[5] for other configurations. Operator and depreciation costs were also added.

Rail

Rail shipment data were obtained for loads in excess of 30,000 lb to various points in the United States from Indianapolis. From these data points, cost per ton-mile curves were developed which were used to determine rail shipping costs.

Air

The Boeing Company, under contract to the government, developed aircraft transportation costs for various aircraft.[6] They also furnished MATS Industrial Fund rates for the C-130E. Their work developed comparable costs for both the C-130E and the new C-5A type aircraft on a $/ton-mile basis. Planning Research Corporation, in the study referenced under Military Truck, developed cost variations for cargo transport as a function of flight distance. These data were used to develop air transport costs.

Ship

In the Boeing study,[6] transport cost data and Industrial Fund rates were developed for the C4-S-1a (Mariner) class MSTS vessel. These data were used to establish ship transport costs.

PREDICTED COST

Tables 3.10-I through 3.10-XI present the cost summary for each of the eleven systems. Research, development, test, and evaluation costs of the contractor are shown itemized both by subsystem and by fiscal year. Production costs, itemized by subsystem, are presented for production levels of 1, 10, and 100 plants per year. Operating costs per pound of ammonia are furnished for each system.

Figure 3.10-1 gives a graphical summary of development cost as a function of ammonia production capability of the plant, and Figure 3.10-2 similarly presents capital costs as a function of ammonia production capability of the plant.

Table 3.10-XII shows the relative costs of major components of the subsystems. In each case the percent of total subsystem cost of the component(s) is shown. The percentages vary from system to system because of such things as the increased relative cost of power conditioning equipment for larger plants and the shifting of nitrogen compressors from components of the ammonia synthesis subsystem for smaller plants to components of the nitrogen generation subsystem for the larger plants.

Table 3.10-XIII presents a summary of transportation costs for various distances by alternative means of transportation.

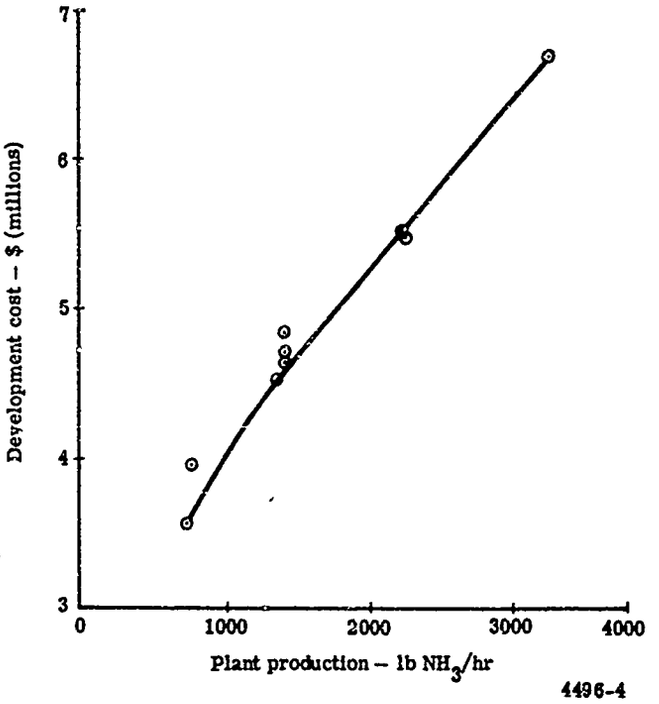


Figure 3.10-1. Development cost summary.

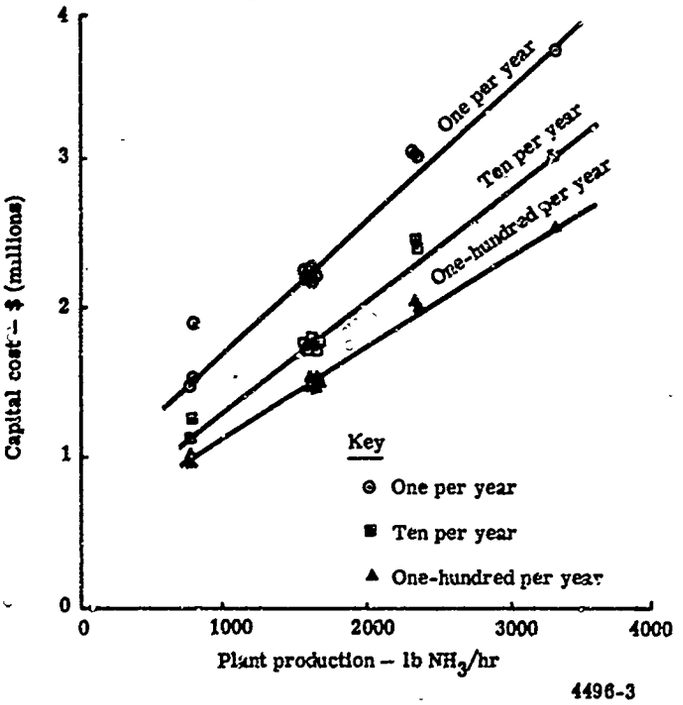


Figure 3.10-2. Capital cost summary.

Table 3.10-I.

Cost summary—system 1.*

Research, Development, Test, and Evaluation

	By subsystem		By fiscal year
Water purification	$ 31,890	FY 66	$ 70,000
Power conditioning	75,905	FY 67	695,000
Nitrogen generation	207,815	FY 68	1,625,000
Hydrogen production	2,337,600	FY 69	1,025,000
Ammonia synthesis	687,400	FY 70	160,000
Control	3,300		
		Total	$3,575,000
System integration	231,090		
Total	$3,575,000		

Production

	1/yr	10/yr	100/yr
Water purification	$ 31,890	$ 17,490	$ 14,735
Power conditioning	75,905	65,320	60,560
Nitrogen generation	152,195	129,360	122,430
Hydrogen production	902,580	677,675	544,905
Ammonia synthesis	314,000	276,000	248,000
Control	3,000	2,400	2,000
System integration	25,430	16,755	12,370
Total	$1,505,000	$1,185,000	$1,005,000

Operation

	¢/lb NH_3	$/yr
Operating labor	1.81	$ 66,000
Maintenance labor	0.15	5,500
Materials & repair parts	0.83	30,000
Total	2.79	$101,500

*3 Mw_e; Class 1 mobility; 729 lb NH_3/hr; air cooled; 60-cps power.

Table 3.10-II.
Cost summary—system 2.*

Research, Development, Test and Evaluation

	By subsystem		By fiscal year
Water purification	$ 32,600	FY 66	$ 70,000
Power conditioning	75,905	FY 67	695,000
Nitrogen generation	207,815	FY 68	1,635,000
Hydrogen production	2,319,175	FY 69	1,030,000
Ammonia synthesis	717,100	FY 70	160,000
Control	3,300		
		Total	$3,590,000
System integration	234,105		
Total	$3,590,000		

Production

	1/yr	10/yr	100/yr
Water purification	32,600	17,490	14,735
Power condition	75,905	65,320	60,560
Nitrogen generation	152,195	129,360	122,430
Hydrogen production	898,880	672,495	542,075
Ammonia synthesis	321,000	281,000	253,000
Control	3,000	2,400	2,000
System integration	31,420	21,935	15,200
Total	$1,515,000	$1,190,000	$1,010,000

Operation

	¢/lb NH_3	$/yr
Operating labor	1.81	$ 66,000
Maintenance labor	0.15	5,500
Materials & repair parts	0.83	30,300
Total	2.79	$101,800

*3 Mw_e; Class I mobility; 730 lb NH_3/hr; H_2O cooled; 60-cps power.

Table 3.10-III.
Cost summary—system 3.*

Research, Development, Test, and Evaluation

	By subsystem		By fiscal year
Water purification	$ 31,890	FY 66	$ 70,000
Power conditioning	82,080	FY 67	695,000
Nitrogen generation	207,815	FY 68	1,930,000
Hydrogen production	2,737,185	FY 69	1,115,000
Ammonia synthesis	687,400	FY 70	175,000
Control	3,300		
		Total	$3,985,000
System integration	235,330		
Total	$3,985,000		

Production

	1/yr	10/yr	100/yr
Water purification	31,890	17,490	14,735
Power conditioning	82,080	63,555	51,785
Nitrogen generation	152,195	129,360	122,430
Hydrogen production	1,305,050	1,002,755	817,060
Ammonia synthesis	314,000	276,000	248,000
Control	3,000	2,400	2,000
System integration	26,785	18,440	13,990
Total	$1,915,000	$1,510,000	$1,270,000

Operation

	¢/lb NH_3	$/yr
Operating labor	1.74	$ 66,000
Maintenance labor	0.16	6,000
Materials & repair parts	1.01	38,000
Total	2.91	$110,000

*3 Mw_e; Class II mobility; 758 lb NH_3/hr; air cooled; 60-cps power.

Table 3.10-IV.
Cost summary—system 4.*

Research, Development, Test, and Evaluation

	By subsystem		By fiscal year
Water purification	$ 40,490	FY 66	$ 70,000
Power conditioning	128,390	FY 67	695,000
Nitrogen generation	241,670	FY 68	2,335,000
Hydrogen production	2,713,925	FY 69	1,230,000
Ammonia synthesis	1,183,800	FY 70	190,000
Control	3,300		
		Total	$4,520,000
System integration	208,425		
Total	$4,520,000		

Production

	1/yr	10/yr	100/yr
Water purification	$ 40,490	$ 31,350	$ 25,545
Power conditioning	128,390	99,615	81,345
Nitrogen generation	182,970	156,735	148,995
Hydrogen production	1,316,585	1,011,670	824,315
Ammonia synthesis	518,000	450,000	405,000
Control	3,000	2,400	2,000
System integration	30,565	23,230	17,800
Total	$2,220,000	$1,775,000	$1,505,000

Operation

	¢/lb NH_3	$/yr
Operating labor	0.96	$ 66,000
Maintenance labor	0.10	7,000
Materials & repair parts	0.65	45,000
Total	1.71	$118,000

*6 Mw_e; Class I mobility; 1380 lb NH_3/hr; air cooled 60-cps power.

Table 3.10-V.
Cost summary--system 5.*

Research, Development, Test, and Evaluation

	By subsystem		By fiscal year
Water purification	$ 40,100	FY 66	$ 70,000
Power conditioning	128,390	FY 67	695,000
Nitrogen generation	251,670	FY 68	2,415,000
Hydrogen production	2,787,210	FY 69	1,250,000
Ammonia synthesis	1,174,000	FY 70	190,000
Control	3,300		
		Total	$4,620,000
System integration	235,330		
Total	$4,620,000		

Production

	1/yr	10/yr	100/yr
Water purification	$ 40,100	$ 31,110	$ 25,400
Power conditioning	128,390	99,615	81,345
Nitrogen generation	192,970	166,735	158,995
Hydrogen production	1,358,700	1,046,455	854,535
Ammonia synthesis	510,000	442,000	398,000
Control	3,000	2,400	2,000
System integration	31,840	21,685	14,725
Total	$2,265,000	$1,810,000	$1,535,000

Operation

	¢/lb NH_3	$/yr
Operating labor	0.94	$ 66,000
Maintenance labor	0.10	7,000
Materials & repair parts	0.65	46,000
Total	1.69	$119,000

*6 Mw_e: Class II mobility; 1410 lb NH_3/hr; air cooled; 60-cps power.

Table 3.10-VI.
Cost summary—system 6.*

Research, Development, Test, and Evaluation

	By subsystem		By fiscal year
Water purification	$ 41,050	FY 66	$ 70,000
Power conditioning	128,390	FY 67	695,000
Nitrogen generation	251,670	FY 68	2,430,000
Hydrogen production	2,765,930	FY 69	1,255,000
Ammonia synthesis	1,214,500	FY 70	195,000
Control	3,300		
		Total	$4,645,000
System integration	240,160		
Total	$4,645,000		

Production

	1/yr	10/yr	100/yr
Water purification	$ 41,050	$ 31,850	$ 26,010
Power conditioning	128,390	99,620	81,350
Nitrogen generation	192,970	166,735	158,995
Hydrogen production	1,331,730	1,025,525	837,435
Ammonia synthesis	505,000	432,000	389,000
Control	3,000	2,400	2,000
System integration	32,860	26,870	20,210
Total	$2,235,000	$1,785,000	$1,515,000

Operation

	¢/lb NH_3	$/yr
Operating labor	0.93	$ 66,000
Maintenance labor	0.10	7,000
Materials & repair parts	0.64	45,500
Total	1.67	$118,500

*6 Mw_e: Class II mobility: 1420 lb NH_3/hr; H_2O cooled; 60-cps power.

Table 3.10-VII.
Cost summary—system 7.*

Research, Development, Test, and Evaluation

	By subsystem		By fiscal year
Water purification	$ 40,490	FY 66	$ 70,000
Power conditioning	56,200	FY 67	795,000
Nitrogen generation	241,670	FY 68	2,435,000
Hydrogen production	2,788,795	FY 69	1,295,000
Ammonia synthesis	1,463,000	FY 70	230,000
Control	3,300		
		Total	$4,825,000
System integration	231,545		
Total	$4,825,000		

Production

	1/yr	10/yr	100/yr
Water purification	$ 40,490	$ 31,415	$ 25,650
Power conditioning	56,200	43,605	35,605
Nitrogen generation	182,970	156,735	148,995
Hydrogen production	1,360,225	1,047,640	855,500
Ammonia synthesis	527,200	455,400	411,000
Control	3,000	2,400	2,000
Syst. integration	34,915	27,805	21,250
Total	$2,205,000	$1,765,000	$1,500,000

Operation

	¢/lb NH_3	$/yr
Operating labor	0.94	$ 66,000
Maintenance labor	0.10	7,000
Materials & repair parts	0.64	45,000
Total	1.68	$118,000

*6 Mw_e; Class II mobility; 1410 lb NH_3/hr; air cooled; 400 cps power.

Table 3.10-VIII.
Cost summary—system 8.*

Research, Development, Test, and Evaluation

	By subsystem		By fiscal year
Water purification	$ 40,490	FY 66	$ 70,000
Power conditioning	128,390	FY 67	695,000
Nitrogen generation	241,670	FY 68	2,410,000
Hydrogen production	2,797,040	FY 69	1,255,000
Ammonia synthesis	1,174,000	FY 70	190,000
Control	3,300		
		Total	$4,620,000
System integration	235,110		
Total	$4,620,000		

Production

	1/yr	10/yr	100/yr
Water purification	$ 40,490	$ 31,415	$ 25,650
Power conditioning	128,390	99,615	81,345
Nitrogen generation	182,970	156,735	148,995
Hydrogen production	1,367,455	1,053,245	860,070
Ammonia synthesis	510,000	442,000	398,000
Control	3,000	2,400	2,000
System integration	32,695	24,590	18,940
Total	$2,265,000	$1,810,000	$1,535,000

Operation

	¢/lb NH_3	$/yr
Operating labor	0.94	$ 66,000
Maintenance labor	0.10	7,000
Materials & repair parts	0.65	46,000
Total	1.69	$119,000

*6 Mw_e: Class III mobility; 1410 lb NH_3/hr; air cooled; 60-cps power.

Table 3.10-IX.
Cost summary—system 9.*

Research, Development, Test, and Evaluation

	By subsystem		By fiscal year
Water purification	$ 48,600	FY 66	$ 70,000
Power conditioning	213,720	FY 67	695,000
Nitrogen generation	318,950	FY 68	3,020,000
Hydrogen production	3,270,050	FY 69	1,435,000
Ammonia synthesis	1,352,900	FY 70	225,000
Control	3,300		
		Total	$5,445,000
System integration	237,480		
Total	$5,445,000		

Production

	1/yr	10/yr	100/yr
Water purification	$ 48,600	$ 37,785	$ 30,920
Power conditioning	213,720	166,170	136,000
Nitrogen generation	253,230	220,430	209,055
Hydrogen production	1,831,895	1,423,720	1,161,555
Ammonia synthesis	629,000	556,000	500,000
Control	3,000	2,400	2,000
System integration	45,555	33,495	25,470
Total	$3,025,000	$2,440,000	$2,065,000

Operation

	¢/lb NH_3	$/yr
Operating labor	0.58	$ 66,000
Maintenance labor	0.07	8,000
Materials & repair parts	0.54	62,000
Total	1.19	$136,000

*10 Mw_e; Class II mobility; 2280 lb NH_3/hr; air cooled; 60-cps power.

Table 3.10-X.
Cost summary—system 10.*

Research, Development, Test, and Evaluation

	By subsystem		By fiscal year
Water purification	$ 48,600	FY 66	$ 70,000
Power conditioning	248,940	FY 67	695,000
Nitrogen generation	318,950	FY 68	3,020,000
Hydrogen production	3,225,520	FY 69	1,430,000
Ammonia synthesis	1,352,900	FY 70	220,000
Control	3,300		
		Total	$5,435,000
System integration	236,790		
Total	$5,435,000		

Production

	1/yr	10/yr	100/yr
Water purification	$ 48,600	$ 37,785	$ 30,920
Power conditioning	248,940	193,550	158,400
Nitrogen generation	253,230	220,430	209,055
Hydrogen production	1,781,885	1,378,755	1,124,285
Ammonia synthesis	639,000	566,000	510,090
Control	3,000	2,400	2,000
System integration	45,345	36,080	25,340
Total	$3,920,000	$2,435,000	$2,060,000

Operation

	¢/lb NH_3	$/yr
Operating labor	0.58	$ 66,000
Maintenance labor	0.07	8,000
Materials & repair parts	0.54	62,000
Total	1.19	$136,000

*10 Mw_e; Class III mobility; 2290 lb NH_3/hr; air cooled; 60-cps power.

Table 3.10-XI.
Cost summary—system 11.*

Research, Development, Test, and Evaluation

	By subsystem		By fiscal year
Water purification	$ 54,300	FY 66	$ 70,000
Power conditioning	306,010	FY 67	695,000
Nitrogen generation	389,770	FY 68	3,950,000
Hydrogen production	3,509,420	FY 69	1,695,000
Ammonia synthesis	2,164,700	FY 70	260,000
Control	3,300		
		Total	$6,670,000
System integration	242,500		
Total	$6,670,000		

Production

	1/yr	10/yr	100/yr
Water purification	$ 54,300	$ 29,630	$ 25,440
Power conditioning	306,010	263,380	244,205
Nitrogen generation	317,610	279,630	265,650
Hydrogen production	2,063,480	1,586,925	1,266,580
Ammonia synthesis	947,000	822,000	740,000
Control	3,000	2,400	2,000
System integration	53,600	36,035	31,125
Total	$3,745,000	$3,020,000	$2,575,000

Operation

	¢/lb NH_3	$/yr
Operating labor	0.40	$ 66,000
Maintenance labor	0.07	12,000
Materials & repair parts	0.47	77,000
Total	0.94	$155,000

*15 Mw_e; Class III mobility; 3300 lb NH_3/hr; air cooled; 60-cps power.

Table 3.10-XII.

Major component cost as percentage of subsystem cost.

Ammonia Synthesis Subsystem	
Compressor, inter- and aftercooler	30-40%
Nitrogen Generation Subsystem	
Heat exchangers	16-18%
Turbo expanders	7-16%
Compressors, motors, and blowers	7-21%
Hydrogen Production, Water Purification, and Power Conditioning	
Modules	69-81%
Power conditioning	6-18%
Liquid-gas separators	2-4 %

Table 3.10-XIII.

Transportation costs (dollars).*

	System										
	1	2	3	4	5	6	7	8	9	10	11
	$	$	$	$	$	$	$	$	$	$	$
Ten mile distance											
Military truck	140	140	280	230	290	280	280	290	480	380	480
Hundred mile distance											
Military truck	1,140	1,140	2,400	1,900	2,400	2,400	2,400	2,400	4,000	3,200	4,000
Rail	800	780	980	1,250	1,330	1,290	1,170	1,340	2,120	2,070	2,880
Air	1,130	1,100	1,360	1,770	1,890	1,820	1,860	1,900	3,000	2,930	4,080
Thousand mile distance											
Rail	3,600	2,500	3,100	4,100	4,300	4,200	3,800	4,400	6,900	6,700	9,400
Air	8,500	8,300	10,200	13,300	14,200	13,700	12,600	14,300	22,500	22,000	30,600
Ship	1,020	970	1,200	1,570	1,670	1,610	1,460	1,680	2,650	2,590	3,600

*Dollars/move.

DISCUSSION

Plant Size

A review of the cost data outlined in Tables 3.10-I through 3.10-XI portrays an incentive to select for development as large a plant as practical within the Army's mobility constraints. The following tabulation (Table 3.10-XIV) of specific costs of the fuel plant (not including the reactor

powerplant) for both development and production for systems 1, 6, 10, and 11 illustrates this point.

Table 3.10-XIV.
Specific cost (thousand \$/lb NH_3/hr production capability).

System	R, D, T & E*	Production (100/yr)
1	4.9	1.4
6	3.3	1.1
10	2.4	0.9
11	2.0	0.8

*Research, development, test, and evaluation.

The cost optimization data presented in Supplement I, Figures 1 through 4, which include the estimated influence of the powerplant, further reinforce this conclusion.

60- Versus 400-cps Power

At any significant production level, it appears to be economically advantageous to develop a 400-cps system. In the one case considered (system 5 compared with system 7), the additional development cost of approximately \$205,000 which would be incurred by developing a 400-cps system, would be recovered after production of only 5 units, considering a production level of 10 units per year. There is, in addition, a significant size and weight advantage described in the Data Presentation section (3.11) of this report.

Mobility Constraints

The mobility constraints do not significantly affect the ammonia output design point of the eleven systems analyzed. The electrical power input is the primary constraint on the output of the optimized systems discussed in the Supplement to this report. (It should be recognized, however, that the mobility constraints imposed in system 4 and system 9 are relieved by utilizing five skids.)

Component Costs

Table 3.10-XII points up the relative significance of the major components of the subsystems. It can be seen that the electrolyzer modules constitute a major cost of the total system. Accounting for approximately 70 to 80% of the hydrogen production subsystem cost, they therefore amount to almost one-half of the total system cost. The other important component cost is the

ammonia compressor and intercooler, accounting for 30 to 40% of the ammonia synthesis subsystem cost, or approximately 10 to 15% of total system cost.

Potential Cost Reductions

A number of areas for potential system cost reduction exist. The following are some of the more important of these.

- Further inclusion of automated machinery in production line processing of the electrolyzer module.
- Development of a new ammonia compressor for systems 9, 10, and 11 in lieu of multiple off-the-shelf units.
- Development of a lightweight centrifugal nitrogen compressor.
- Development of a larger electrolyzer module size.

Electrolyzer module costs were determined by applying learning curves to data based on current electrode-processing procedures. Current processing techniques, though they incorporate a considerable number of labor-saving measures, would be automated for production levels of 100 plants per year. It is estimated that final production cost estimates for the entire system could be reduced by 15% at 100 per year production levels by incorporation of increased automation in production.

Current cost estimates for systems 9, 10, and 11 include multiple ammonia compressors of existing design, two for systems 9 and 10, and three for system 11. There would appear to be a cost incentive to develop a large single compressor for either of these systems. Joy Manufacturing Company estimated a development program cost of approximately $1 million for such a large compressor including pattern and tooling costs. Production cost estimates indicate that this investment should be recovered after production of 10 to 25 units based on anticipated savings in capital cost. There thus appears an incentive to develop a new compressor for the three larger systems.

The current design for the nitrogen generation subsystem was based on the use of existing off-the-shelf reciprocating nitrogen compressors. There appears to be an incentive to develop a lightweight centrifugal compressor for this service. Mechanical Technology of Latham, New York, estimated, as an example, that a 50,000-rpm centrifugal compressor could be developed for systems 4 through 8 at a cost of $35,000. In production, they estimate a cost of $2500 per unit. Assuming a reasonable gearbox cost, this is approximately 60% of the current compressor cost.

As a final example, for the larger size systems, a revised electrolyzer module design could prove to be more economical. An increased electrode size would allow economy in processing of electrodes and would decrease assembly costs for the module.

All of the above potential cost reductions developed during the course of the analysis of the eleven system designs. Once a system design point is selected, other economies might well be revealed during a value engineering analysis of the specific design selected.

CALCULATIONS

The detailed calculations and accounting used in arriving at the final cost data cannot be reduced to several pages of computation methods. An example of the derivation of learning curves used to estimate price of individual components under varying quantity production conditions, is illustrated in the following.

Learning Curves

The method used to arrive at price variations as a function of quantity for components of the subsystems (allowing subsystem costs to be accumulated) was to use learning curves.* The learning curve formula is

$$C_A = C_F Q^n$$

where

C_A = average cost per unit for quantity Q
C_F = cost of first unit
Q = quantity produced
n = learning curve exponent

The learning curve exponent, n, is representative of the cost reduction which results from each doubling of production, that is:

$$n = \frac{\log P}{\log 2}$$

where P is the ratio of the cumulative average price with doubling of production to the average price with the reference production quantity. For example, a 90% learning curve (P = 0.9) would be represented by

*Because the total system cost is an accumulation of many component costs with different learning curves, it will not follow this learning curve formulation.

Quantity produced (Q)	Cumulative average price per unit (C_A) - [% of C_F]
1	100
2	90
4	81
8	72.9
16	65.6
32	59

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3. A Cost Study of POL Logistics in the Congo. Planning Research Corporation. 3 December 1963.

4. Army truck, tractor, 10 ton, 6 X 6, carries 35,000 lb payload for highway operations (TM9-500, Ordnance Corps Equipment Data Sheets, pages 21-99).

5. ATAC developed heavy equipment transporter described in Trackporter. U.S. Army Tank Automotive Center. April 1963.

6. Evaluation Study of Transport Vehicles. The Boeing Company. 17 November 1964.

Data Presentation

DESCRIPTION

Pertinent major system design point data are presented in this subsection in parametric form. The parametrics demonstrate the interdependence of significant system variables and illustrate the effects of making trade-offs between the variables. The parametrics were also developed to provide information in a convenient form for operational and cost effectiveness studies.

INVESTIGATION

The most important single item obtained from the present study was the maximum ammonia yield resulting from the optimization work for a given total input power, system weight, cost, and response time. Consequently, the interdependence of these variables was related graphically. In addition, a computer program was written to describe the relationships between variables.

The data required to perform the parametric study were obtained from the results presented in the Integration and Packaging, the Performance Analysis, and the Cost Analysis subsections.

PREDICTED PERFORMANCE

Graphical Presentations

Production Rate as a Function of Input Power

The functional dependence of the ammonia production rate on system total input power is shown in Figure 3.11-1. For convenience, and to conform to later figures, the ammonia production rate is plotted on the abscissa, and the total input power to the system is plotted on the ordinate. Data from all eleven systems are plotted and grouped according to system similarities, i.e., the two water-cooled systems were treated as one group, the three mobility class systems were treated as three individual groups, and the 400-cps system was treated as a separate entity.

The data in Figure 3.11-1 show that the differences between system constraints have a relatively small effect on the ammonia production rate. For example, at the 3-Mw_e input power level, the largest difference in production between system is about 4%. Similarly, at the 6-Mw_e level, the largest difference is about 3%.

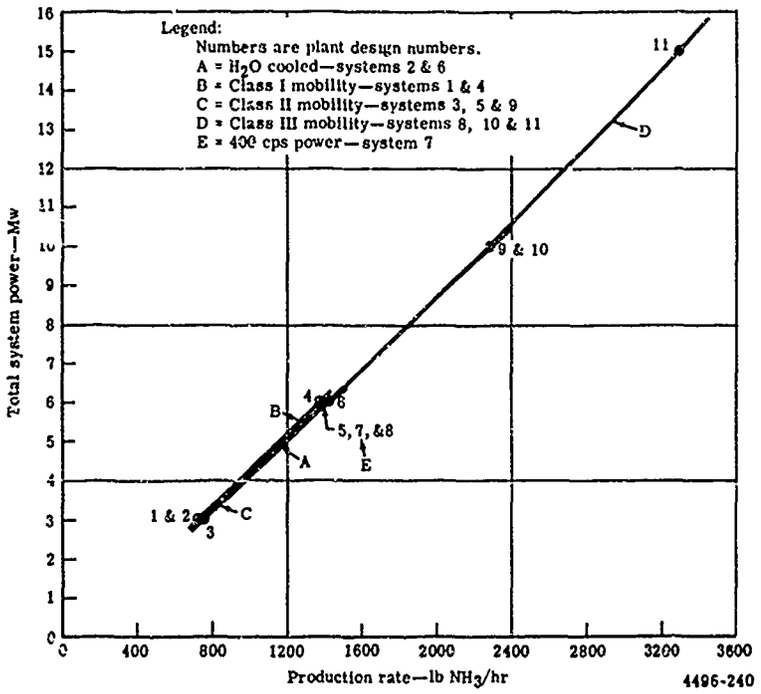


Figure 3.11-1. Ammonia production rate as a function of total system power.

System Weight as a Function of Production Rate

The relationship between ammonia production rate and fuel production system weight is shown in Figure 3.11-2. The data are plotted and grouped as described for Figure 3.11-1.

The system 3 point in Figure 3.11-2 requires some clarification. The total system weight for system 3 was determined based on preliminary cost optimization data. Final design data suggest that system 3 should be designed identically to system 1. The point in Figure 3.11-2 representing system 3 would then be coincident with the point for system 1, and the shape of curve C would be altered. The dashed curve in Figure 3.11-2 represents this alteration.

The data in Figure 3.11-2 indicate that there is a significant weight advantage in using 400-cps power as illustrated by point E for system 7. The weight of fuel production system 7 is 12% less than the comparable 60-cps system 5.

The data also indicates that water-cooled systems are slightly lighter than comparable air-cooled systems. The difference is approximately 5%, as illustrated by comparing systems 1 and 2 and systems 5 and 6.

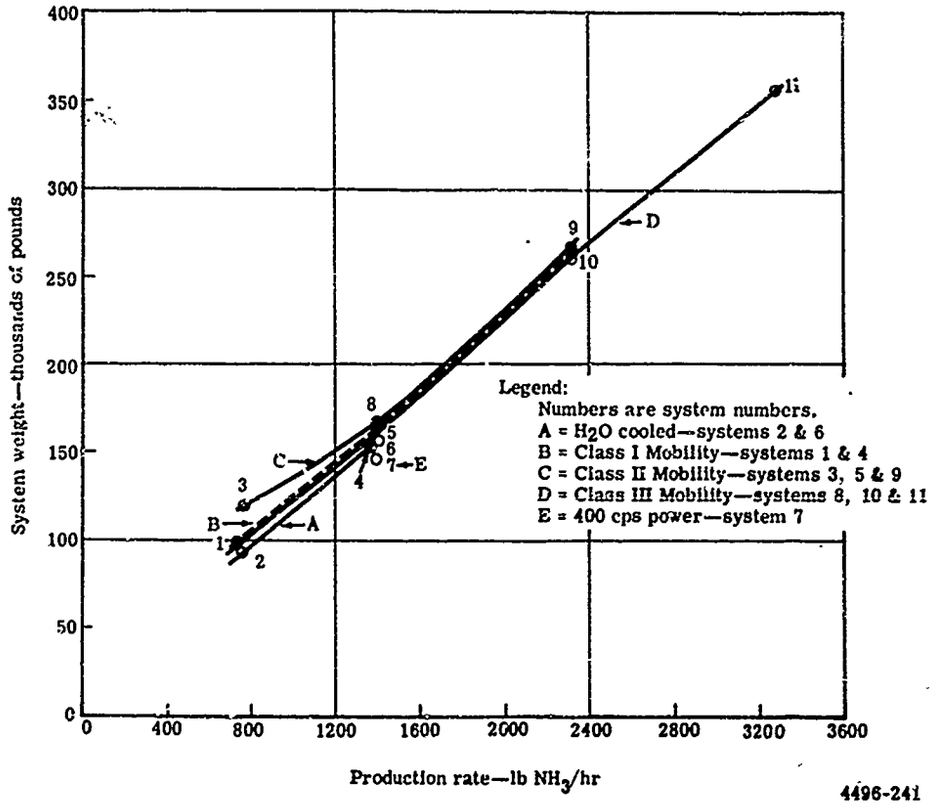


Figure 3.11-2. Fuel production system weight as a function of ammonia production rate.

The variations in mobility classes appear to have little influence on total system weight. Where different mobility class data overlap, there is only a slight difference in system weight between the mobility classes. This is illustrated by curve A and the dashed curve, and the dashed curve and curve D where the curves overlap for a range of ammonia production rates.

System Cost as a Function of Production Capacity

Fuel production system cost is plotted in Figure 3.11-3 as a function of the ammonia production capacity. Data are presented for the cost of producing one, ten, and one-hundred of each of the eleven systems per year. The grouping of data points is the same as discussed earlier for Figure 3.11-1.

As in the case of system weight, system 3 was designed using preliminary cost optimization data. The final design data suggest that system 3 should be coincident with system 1. The dashed curve in Figure 3.11-3 represents this alteration.

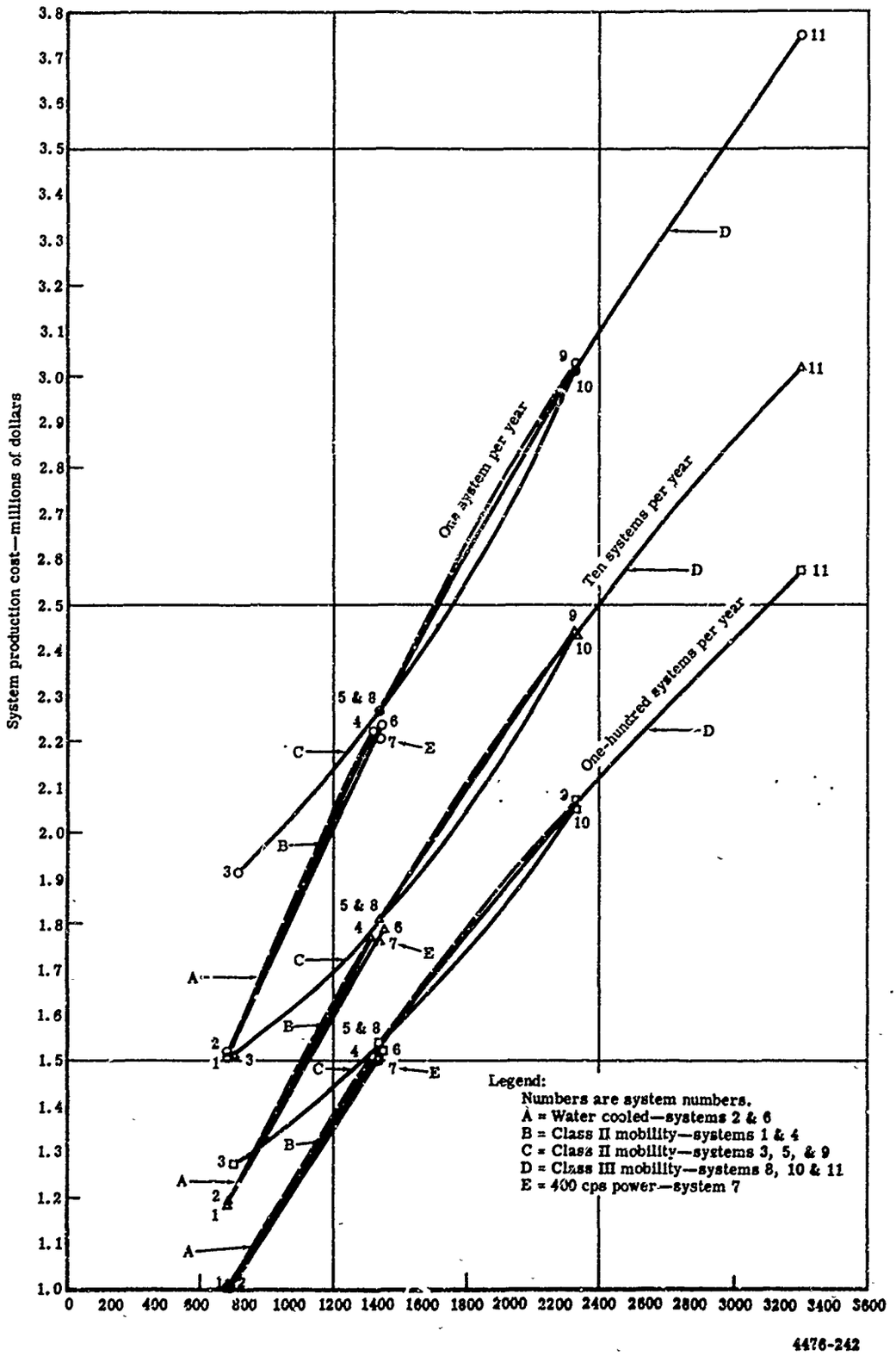


Figure 3.11-3. Total system production cost as a function of ammonia production rate and number of systems produced.

The data in Figure 3.11-3 show that the use of 400-cps power (system 7, point E) results in a slightly lower system cost than similar capacity 60-cps power systems (e.g., systems 4, 5, 6, and 8). The difference, however, is small, i.e., about 2%.

An analysis of the data similar to that discussed for system weight suggests that mobility class variations have little effect on system costs. This is again illustrated by comparing curve A and the dashed curve, and the dashed curve and curve D where the respective curves overlap for a range of ammonia production rates.

Comparing the data for the air-cooled system 1 and the water-cooled system 2 shows that the air-cooled system costs less than the water-cooled system. However, at a higher production rate the most advantage is reversed, i.e., system 6 (water-cooled) costs less than system 5 (air-cooled). However, the cost differences are small, viz, about 1%.

Response Time as a Function of Production Rate and System Weight

System response times are plotted in Figure 3.11-4 as a function of ammonia production rate. These data show that, in general, the water-cooled systems have longer response times than the comparable air-cooled systems, and the greater the number of system packages, the longer the response time. The latter effect is demonstrated by the large difference in response time between systems 4 (five Class I packages) and system 5 (three Class II packages). For this case, the difference in response time is about 22%.

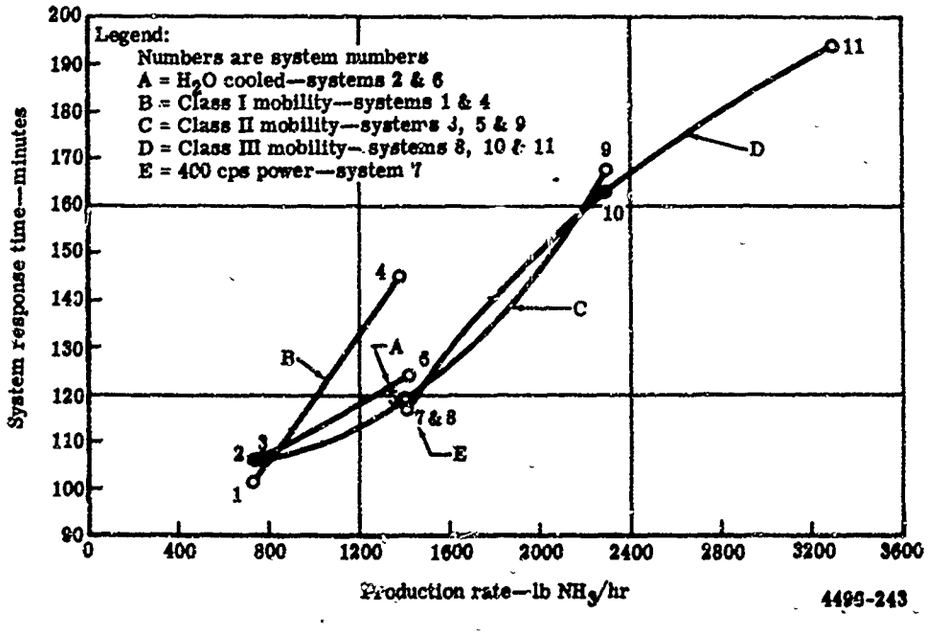


Figure 3.11-4. System response time as a function of ammonia production rate.

System response times are plotted as a function of the total system weight in Figure 3.11-5. In general, the relationship between response time and total system weight is similar to the relationship between response time and production rate discussed previously.

Computer Presentation

The preceding parametric data were used to write a digital computer program to describe the relationships between the selected major variables. The program listing is given in Table 3.11-I. The program is capable of handling 25 systems per computer run. The computer input and output parameters for system 1 are given in Table 3.11-II. The digital computer program requires the following pieces of input information:

- Mobility class
- Electrical power frequency
- System cooling fluid (air is denoted by a ϕ and water is denoted by a 1)
- Total system input power

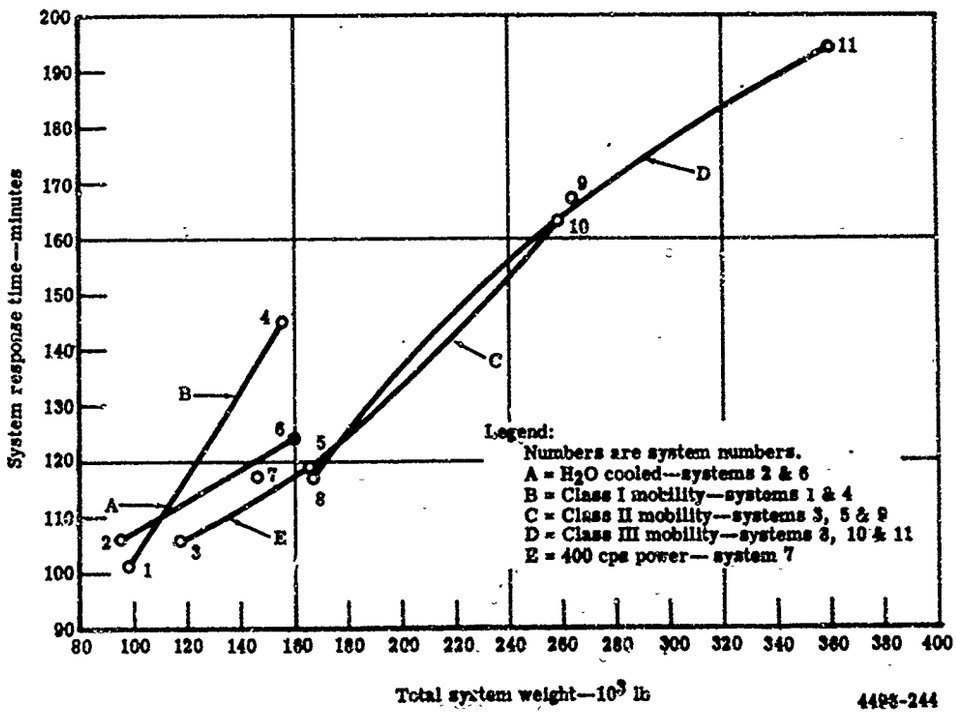


Figure 3.11-5. Relationship between total system weight and response time.

Table 3.11-I.

Digital computer program listing for computerized system parametrics.

```
      DIMENSION IC(25),JF(25),IS(25),P(25)                                  10
    5 READ 10,J                                                             20
   10 FORMAT (I2,I3,I2,E10.0)                                               30
      DO 20 I=1,J                                                           40
   20 READ 10,IC(I),JF(I),IS(I),P(I)                                        50
      PRINT 30                                                              60
   30 FORMAT(1X6HRUN NO3X8HMODEL17/3X4HFREQ3X7HCOOLING3X5HPOWER)            70
      PRINT 40                                                              80
   40 FORMAT(1X5HCLASS5X3HCPS5X5HFLUID5X2HMW/)                              90
      DO 50 I=1,J                                                           100
   50 PRINT 60,I,IC(I),JF(I),IS(I),P(I)                                     110
   60 FORMAT(2X13,7X12,6X14,6X12,5XF6.2)                                    120
   62 PRINT 63                                                              130
   63 FORMAT(///)                                                           140
      PRINT 64                                                              150
   64 FORMAT(2X3HRUN3X5HYIELD8X2HWT6X8HCOST 1/Y5X4H10/Y7X5H100/Y7X3HRES)    160
      PRINT 65                                                              170
   65 FORMAT(2X3HNO.3X4H#/HR8X4HLBS.8X1H$11X1H$10X1H$6X8HTIME MIN/)         180
      K=0                                                                   190
      I=1                                                                   200
   70 IF(J-I) 340,80,80                                                     210
   80 IF(JF(I)-60) 90,90,220                                                220
   90 IF(IS(I)-1) 100,170,170                                               230
  100 IF(IC(I)-2) 110,130,150                                               240
  110 PR=217.0*P(I)+78.0                                                    250
      IF(P(I)-3.0) 117,120,115                                              260
  115 IF(P(I)-6.0) 120,120,117                                              270
  117 K=1                                                                   280
  120 W=87.09*PR+34800.0                                                    290
      CH=768.0*PR+445100.0                                                  300
      CT=906.3*PR+524300.0                                                  310
      CO=1098.0*PR+704300.0                                                 320
      RT=0.06759*PR+51.73                                                   330
      GO TO 260                                                             350
  130 PR=220.000*P(I)+90.0                                                  360
      IF(P(I)-3.0) 137,140,135                                              370
  135 IF(P(I)-10.0) 140,140,137                                             380
  137 K=1                                                                   390
  140 W=0.007655*PR*PR+83.03*PR+33690.0                                     400
      CH=-.1127*PR*PR+1019.0*PR+321700.0                                    410
      CT=-.1293*PR*PR+1154.0*PR+383000.0                                    420
      CO=-.1616*PR*PR+1461.0*PR+525200.0                                    430
      RT=0.00001801*PR*PR-0.01209*PR+100.2                                  440
      GO TO 260                                                             460
  150 PR=211.33*P(I)+150.0                                                  470
      IF(P(I)-6.0) 157,160,155                                              480
  155 IF(P(I)-15.0) 160,160,157                                             490
  157 K=1                                                                   500
  160 W=-0.002405*PR*PR+113.4*PR+11820.0                                    510
      CH=-0.04586*PR*PR+766.3*PR+545700.0                                   520
      CT=-0.06932*PR*PR+963.7*PR+684700.0                                   530
      CO=-0.07414*PR*PR+1132.0*PR+815800.0                                  540
      RT=-0.00001141*PR*PR+0.09451*PR+6.428                                 550
      GO TO 260                                                             570
  170 IF(IC(I)-3)  175,303,303                                              575
  175 IF(IC(I)-2) 180,200,200                                               580
  180 PR=218.333*P(I)+75.1                                                  590
      IF(P(I)-3.0) 185,190,185                                              600
```

Table 3.11-I. (cont)

```
185 K=1                                                                  610
190 W=96.25*PR+24660.0                                                   620
    CH=785.0*PR+634000.0                                                 630
    CT=921.4*PR+513000.0                                                 640
    CO=1131.0*PR+684000.0                                                650
    RT=0.06769*PR+56.59                                                  660
    GO TO 260                                                            680
200 PR=225.0*P(I)+70.0                                                   690
    IF(P(I)-6.0) 205,210,205                                             700
205 K=1                                                                  710
210 W=97.5*PR+18700.0                                                    720
    CH=758.3*PR+416000.0                                                 730
    CT=887.5*PR+500000.0                                                 740
    CO=1100.0*PR+640000.0                                                750
    RT=0.03618*PR+72.625                                                 760
    GO TO 260                                                            780
220 PR=220.0*P(I)+90.0                                                   790
    IF(IS(I)-1) 221,303,303                                              800
221 IF(IC(I)-2) 303,223,303                                              802
223 IF(P(I)-6.0) 225,230,225                                             805
225 K=1                                                                  810
230 W=103.12*PR+700.0                                                    820
    CH=667.7*PR+474000.0                                                 830
    CT=790.9*PR+650000.0                                                 840
    CO=922.7*PR+990000.0                                                 850
    RT=0.02643*PR+79.73                                                  860
260 IF(K) 270,270,290                                                    880
270 PRINT 280,I,PR,W,CO,CT,CH,RT                                         890
280 FORMAT(1X13,7XE10.4,1XE10.4,1XE10.4,1XE10.4,1XE10.4,1XE10.4)         1000
    GO TO 320                                                            1010
290 PRINT 300,I,PR,W,CO,CT,CH,RT                                         1020
300 FORMAT(  1H*I3,2XF10.4,1XE10.4,1XE10.4,1XE10.4,1XE10.4,1XE10.4)      1030
    GO TO 320                                                            1031
303 PRINT 305                                                            1035
305 FORMAT (3X44' INPUT DATA ERROR-SYSTEM NOT 1 OF 11 ANALYZED)          1037
320 I=I+1                                                                1040
    GO TO 70                                                             1050
340 PRINT 350                                                            1060
350 FORMAT(//1X29H* INDICATES EXTRAPOLATED DATA)                         1070
    GO TO 5                                                              1080
    END                                                                  1090
```

Table 3.11-II.

Input and output data for computerized system parametrics, system 1.

Input data

Run No.	Mobility class	Frequency (cps)	Cooling fluid	Power (Mw)
1	1	60	ϕ	3.00

Run No.	Yield (lb/hr)	Weight (lb)	Cost 1/year ($)	10/year ($)	100/yr ($)	Response time (min)
1	0.7290E+03	0.9828E+05	0.1504E+07	0.1184E+07	0.1004E+07	0.1010E+03

Using these input data, the computer program performs the necessary calculations and yields the following output information:

- Ammonia production rate
- Fuel production system weight
- Total cost of producing one of the specified systems per year
- Total cost of producing ten of the specified systems per year
- Total cost of producing one-hundred of the specified systems per year
- Fuel production system response time

If combinations of input parameters are chosen that are different from the combinations studied, the program is designed to print out a statement to this effect. For example, if the computer was instructed to analyze a 400-cps electrical power, water-cooled, Class III system, the computer would print out the statement that this specific combination of variables was not studied.

If the computer program is used to extrapolate the system data beyond the range of input powers studied, the output data will be preceded by an asterisk. The note on the asterisk will indicate that the output data are the result of an extrapolation beyond the range of the original study.

DISCUSSION

The graphical and computerized system parametric relationships presented are useful for estimating characteristics of systems other than those studied. The data may be interpolated to obtain information for systems with production capacities within the range of the system

capacities investigated. In addition, the data may be extrapolated to obtain information for systems with production capacities greater than the system capacities investigated. Interpolation and extrapolation results should be treated judiciously as estimates since each specific system studied was designed using unique constraints.

As indicated in the Supplement to this report, system 3 was designed using preliminary cost optimization data. Final data indicated that a design identical to system 1 is preferred. This difference was reflected in the graphical parametrics by presenting two data points for system 3 in each of the figures shown. For the computer work, however, the final design for system 3 was used. This obviates the need for analyzing two system 3 alternates and the effects of these alternates on other nondesign systems which may be examined by interpolation or extrapolation.

CALCULATIONS

The data presented in this subsection were collected from previous subsections of this report. Therefore, no calculations were required.

IV. CONCLUSIONS

The following conclusions were derived from this ammonia production system integration study.

- Ammonia production systems can be designed and fabricated to meet all the input electrical power level and frequency requirements and cooling medium variations and to conform to the mobility class limitations on size and weight of individual packages specified in the Introduction, Figure 2-2.
- Ammonia fuel production system designs require a minimum of three and a maximum of five separately-transportable packages depending on input power level and mobility class limitations. Relaxation of mobility class limitations reduces the number of packages, number of package interconnections, and response time of the system, but does not significantly affect system cost or output.
- The fuel production systems, when designed to minimize the specific cost and weight of the combined nuclear powerplant and ammonia production system, will produce from 729 to 3300 lb NH_3/hr. The primary limitation on output is electrical power input. There is no apparent technical reason, other than power input, why systems with greater ammonia production capabilities cannot be designed and constructed.
- The specific cost of the ammonia production systems vary from approximately $800 to $1400 per lb NH_3 per hour. The specific cost decreases significantly as plant size increases. The ammonia production system is a minor part of the combined nuclear powerplant and fuel production plant cost, approximately 25 to 35%.
- The hydrogen production subsystem is a major portion, approximately 50 to 55%, of the ammonia production system cost. The electrolyzer modules are approximately 80% of the hydrogen production subsystem cost.
- Water-cooled ammonia production systems are slightly lower in weight and cost than air-cooled systems, but have longer assembly times because of the greater number of package interconnections.
- Ammonia production systems designed for 400-cps electrical power input are significantly lower in weight than 60-cps systems, but will require development of 400-cps components in the ratings required. The effects on the nuclear powerplant were not considered in this study.
- Ammonia production capability is sensitive to ambient air temperatures and pressures. An increase in ambient air temperature decreases heat rejection capability of the various heat exchangers in the system and reduces system output. A decrease in ambient air pressure also decreases heat rejection capability of heat exchangers, but, more significantly, the air compressor capability is reduced and limits system output. However, it should be noted that nuclear powerplant output is also limited by ambient conditions and is the controlling influence on system production.

The following conclusions were derived from the results of the subsystem design studies.

- The most attractive ammonia synthesis subsystem process has a 6000 psig operating pressure and an ejector-driven refrigeration system. This process requires less power than any other investigated in the 1900 to 15,000 psig pressure range.
- The best nitrogen generation subsystem process utilizes a split air-fractionation column and refrigeration from waste oxygen gas from the hydrogen production subsystem. This process has the lowest power requirement of any investigated but is dependent on operation of the hydrogen production subsystem.
- Hydrogen production subsystems operating with power requirements of 19.9 to 23.2 kwh/lb H_2 can be designed and fabricated using as the basic component the electrolyzer module developed by Allison under Contract 1021(X). Most of the other subsystem components, such as heat exchangers and pumps, can be procured with a minimum of development effort required.
- The best water purification process to provide makeup water for the hydrogen production subsystem is vapor compression. This process requires a moderate amount of power, is relatively inexpensive, and is neither excessively large nor heavy. It is capable of handling the range of possible feedwater impurities.
- The pneumatic control subsystem selected will provide the subsystem coordination and control necessary for automatic operation. In addition, the control scheme allows system production rates to be maximized for the total available power and operating parameter restrictions.
- The power conditioning subsystem losses are minimized by using relatively large, efficient transformers.

Potential improvements to the system can be made through the development of components not available as state-of-the-art today. These developmental components include: high-speed, centrifugal nitrogen compressors and large air-cooled compressors for the ammonia synthesis subsystem.

V. RECOMMENDATIONS

The ammonia production system integration study confirmed the technical feasibility of producing ammonia fuel from air and water with electrical power in mobile plants. It is therefore recommended that the design, development, fabrication, and test of one or more ammonia production systems be started at once based on design criteria developed from the NPEL cost-effectiveness study results.

The design of the ammonia production system should be coordinated with the nuclear powerplant so as to achieve maximum output over the range of ambient operating conditions expected. Further consideration should be given to refinements in subsystem design, such as recovery of useful work from the oxygen expansion turbine in the nitrogen generation subsystem.

The recommended development program should include the following areas.

- Larger electrolyzer modules should be investigated for the higher power level plants. The investigation should include exploring the feasibility of fabricating larger electrodes, and modules containing approximately 300 cells, as well as an analysis of the performance of larger modules.
- High-speed centrifugal nitrogen compressors should be developed.
- A large air-cooled compressor unit should be studied. This should include the preliminary design of a lightweight unit to fit specific dimensional mobility restraints and a trade-off analysis which considers the effect of this unit on overall system weight, mobility, cost, and reliability.

APPENDIX A

Computer Program and Sample Calculation

A digital computer program was used to analyze the hydrogen production subsystem and to integrate the overall fuel production system. This computer program is discussed briefly in the Hydrogen Production Subsystem subsection and in detail in Reference 1. The computer program listing is given in Table A-I. The definitions and units of the input data are given in Table A-II.

As a sample calculation, the data for system 10 is included in Table A-II and is shown in parentheses. The computer print-out for system 10 is given in Tables A-III and A-IV. Table A-III is a tabulation of the input data used by the computer in making the calculations. The "AKS" value printed in column 4, row 5, is of special interest. It represents the power consumed by the ammonia, nitrogen, and water purification subsystems as a function of hydrogen production rate. (See Figure 3.7-2 and the Integration and Packaging portion of Section III.) When specified as input data, "AKS" allows the computer program to be used to integrate the hydrogen, ammonia, nitrogen, and water purification subsystems. This is done using an iterative calculation in which the total power available is apportioned among the subsystems to achieve the proper production rate from each subsystem. When the program is used to analyze only the hydrogen subsystem, "AKS" is given a value of zero.

Table A-IV is the solution for hydrogen subsystem, system 10. The hydrogen production rate is given in pounds per hour. The cell current and current density are given in amps and milliamps per square centimeter, respectively. The flow rates, except where noted, are given in cubic feet per minute. Voltages are given in volts, power values are in kilowatts, and weights are in pounds. The total cell area is given in square centimeters.

Once the units are known, the output is largely self-explanatory. The figures shown in rows 4 through 7 are values of internal program variables and, in general, are of no concern. Two exceptions exist—"Q" and "Power HR." "Q" is the gas-to-liquid volume ratio of the effluent module streams. The other term is the radiator fan power in kilowatts. Two other entries warrant special discussion also. In column 4, row 9, the design point bank voltage drop is given. When 25% of the banks are shut down, the voltage increases to the value shown in column 2, row 10. This latter value was limited to 619 v, as discussed in the Hydrogen Production Subsystem portion of Section III.

REFERENCES

1. Final Report Energy Depot Electrolysis Systems Study. Allison Division of General Motors, EDR 3714, 1 June 1964.

2. General Electric Bulletins GEA-8089, GEA-8090, GEA-8095, GEA-8096, and GED-5048. Schenectady, New York: Small AC Motor and Generator Department, General Electric Company.

3. Knowlton, A. E. Electrical Engineering Handbook. McGraw-Hill, 1939.

4. MCR Electrical Power Parametric Studies. Norwood Development Laboratories, Allis-Chalmers Manufacturing Co. Report No. 328. 15 February 1965. (with revisions)

Table A-I.

Computer program.

```
      DIMENSION DATA(50)                                                    0001
      EQUIVALENCE (DATA(1),WTOT),(DATA(2),WOA),(DATA(3),POWER),(DATA(4),    0002
     1WOCON),(DATA(5),PRESS),(DATA(6),RHOH2),(DATA(7),AKS),(DATA(8),DELT    0003
     2AP),(DATA(9),AKCELL),(DATA(10),WOCEL),(DATA(11),WOHR),(DATA(12),TE    0004
     3MP),(DATA(13),C1),(DATA(14),C2),(DATA(15),C3),(DATA(                  0005
     416),X),(DATA(17),VZERO),(DATA(18),VAD),(DATA(19),AKAUXL),(DATA(20)    0006
     5,AKAUXG),(DATA(21),AKDC),(DATA(22),AKAC),(DATA(23),ETADC),(DATA(24    0007
     6),ETAAC),(DATA(25),AKZERO),(DATA(26),AKRES),(DATA(27),AKHTR),(DATA    0008
     7(28),AKBUS),(DATA(29),AKPHR),(DATA(30),ETAPUM),(DATA(31),QSTART),(    0009
     8DATA(32),QMAX),(DATA(33),DIFLIM),(DATA(34),TRICON),(DATA(35),TOTVO    0010
     9L),(DATA(36),BANK),(DATA(37),DHH),(DATA(38),DVV),(DATA(39),BPB)    *  0011
  12 CALL RREAD (DATA)                                                      0012
      XULIM = QMAX                                                          0013
      XLLIM = 0.0                                                           0014
      J=0                                                                   0015
      ICHECK = 1                                                            0016
      POWER=POWER*1000.                                                     0017
      A=.025*AKAUXG/RHOH2+12062.*AKHTR*(VZERO-VAD)+AKDC*12062.*VZERO        0018
     *+AKAC*AKS+12062.*AKBUS                                                0019
      B=1.5/(60.*RHOH2)*(AKAUXL+(AKAC*DELTAP*3.25/ETAPUM))                  0020
      D=12062.*(AKHTR+AKDC)                                                 0021
      E=12062.*AKCELL                                                       0022
      F=12062.*VZERO/(ETAAC*ETADC)+AKS/ETAAC+12062.*(VZERO-VAD)*AKPHR       0023
     */ ETAAC                                                               0024
      G=1.5*3.25*DELTAP/(60.*RHOH2*ETAAC*ETAPUM)                            0025
      H=12062./(ETAAC*ETADC)+12062.*AKPHR/ETAAC                             0026
      AL=(WTOT-WOCEL-WOA-WOHR-WOCON)/POWER                                  0027
      Q=QSTART                                                              0028
   6  RHO=C1+C2*Q+C3*Q*Q                                                    0029
      AJ=(H*AL-D)*(AKZERO+X*RHO)                                            0030
      IF(H*AL - D)50,50,52                                                  0031
  50  TOTW =(D*POWER / H)+ WOCEL + WOA + WOHR + WOCON                       0032
      POWER = POWER / 1000.                                                A0032
      ICHECK = 2                                                            0033
  57  FORMAT(1H1)                                                           0034
      WRITE OUTPUT TAPE 6,57                                                0035
  58  FORMAT(///55X10HINPUT DATA///)                                        0036
      WRITE OUTPUT TAPE 6,58                                                0037
      WRITE OUTPUT TAPE 6,51,TOTW                                           0038
  51  FORMAT(15X47HTHE APPROX. MIN. WT. SKID FOR THIS CONDITION IS,E15.8    0039
     *,1X4HLBS.,3X5HSORRY)                                                  0040
      GO TO 59                                                              0041
  52  AK=F*AL-A+((G*AL-B)/Q)                                                0042
      EA = E + 12062. * AKRES * Q                                           0043
      SI = (-AK+(AK*AK+4.*EA*AJ)**0.5)/(2.*AJ)                              0044
      DRHODQ=C2+2.*C3*Q                                                     0045
      DAJDQ=(H*AL-D)*X*DRHODQ                                               0046
      DAKDQ=(B-G*AL)/(Q*Q)                                                  0047
      DSIDQ =(-DAKDQ + (AK*DAKDQ + 2.*EA*DAJDQ + 24124.*AJ*AKRES)/(AK*AK    0048
     *+4.*EA*AJ)**0.5)/(2.*AJ) - DAJDQ * (-AK + (AK*AK + 4.*EA*AJ)**0.5    0049
     *) / (2.*AJ*AJ)                                                        0050
      DPOWDQ=-G/(Q*Q)+H*(DSIDQ*(AKZERO+X*RHO)+SI*X*DRHODQ)                  0051
      IF(ABSF(DPOWDQ)-DIFLIM)3,3,2                                          0052
   2  IF(J-14)4,5,3                                                         0053
   4  J = J + 1                                                             0054
```

Table A-I. (cont)

```
      X1= Q                                                          0055
      Y1= DPOWDQ                                                     0056
      Q = Q + 0.2                                                    0057
      GO TO 6                                                        0058
    5 J = J + 1                                                      0059
      IF(J-50)1001,1001,10                                           0060
   10 WRITE OUTPUT TAPE 6,1002,X1,Y1,Q,DPOWDQ                        0061
 1002 FORMAT(1H1,4E15.8)                                            0062
      GO TO 3                                                        0063
 1001 X2=Q                                                          0064
      Y2=DPOWDQ                                                      0065
      BDELT = ABSF(Y2) - ABSF(Y1)                                    0066
      SDELT = X2 - X1                                                0067
      IF(SDELT)99,999,99                                             0068
   99 IF(Y2 / Y1)100,100,101                                         0069
  101 IF(BDELT)102,103,103                                           0070
  102 DUM = X2                                                       0071
      IF(SDELT)104,999,105                                           0072
  104 XULIM = DUM                                                    0073
      GO TO 106                                                      0074
  105 XLLIM = DUM                                                    0075
      GO TO 106                                                      0076
  103 DUM = X1                                                       0077
      IF(SDELT)105,999,104                                           0078
  100 IF(Y2)107,107,108                                              0079
  107 ITEST = 1                                                      0080
      GO TO 109                                                      0081
  108 ITEST = 2                                                      0082
  109 IF((Y2 - Y1) / SDELT)110,999,111                               0083
  110 GO TO (112 , 113),ITEST                                        0084
  111 GO TO (113 , 112),ITEST                                        0085
  112 XULIM = X2                                                     0086
      XLLIM = X1                                                     0087
      GO TO 106                                                      0088
  113 XULIM = X1                                                     0089
      XLLIM = X2                                                     0090
  106 IF(BDELT)114,115,115                                           0091
  114 Q = X2 +(Y2 * (X2 - X1))/ (Y1 - Y2 )                           0092
      IF(Q - XULIM)116,999,117                                       0093
  116 IF(Q - XLLIM)117,999,118                                       0094
  117 Q = (XULIM + XLLIM) / 2.0                                      0095
  118 Y1 = Y2                                                        0096
      X1 = X2                                                        0097
      GO TO 6                                                        0098
  115 Q = (XULIM + XLLIM) / 2.0                                      0099
      GO TO 6                                                        0100
  999 WRITE OUTPUT TAPE 6,1000,J,X1,Y1,X2,Y2                         0101
 1000 FORMAT(5X21HTROUBLE IN ITER. J = ,I3,4E14.6)                   0102
      GO TO 3                                                        0103
    3 WH2=POWER/(F+G/Q+H*SI*(AKZERO+X*RHO))                          0104
      BI=12062.*WH2                                                  0105
      V=VZERO+SI*(AKZERO+X*RHO)                                      0106
      PCELLO=BI*V/1000.                                              0107
      PCELLA=PCELLO/(ETADC*ETAAC)                                    0108
      PSYN=AKS*WH2/(ETAAC*1000.)                                     0109
      OG=1.5*WH2/(60.*RHOH2)                                         0110
```

Table A-I. (cont)

```
      QL=QG/Q                                                                  0111
      AAJ=J                                                                    0112
      PAUX=QL*DELTAP*3.25/(ETAPUM*ETAAC*1000.)                                 0113
      ACELL=12062.*HH2/SI                                                      0114
      QREJ=BI*(V-VAD)/1000.                                                    0115
      PHR = AKPHR * QREJ / ETAAC                                               0116
      WCELL=WOCEL+AKCELL*ACELL                                                 0117
      WSEP = AKAUXG * ( QL + QG)                                               0118
      WPUMP = QL * ( AKAUXL - AKAUXG)                                          0119
      WRES = AKRES * ACELL * Q                                                 0120
      WBUS = AKBUS * BI                                                        0121
      WAUX=WOA+AKAUXL*QL+AKAUXG*QG+AKRES*ACELL*Q+AKBUS*BI                      0122
      WHR=WOHR+AKHTR*QREJ*1000.                                                0123
      WCONV=WOCON+(AKDC*PCELLD+AKAC*(PAUX+PSYN)*ETAAC)*1000.                   0124
      SIA=SI*1000.                                                             0125
      POWER=POWER/1000.                                                        0126
   13 FORMAT(1H1,55X10HINPUT DATA///)                                          0127
      WRITE OUTPUT TAPE 6,13                                                   0128
    9 FORMAT(42X36HAMMONIA SYNTHESIS PLANT OPTIMIZATION///)                    0129
   59 WRITE OUTPUT TAPE 6,9                                                    0130
      IF(TOTVOL)53,53,54                                                       0131
   54 WRITE OUTPUT TAPE 6,55                                                   0132
   55 FORMAT(35X47HNOTE ---THINGS CALLED WEIGHT ARE REALLY VOLUMES///)         0133
   56 FORMAT(30X22HOVERALL SKID VOLUME IS,E15.8,3X10HCUBIC FEET//)             0134
      WRITE OUTPUT TAPE 6,56,TOTVOL                                            0135
   15 FORMAT(3X17HTOTAL SKID WEIGHT,6X21HTOTAL POWER AVAILABLE, 5X16HBAS       0136
     *E CELL WEIGHT, 3X24HCELL WT. PROPORTIONALITY, 3X21HBASE AUXILLARY        0137
     *WEIGHT)                                                                 0138
   40 FORMAT(30X51HTHE SYSTEM PRESSURE USED IN THIS CALCULATION WAS   ,       0139
     *F10.1,3X4HPSIA//)                                                       0140
   53 WRITE OUTPUT TAPE 6,40,PRESS                                             0141
   41 FORMAT(30X37HTHE ELECTROLYSIS CELL TEMPERATURE WAS,F10.1,3X6HDEG.        0142
     *C//)                                                                    0143
      WRITE OUTPUT TAPE 6,41,TEMP                                              0144
      IF(TRICON)42,43,42                                                       0145
   43 WRITE OUTPUT TAPE 6,44                                                   0146
   44 FORMAT(30X23HOPTIMIZED ON DRY WEIGHT////)                                0147
      GO TO 45                                                                 0148
   42 WRITE OUTPUT TAPE 6,46                                                   0149
   46 FORMAT(30X23HOPTIMIZED ON WET WEIGHT////)                                0150
   45 WRITE OUTPUT TAPE 6,15                                                   0151
   14 FORMAT(5E24.8//)                                                         0152
      WRITE OUTPUT TAPE 6,14,WTOT,POWER,WOCEL,AKCELL,WOA                       0153
   16 FORMAT(4X15HLIQ. AUX. PROP.,9X14HGAS AUX. PROP.,8X21HBASE HEAT REJ       0154
     *. WEIGHT,2X22HHEAT REJ. EQUIP. PROP.,1X22HPR. COND. EQ. BASE WT.)       0155
      WRITE OUTPUT TAPE 6,16                                                   0156
      WRITE OUTPUT TAPE 6,14,AKAUXL,AKAUXG,WOHR,AKHTR,WOCON                    0157
   17 FORMAT(5X15HD-C COND. PROP.,9X15HA-C COND. PROP.,6X20HA-C CONV. EF       0158
     *FICIENCY,4X20HD-C CONV. EFFICIENCY,12X5HVZERO)                           0159
      WRITE OUTPUT TAPE 6,17                                                   0160
      WRITE OUTPUT TAPE 6,14,AKDC,AKAC,ETAAC,ETADC,VZERO                       0161
   18 FORMAT(11X2HKO,23X1HX,22X2HC1,22X2HC2,22X2HC3)                           0162
      WRITE OUTPUT TAPE 6,18                                                   0163
      WRITE OUTPUT TAPE 6,14,AKZERO,X,C1,C2,C3                                 0164
   19 FORMAT(3X19HDENSITY OF HYDROGEN,6X16HSYS. PRESS. DROP,9X15HPUMP EF       0165
     *FICIENCY,15X3HAKS,15X17HADIABATIC VOLTAGE)                              0166
```

Table A-I. (cont)

```
      WRITE OUTPUT TAPE 6,19                                                    0167
      WRITE OUTPUT TAPE 6,14,RHOM2,DELTAP,ETAPUM,AKS,VAD                        0168
   27 FORMAT(1X25HRESERVOIR PROPORTIONALITY,10X6HDIFLIM,19X5HAKBUS,15X6H        0169
     *QSTART,19X4HQMAX)                                                       0170
      WRITE OUTPUT TAPE 6,27                                                    0171
      WRITE OUTPUT TAPE 6,14,AKRES,DIFLIM,AKBUS,QSTART,QMAX                     0172
   28 FORMAT (10X5HAKPHR,21X12HNO. OF BANKS,12X15HELEC. BASE (IN),10X14H'
     1ELEC. HT. (IN),11X13HMOD. PER BANK)
      WRITE OUTPUT TAPE 6,28                                                    0174
      WRITE OUTPUT TAPE 6,14,AKPHR,BANK,DHH,DVV,BPB
      GO TO(60,12),ICHECK                                                       0176
   20 FORMAT(1H1)                                                               0177
   60 WRITE OUTPUT TAPE 6,20                                                    0178
   21 FORMAT(57X7HRESULTS///)                                                   0179
      WRITE OUTPUT TAPE 6,21                                                    0180
   22 FORMAT(43X18HHYDROGEN PRODUCED=,E15.8///)                                 0181
      WRITE OUTPUT TAPE 6,22,WH2                                                0182
   23 FORMAT(9X15HCURRENT TO CELL,9X15HVOLTAGE TO CELL,9X14HCELL D-C POW        0183
     *ER,10X14HCELL A-C POWER,10X15HSYNTHESIS POWER)                           0184
      WRITE OUTPUT TAPE 6,23                                                    0185
      WRITE OUTPUT TAPE 6,14,BI,V,PCELLD,PCELLA,PSYN                            0186
   24 FORMAT(7X15HAUXILLARY POWER,9X13HGAS FLOW RATE,10X16HLIQUID FLOW R        0187
     *ATE,9X15HTOTAL CELL AREA,7X19HCELL HEAT REJECTION)                       0188
      WRITE OUTPUT TAPE 6,24                                                    0189
      WRITE OUTPUT TAPE 6,14,PAUX,QG,QL,ACELL,QREJ                              0190
   25 FORMAT(9X11HCELL WEIGHT,9X16HAUXILLARY WEIGHT,5X23HHEAT REJECTION         0191
     *EQIP WT.,3X20HPR. COND. EQUIP. WT.,4X20HCELL CURRENT DENSITY)            0192
      WRITE OUTPUT TAPE 6,25                                                    0193
      WRITE OUTPUT TAPE 6,14,WCELL,WAUX,WHR,WCONV,SIA                           0194
   26 FORMAT(15X3HRHO,21X2HSI,21X5HDAJDQ,19X5HDAKDQ,19X5HDSIDQ)                 0195
      WRITE OUTPUT TAPE 6,26                                                    0196
      WRITE OUTPUT TAPE 6,14,RHO,SI,DAJDQ,DAKDQ,DSIDQ                           0197
   30 FORMAT(13X1HA,23X1HB,23X1HD,23X1HE,23X1HF)                                0198
      WRITE OUTPUT TAPE 6,30                                                    0199
      WRITE OUTPUT TAPE 6,14,A,B,D,E,F                                          0200
   31 FORMAT(13X1HG,23X1HH,23X1HL,23X1HQ,23X1HJ)                                0201
      WRITE OUTPUT TAPE 6,31                                                    0202
      WRITE OUTPUT TAPE 6,14,G,H,AL,Q,AJ                                        0203
   32 FORMAT(9X6HDPOWDQ,20X2HIJ,23X1HK,22X2HEA,19X9HPOWER H R)                  0204
      WRITE OUTPUT TAPE 6,32                                                    0205
      WRITE OUTPUT TAPE 6,14,DPOWDQ,AAJ,AK,EA,PHR                               0206
      CNUM=ACELL*0.001076/DHH/DVV*144.0                                         1000
      CPB=CNUM/BANK                                                             1001
      CPM=CPB/3.0
      VDPB=V*CPB                                                                1003
      AOP=7.15*WH2/100.298                                                      1004
      CFV=QL/60./CNUM*144./.3937/DHH                                            1005
      EGPM=QL*7.481                                                             1006
      TWPC=4.425*CPM+0.003051*C.M*SIA*BANK/BPB
   33 FORMAT (10X4HWSEP,20X5HWPUMP,19X4HWRES,20X4HWBUS,20X15HEST NH3 OUT       1008
     1 T/D)                                                                    1009
      WRITE OUTPUT TAPE 6,33                                                    0208
      WRITE OUTPUT TAPE 6,14,WSEP,WPUMP,WRES,WBUS,AOP                           1010
   34 FORMAT (11X12HNO. OF CELLS,12X14HCELLS PER BANK,11X13HCELLS PER MO       1011
     1D,12X14HVOLTS PER BANK,11X14HCELL VEL (FPS))                             1012
      WRITE OUTPUT TAPE 6,34                                                    1013
```

Table A-I. (cont)

```
      WRITE OUTPUT TAPE 6,14,CNUM,CPB,CPM,VDPB,CFV                    1014
   35 FORMAT (10X13HELEC FLOW GPM,12X15HWT PER CELL LBS)              1015
      WRITE OUTPUT TAPE 6,35                                          1016
      WRITE OUTPUT TAPE 6,14,EGPM,THPC                                1017
      GOTO 12                                                         0210
      END(1,0,0,0,0,0,1,0,0,0,0,0,0,0,0,0)
```

Table A-II.

Computer program nomenclature.

WTOT	Total weight, lb (83, 960)
WOA	Constant portion of auxiliary equipment weight, lb (14, 860)
POWER	Total power, kw (9, 993.3)
WOCONV	Constant portion of electrical conversion unit weight, lb (800)
PRESS	Nominal operating pressure, psia (314.7)
$RHOH_2$	Density of hydrogen, lb/ft^3 (0.0937)
AKS	Power required by synthesis plant per unit hydrogen flow, w-hr/lb (2, 220)
DELTAP	Pump pressure rise, psi (25.0)
AKCELL	Cell weight per unit area, lb/cm^2 (0.00314)
WOCELL	Constant portion of cell weight, lb (8, 500)
WOHR	Constant portion of heat rejection equipment weight, lb (1, 750)
TEMP	Operating temperature, °C (75.0)
C1	Coefficient of resistivity equation, ohm-cm (0.965)
C2	Coefficient of resistivity equation, ohm-cm (0.3135)
C3	Coefficient of resistivity equation, ohm-cm (0.367)
X	Cell spacing, cm (0.446)
VZERO	Cell voltage extrapolated to zero current from linear part of V-I curve, v (1.485)
VAD	Adiabatic cell voltage, v (1.48)
AKAUXL	Weight of liquid carrying auxiliaries per unit liquid flow, $lb/ft^3/min$ (5.6)
AKAUXG	Weight of gas carrying auxiliaries per unit gas flow, $lb/ft^3/min$ (0.464)
AKDC	Weight of d-c power conversion equipment per unit power, lb/w (0.00017)
AKAC	Weight of a-c power conversion equipment per unit power, lb/w (0)
ETADC	Efficiency of d-c conversion (0.995)
ETAAC	Efficiency of a-c conversion (0.991)
AKZERO	Rate of change of cell overvoltage with current density, $ohm-cm^2$ (0.51)
AKRES	Reservoir weight per unit cell area times q, lb/cm^2 (0.000603)
AKHTR	Heat rejection weight per unit heat rejection rate, lb/w (0.00258)
AKBUS	Bus weight per unit current, lb/amp (0.00025)
AKPHR	Rate of change of power with heat rejection (0.03)
ETAPUM	Pump efficiency (0.65)
QSTART	Initial estimate of ratio of gas-to-liquid flow leaving cell (0.2)
QMAX	Maximum value of gas-to-liquid ratio (10.0)
DIFLIM	Iteration tolerance (0.0001)
TRICON	Trigger to print whether optimizing on a dry or wet basis—0.0 = dry and 1.0 = wet (1.0)
TOTVOL	Total skid volume for volume optimization analysis, ft^3 (must be 0 for weight analysis) (0.0)

Table 2-II. (cont)

BANK	Number of module banks (18.0)
DHH	Horizontal electrode dimension, in. (22.0)
DVV	Vertical electrode dimension, in. (18.0)
BPB	Minimum number of module banks that can be operated without shutting down subsystem (13.5)

Table A-III.
Computer input data.

INPUT DATA

AMMONIA SYNTHESIS PLANT OPTIMIZATION

THE SYSTEM PRESSURE USED IN THIS CALCULATION WAS 314.7 PSIA

THE ELECTROLYSIS CELL TEMPERATURE WAS 75.0 DEG. C

OPTIMIZED ON WET WEIGHT

TOTAL SKID WEIGHT 0.83959999E 05	TOTAL POWER AVAILABLE 0.99932994E 04	BASE CELL WEIGHT 0.85000000E 04	CELL WT. PROPORTIONALITY 0.31400000E-02	BASE AUXILLARY WEIGHT 0.14659999E 05
LIQ. AUX. PROP. 0.55999999E 01	GAS AUX. PROP. 0.46399999E-00	BASE HEAT REJ. WEIGHT 0.17500000E 04	HEAT REJ. EQUIP. PROP. PR. 0.25799999E-02	COND. EQ. BASE WT. 0.60000000E 03
D-C COND. PROP. 0.16999999E-03	A-C COND. PROP. 0.	A-C CONV. EFFICIENCY 0.99100000E 00	D-C CONV. EFFICIENCY 0.99500000E 00	VZERO 0.14850000E 01
KO 0.51000000E 00	X 0.44599999E-00	C1 0.96500000E 00	C2 0.31349999E-00	C3 0.36700000E-00
DENSITY OF HYDROGEN 0.93699999E-01	SYS. PRESS. DROP 0.25000000E 02	PUMP EFFICIENCY 0.65000000E 00	AKS 0.22200000E 04	ADIABATIC VOLTAGE 0.14799999E 01
RESERVOIR PROPORTIONALITY 0.60299999E-03	DIFLIM 0.99999999E-04	AKBUS 0.23000000E-03	QSTART 0.20000000E-00	QMAX 0.99999999E 02
AKPHR 0.30000000E-01	NO. OF BANKS 0.18000000E 02	ELEC. BASE (IN) 0.22000000E 02	ELEC. HT. (IN) 0.18000000E 02	BANK 0.13500000E 02

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Table A-IV.
Computer results.

RESULTS

HYDROGEN PRODUCED= 0.40564854E 03

CURRENT TO CELL 0.48929326E 07	VOLTAGE TO CELL 0.1R086314E 02	CELL D-C POWER 0.88495117E 04	CELL A-C POWER 0.89747545E 04	SYNTHESIS POWER 0.9087'821E 03
AUXILLARY POWER 0.61149567E 02	GAS FLOW RATE 0.10823067E 03	LIQUID FLOW RATE 0.48479377E 03	TOTAL CELL AREA 0.14812884E 08	CELL HEAT REJECTION 0.16079716E 04
CELL WEIGHT 0.55012458E 05	AUXILLARY WEIGHT 0.20684553E 05	HEAT REJECTION EQIP WT. 0.58985665E 04	PR. COND. EQUIP. WT. 0.23044 04	CELL CURRENT DENSITY 0.33031598E 03
RHO 0.10532808E 01	SI 0.33031598E-00	DAJDQ 0.85181446E 01	DAK 0.26055618E 02	DSIDQ -0.17072918E-01
A 0.60987113E 01	B 0.14941302E 01	D 0.33170499E 02	E 0.37874679E 02	F 0.20407557E 05
G 0.39654007E 02	H 0.12597854E 05	L 0.58088922E-02	Q 0.22325094E-00	J 0.39199419E 02
DPDWDQ -0.99182129E-04	FJ .0000000E 01	K 0.10662965E 03	EA 0.39498469E 02	POWER H 2 0.48677242E 02
WSEP 0.27516333E 03	WPUMP 0.243 9007E 04	WRES 0.19941152E 04	WBUS 0.11253744E 04	EST NH3 OUT T/D 0.26781537E 02
NO. OF CELLS 0.57959777E 04	CELLS PER BANK 0.32199320E 03	CELLS PER MOD 0.10733107E 03	VOLTS PER BANK 0.582367038 03	CELL VEL (ERS) 0.51966813E-01
ELEC FLOW GPM 0.36267421E 04	VOLTS PER BANK OD 0.61916346E 03			

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APPENDIX B

Derivation of Analog Computer Program for Control Analysis of Fuel Production System

This appendix describes derivation of the analog computer program required for control analysis of the fuel production system. Results obtained from the supporting subsystem computer programs are presented. This appendix contains the following subsections:

- Fuel Production System Analog Computer Program
- Hydrogen Production Subsystem
- Nitrogen Generation Subsystem Analog Computer Program
- Ammonia Synthesis Subsystem Start-up Analysis
- Nitrogen Generation Subsystem Start-up Analysis
- Hydrogen Production Subsystem Start-up Analysis
- Ammonia Synthesis Subsystem Operating Point Optimization

FUEL PRODUCTION SYSTEM ANALOG COMPUTER PROGRAM

The fuel production system computer program consists of a simplified simulation of the hydrogen production and nitrogen generation subsystems and a detailed simulation including each major component of the ammonia synthesis subsystem. These components are shown in Figure B-1 and include the synthesis gas and recycle compressor; the primary, secondary, and product condensers and separators; the converter heat exchanger; the ammonia vaporizing heat exchanger; the refrigeration ejector; and the ammonia synthesis converters. The simplified hydrogen subsystem simulation includes the liquid-gas separator tank volume and the hydrogen pressure control valve which controls hydrogen flow into the ammonia production subsystem. The nitrogen subsystem simulation includes the control valve which controls nitrogen flow into the ammonia synthesis subsystem. Figure B-1 also includes a numerical coding which will be followed in discussing the analog program.

The major equations and techniques used in generating the fuel manufacturing system computer program follow. The general approach taken was to provide for conservation of mass throughout the system. This implies that continuity of flow is preserved. Also, conservation of energy is provided by balancing heat loads throughout the system.

Compressor

The synthesis gas compressor simulation assumes a constant volume flow rate through the machine.

The weight flow through the compressor is directly dependent on the pressure and temperature at the input. Because the hydrogen line connecting the hydrogen production subsystem liquid-gas separator-reservoir and the synthesis gas compressor is unrestricted, the liquid-gas separator-reservoir void volume, pipe volume, and input volume of the synthesis gas separator were lumped in a single volume for the simulation. Basically, the simulation consists of

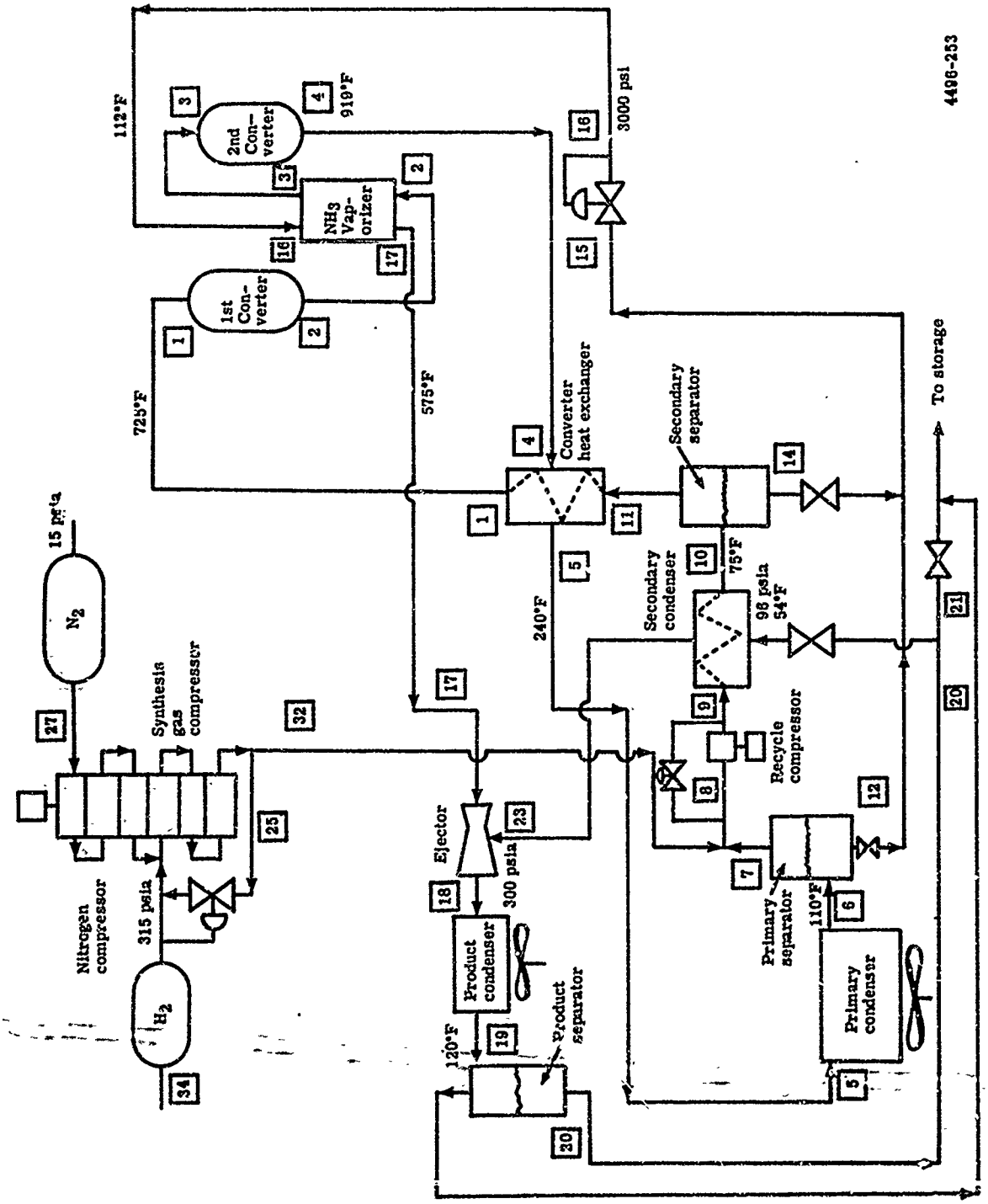


Figure B-1. Fuel production system flow diagram and coding.

the determination of the weight of gas in the plenum and a perturbation form of the gas law equation. These equations are

$$\omega = \int \left[\dot{\omega}_{34} + \dot{\omega}_{27} + \dot{\omega}_{25} - \dot{\omega}_{32} \right] dt \tag{1}$$

$$P_p = \frac{\partial P_p}{\partial \omega} \Delta\omega + \frac{\partial P_p}{\partial T} \Delta T \tag{2}$$

$$\dot{\omega} = K_{1p} (P_p) + K_{2p} \int P_p \, dt \tag{3}$$

where

ω = mass in plenum, lb-moles
$\dot{\omega}$ = gas flow rate, lb-moles/sec
P_p = plenum pressure, psia
K_{1p} and K_{2p} = control constants

Subscripts refer to points in Figure B-1

The flow of the nitrogen into the plenum is regulated to correspond to the hydrogen flow as follows.

$$\dot{\omega}_{27} = K_{1\omega} (\dot{\omega}_{34} - C\dot{\omega}_{27}) + K_{2\omega} \int (\dot{\omega}_{34} - C\dot{\omega}_{27}) \, dt \tag{4}$$

where

C = hydrogen-nitrogen molar ratio
$K_{1\omega}$ and $K_{2\omega}$ = control constants

The recycle compressor was programmed on the computer in a manner similar to the synthesis gas compressor. Constant volume flow was assumed, and weight flow is a function of the temperature and pressure at the compressor inlet.

Converters

The ammonia synthesis converters were simulated as a single unit to use operational information provided by Girdler Corporation. The basic equation to simulate performance of the converters is

$$\Delta\% \, NH_{3\ out} = F(T) \left[K_p \, \Delta P + K_{S.V.} \, \Delta S.V. + K_A \, \Delta\% A + K_{NH_3\ in} \, \Delta\% \, NH_{3\ in} + K_{H_2/N_2} \, \Delta(H_2/N_2)_{in} \right] \tag{5}$$

where

ΔP = change in pressure in the converters, psi
$\Delta S.V.$ = change in space velocity, hr^{-1}
$\Delta \%\ A$ = change in percent inerts in the synthesis gas, %
$\Delta \%\ NH_{3\ in}$ = change in percent ammonia concentration in the synthesis gas, %
$\Delta [H_2/N_2]_{in}$ = change in synthesis gas ratio at the converter inlet

Equation (5) describes the effect of each variable on the percent ammonia produced in the converters. The effects are computed from curves shown in Figures 3.1-6 through 3.1-11 of the Ammonia Synthesis Subsystem portion of Section III. Each variable was linearized except the effects of H_2/N_2 ratio and the average converter temperature which were set up on function generators. The converter temperature is included as a separate function because it is an important controlling factor in ammonia synthesis. Thus, the amount of ammonia produced is found by

$$\dot{\omega}_{4NH_3} = \%\ NH_{3\ out} \cdot \dot{\omega}_{4T} \tag{6}$$

where

$\dot{\omega}_{4NH_3}$ = flow of ammonia from the converters, lb-moles/sec
$\dot{\omega}_{4T}$ = total flow leaving the converter, lb-moles/sec

In normal operation about 20% of the moles are converted to ammonia.

The heat transfer rate to the synthesis gas flow through the converters is calculated by

$$Q_{1-4} = Q_{gen} - Q_{2-3} + Q_{trans} = \dot{\omega}_{4T}\, C_p\, (T_4 - T_1) \tag{7}$$

where

Q_{1-4} = heat transfer rate to the gas flow through converters, BTU/sec
Q_{gen} = heat of formation of ammonia, $\dot{\omega}_{NH_3(gen)} \times H^{\circ}_{form}$, BTU/sec
Q_{2-3} = heat removal rate by the vaporizing heat exchanger, BTU/sec
Q_{trans} = heat transfer rate, BTU/sec, from the catalyst bed to the gas stream (through the converters) during transient operation

Thus, the converter outlet temperature is found by

$$T_4 = \frac{Q_{1-4}}{\dot{\omega}_{4T}\,(C_{p_4})} + T_1 \tag{8}$$

where

T_4 = converter outlet gas temperature (°F)
T_1 = converter inlet temperature which is also the converter heat exchanger exit temperature

The average converter temperature is then calculated from

$$T_{conv} = \frac{T_1 + B\,T_4}{1 + B} \tag{9}$$

where

T_{conv} = average converter temperature, °F
B = constant weighting factor

Converter Heat Exchanger

The converter heat exchanger is a tube and shell heat exchanger for heating the incoming synthesis gases to reaction temperature while cooling the converter efflux gases. The heat transferred to the converter inlet gases is expressed by

$$Q_{11-1} = \dot{\omega}_{11}\, C_{p11}\,(T_1 - T_{11}) = UA\left(T_S - \frac{T_{11} + T_1}{2}\right) \tag{10}$$

and

$$T_S = \frac{1}{\omega C_p}\int \left[Q_{4-5} - Q_{11-1}\right] dt \tag{11}$$

On the hot side of the converter-heat exchanger the heat transfer rate from the converter exit gases is

$$Q_{4-5} = \dot{\omega}_4\, C_{p_4}\,(T_4 - T_5) = UA\left(\frac{T_4 + T_5}{2} - T_S\right) \tag{12}$$

where

Q_{11-1} = heat transfer rate to the converter inlet gases, BTU/sec
T_S = average wall temperature of the heat exchanger, °F
ωC_p = the thermal inertia, BTU/°F, of the heat exchanger (The fluids in the heat exchanger contribute to this term; however, for a gas, the ωC_p is negligible compared to the heat exchanger itself.)
Q_{4-5} = heat transfer rate from converter exit gases, BTU/sec
Q_{11-1} = Q_{4-5} under steady-state conditions

Ammonia Vaporizer

The ammonia vaporizer is used to transfer heat from the synthesis gas stream exiting from converter 1 to vaporize a 3000-psi ammonia stream. Because the ammonia being vaporized is above the critical pressure, no heat of vaporization is involved; instead, vaporization takes place at the driving temperature. Thus, the heat transfer rate to the ammonia stream is expressed by

$$Q_{16-17} = \dot{\omega}_{16} C_p (T_{17} - T_{16}) = UA \left(T_S - \frac{T_{16} + T_{17}}{2} \right) \tag{13}$$

and

$$T_S = \frac{1}{\omega C_p} \int \left[Q_{2-3} - Q_{6-17} \right] dt \tag{14}$$

On the hot side of the heat exchanger, the heat transfer rate from the synthesis gas stream is

$$Q_{2-3} = UA \left(\frac{T_1 + T_4}{2} - T_S \right) \tag{15}$$

where

Q_{16-17} = heat transfer rate to ammonia stream, BTU/sec
T_S = average wall temperature of the ammonia vaporizer, °F
Q_{2-3} = heat transfer rate from synthesis gas stream, BTU/sec

In Equation (15), UA is an effective value because T_1 and T_4 were used instead of T_2 and T_3 which are not generated by the computer program. This approach produces negligible error because the system is simulated on a perturbation basis about a steady-state point.

Primary Condenser

The primary condenser liquefies ammonia from the synthesis stream. The condenser cools the stream and condenses ammonia. The heat transfer rate is given by the following relationship.

$$Q_{5-6} = Q_{\Delta T_{5-6}} + Q_{\Delta H_{5-6}} \tag{16}$$

(The heat transferred from the synthesis gas, Q_{5-6}, equals the heat transferred to the air (or water) during steady-state operation. For the simulation, the temperature rise across the cooling fluid was constant.)

$$Q_{\Delta T_{5-6}} = \dot{\omega}_{5T}\, C_p\, (T_5 - T_6) = UA\left(\frac{T_5 + T_6}{2} - T_S\right) \quad (17)$$

$$Q_{\Delta H_{5-6}} = \dot{\omega}_{NH_3\,(l)_6} \times \Delta H_{cond} \quad (18)$$

where

Q_{5-6} = heat transfer rate from synthesis gas stream, BTU/sec
$Q_{\Delta T_{5-6}}$ = heat transfer rate to cool stream, BTU/sec
$\dot{\omega}_{NH_3\,(l)_6}$ = the ammonia condensation rate, lb mole/sec
$Q_{\Delta H_{5-6}}$ = heat transfer rate to condense ammonia, BTU/sec

The quantity of ammonia remaining as a gas in the stream is a function of both the pressure and exit temperature of the stream from the condenser.

$$\%\ NH_3\ (gas) = \frac{\partial NH_3}{\partial T_6}\ \Delta T_6 + \frac{\partial NH_3}{\partial P}\ \Delta P \quad (19)$$

where

ΔT_6 = change in condenser efflux temperature, °F
ΔP = change in loop pressure, psia
$\frac{\partial NH_3}{\partial T_6}$ = change of percent gaseous ammonia with respect to temperature, %/°F
$\frac{\partial NH_3}{\partial P}$ = change of percent gaseous ammonia with respect to system pressure, %/psia

The $\frac{\partial NH_3}{\partial T}$ and $\frac{\partial NH_3}{\partial P}$ terms are found from curves in Nielsen.[1]* The flow rate of gaseous ammonia remaining in the exit stream is

$$\dot{\omega}_{NH_3(g)_6} = \%\ NH_3\ (gas) \times \dot{\omega}_{T_6(g)} \quad (20)$$

where

$\dot{\omega}_{NH_3\,(g)_6}$ = flow rate of gaseous ammonia from the primary condenser, lb-mole/sec
$\dot{\omega}_{T_6\,(g)}$ = flow rate of all gases from the primary condenser, lb-mole/sec

The amount of ammonia condensed is then the difference between the total ammonia in the stream and the amount which remains a gas.

$$\dot{\omega}_{NH_3\,(l)_6} = \dot{\omega}_{NH_3\,(g)_4} - \dot{\omega}_{NH_3\,(g)_6} \quad (21)$$

*Superscripts indicate references at the end of this appendix.

Secondary Condenser

The secondary condenser cools the synthesis gas stream from the recycle compressor to condense ammonia. This condenser is cooled by low pressure, low temperature ammonia from the product stream. Condensation of ammonia in the synthesis stream causes evaporation of the ammonia in the product stream. The heat transfer rate from the synthesis stream to cool the gas and condense ammonia is determined by the same equations as for the primary condenser; the equations will not be repeated. On the cooling side, the heat transferred to the ammonia stream, Q_{12-23}, causes ammonia vaporization according to

$$\dot{w}_{NH_3\,(g)} = \frac{Q_{12-23}}{\Delta H_{evap}} \tag{22}$$

where

$\dot{w}_{NH_3\,(g)}$ = mass rate of evaporation, lb mole/sec
ΔH_{evap} = heat of vaporization for ammonia at 96 psia and 54°F, BTU/lb mole

The liquid ammonia level in the condenser is controlled by a valve which operates by proportional control.

Product Condenser

The product condenser condenses the ammonia from the ejector to form the product stream. It operates according to the classical heat exchange equations for condensers presented previously for the primary and secondary condensers.

Primary, Secondary, and Product Separators

The equations for the three liquid-gas separators are identical. The separators each have a liquid level which is maintained to a preset level by a controlled valve in the liquid efflux. The equations describe the mass of liquid ammonia in the separators, the liquid level, and the valve response.

The mass of liquid in any separator is determined by

$$\omega(l) = \int \left[\dot{w}(l)_{in} - \dot{w}(l)_{out}\right] dt \tag{23}$$

Knowing the physical dimensions of the separators allows calculation of the liquid level from the mass and density of the liquid. The valve then controls the liquid level to a set level by

$$A = K_{1H}(h - h_{set}) + K_{2H}\int (h - h_{set})\,dt \qquad (24)$$

where

A = valve area (in.^{2})
K_{1H} = the proportional control gain (in.^{2}/inch)
K_{2H} = the integral gain (in.^{2}/in.-sec)
h = actual liquid level in the separator (in.)
h_{set} = separator liquid level control point (in.)

The flow rate of ammonia out of the separator, $\dot{\omega}(1)_{out}$, is controlled by the valve according to the relation:

$$\dot{\omega}(1)_{out} = \frac{\partial \dot{\omega}(1)_{out}}{\partial A}\,\Delta A \qquad (25)$$

Ejector

The ejector is a jet pump which combines the high pressure (3000-psi) ammonia flow from the vaporizing heat exchanger with the low pressure (96-psia) ammonia flow from the secondary condenser to produce a flow at an intermediate pressure (300-psia). The ejector operates according to a curve of mass ratio (M_R) vs pressure ratio (P_R)^2 where

$$M_R = \frac{\dot{\omega}_{23}}{\dot{\omega}_{17}} \qquad (26)$$

and

$$P_R = \frac{P_{17} - P_{18}}{P_{18} - P_{23}} \qquad (27)$$

This curve defines the ejector for steady-state operation and for small perturbations from steady state.

Analog Computer Diagram for Fuel Production System Simulation

The equations presented in this subsection plus others required to provide continuity were rearranged as required to complete the simulation. The analog diagram, Figure B-2, shows the equations as they were programmed on the computer.

HYDROGEN PRODUCTION SUBSYSTEM

The hydrogen production subsystem was described by perturbation equations to simulate pressure balance throughout the subsystem and to describe the liquid level behavior within the subsystem.

The approach used to develop the describing equations involved the classical approach of conducting a force balance within the system and conducting a mass balance on the flowing liquid. The gaseous phase was described by the equation of state. The pressures were described at specific points in the system as a function of the fluid flowing past the point. Continuity was then provided by balancing the fluid mass to and from the point. The pressure equations were derived using two principles analogous to those of Kirchhoff's electrical equations, namely, the sum of the flows at a point must equal zero and the pressure drops around a closed loop must be zero.

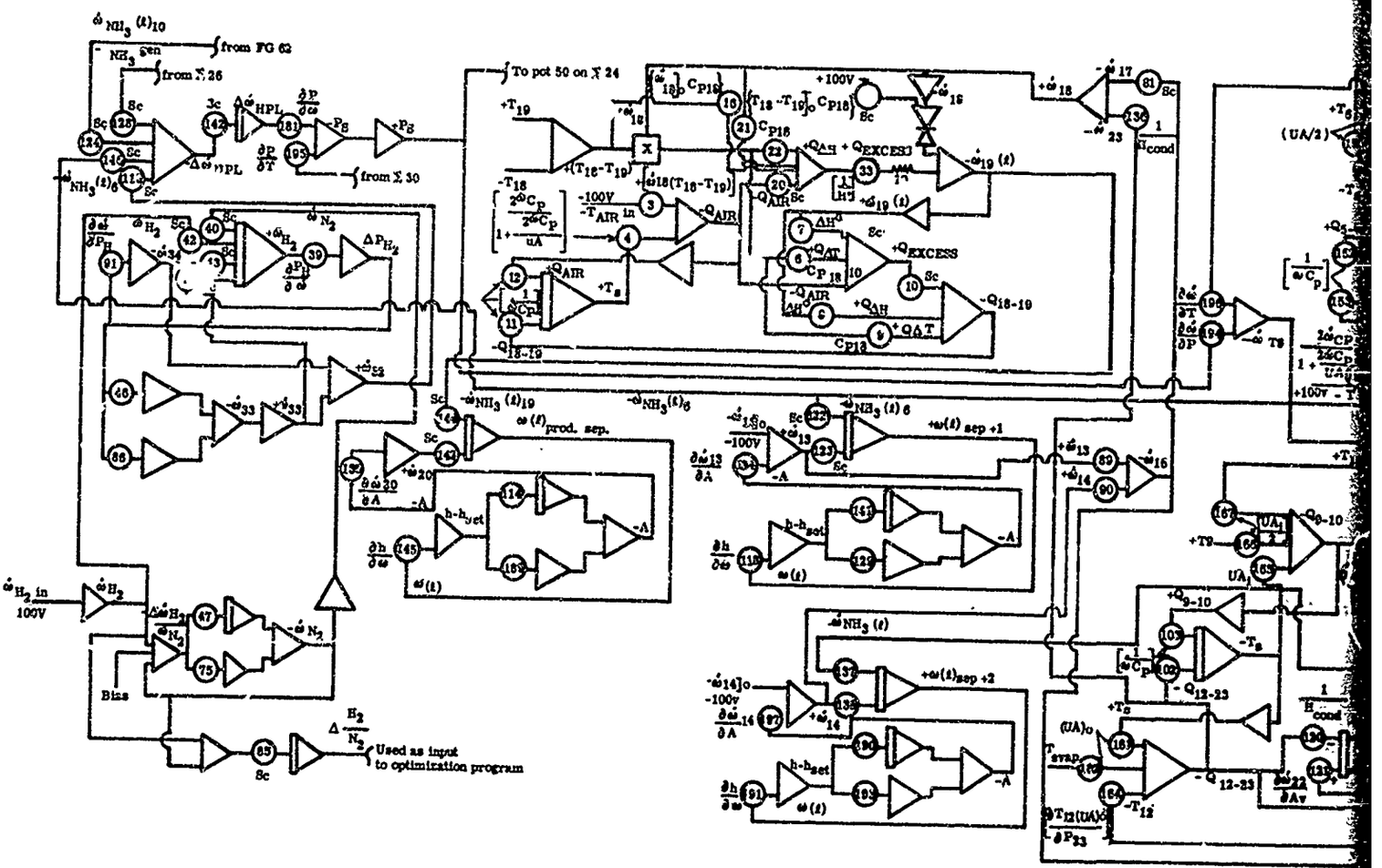
The hydrogen production subsystem flow diagram, Figure B-3, references the flow points used in the following discussion.

The equation for dynamic pressure balance from points 1 to 2 through point C is written in the following form, using perturbation techniques.

$$P_1 + \gamma_{KOH}\, h_1\Big]_o + \gamma_{KOH}\,\Delta h_1 - \Delta P_{1-c}\Big]_o - \partial_{1-c}\,\Delta\dot{\omega}_1 - C_{1-c}\,\frac{d\dot{\omega}_1}{dt} = P_2 + \gamma_{KOH}\, h_2\Big]_o$$

$$+ \gamma_{KOH}\,\Delta h_2 - \Delta P_{2-c}\Big]_o - \partial_{2-c}\,\Delta\dot{\omega}_2 - C_{2-c}\,\frac{d\dot{\omega}_2}{dt}$$

In this equation P_1 and P_2 are the pressures existing within the separator tanks, and $\Delta P_{1-c}\Big]_o$ and $\Delta P_{2-c}\Big]_o$ are the steady-state pipe pressure drops from the oxygen and hydrogen separators, respectively, to the common point C. The Δh terms are deviations in the liquid level from a nominal level $h\Big]_o$. (See Figure B-3.) The terms denoted as ∂_{1-c} and ∂_{2-c} are the $\frac{\partial P_1}{\partial \dot{\omega}_1}$ and $\frac{\partial P_2}{\partial \dot{\omega}_2}$ terms, respectively. When multiplied by the deviation in liquid velocity ($\Delta\dot{\omega}_1$ and $\Delta\dot{\omega}_2$) from the steady-state velocities, these terms result in the increase or decrease in pipe pressure drop due to flow variations. The terms C_{1-c} and C_{2-c} are $\frac{\partial P_1}{\partial \ddot{\omega}_1}$ and $\frac{\partial P_2}{\partial \ddot{\omega}_2}$; they are used to describe the change in pressure drop resulting in the acceleration of fluid in a channel of constant cross section.


from FG 62
from Σ 26
from Σ 30
To pot 50 on X 24
Used as input to optimization program
Bias
100V
ω(t) prod. sep.
ω(t) sep +1
ω(t) sep +2

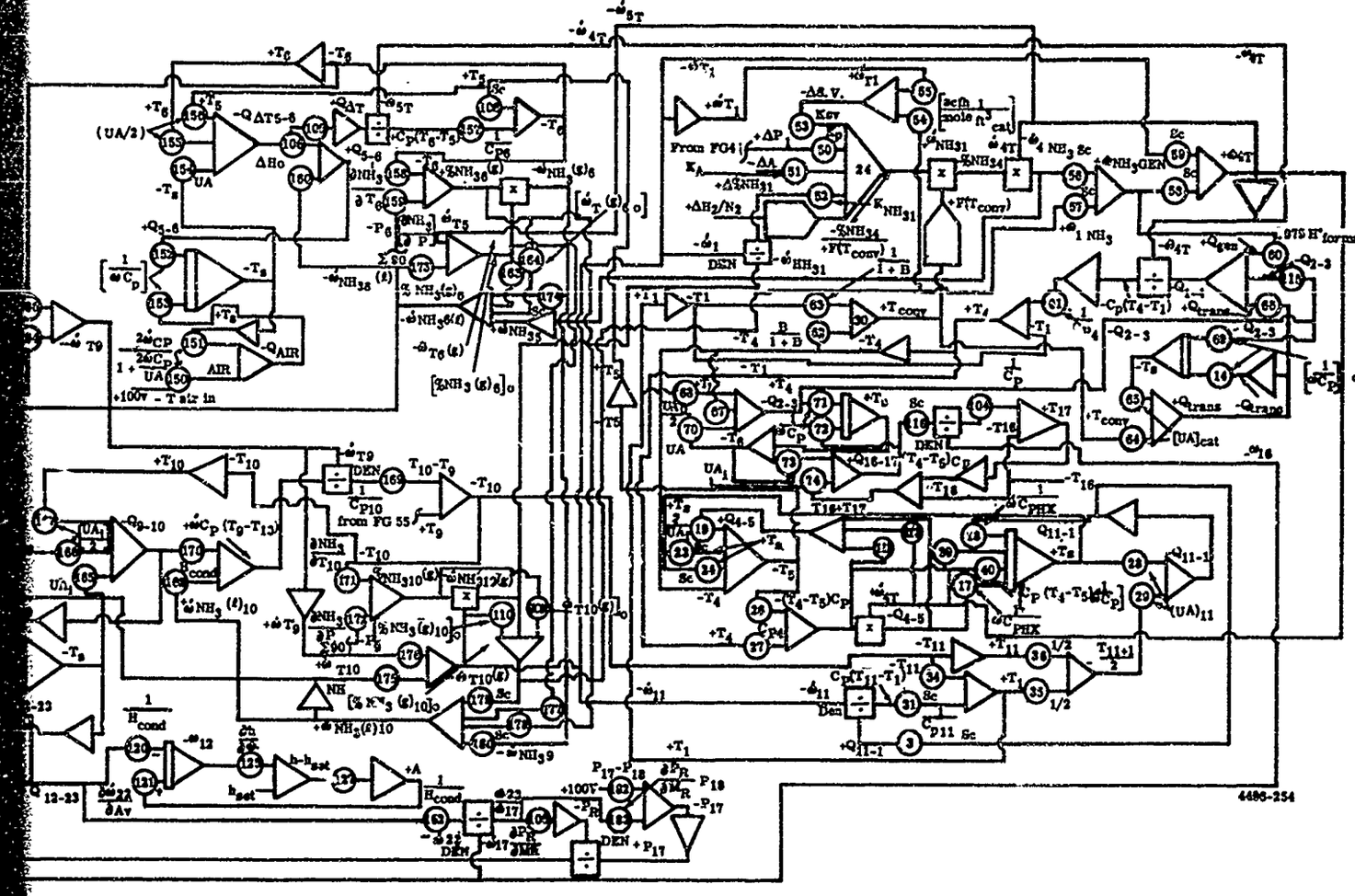


Figure B-2. Analog computer diagram for fuel production system simulation.

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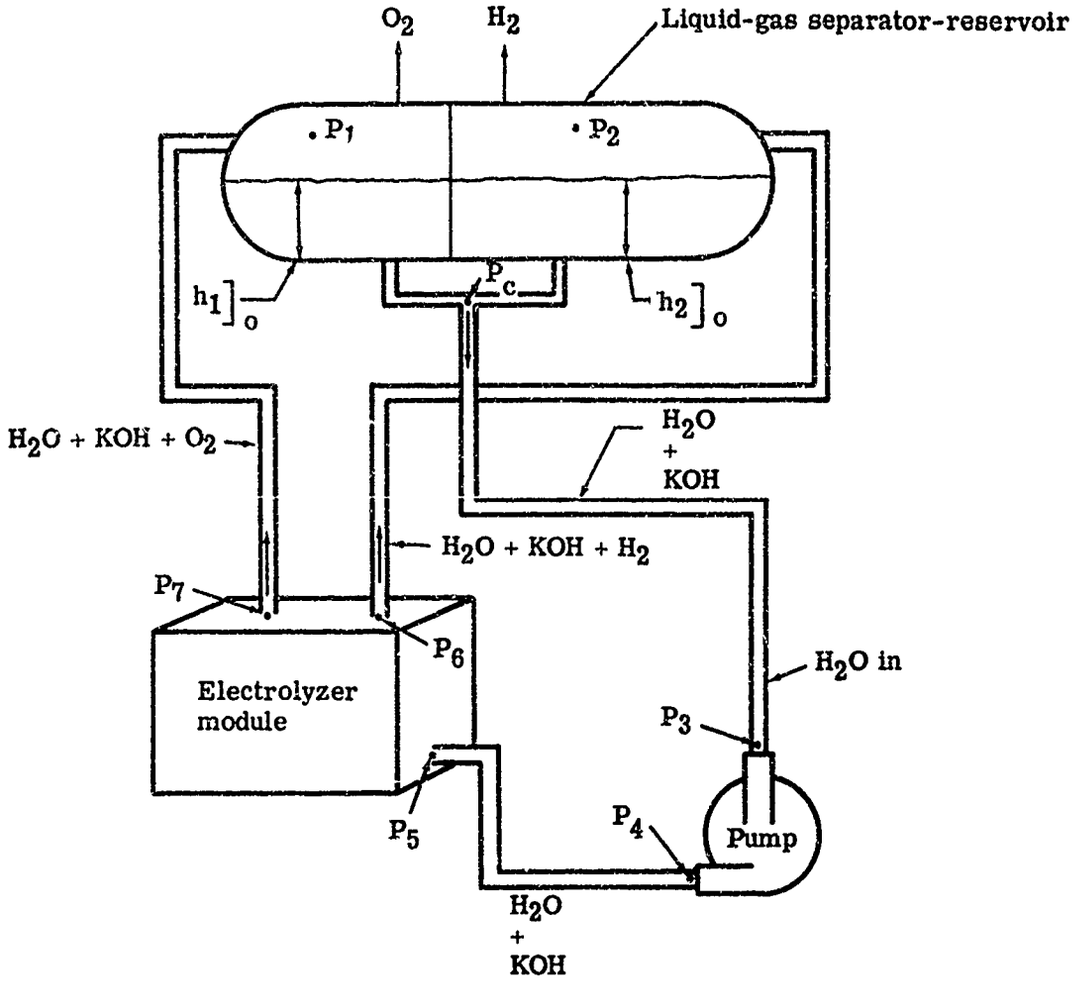


Figure B-3. Hydrogen production subsystem description points.

The C values are referred to as "Burgreen constants"[3] and are functions of the physical dimensions of the fluid container. They are derived by the following formula.

$$C_{x-y} = \frac{L}{Ag}$$

where

L = pipe or channel length, in.
A = pipe diameter or equivalent channel hydraulic diameter, $in.^2$
g = gravitational constant, 384 $in./sec^2$

The pressure in the cells, P_5, is described by the following equation.

$$P_2 + \gamma_{KOH}\, h_2\big]_o + \gamma_{KOH}\, \Delta h_2 - \Delta P_{2\text{-}c}\big]_o - \partial_{2\text{-}c}\, \Delta\dot{\omega}_2 - C_{2\text{-}c}\, \frac{d\dot{\omega}_2}{dt} - \Delta P_{c\text{-}3}\big]_o - \partial_{c\text{-}3}\, \Delta\dot{\omega}_c$$

$$- C_{c\text{-}3}\, \frac{d\dot{\omega}_c}{dt} + \Delta P_{3\text{-}4}\big]_o - \partial_{3\text{-}4}\, (\Delta\dot{\omega}_c + \Delta\dot{\omega}_{H_2O}) - \Delta P_{4\text{-}5}\big]_o - \partial_{4\text{-}5}\, (\Delta\dot{\omega}_c + \Delta\dot{\omega}_{H_2O})$$

$$-C_{4\text{-}5}\, \frac{d\,(\dot{\omega}_c + \dot{\omega}_{H_2O})}{dt} = P_5$$

The procedure required to develop the cell pressure equation is similar to that used for the previous equation. The pressure change $\Delta P_{3\text{-}4}\big]_o$ is defined as the steady-state pressure rise provided by the system pump and is therefore a positive term. It should also be noted that the water is added to the system immediately prior to the pump and thus must be considered when equating pressures from points 4 to 5.

The next two equations describe the pressure variations from the cell to the hydrogen or oxygen separator tanks.

$$P_5 - \Delta P_{5\text{-}2}\big]_o - \partial_{5\text{-}2}\, \Delta\dot{\omega}_6 - C_{5\text{-}2}\, \frac{d\dot{\omega}_6}{dt} = P_2$$

$$P_5 - \Delta P_{5\text{-}1}\big]_o - \partial_{5\text{-}1}\, \Delta\dot{\omega}_7 - C_{5\text{-}1}\, \frac{d\dot{\omega}_7}{dt} = P_1$$

Thus, the pressure variations are described at all points within the liquid flow paths. The following equations are required to complete the liquid flow description.

$$\dot{\omega}_c = \dot{\omega}_1 + \dot{\omega}_2$$

$$\frac{d\dot{\omega}_c}{dt} = \frac{d\dot{\omega}_1}{dt} + \frac{d\dot{\omega}_2}{dt}$$

$$\Delta\omega_1 = \int (\Delta\dot{\omega}_7 - \Delta\dot{\omega}_1)\, dt$$

where

$\Delta\omega_1$ = change in liquid mass in KOH-C_2 separator, lb_m
$\Delta\dot{\omega}_7$ = mass rate entering separator, lb_m/sec
$\Delta\dot{\omega}_1$ = mass rate leaving separator, lb_m/sec

$$\Delta\omega_2 = \int (\Delta\dot{\omega}_6 - \Delta\dot{\omega}_2)\, dt$$

$$h_1 = h_1\Big]_o + \frac{\partial h_1}{\partial \omega_1}\ \Delta\omega_1$$

where

h_1 = actual liquid level in KOH-O_2 separator tank, in.
$h_1\Big]_o$ = nominal steady-state liquid level, in.

$$h_2 = h_2\Big]_o + \frac{\partial h_2}{\partial \omega_2}\ \Delta\omega_2$$

The dynamic conditions related to the oxygen and hydrogen gases in their respective separators are now described. The volume of the gas is assumed to be calculated from the system operating conditions and the equation of state, PV = ωRT. Any deviation in available gas volume will be inversely proportional to the liquid mass in the separator. Thus, the gas volume may be expressed by

$$V_1 = V_1\Big]_o + \frac{\partial V_1}{\partial \omega_1}\ \Delta\omega_1$$

where $V_1\Big]_o = \dfrac{\omega_{O_2} R T_1}{P_1}$ and $\dfrac{\partial V_1}{\partial \omega_1}$ is the partial derivative of gas volume with respect to liquid mass in the tank. Also, using similar reasoning

$$V_2 = V_2\Big]_o + \frac{\partial V_2}{\partial \omega_2}\ \Delta\omega_2$$

The gas pressure in the tank is expressed by considering a steady-state pressure and the perturbation of variables affecting the pressure. Thus, the oxygen gas pressure, P_1, is expressed by

$$P_1 = P_1\Big]_o + \frac{\partial P_1}{\partial \omega_{O_2}}\Delta\omega_{O_2} + \frac{\partial P_1}{\partial V_1}\ \Delta V_1 + \frac{\partial P_1}{\partial T_1}\ \Delta T_1$$

where $P_1\Big]_o$ is the steady-state oxygen pressure. The partial derivatives in this expression are found by considering the equation of state as affected by small perturbations.

$$(P_o + \Delta P)(V_o + \Delta V) = (\omega_o + \Delta\omega)\, R\, (T_o + \Delta T)$$

and

$$\frac{\partial P_1}{\partial \omega_{O_2}} = \frac{RT_1}{V_1}$$

$$\frac{\partial P_1}{\partial V_1} = \frac{\omega_1 RT_1}{V_1^2}$$

$$\frac{\partial P_1}{\partial T_1} = \frac{\omega_1 R}{V_1}$$

By similar development, the hydrogen gas pressure perturbation is expressed by

$$P_2 = P_2\Big]_o + \frac{\partial P_2}{\partial \omega_{H_2}} \Delta\omega_{H_2} + \frac{\partial P_2}{\partial V_2} \Delta V_2 + \frac{\partial P_2}{\partial T_2} \Delta T_2$$

which completes the equations required to describe the oxygen and hydrogen gas behavior.

An important capability of analog equipment is the ability to simulate the time required for a particle to be transported from one point to another. In the electrolysis simulation, the time required for a gas bubble formed in an electrolyzer cell to emerge from the cell is significant when compared to other transport times in the system. Using the flow pattern shown in Figure B-4, for reference, an analysis can be made of the delay required. Electrolyte enters the bottom of the cathodic side of an electrolytic cell at a velocity on the order of 0.03 to 0.06 ft/sec. As the electrolyte moves upward, hydrogen gas is produced by electrolysis, accelerating the electrolyte because of the decreased flow volume available. At the top of the cell the flow velocity will have increased by about 15% which is the void fraction* of the gas produced. Thus, as shown on Figure B-4, an "average bubble" (defined as a bubble produced at the midpoint of the cell) would require approximately 12 to 23 sec to reach the cell exit.

To describe the gas produced by electrolysis, the equivalent amount of gas produced from water by electrolysis may be determined. The volume of gas produced in the cells may be expressed by balancing the quantities entering and leaving the cells Thus

$$V_5 = \int \left[\frac{1}{\rho_{KOH}} \Delta\dot{\omega}_c + \frac{1}{\rho_{H_2O}} \Delta\dot{\omega}_{H_2O} - \frac{1}{\rho_{KOH}} \Delta\dot{\omega}_6 - \frac{1}{\rho_{KOH}} \Delta\dot{\omega}_7 - \frac{1}{\rho_{H_2}} \Delta\dot{\omega}_{EG} \right] dt$$

where ΔV_5 is the volume of gas in the cells and all other terms have been defined previously except $\Delta\dot{\omega}_{EG}$. The quantity $\Delta\dot{\omega}_{EG}$ is the equivalent amount of hydrogen gas produced from

*Void fraction is $\frac{V_{gas}}{V_{KOH} + V_{gas}}$.

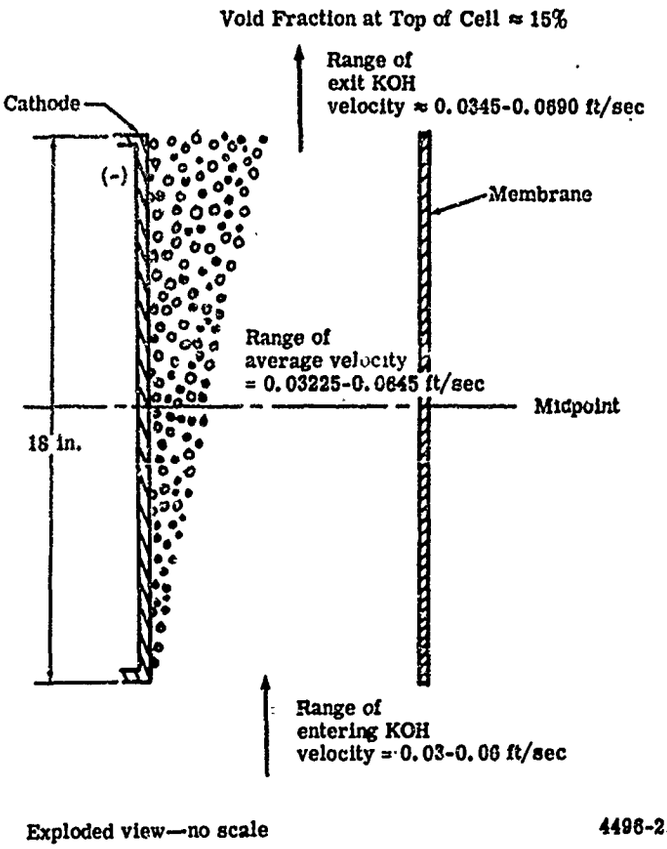


Figure B-4. Electrolytic cell gas and liquid flow pattern.

water by electrolysis if all the gas produced were hydrogen instead of one-third, on a volume basis, being oxygen. Thus the actual amount of hydrogen produced is $2/3 \times \Delta\dot{\omega}_{EG}$, while the actual amount of oxygen produced is $1/3 \times \Delta\dot{\omega}_{EG}$.

The equivalent mass of gas appearing in the electrolytic cells at any time can thus be expressed, using the time delay previously discussed.

$$\Delta\dot{\omega}_5 = \int \left[\Delta\dot{\omega}_{EG}(t) - \Delta\dot{\omega}_{EG}(t - \tau)\right] dt$$

where τ is the transport delay time. Next, the pressure existing in the cells may be expressed.

$$P_5 = P_5\Big]_o + \frac{\partial P_5}{\partial \omega_5} \Delta\omega_{EG} + \frac{\partial P_5}{\partial V_5} \Delta V_5 + \frac{\partial P_5}{\partial T_5} \Delta T_5$$

Finally, by using the time delay, the mass of oxygen and hydrogen in the separator tanks may be expressed by differing the masses entering and leaving each tank.

$$\Delta\omega_{O_2} = \int \left[\dot{\omega}_{O_2}(t-\tau) - \dot{\omega}_{O_2}(t)\right] dt$$

$$\Delta\omega_{H_2} = \int \left[\dot{\omega}_{H_2}(t-\tau) - \dot{\omega}_{H_2}(t)\right] dt$$

The electrolysis unit has now been described during dynamic operation. A control mode must be chosen before analog simulation can be accomplished.

Control Equations

The controls for the hydrogen production subsystem must provide regulation for the water input, the liquid level in both the oxygen and hydrogen separator tanks, and the operating pressure level.

The pressure level is controlled by a servo-driven valve which acts on the efflux of hydrogen gas from the separator-reservoir. The control is located on the ammonia package and recirculates synthesis gas around the synthesis gas compressor to maintain hydrogen pressure.

The liquid level in the hydrogen side of the separator-reservoir is controlled by adding water to the electrolyte flow as required. The liquid level is sensed in the hydrogen separator-reservoir and a low-limit switch operates a feedwater pump to initiate water flow when a predetermined low liquid level limit is reached. A high-limit switch stops the feedwater pump when the high level limit is reached.

The liquid level in the oxygen side of the separator-reservoir is controlled by sensing the differential level between the two sides of the separator, h_2-h_1, and actuating a valve controlling oxygen efflux from the tank. This valve responds to the level differential.

Additional controls, which the system would use but did not require simulation, include mechanically operated absolute high- and low-limit valves on each separator tank.

The rate at which the oxygen is released from the separator tanks was determined by considering perturbations from a steady-state flow rate across the valve from the liquid-gas separator tank.

$$\dot{\omega} = \dot{\omega}\Big]_o + \frac{\partial\dot{\omega}}{\partial P}\,\Delta P + \frac{\partial\dot{\omega}}{\partial A}\,\Delta A + \frac{\partial\dot{\omega}}{\partial T}\,\Delta T$$

The partial derivatives required were found by considering the valve to allow isentropic flow and behave according to the equation[4]

$$\dot{\omega} = 223.7\,\frac{P_1 A_{valve}}{R}\left\{\frac{C_p}{T}\left[\left(\frac{P_2}{P_1}\right)^{2/k} - \left(\frac{P_2}{P_1}\right)^{(k+1)/k}\right]\right\}^{1/2}$$

where

$\dot{\omega}$ = flow rate, lb/sec
P_1 = upstream pressure, psia
P_2 = downstream pressure, psia
A = valve operating flow area, $in.^2$
R = gas constant
C_p = gas specific heat, $BTU/lb_m\text{-}°R$
T = operating temperature, °R
k = specific heat ratio

where the partial derivative of flow with respect to pressure would be the upstream pressure, P_1.

The control equation for a proportional-plus-reset valve has the following general form.[5]

$$m = Se + SI \int_0^t e\, dt$$

where

m = change in output variable
e = input variable
S = proportional sensitivity
I = reset rate

The oxygen valve controls the liquid level differential, $h_2\text{-}h_1$, by changing the valve area, A_o. Therefore the oxygen valve equation is

$$A_o = S\,(h_2\text{-}h_1) + SI \int_0^t (h_2\text{-}h_1)\, dt$$

where S is the proportional gain required for control ($in.^2/in.$) and the product SI is the integral gain required ($in.^2/in.\text{-}sec$). For the hydrogen valve, the gas pressure in the hydrogen separator tank, P_2, is kept at a predetermined level, P_{set}, by changing the valve area, A_H. The hydrogen valve equation is

$$A_H = S\,(P_{set} - P_2) + SI \int_0^t (P_{set} - P_2)\, dt$$

where S is the proportional gain ($in.^2/psi$) and SI is the integral gain ($in.^2/psi\text{-}sec$).

The equations were rearranged for the simulation resulting in the analog computer diagram shown in Figure B-5. The numbers on the diagram refer to amplifier numbers and potentiometers used for the simulation. The feedwater control circuit is operated by an on-off relay as shown in the lower left-hand corner of the diagram. The hydrogen and oxygen efflux valve circuits are shown on the right-hand side of the diagram.

The analog computer program developed for the hydrogen production subsystem was used to determine the steady-state stability and transient response for systems 1, 2, 3, and 11. The result of this study was used to ensure the adequacy of the lumped models used in the fuel production system simulation.

Variations in system parameters were determined for steady-state operation and for response to a step change in power supplied to the hydrogen production subsystem.

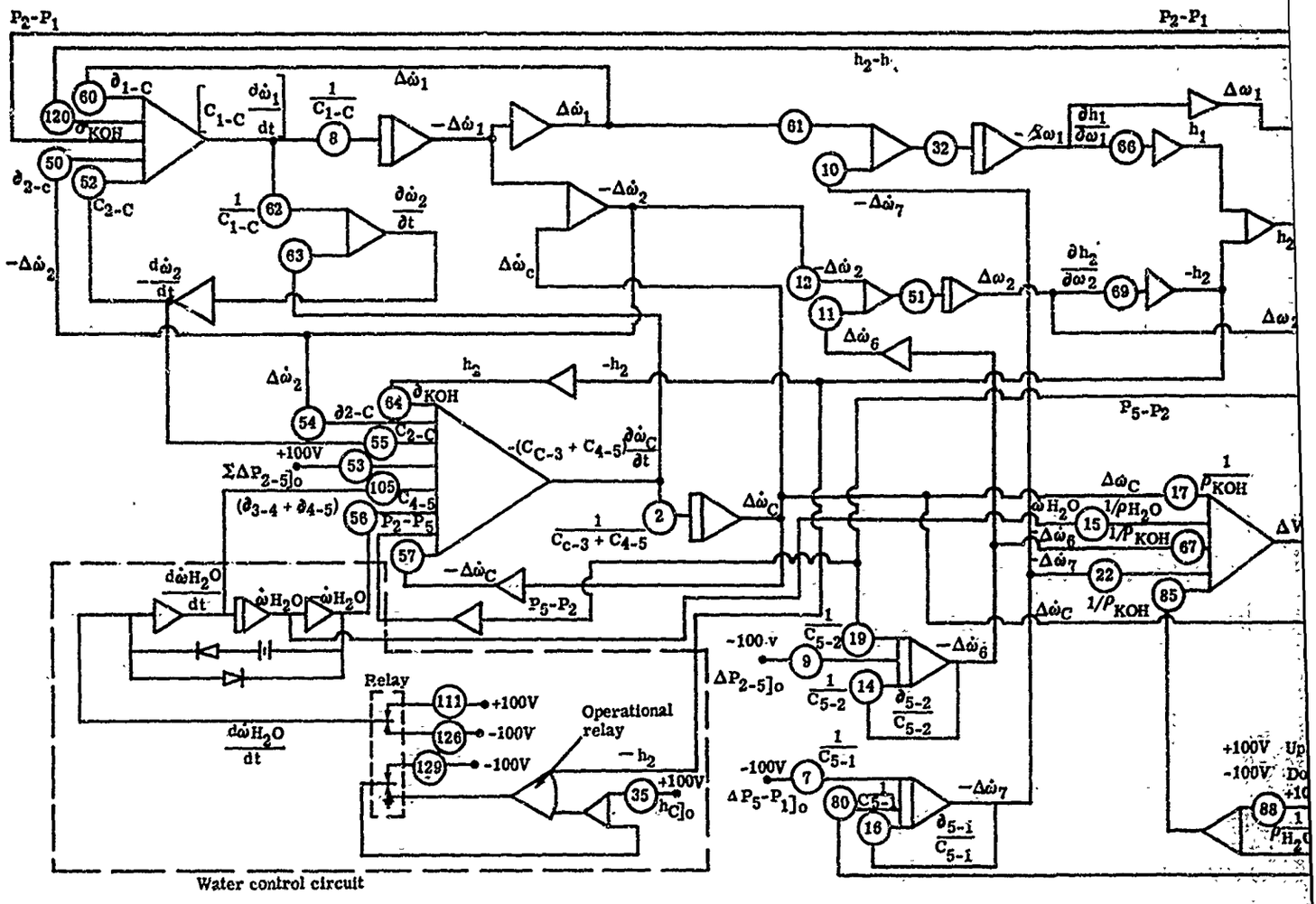
The data shown in Figure B-6 indicate normal hydrogen production subsystem operation at a hydrogen gas pressure of 315 psia for systems 1, 2, and 3. Hydrogen gas is produced at a constant rate. The liquid level in the separator-reservoir decreases as H_2O is consumed by electrolysis. The level decreases at a rate of 0.23 in./min. When the level reaches a preset lower limit, the feedwater pump is started and the level rises until a preset high limit is attained. Feedwater flow, off-on actuation, was set to occur at a deviation of ± 1.0 in. from the normal operating level of the separator tank. The maximum and minimum steady-state liquid levels are determined by the level deviation caused by the changes in void fraction which occur at different power levels. However, the level deviation shown by the data does illustrate the degree of control attainable. Feedwater flow rate was set at 1.0 lb H_2O/sec. At this rate, level increases by 0.65 in./min.

The response of the oxygen efflux valve, which controls the differential in liquid level between the hydrogen and oxygen separator-reservoirs is shown in Figure B-6(A).

The variation of liquid level in the hydrogen separator-reservoir is shown in Figure B-6(B). The level deviates 1 in. from the tank steady-state level before the feedwater valve is actuated. Level variation is linear with respect to time.

The differential in gas pressure between the hydrogen and oxygen separator-reservoir, P_2-P_1, is shown in Figure B-6(C). The pressure differential is essentially zero, indicating differential level control is adequate for controlling differential pressure and thus simplifying the control system.

The differential in liquid level between the hydrogen and oxygen separator-reservoir, h_2-h_1, which is controlled by the oxygen efflux valve is shown in Figure B-6(D). The negligible differential in liquid level indicates that satisfactory control is achieved.


Water control circuit
Relay
Operational relay
+100V
-100V
-100V
+100V
-100V
+100V
Up
Down

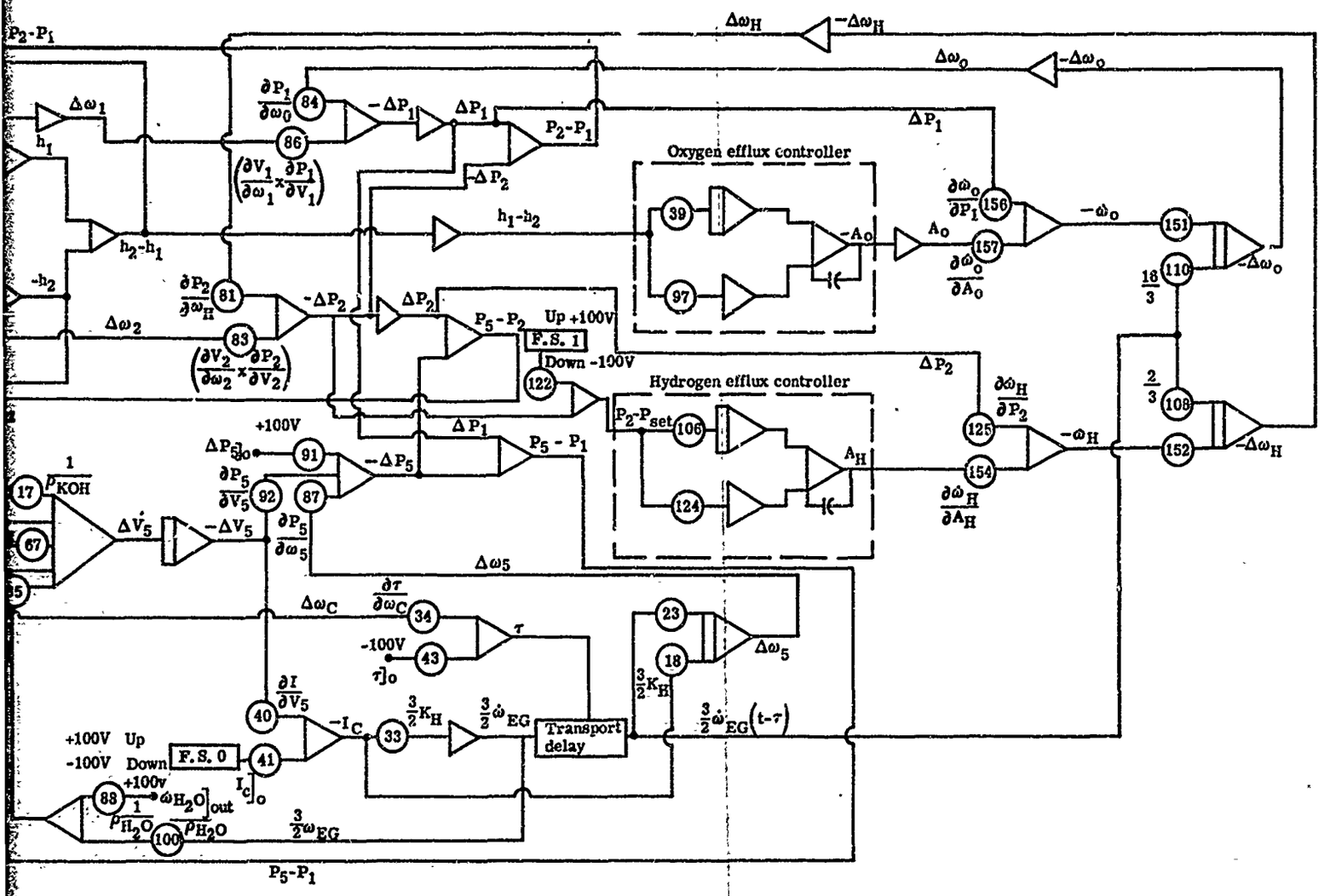


Figure B-5. Analog computer diagram for Energy Depot electrolysis unit.

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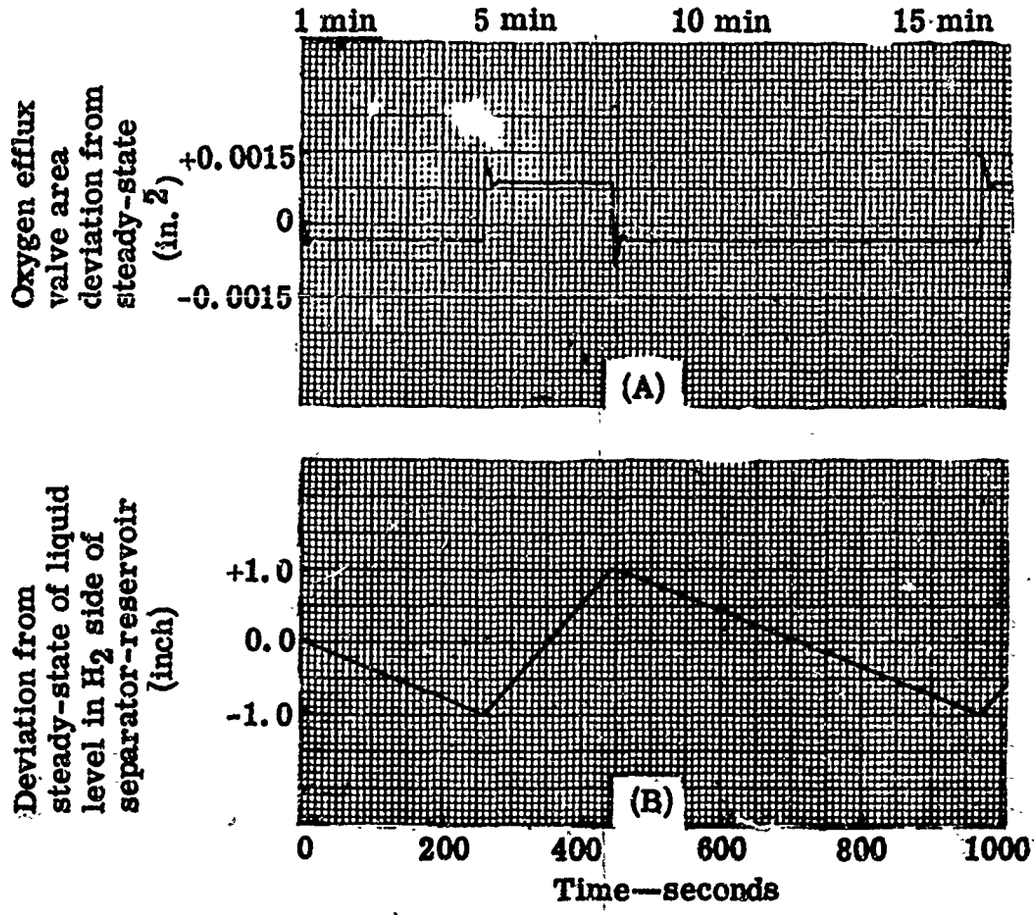


Figure B-6. Hydrogen production subsystem steady-state operating performance—systems 1, 2, and 3.

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The hydrogen gas pressure is controlled by the synthesis gas recirculation valve. Good control is achieved. The small pressure increase, which occurs when the feedwater pump is started, is caused by the time required for the valve to respond.

An electrical power input perturbation is important in analyzing dynamic performance of the hydrogen production subsystem because changing power will perturb the production rate of hydrogen and oxygen gas.

To determine the response of an increase in hydrogen production, a 40% increase in electrolyzer current from 80 to 120% of design was simulated. The results of this transient are shown in Figure B-7.

The increase in electrolyzer current increases the hydrogen and oxygen production rate. The volume of gas in the electrolyzer and the electrolyte lines to the separator-reservoirs increases. Initially, the liquid level in the separator-reservoir and the gas pressure in the separator-reservoir increase. The hydrogen gas pressure is quickly reduced with the maximum deviation being less than 4 psi (Figure B-7(E)). The oxygen efflux valve increases the oxygen flow rate and maintains the liquid level differential to within 0.1-in. (Figure B-7(D)). The liquid levels in the separator-reservoirs continue to increase for approximately 23 sec, the average time for a gas bubble to reach the separator-reservoir after formation in the electrolyzer (Figure B-7(B)). The maximum deviation in level is approximately 5 in., 4 in. above the level at initiation of transient.

Except for the liquid levels in the separator-reservoirs, all parameters reach new, stable operating points within 30 sec. The liquid levels are decreasing at a normal rate, approximately 0.41 in./min, as water is being consumed in electrolysis. A new steady-state liquid level is established at the new power level, and makeup water addition is governed at the new operating level.

Steady-state operation, subsystem response to a step change in power supplied to the subsystem, and response to a decrease in subsystem pressure for the hydrogen production subsystem (system 11) are shown in Figures B-8, B-9, and B-10.

Figure B-8 indicates performance of the hydrogen subsystem (system 11) under normal operating conditions. Subsystem operation is similar to that shown for systems 1, 2, and 3 shown in Figure B-6.

The feedwater pump cycles on and off to maintain the liquid level in the separator-reservoir within 1 in. of the desired level. Cycling of the feedwater flow induces minor oscillations in the hydrogen pressure and oxygen flow which are quickly damped.

Figure B-9 indicates the hydrogen subsystem (system 11) response to a decrease in power input to the hydrogen subsystem electrolyzer. Except for the liquid level in the separator-reservoir, all subsystem parameters stabilize at new steady-state values within approximately 3 min. The liquid level in the separator-reservoir drops approximately 5 in. Approximately 15 to 20 min are required to return the liquid level to the normal control range. The damped oscillations in the cell pressure and the differential in liquid level between the hydrogen and oxygen separator-reservoirs indicate that further optimization of the control settings is desirable.

Figure B-10 indicates system response to a 50-psi decrease in hydrogen production subsystem pressure such as would occur with a broker hydrogen line between the hydrogen package and ammonia package. In this situation the hydrogen valve on the liquid-gas separator tanks controls hydrogen pressure at 265 psia, 50 psi below normal subsystem operating pressure. The parameters shown indicate that good control is achieved. Initially, the liquid level in the liquid-gas separator-reservoir (Figure B-10b) rises because of the increased void fraction caused by the lower system pressure. In approximately 90 sec, the liquid level returns to the normal control range.

NITROGEN GENERATION SUBSYSTEM ANALOG COMPUTER PROGRAM

The nitrogen generation subsystem was represented on the analog computer in perturbation form. This approach to programming the subsystem offers the advantage of accurate simulation which makes use of the steady-state analysis of the system operating point. In addition, the perturbation representation reduces the complexity of the representation so that the simulation can be achieved with a reasonable number of computer components. The nitrogen generation subsystem was represented by simulating the major components which are germane to the subsystem dynamics. (See Figure 3.6-1 in Section III.) These components are:

- The main heat exchanger (NE-1) which cools the incoming air with nitrogen and waste gas streams from the low pressure column
- The refrigerated heat exchanger (NE-2) which further cools and liquefies the incoming air (The heat is removed by the nitrogen and waste gas stream from the low pressure column and the refrigeration supplied by the waste oxygen from the hydrogen production subsystem.)
- The oxygen reversing heat exchanger (NE-3) which cools the incoming oxygen from the hydrogen production subsystem with the discharge oxygen after it has been expanded and passed through the refrigerated heat exchanger
- The expansion turbine for the oxygen from the hydrogen production subsystem which provides the additional refrigeration needed in the air fractionation process
- The high pressure air fractionation column (NT-1) which separates part of the nitrogen in the process air from the argon and oxygen
- The low pressure air fractionation column (NT-2) which extracts additional nitrogen from the incoming air

0
1 min
2 min
+0.03
0
-0.03
Oxygen efflux valve area varia 'on
(in.^{2})
(A)
4.0
2.0
0.0
-2.0
-4.0
Deviation from steady-state of liquid level in H_2 side of separator-reservoir
(inch)
(B)
0
50
100
Transient initiation
Time—seconds

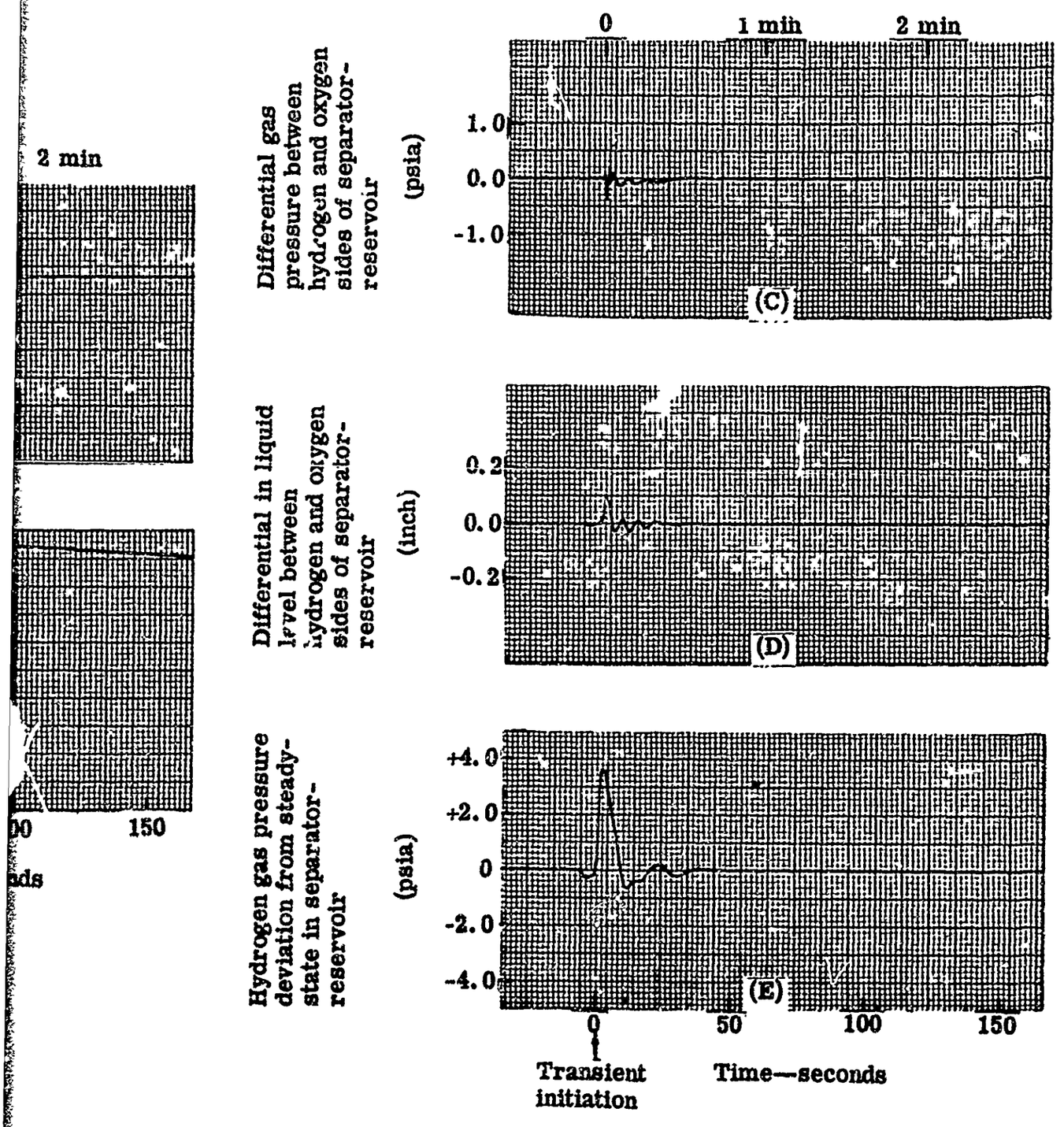


Figure B-7. Hydrogen production subsystem perturbation due to an increase in electrical power input from 80 to 120% of design rating—systems 1, 2, and 3.

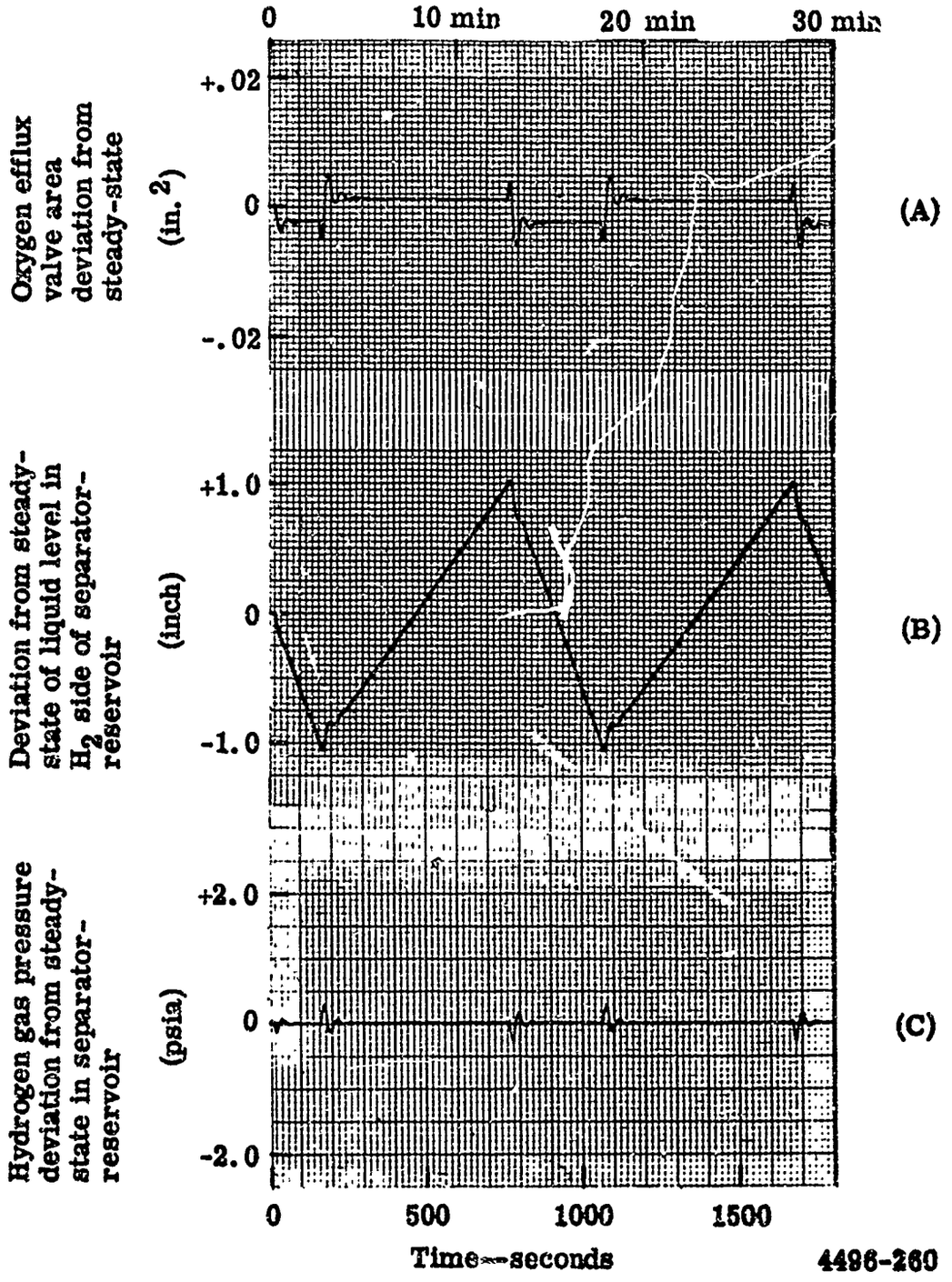


Figure B-8. Hydrogen production subsystem steady-state performance—system ii.

These major components are coupled together by controls. The controls were simulated as an integral part of the control component.

The mathematical equations which were used to represent the heat exchangers are of the same form used for heat exchanger representation for the start-up simulations. The air fractionation column perturbation equations follow.

For the two trays lumped as one effective tray (used for the high pressure column):

$$2H \frac{d X_n}{dt} = \Delta V (Y_{o\, n-2} - Y_{o\, n}) + \Delta L (X_{o\, n+2} - X_{o\, n})$$

$$- \Delta X_n (L_o + \alpha^*_{n \text{ to } n+2} V_o) + L_o \Delta X_{n+2} + \alpha^*_{n-2 \text{ to } n} V_o \Delta X_{n-2}$$

where

- H = the liquid held by each tray, lb-moles
- ΔV = the change in flow rate of the vapor up the column, lb-moles/sec
- L_o = the steady-state design liquid flow rate down the column, lb-moles/sec
- ΔL = the change in liquid flow rate, lb-moles/sec
- Y_o = the design concentration of the more volatile product (nitrogen) leaving the n^{th} tray
- X_o = the design concentration of nitrogen in the liquid leaving the n^{th} tray in parts
- ΔX = the change in this concentration from design point
- α^* = the ratio between change in the concentration of nitrogen in the vapor above an effective tray and the nitrogen concentration in the tray.

The low pressure column was represented in the same manner as the high pressure column except that three trays were lumped as one effective tray. The column digital computer design information (discussed in Nitrogen Generation Subsystem portion of Section III) was used to establish the steady-state values (X_o, Y_o, and α^*) for the columns.

The basic tray equations were modified for feed and reboiler trays. For the high pressure column reboiler and feed tray, since none of the vapor boils from this reboiler

$$H_R \frac{d X_R}{dt} = \Delta L_o X_{2\, o} + L_o \Delta X_2 - X_{o\, R} \Delta L_R - L_{o\, R} \Delta X_R + X_{fo} (q_o \Delta \dot{\omega}_a + \dot{\omega}_{ao} \Delta q)$$

where

- q_o = the ratio of liquid to total air input
- $\dot{\omega}_a$ = the total flow of air into the column, lb-moles/sec

Subscript R is used for the reboiler and f for feed

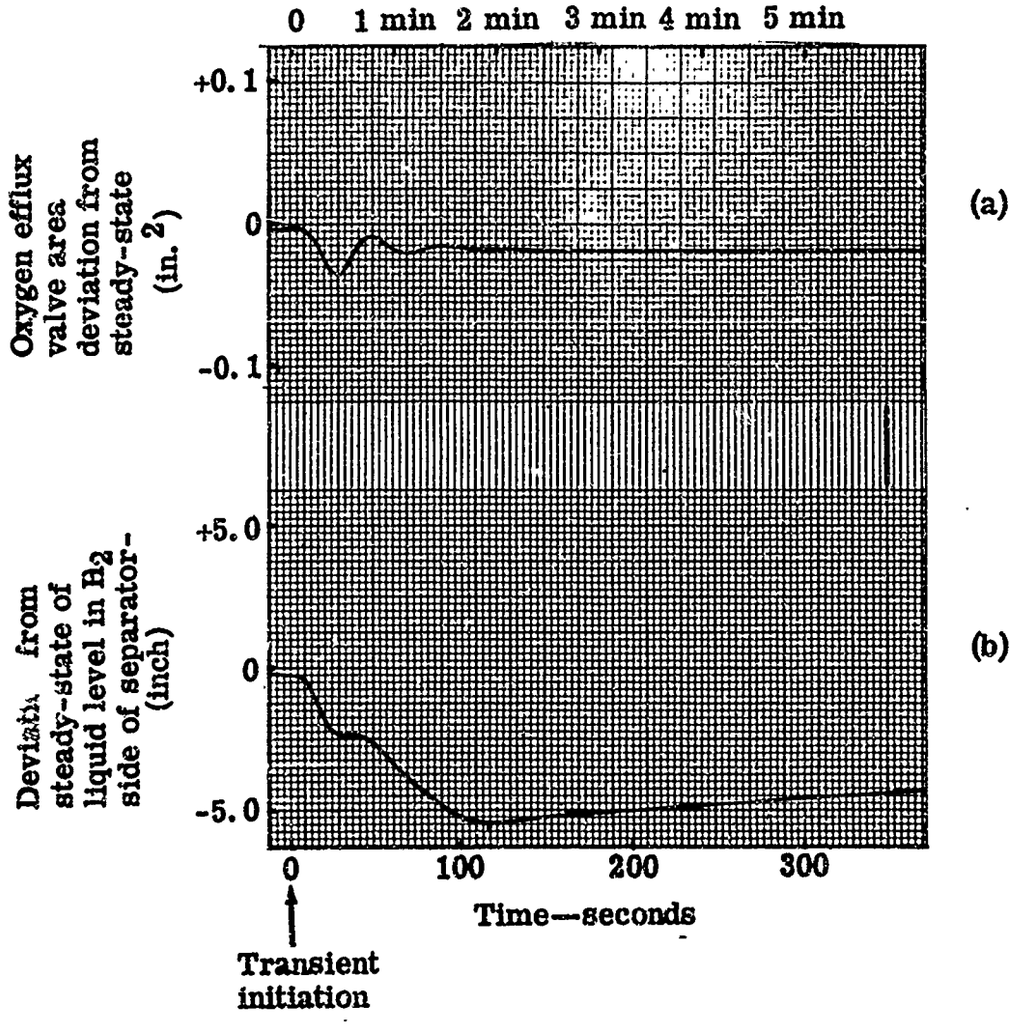

0
1 min
2 min
3 min
4 min
5 min
+0.1
0
-0.1
+5.0
0
-5.0
Oxygen efflux valve area deviation from steady-state (in.²)
Deviation from steady-state of liquid level in H2 side of separator (inch)
(a)
(b)
0
100
200
300
Time—seconds
Transient initiation

Figure B-2

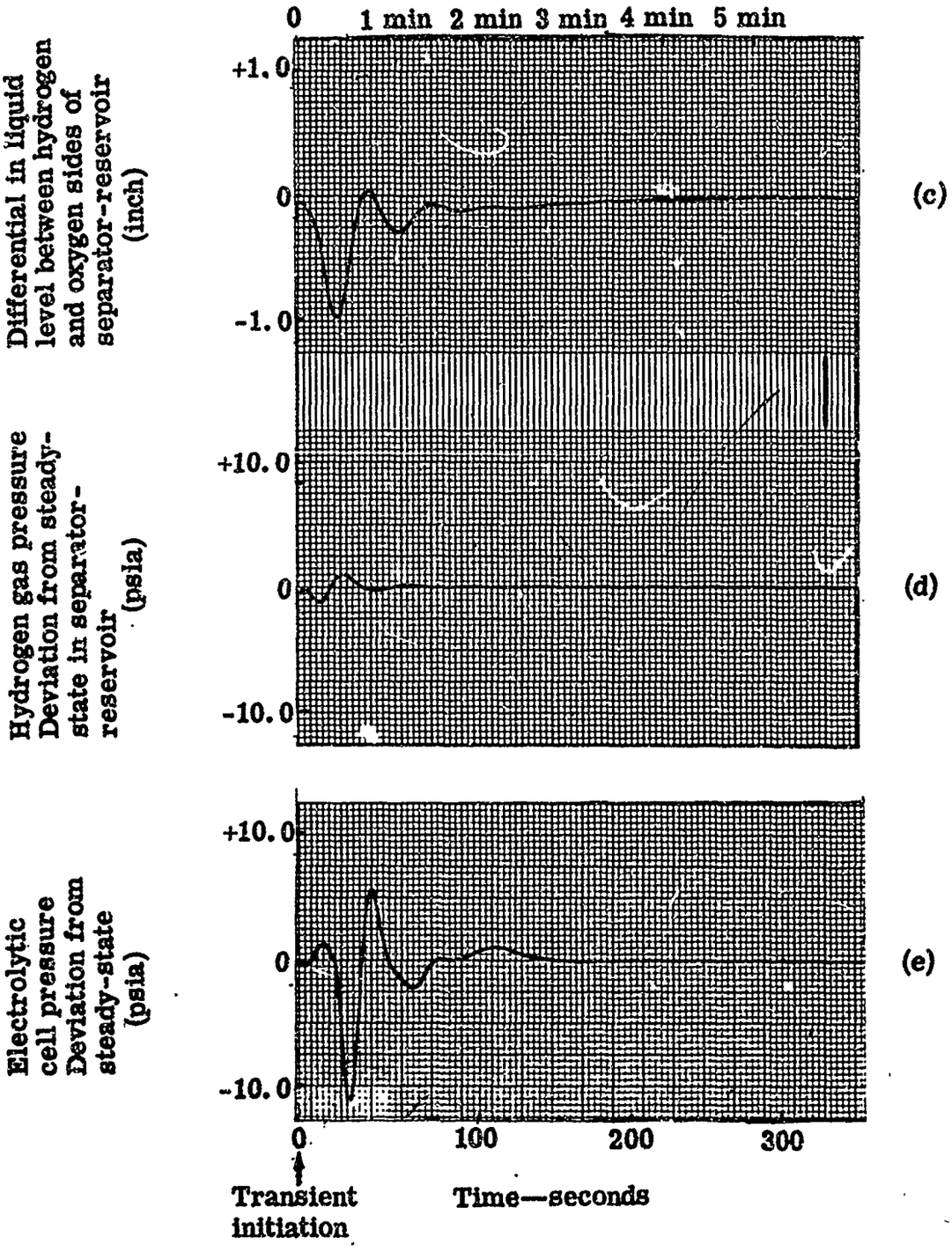


Figure B-9. Hydrogen production subsystem power decrease from 100 to 90% of design level—system 11.

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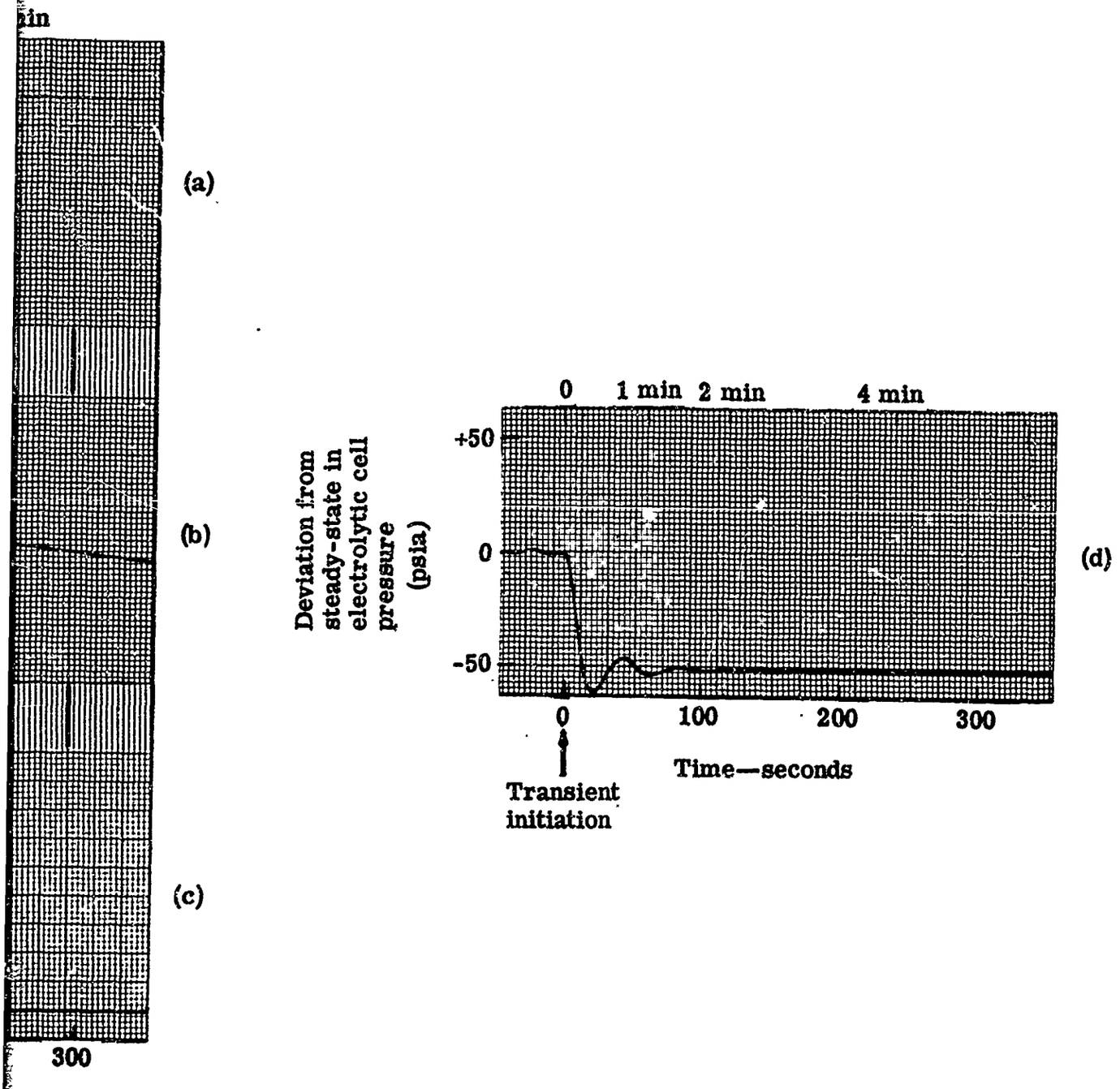


Figure B-10. Hydrogen production subsystem perturbation due to a 50-psi decrease in system pressure—system 11.

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The level in the high pressure column reboiler is regulated by the amount removed to the high pressure column condenser cooler. The equation for the change in level in the reboiler is

$$\Delta l = \frac{1}{\rho} \frac{\partial l}{\partial vol} \int (q_o \Delta \dot{\omega}_a + \dot{\omega}_a \Delta q + \Delta L - \Delta L_R)\, dt$$

where

ρ = the density of the liquid in the reboiler, lb moles/in.^{3}
Δl = the change in reboiler liquid level, in.

The high pressure column condenser equation is as follows.

$$H_C \frac{d X_C}{dt} = \Delta X_{12} \alpha^* V_o + \Delta V Y_{12}\Big]_o - (\Delta L_c + \Delta V_D) X_{Co} - (\bar{L}_{Co} + \bar{V}_{Do}) \Delta X_C$$

where

Subscript C refers to the condenser
Subscript D refers to the overhead from the column

For the low pressure columns the equations which were rewritten for the reboiler, condenser, and feed trays are as follows. For the reboiler

$$H_{R_L} \frac{d X_{R_L}}{dt} = \Delta X_4 L_o + \Delta L X_4\Big]_o - (\Delta V + \Delta V_{W_G}) Y_{Ro} - \alpha^* \Delta X_R (V_o - V_{W_{G_o}})$$

where

Subscript R_L is the low pressure column reboiler
Subscript W_G refers to the waste gas

The equation for the lower feed tray is changed by the addition of the terms resulting from the feed.

$$\Delta V_{f1} X_{f1} \alpha^* + \Delta X_{f1} \alpha^* V_{fo}$$

$$\Delta X_{f1} L_{f1_o} + \Delta L_{f1} X_{f1\,o}$$

The subscript f1 indicates the lower feed for the low pressure column. The equation for the condenser on the low pressure column is

$$H_C \frac{d X_C}{dt} = - \Delta X_C (V_{Co} \alpha_C^* + L_{f2o}) + \Delta X_{f2} L_{f2o} + \Delta V_{f2} (Y_C - Y_{22})$$

$$+ \Delta L_{f2} (X_{fo} - X_C) + \Delta X_{22} (V \alpha^*)$$

where the subscript f2 indicates the upper feed on the low pressure column.

The expansion turbine is represented by a temperature drop as a linear function of temperature at the input to the turbine.

The temperature of the mixed streams of the oxygen expanded through and bypassed around the expansion turbine is represented by the following equation.

$$T_{39} = K_{1T} T_{38} + K_{2T} + D_T \left[(1 - K_{1T}) T_{38} - K_{2T} \right]$$

where

T_{39} = the mixed temperature out of the expansion turbine and bypass
T_{38} = the input temperature
D_T = the position of the bypass valve
K_{1T} and K_{2T} = constants determined by the expansion turbine characteristics

(The subscripts on the temperature values refer to the numbers on the process diagram, Figure 3.9-1 of Section III.)

The process controllers required were simulated in the same manner as the hydrogen subsystem controllers, as discussed previously.

$$m = Se + SI \int_0^t e \, dt$$

The liquid level in the high pressure column reboiler is controlled by the rate at which liquid from the reboiler is sent to the condenser at the top of the high pressure column. The valve equation to accomplish level control is

$$A_R = S_1 \Delta l_H + (SI)_1 \int_0^t \Delta l_H \, dt$$

where

A_R = valve area, $in.^2$
Δl_H = error in the high pressure column reboiler level, in.

S_1 = proportional sensitivity, $in.^2/in.$
$(SI)_1$ = integral gain, $in.^2/in.-sec$

The liquid level in the low pressure column reboiler is controlled by regulating the amount of oxygen which bypasses the expansion turbine. The valve equation is

$$D_T = S_2 \, \Delta l_L + (SI)_2 \int_0^t \Delta l_L \, dt$$

where

D_T = bypass valve position, % open
Δl_L = error in the low pressure column reboiler level, in.
S_2 = proportional sensitivity, %/in.
$(SI)_2$ = integral gain, %/in.-sec

The nitrogen purity is controlled by regulating the amount of nitrogen which is fed to the top of the low pressure column from the high pressure column. The equation for this controller is

$$\Delta f_2 = S_p \, \Delta X_{CLPC}$$

where

Δf_2 = change in feed rate at the top of the low pressure column, %
ΔX_{CLPC} = change in product purity, %
S_p = proportional sensitivity

Purity control is simulated as a proportional controller with no reset action because there is a range of acceptable nitrogen purity.

Figure B-11 is the analog diagram of the nitrogen generation subsystem computer program. Note that the column fractionations are represented by the simulation by the liquid concentration of the more volatile product (nitrogen) on each equivalent tray. Where practical, the group of computer components which represented a system component is labeled as representing that component.

This analog computer program was used to analyze the response of the nitrogen generation subsystem to perturbations during normal operation. Four major sources of perturbations were studied:

- A change in the oxygen flow rate from the hydrogen production subsystem
- A change in the nitrogen product draw-off rate

- A change in the feed airflow rate
- A change in the rate of the liquid nitrogen sent to storage

The analyses were made to determine the influences of these changes on six major subsystem parameters.

The nitrogen generation subsystem response to an increase from 80% to 100% in flow rate of oxygen from the hydrogen production subsystem is shown in Figure B-12. The portion of incoming air which is condensed decreases initially because the oxygen temperature into the refrigerated heat exchanger increases with increased oxygen flow. Subsequently, as the liquid level in the low pressure column drops, control action reduces the amount of oxygen which bypasses the expansion turbine, decreases the oxygen temperature into the refrigerated heat exchanger, and increases the portion of incoming air which is condensed. Subsequent changes in the portion of incoming air condensed were the result of control action. Steady-state conditions were not fully established at the end of the analog trace. The nitrogen product purity was relatively unaffected by the perturbation; control of the subsystem is adequate.

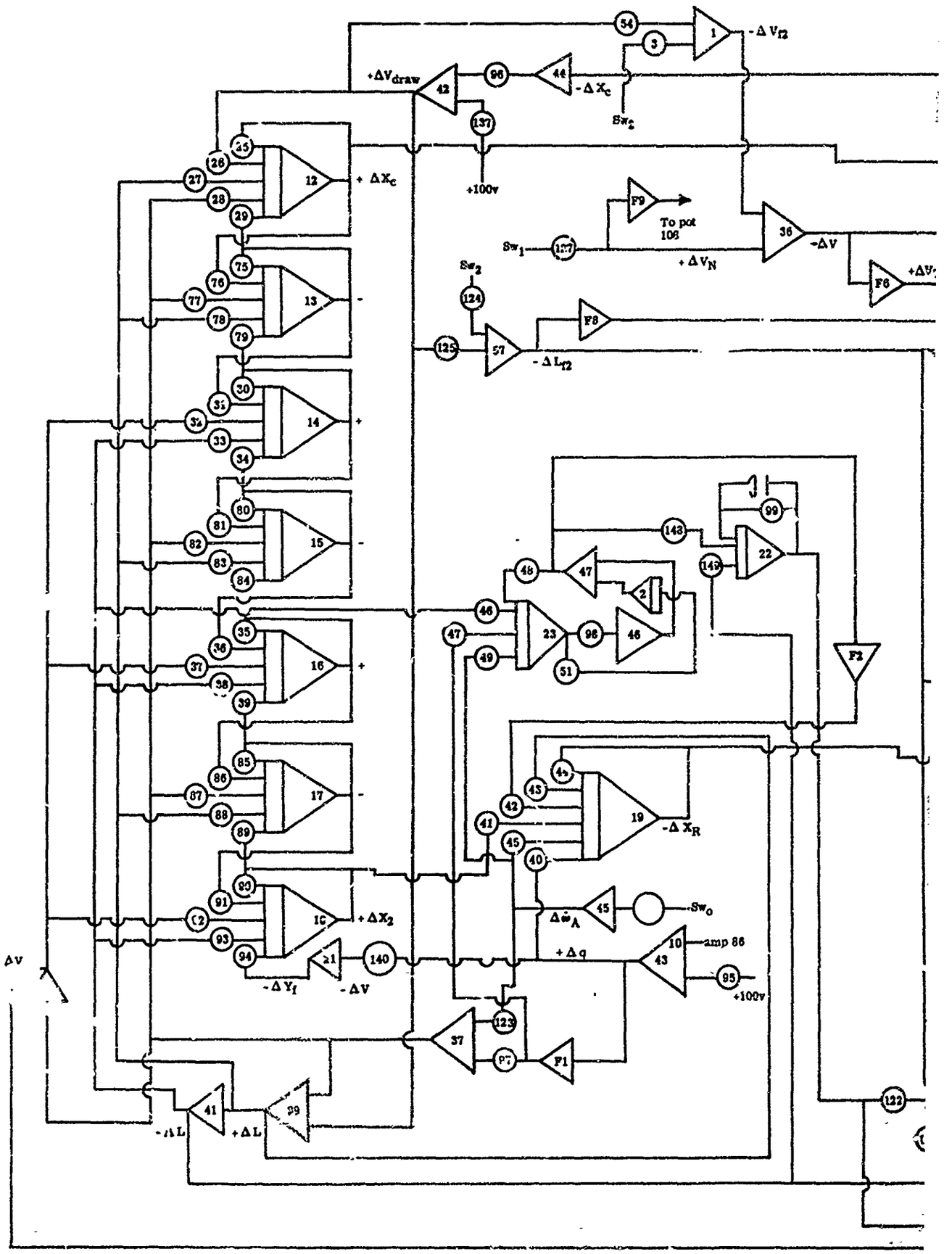
The nitrogen generation subsystem response to a decrease from 100% to 86.5% in product nitrogen draw-off rate is shown in Figure B-13. As expected, the nitrogen product purity increases with a decrease in draw-off rate. Other parameters are perturbed but are controlled within acceptable ranges.

The nitrogen generation subsystem response to a 5% increase in feed airflow rate is shown in Figure B-14. The nitrogen product purity increases. The portion of incoming air condensed decreases initially, but subsequent control action decreases oxygen temperature into the refrigerated heat exchanger and increases the condensation of incoming air.

The nitrogen generation subsystem response to stopping the withdrawal for storage of liquid nitrogen from the overhead product stream of the high pressure air fractionation column is shown in Figure B-15. The product nitrogen purity increases and finally stabilizes at the design point purity. The portion of the incoming air condensed decreases with minor control oscillations and finally stabilizes at design point. The other parameters also eventually stabilize at the design point value.

The response of the nitrogen generation subsystem to the perturbation studies is extremely slow—many minutes to several hours are required to reach a new steady-state operating condition. As a result, the product nitrogen purity is very stable and remains well above the minimum purity acceptable.

The variations described for the nitrogen subsystem are in general agreement with the responses expected based on a qualitative analysis of the operation of distillation processes. The results of the analyses serve to strengthen the validity of the model chosen for the analog simulation.


54
3
1
$-\Delta V_{f2}$
$+\Delta V_{draw}$
42
96
44
$-\Delta X_c$
Sw_2
137
+100v
25
26
27
28
29
12
$+\Delta X_c$
F9
To pot 106
Sw_1
127
$+\Delta V_N$
36
$-\Delta V$
F6
$+\Delta V_{f2}$
Sw_2
124
F8
125
57
$-\Delta L_{f2}$
75
76
77
78
79
13
74
30
31
32
33
34
14
80
81
82
83
84
15
99
143
22
145
48
47
2
46
23
98
46
49
51
F2
35
36
37
38
39
16
85
86
87
88
89
17
43
42
41
45
40
19
$-\Delta X_R$
90
91
92
93
94
18
$+\Delta X_2$
$\Delta\dot{\omega}_A$
45
$-Sw_o$
$+\Delta q$
10
43
amp 86
95
+100v
≥ 1
140
$-\Delta Y_f$
$-\Delta V$
Δv
123
37
97
F1
122
41
$-\Delta L$
39
$+\Delta L$

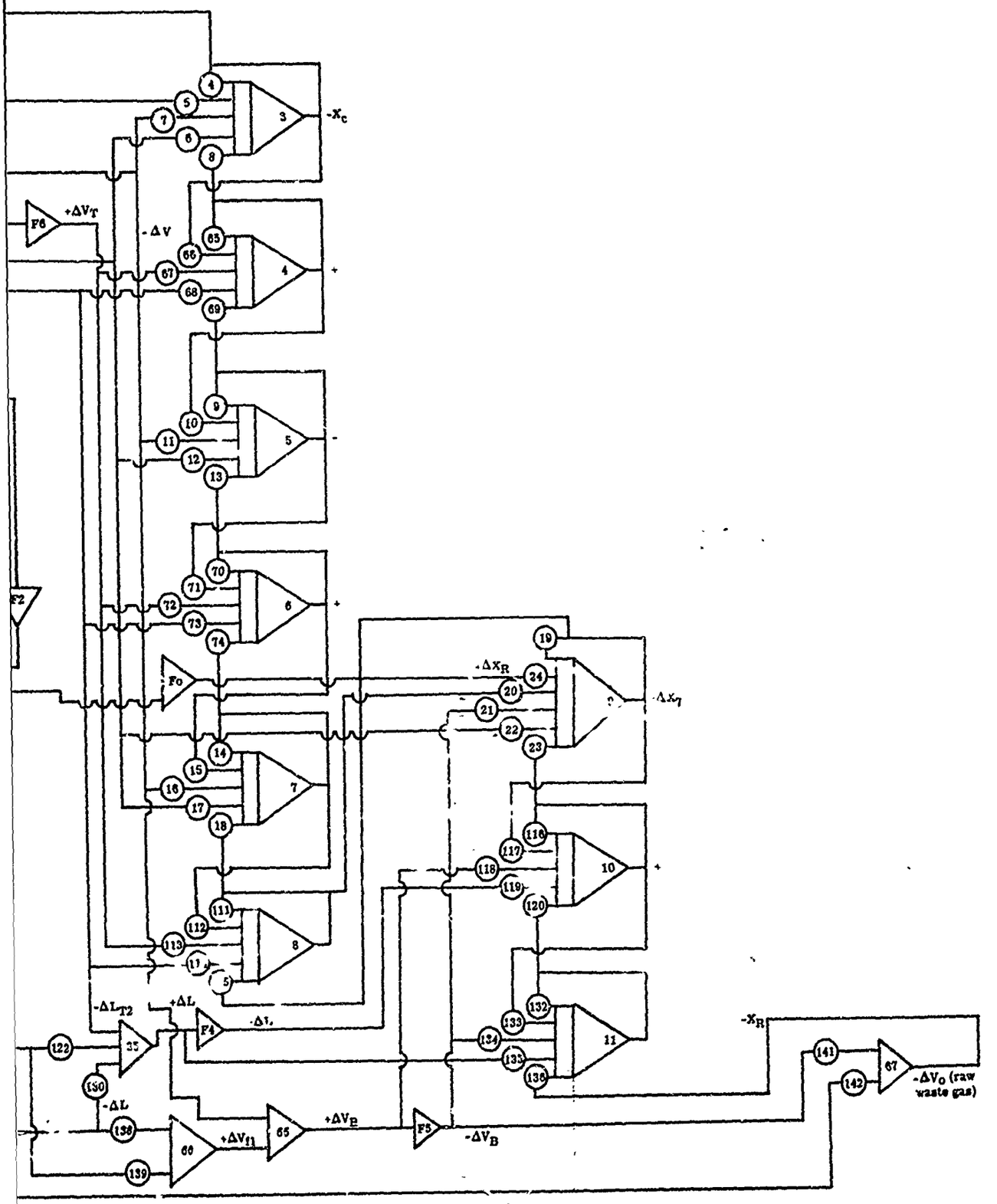
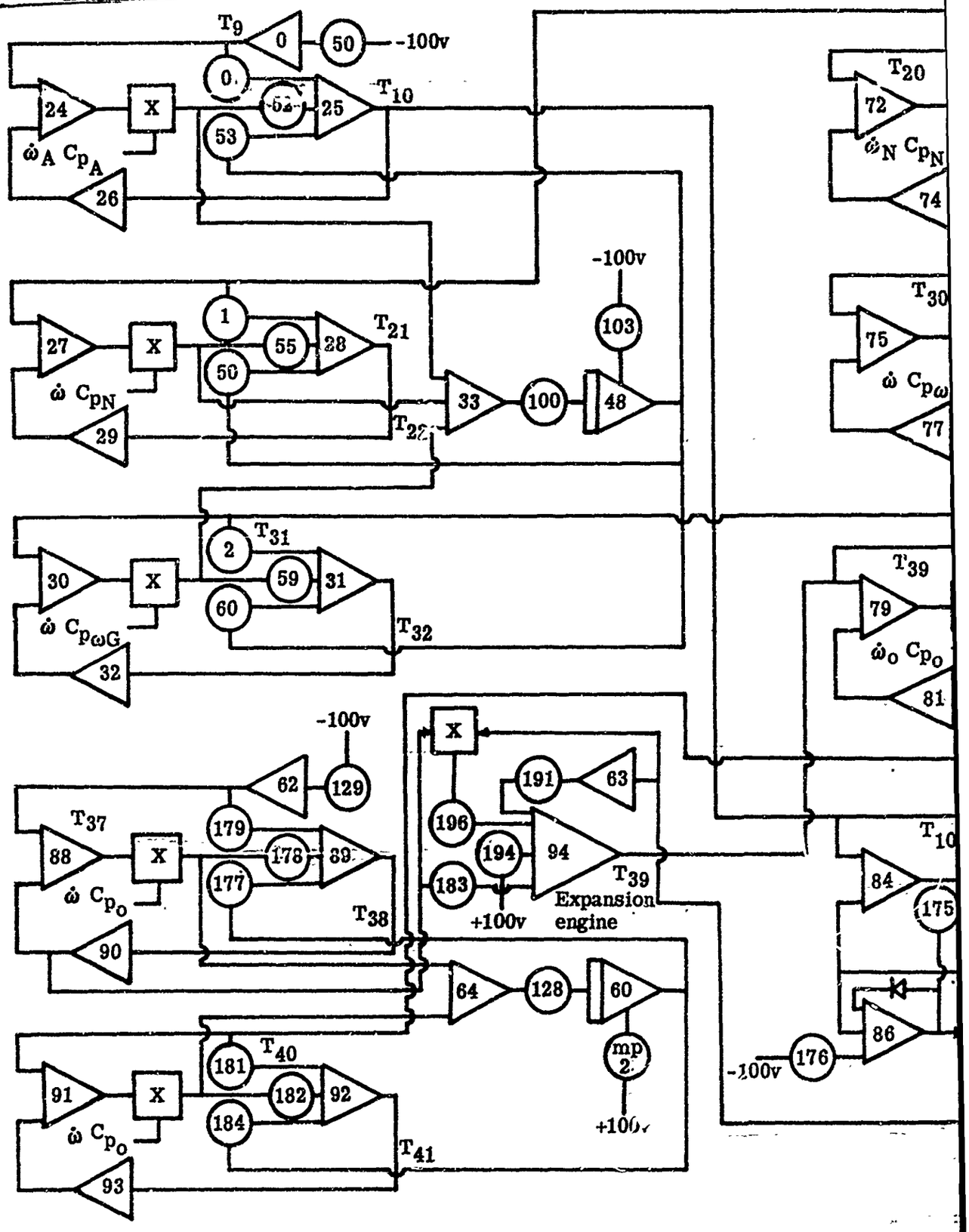


Figure B-11. Nitrogen generation subsystem analog diagram. (Sheet 1 of 2)

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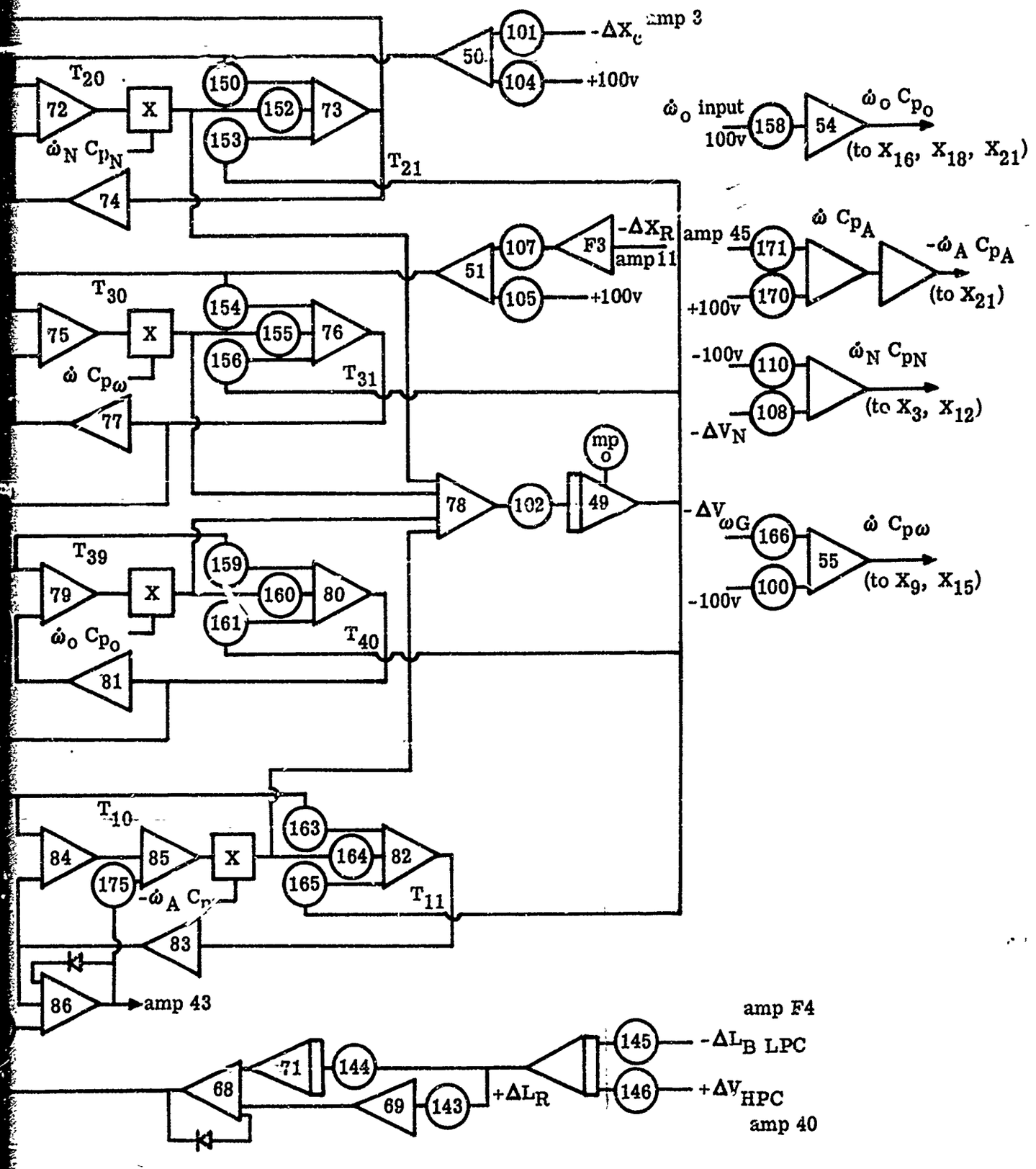
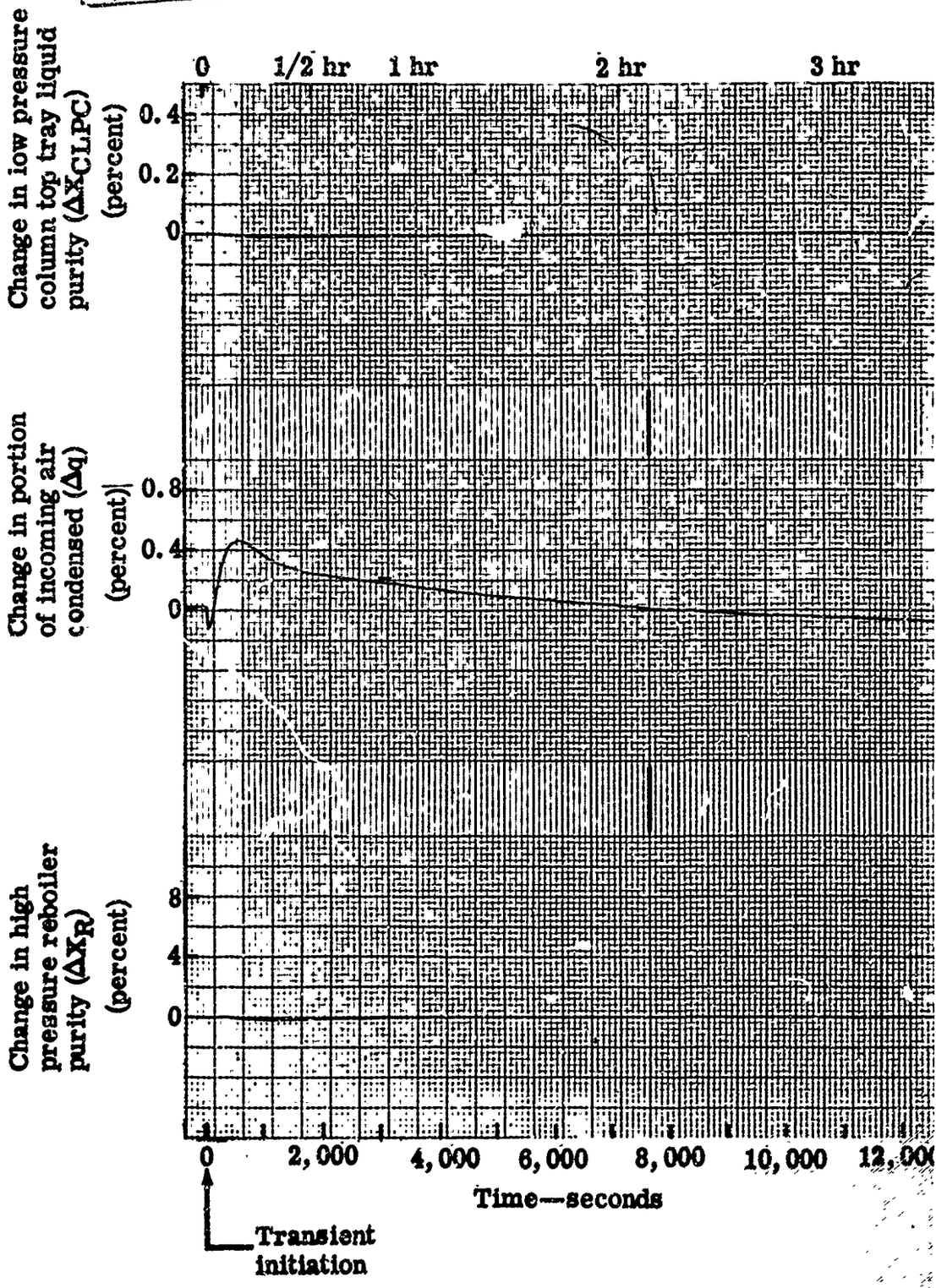


Figure B-11. Nitrogen generation subsystem analog diagram. (Sheet 2 of 2)

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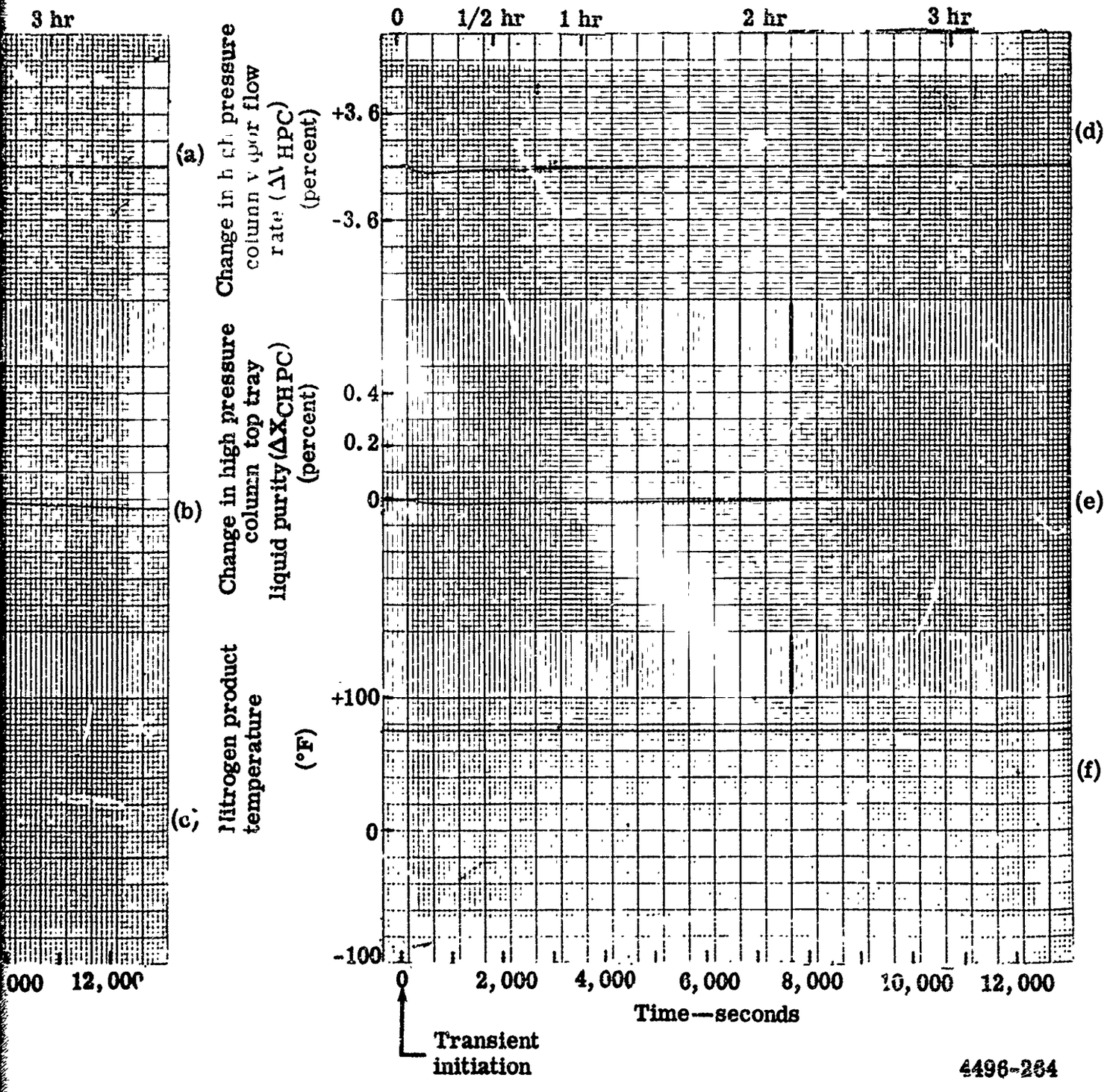


Figure B-12. Nitrogen generation subsystem response to an increase in refrigeration oxygen flow rate from 80% to 120% of design flow rate.

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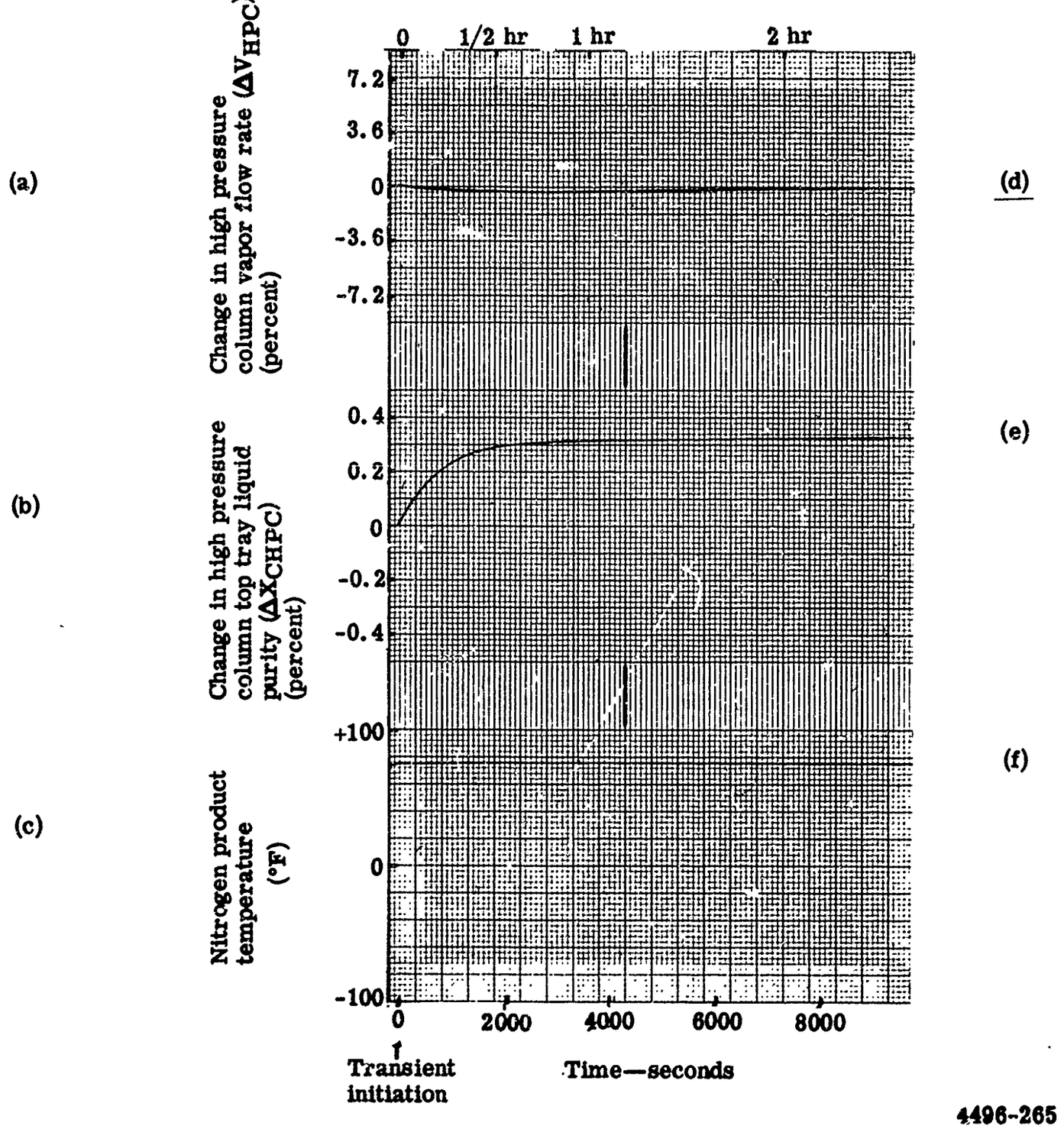


Figure B-13. Nitrogen generation subsystem response to a decrease in product nitrogen draw-off rate from 100% to 86.7%.

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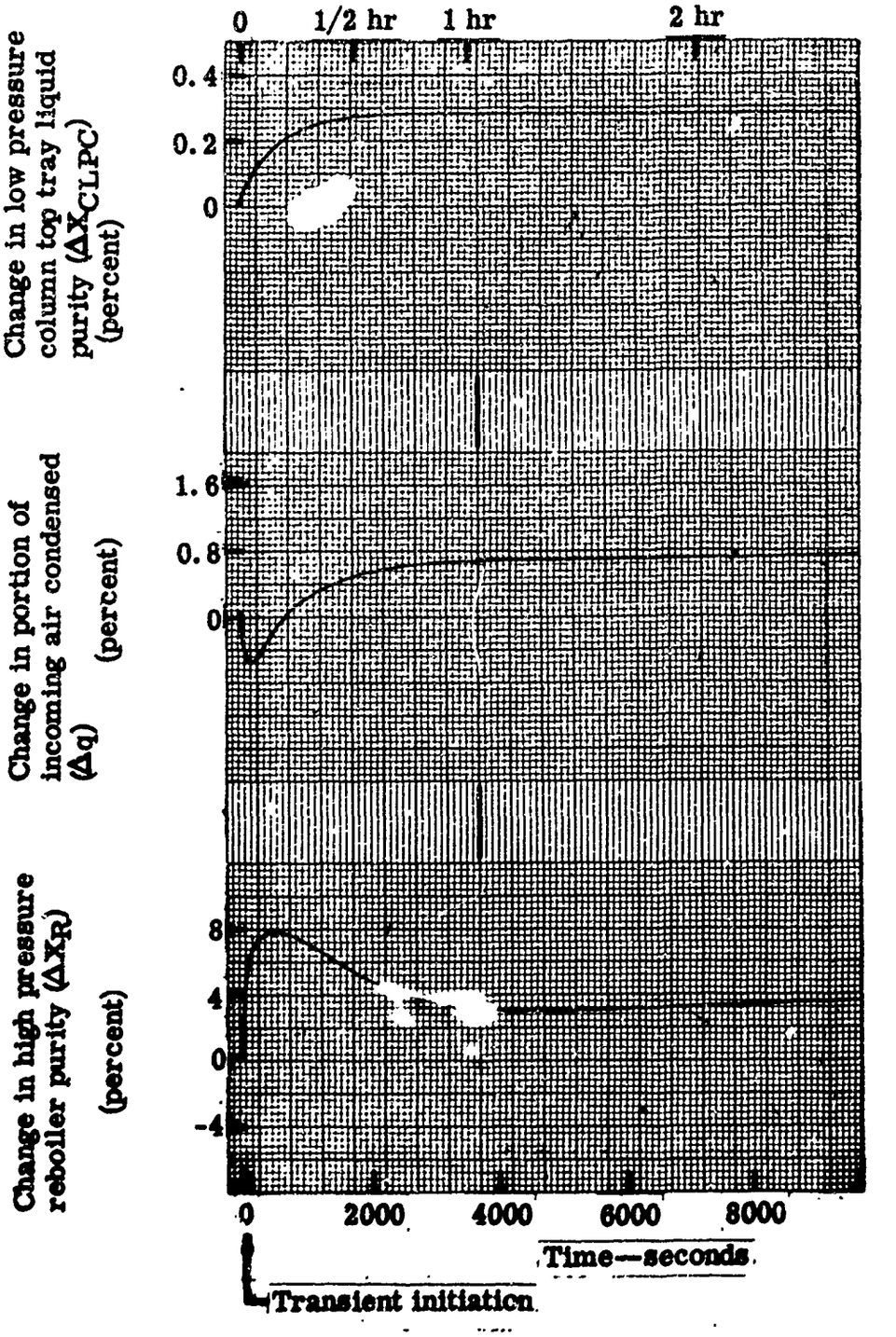


Figure B-1

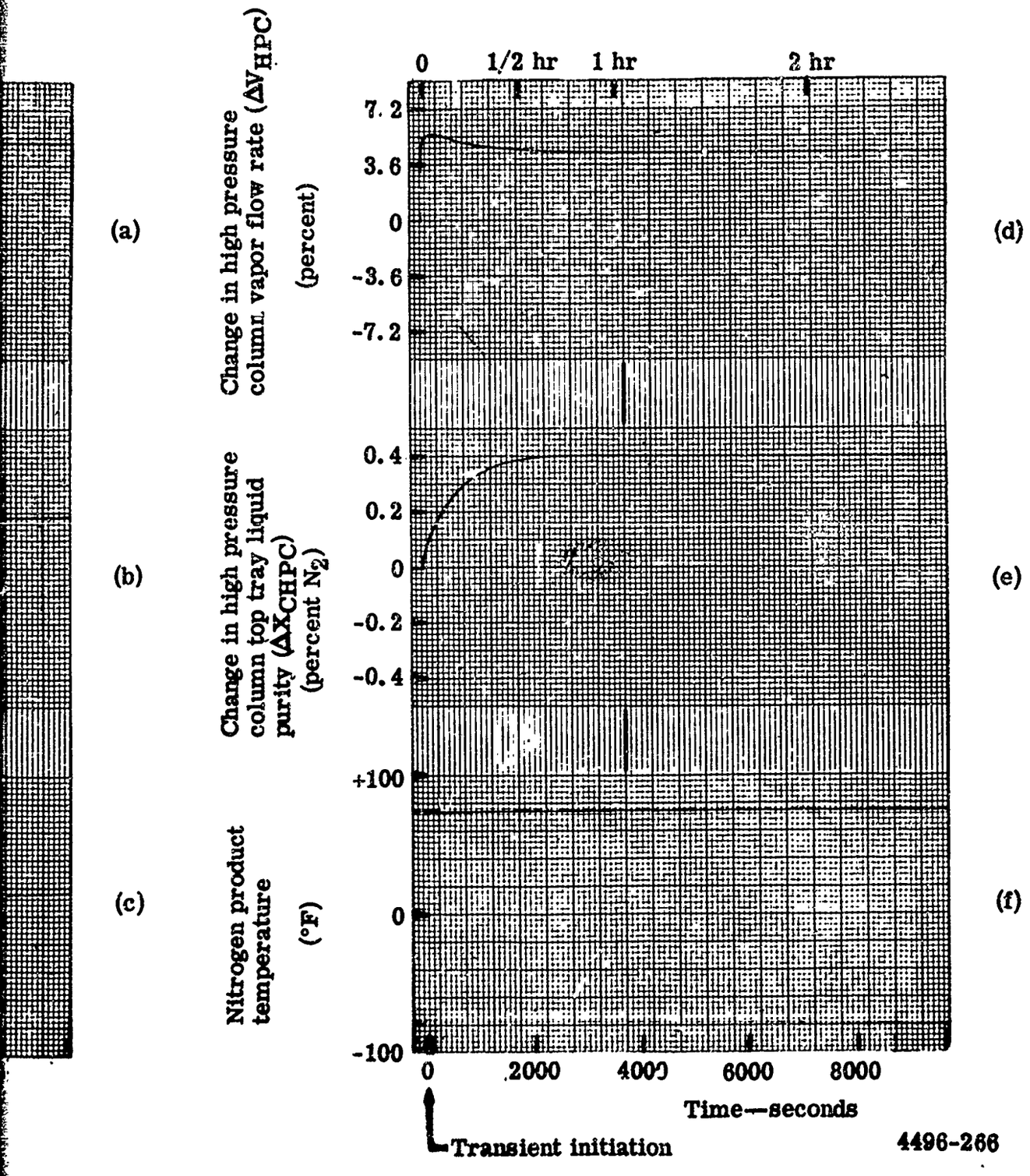
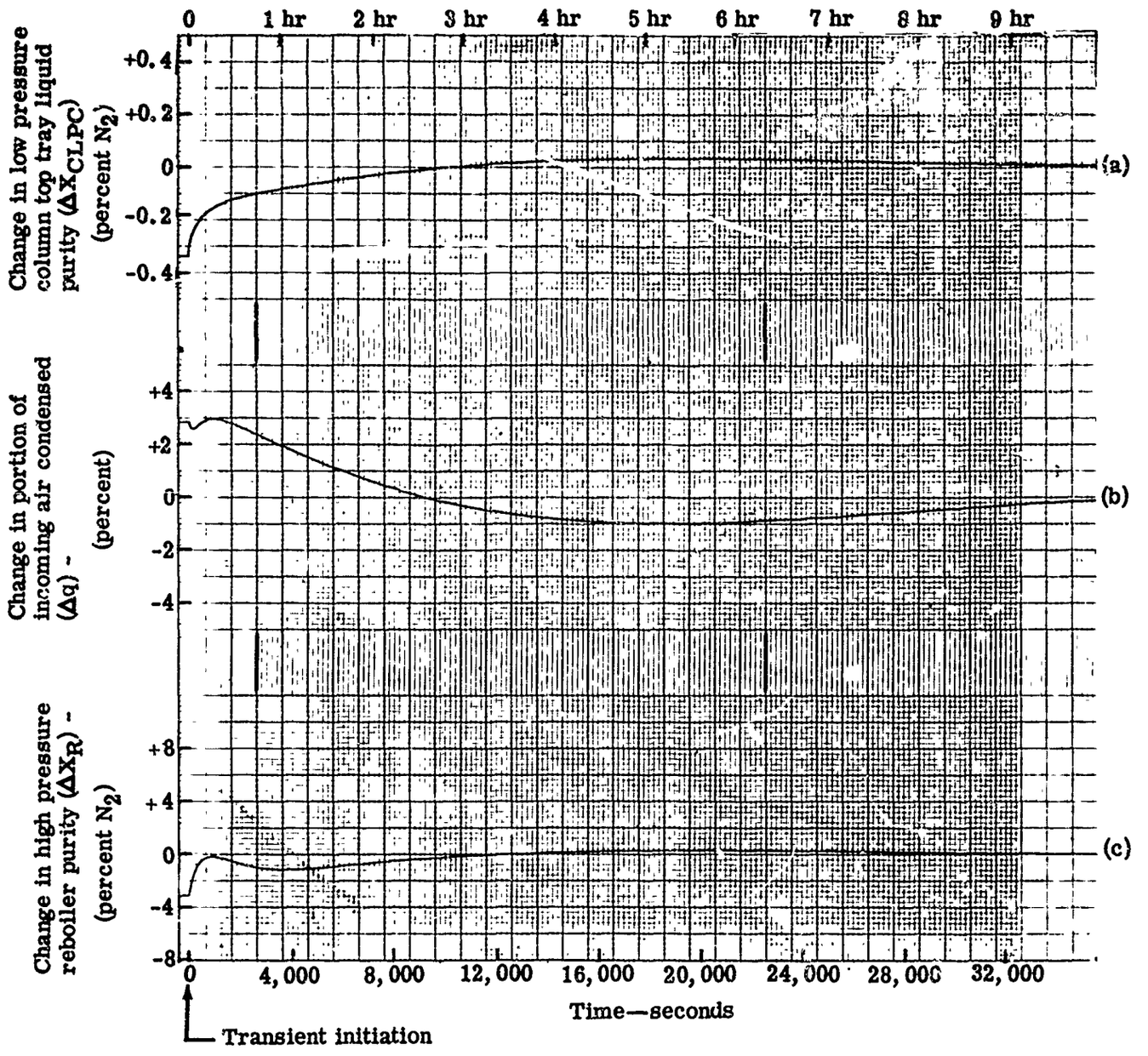


Figure B-14. Nitrogen generation subsystem response to an increase in feed airflow rate from 100% to 105% of design flow.

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Fig

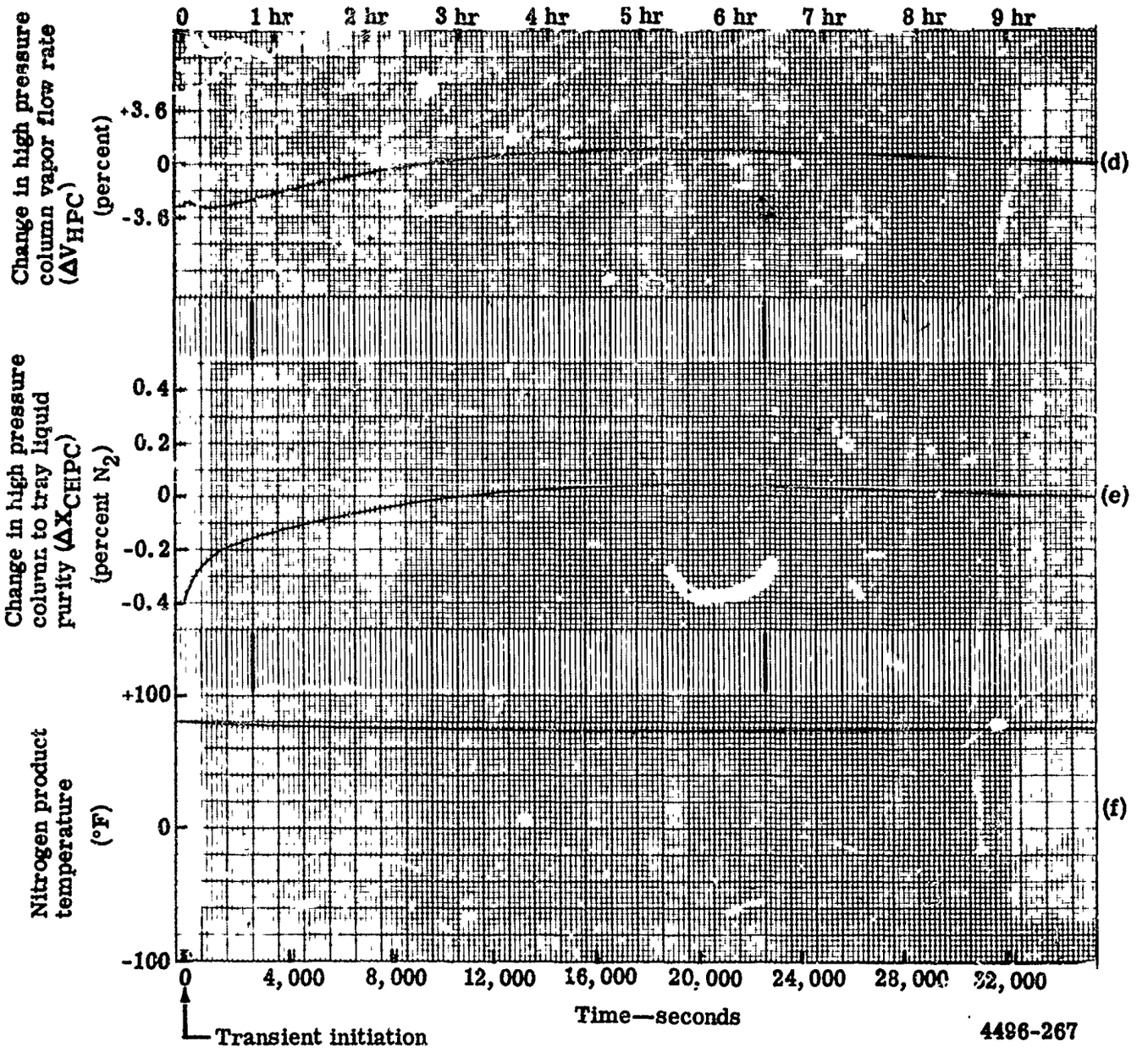


Figure B-15. Nitrogen gene[illegible] ystem response to stopping withdrawal of liquid nitrogen for storage.

AMMONIA SYNTHESIS SUBSYSTEM START-UP ANALYSIS

To assist in the determination of the response time of the fuel production system, an analog computer program was written to determine the characteristics of the ammonia synthesis subsystem during warm-up from ambient temperature to the temperature and pressure where ammonia synthesis begins. Since ammonia synthesis does not take place at these low temperatures during this time, the mathematical model used for the computer program was simplified. Only those portions of the ammonia synthesis subsystem which require a significant change in temperature were simulated. Specifically, the converter heat exchanger, the ammonia synthesis catalyst, and the electric heaters were represented. In addition, a simulation of the ammonia converter wall was represented. The basic heat exchanger equations were changed to the following form for simulation.

For the converter heat exchanger

$$Q_{11-1} = \dot{\omega}\, C_p\, (T_{11} - T_1)$$

$$T_1 = \frac{2}{U_H A_H} (Q_{11-1}) - T_{11} + 2\, T_{S_{11-1}}$$

$$Q_{4-5} = \dot{\omega}\, C_p\, (T_4 - T_5)$$

$$T_5 = \frac{2}{U_C A_C} (Q_{4-5}) - T_4 + 2\, T_{S_{11-1}}$$

$$T_{S_{11-1}} = \frac{1}{\omega_H C_H} \int (Q_{11-1} + Q_{4-5})\, dt$$

For the converter

$$Q_{heater} = \text{constant}$$

$$Q_{1-4} = \dot{\omega}\, C_p\, (T_1 - T_4)$$

$$T_4 = \frac{2}{U_C A_C} (Q_{1-4}) - T_1 + 2\, T_{S_{1-4}}$$

$$T_{S_{1-4}} = \frac{1}{\omega_C C_C} \int (Q_{heater} + Q_{1-4} + Q_{HL})\, dt$$

Figure B-16 is an analog diagram of this computer program.

The response of the ammonia synthesis subsystem was run for each group of systems where the change in plant configuration would affect the heat-up characteristics. Figure B-17 is typical of the heat-up transient for the subsystem in systems 1, 2, and 3.

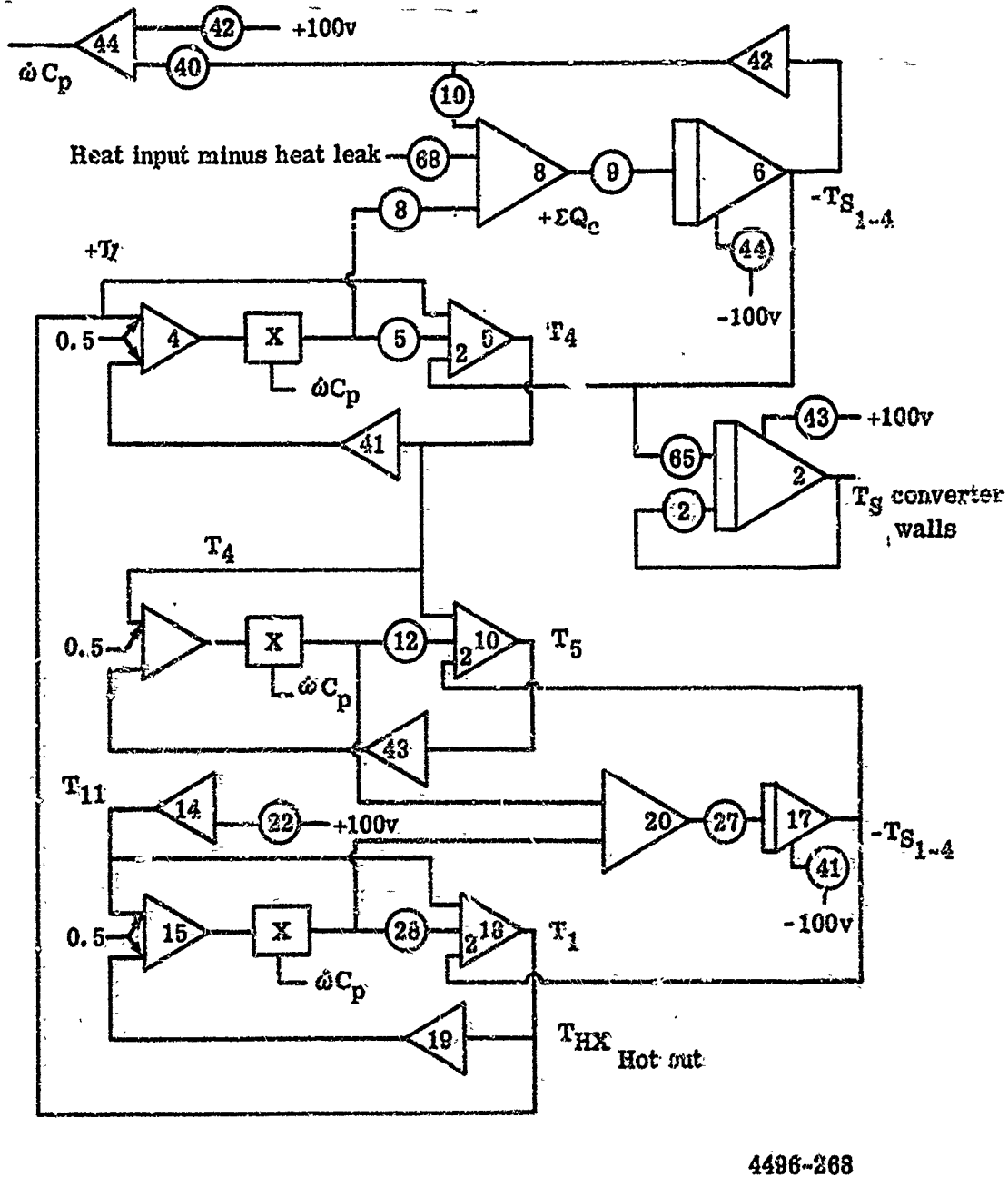


Figure B-16. Ammonia synthesis subsystem start-up simulation.

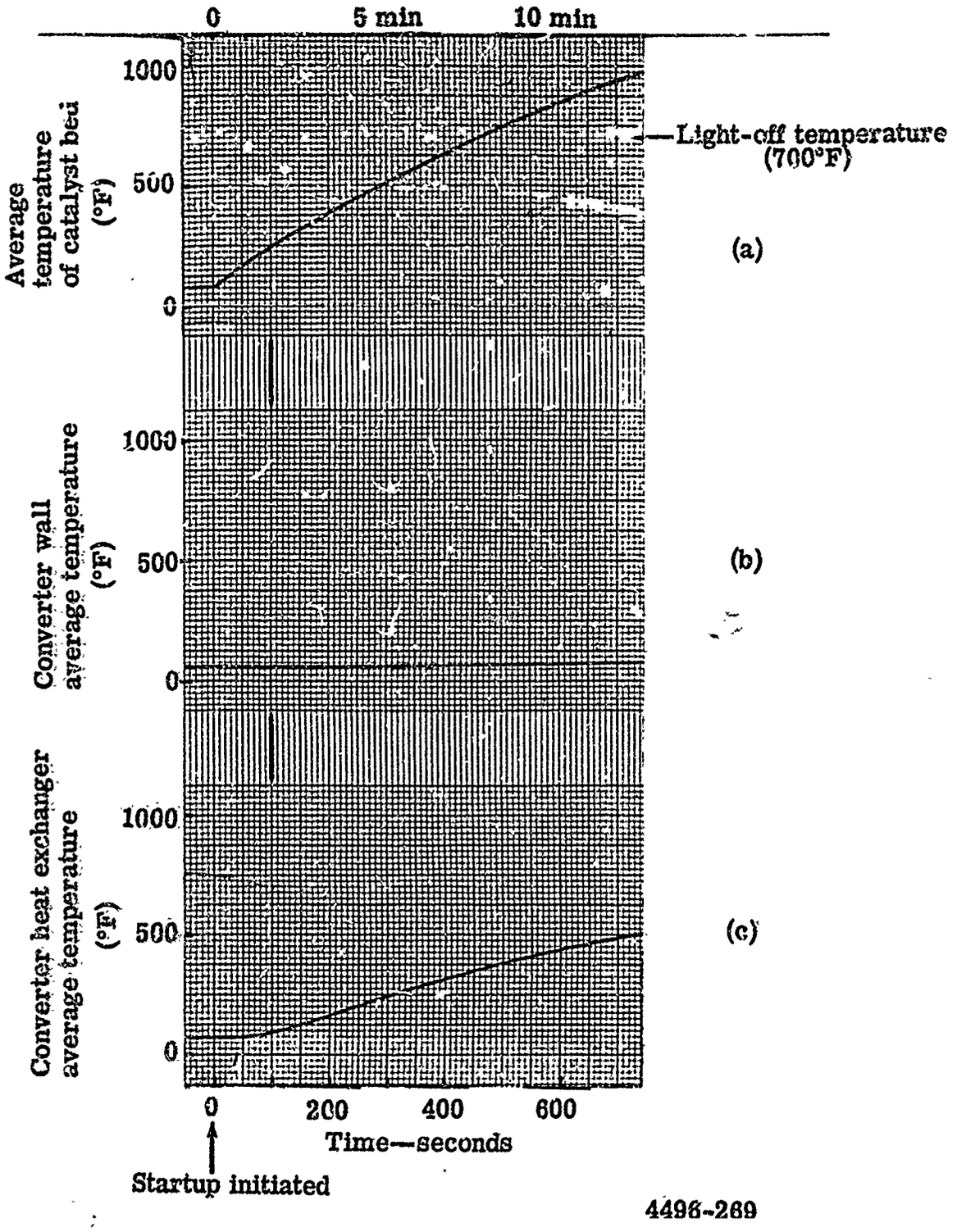


Figure B-17. Ammonia synthesis subsystem start-up—systems 1, 2, and 3.

The average temperature of the catalyst bed during warm-up is shown in Figure B-17(a) with the synthesis gas flow in the ammonia synthesis loop varied as a function of temperature from 3-1/3% to 33-1/3% as the catalyst temperature was increased from ambient temperature to 700°F. In the simulation, the temperature of the catalyst bed continues to increase beyond the 700°F point. In actual practice the start-up heater is cycled as ammonia synthesis begins at approximately 700°F and the subsystem is brought up to operating temperature and pressure by the operating crew.

The converter wall temperature during start-up is shown in Figure B-17(b). The temperature of the converter wall increases slowly because it is insulated from the hot synthesis gases. This insulation allows rapid start-up because of the reduction of heat leakage and the decrease in temperature-related stresses.

The average temperature of the converter heat exchanger is shown in Figure B-17(c). The heat exchanger is heated by the circulation of the synthesis gases and does not rise as rapidly in temperature as the catalyst bed, thus lessening significantly the thermal stress problems.

The start-up transients for the ammonia synthesis subsystems of systems 4 through 11 are shown in Figure B-13 through B-21. The heat-up of ammonia synthesis subsystems in the larger system requires more time than the smaller systems because the thermal inertia, relative to the capacity of the heaters in the catalyst bed, is greater. The electric heater power is 100-, 150-, 200-, and 250-kw for the 3, 6-, 10-, and 15-Mw systems, respectively.

NITROGEN GENERATION SUBSYSTEM START-UP ANALYSIS

During start-up of the nitrogen generation subsystem, the air fractionation columns and the heat exchangers which are contained in an insulated structure (cold box) must cool down to the condensation temperature of air, and the liquid nitrogen and oxygen must be collected in the columns in the proper ratio for each tray, reboiler, and condenser. For the start-up and shutdown analysis, the column dynamics and much of the control dynamics are not involved. The analog program written for start-up represents the oxygen reversing heat exchanger on the oxygen stream from the hydrogen production subsystem, the temperature decrease produced by the expansion engine, and the thermal inertia of the contents of the cold box. All refrigeration supplied by the oxygen stream from the hydrogen production subsystem is assumed to provide condensation of oxygen and nitrogen when the temperature of the cold box decreases to the point that condensation takes place.

The form of the heat exchanger equation used in this simulation is the same as that given for the ammonia synthesis system start-up analysis. For the cold box, the representation is

$$Q_{CB} = \dot{\omega}_o C_p (T_{39} - T_{40})$$

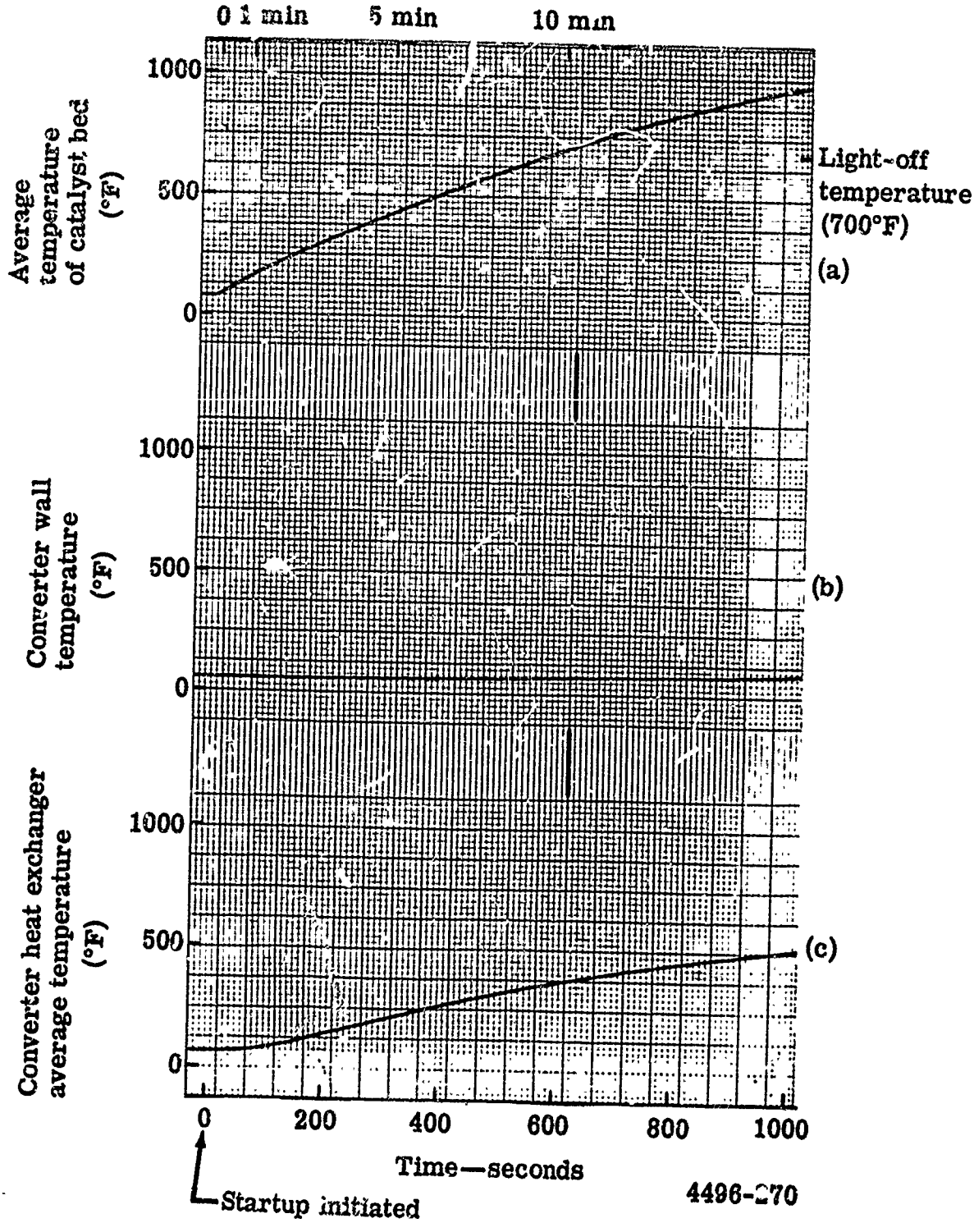


Figure B-16. Ammonia synthesis subsystem start-up—system 4.

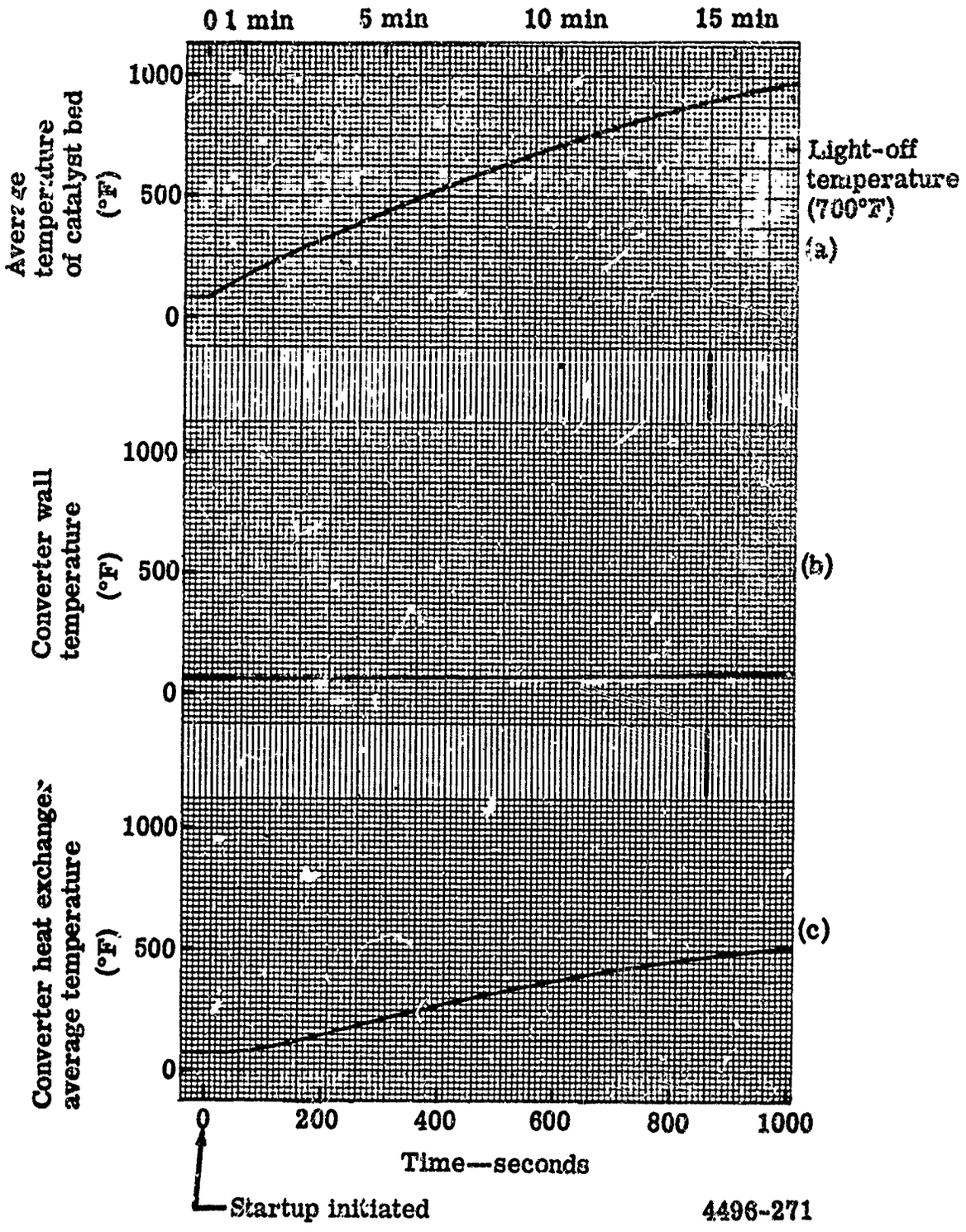


Figure B-19. Ammonia synthesis subsystem start-up—systems 5, 6, 7, and 8.

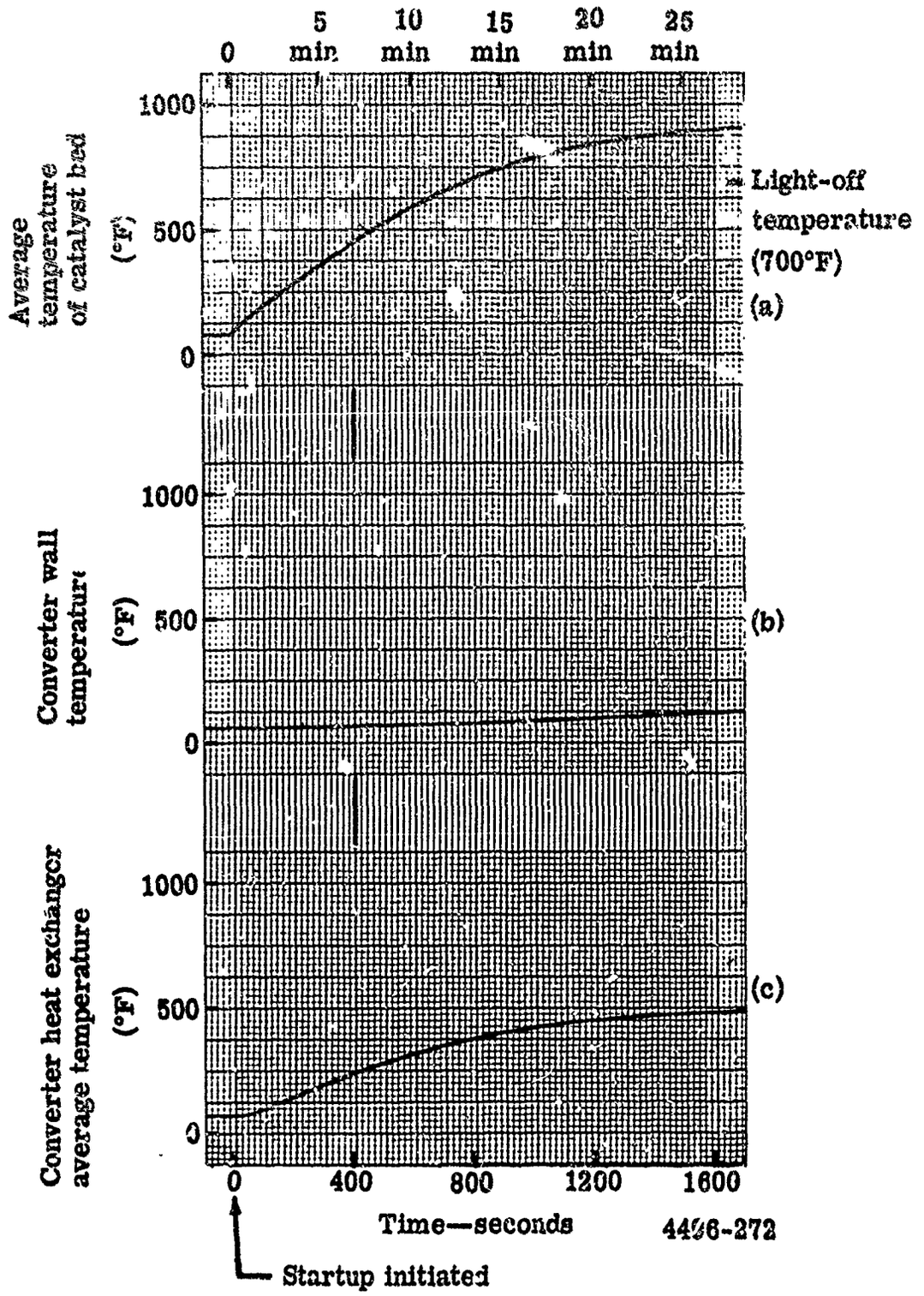


Figure B-20. Ammonia synthesis subsystem start-up—systems 9 and 10.

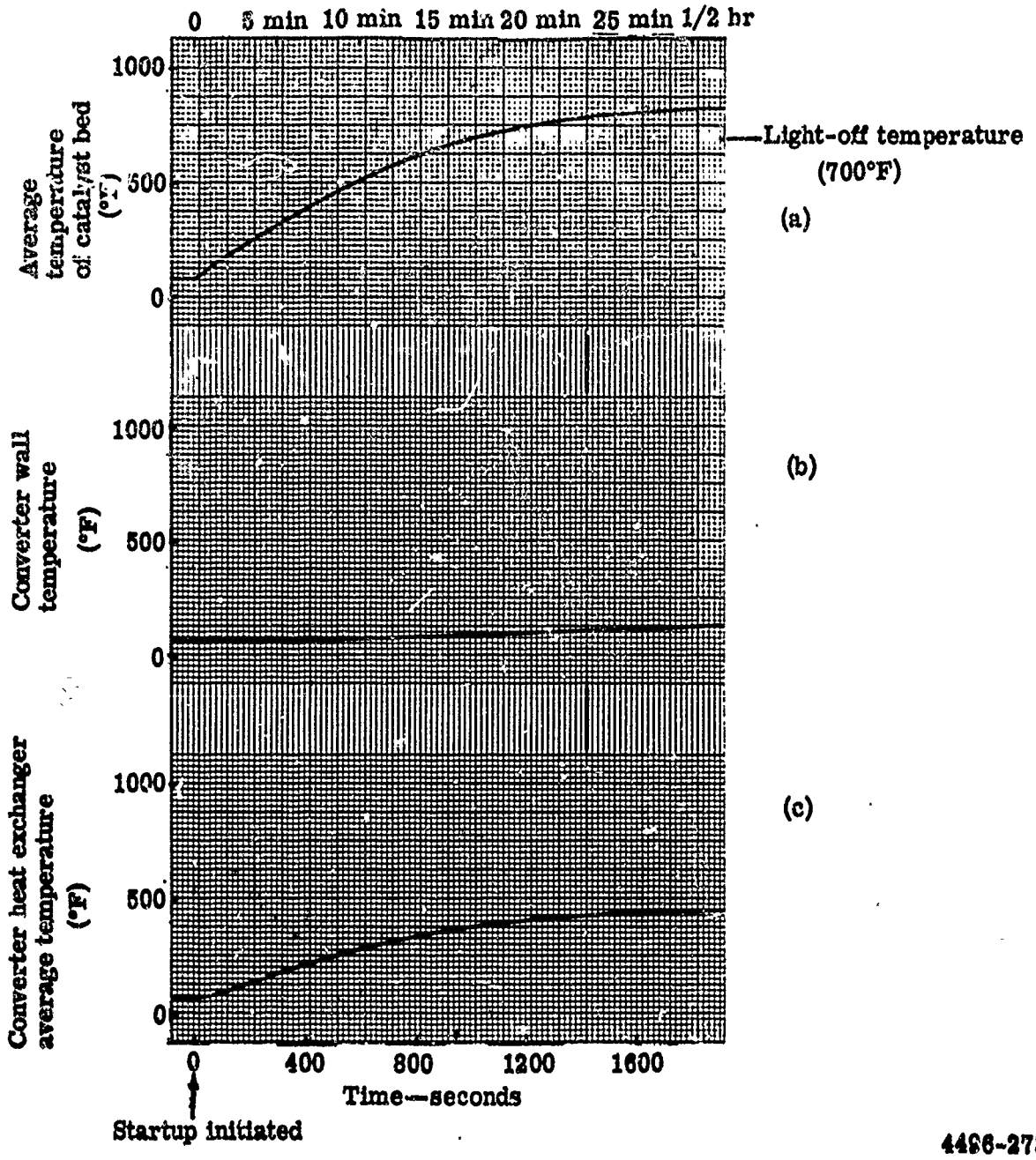


Figure B-21. Ammonia synthesis subsystem start-up—system 11.

$$T_{40} = \frac{2}{U_o A_o} Q_{CB} - T_{39} + 2 T_{S_{CB}}$$

$$T_{S_{CB}} = \frac{1}{\omega_{CB} C_{p_{CB}}} \int (Q_{CB} + Q_{HL} + Q_{LN})\, dt;\ T_{S_{CB}} > T_{cond}$$

$$\omega_{AL} = \frac{1}{\Delta H_{liq}} \int (Q_{CB} + Q_{HL} + Q_{LN})\, dt;\ T_{S_{CB}} < T_{cond}$$

The temperature at the exit of the expansion engine with no oxygen flow bypassed is represented as it is in the control analysis model by

$$T_{39} = K_{1T} (T_{38}) + K_{2T}$$

where

T = temperature at point indicated by subscript, °R
$\dot{\omega}_o$ = oxygen flow rate, lb-moles/sec
C_p = specific heat of oxygen, BTU/lb-mole
Q_{CB} = heat removed by oxygen stream, BTU/sec
Q_{LN} = heat removed by liquid nitrogen, BTU/sec
Q_{HL} = heat leakage into cold box, BTU/sec
ω_{AL} = amount of nitrogen and oxygen condensed, lb-moles
K_{1T} and K_{2T} = constants which define operation of expansion engine

Subscript S = average
Subscript CB = cold box
Subscripts 38, 39, 40 = points indicated in Figure 3.9-1 of Section III

For the start-up analysis, the oxygen flow was approximated by a ramp which represented the results of the hydrogen production system electrolyzer start-up study. The formation of liquid or evaporation of liquid was assumed to occur at a constant temperature with a constant heat of vaporization.

The analog diagram for nitrogen generation subsystem start-up simulation is shown in Figure B-22.

The results of the nitrogen generation subsystem shutdown analyses are shown in Figures B-23 and B-24. After system 1, 2, or 3 is shut off, all of the liquid in the air fractionation column evaporates in 1.7 hr (Figure B-23(b)). After 8 hr of nonoperation, the cold box temperature rises to -160°F. Similarly, for the 15-Mw_e system 11, all the liquid evaporates during the first 1.7 hr (Figure B-24(b)) and the cold box temperature rises to -200°F after an 8-hr shutdown and to +30°F after a 48-hr shutdown (Figure B-24(a)).

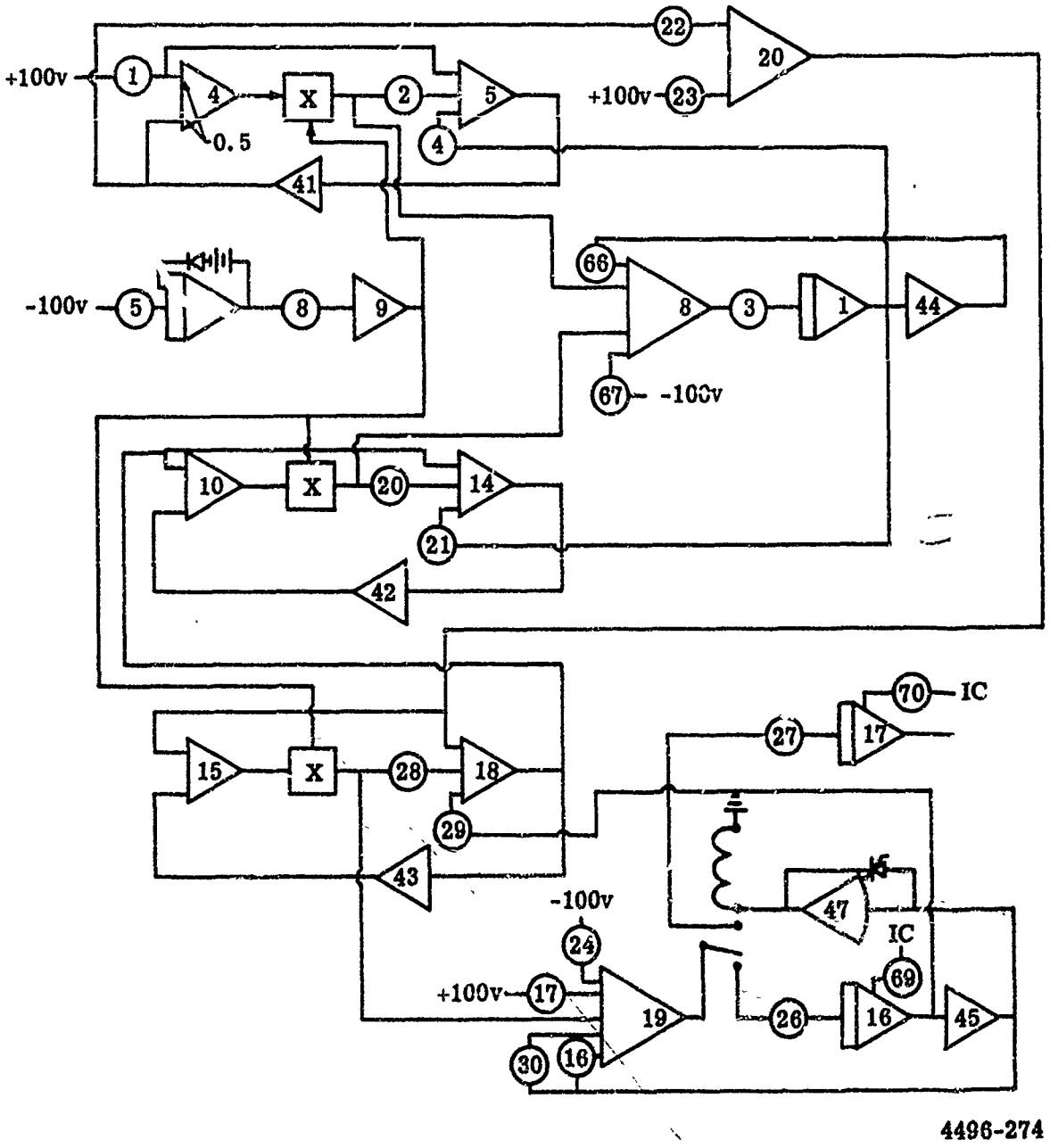


Figure B-22. Nitrogen generation subsystem start-up simulation.

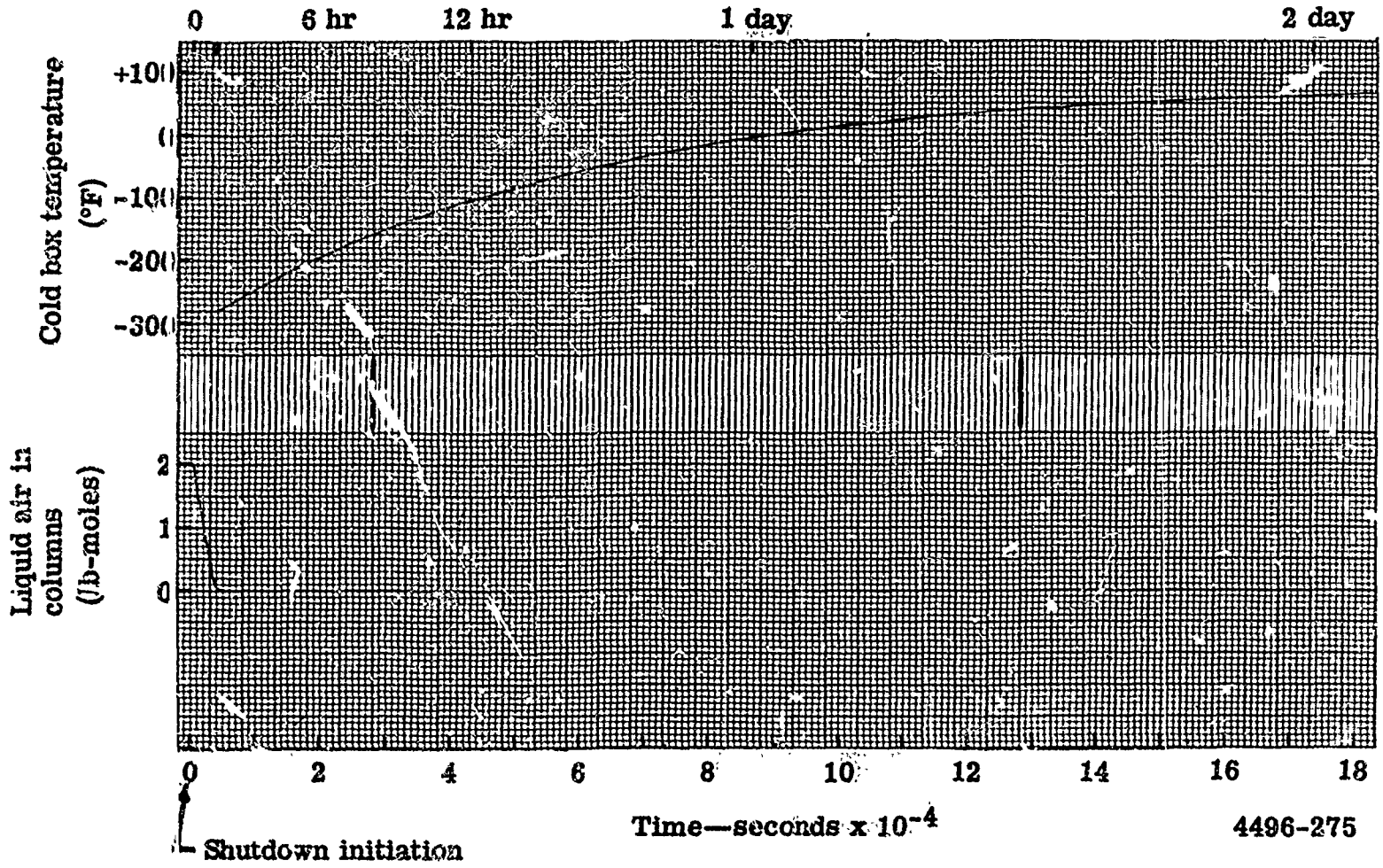


Figure B-23. Nitrogen generation subsystem shutdown—systems 1, 2, and 3.

Allison

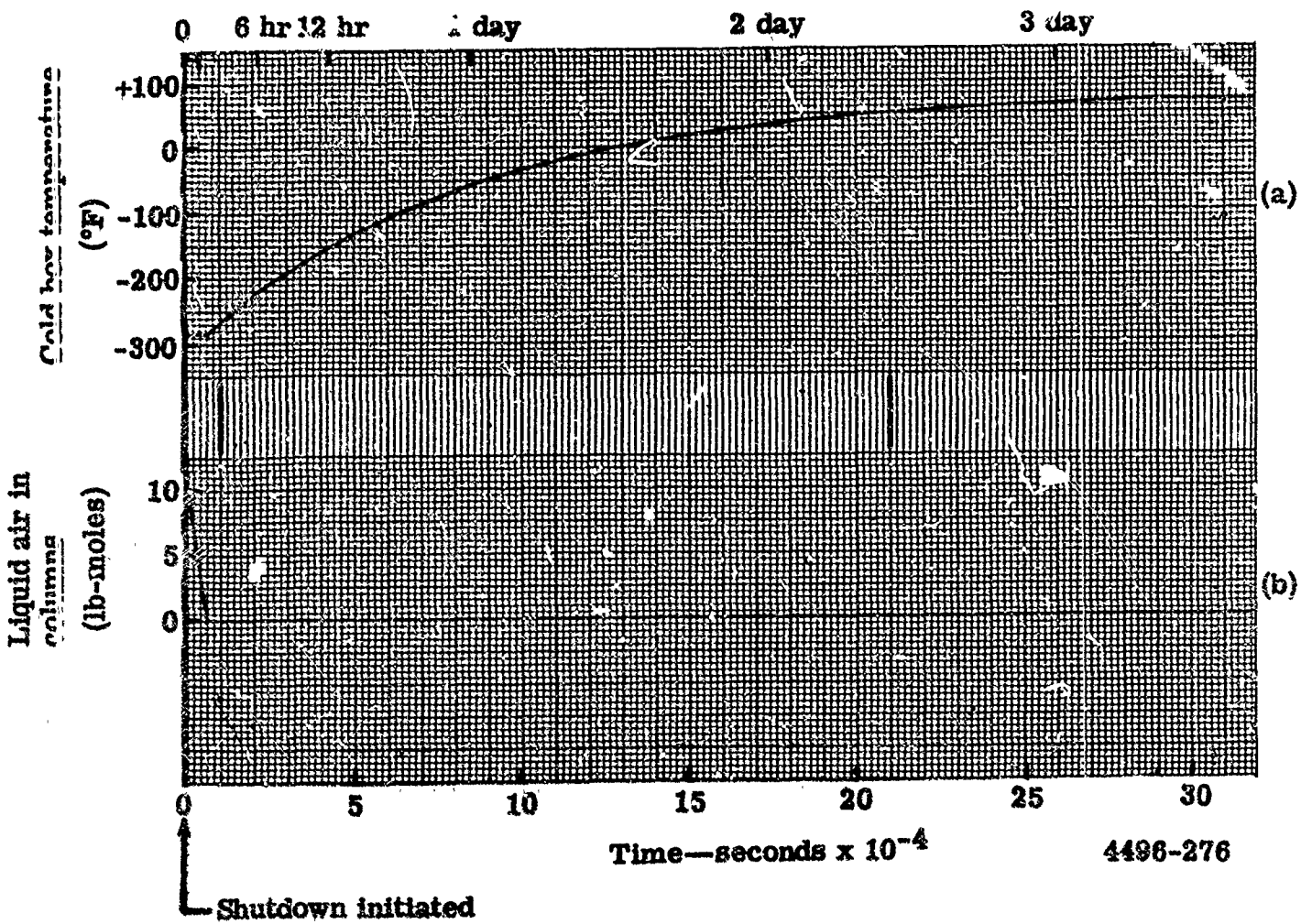


Figure B-24. Nitrogen generation subsystem shutdown—system 11.

Allison

The results of the nitrogen generation subsystem start-up analysis are shown in Figures B-25 through B-32. The nitrogen generation subsystem is considered on-stream when the liquid accumulated reaches the constant design values. However, as pointed out earlier in the Nitrogen Generation Subsystem portion of Section III, when stored liquid nitrogen is used for start-up, product nitrogen is flashed and available almost immediately for delivery to the ammonia synthesis subsystem.

Figure B-25 shows the nitrogen generation subsystem start-up response of the 3-Mw_e systems 1, 2, and 3 after an 8-hr shutdown (cold box temperature is -160°F) using stored liquid nitrogen flowing at a rate equivalent to the full system output. The subsystem will be on-stream in about 0.54 hr. The oxygen reversing heat exchanger temperature increases as warm electrolyzer oxygen is delivered and cools down as the start-up progresses. Figure B-26 indicates that this nitrogen generation subsystem for the 15-Mw_e system 11 will be on-stream in about 0.42 hr after an 8-hr shutdown. If system 11 is shut down for 48 hr (cold box temperature is 40°F), about 0.92 hr is required for the nitrogen generation subsystem to come on-stream (Figure B-27).

Figures B-28 through B-30 present the results of analyses run to determine nitrogen generation subsystem start-up times for the cases where stored liquid nitrogen flow rates equivalent to 20% of the design product rates are used. Figure B-28 shows that the 3-Mw_e systems 1, 2, and 3 require about 2 hr for the nitrogen generation subsystem to come on-stream after an 8-hr shutdown. Figure B-29 shows that the 15-Mw_e system 11 requires about 1.5 hr for the nitrogen generation subsystem to come on-stream after an 8-hr shutdown. After a 48-hr shutdown, the 15-Mw_e system requires about 3.2 hr to bring the nitrogen generation subsystem on-stream, as shown in Figure B-30.

Start-up of the nitrogen generation subsystem was also analyzed for the case where the subsystem was initially at design ambient temperature (77°) and no liquid nitrogen was available for a quick cool-down. For the 3-Mw_e systems 1, 2, and 3 the nitrogen generation subsystem will be on-stream in about 10.6 hr (Figure B-31). For the 15-Mw_e system 11, again about 10.6 hr are required to bring the nitrogen generation subsystem on-stream (Figure B-32).

The start-up data described are in good agreement with estimates made earlier based on operating experience with air fractionation facilities.

HYDROGEN PRODUCTION SUBSYSTEM START-UP ANALYSIS

The hydrogen production subsystem produces hydrogen and oxygen as soon as the power is supplied. The rate of hydrogen production is approximately proportional to the current density on the electrodes. This current density is controlled by the voltage applied to the electrolysis

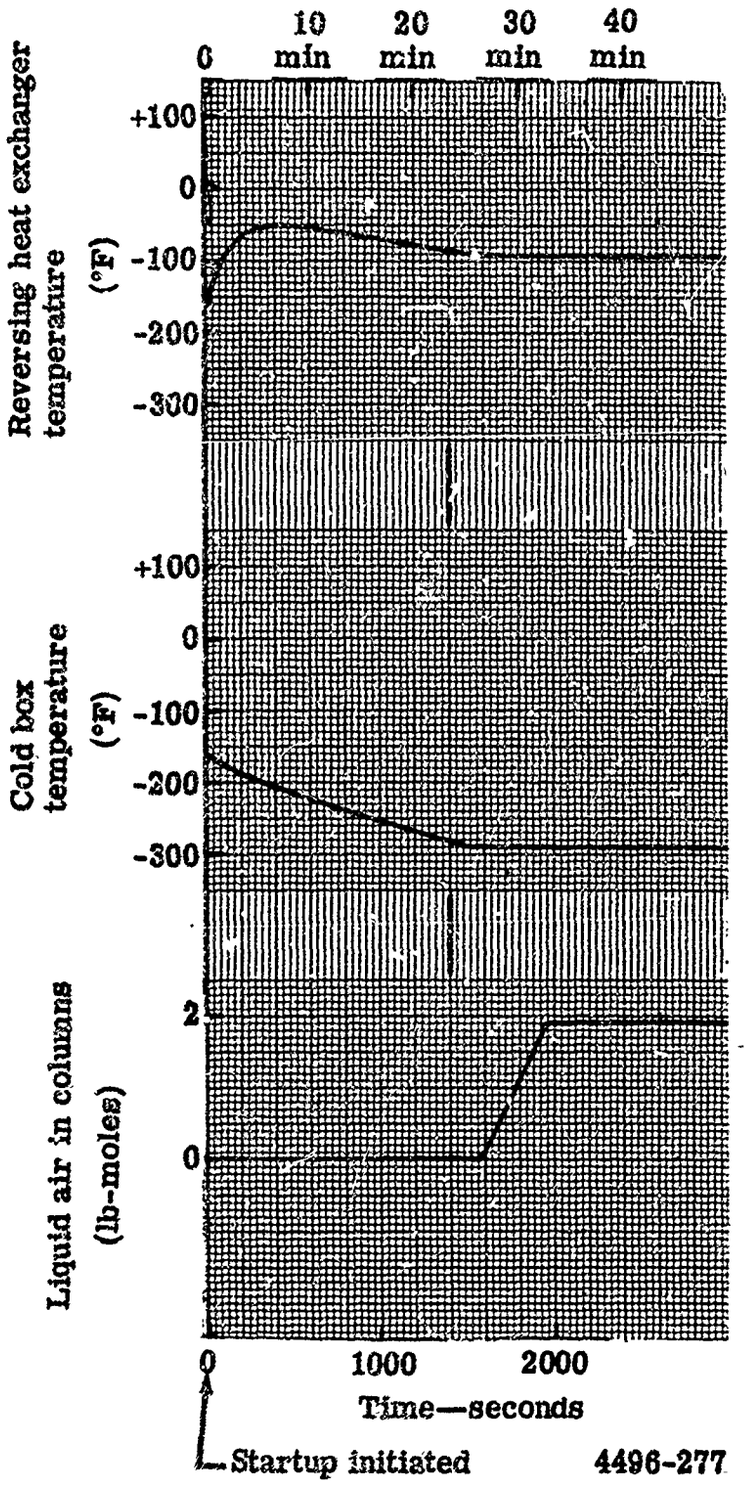


Figure B-25. Nitrogen generation subsystem start-up with full liquid nitrogen flow after 3-hr shutdown—systems 1, 2, and 3.

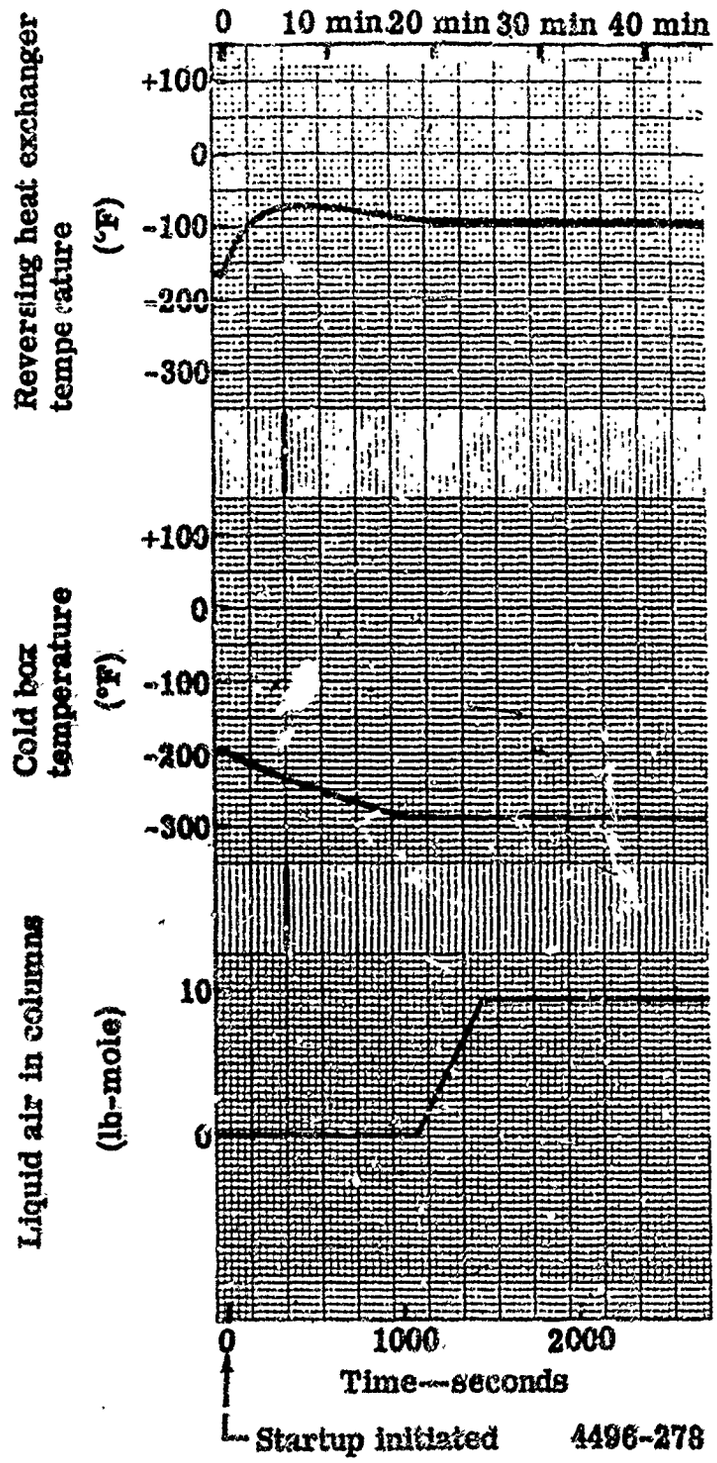


Figure B-26. Nitrogen generation subsystem start-up with full liquid nitrogen flow after 8-hr shutdown—system 11.

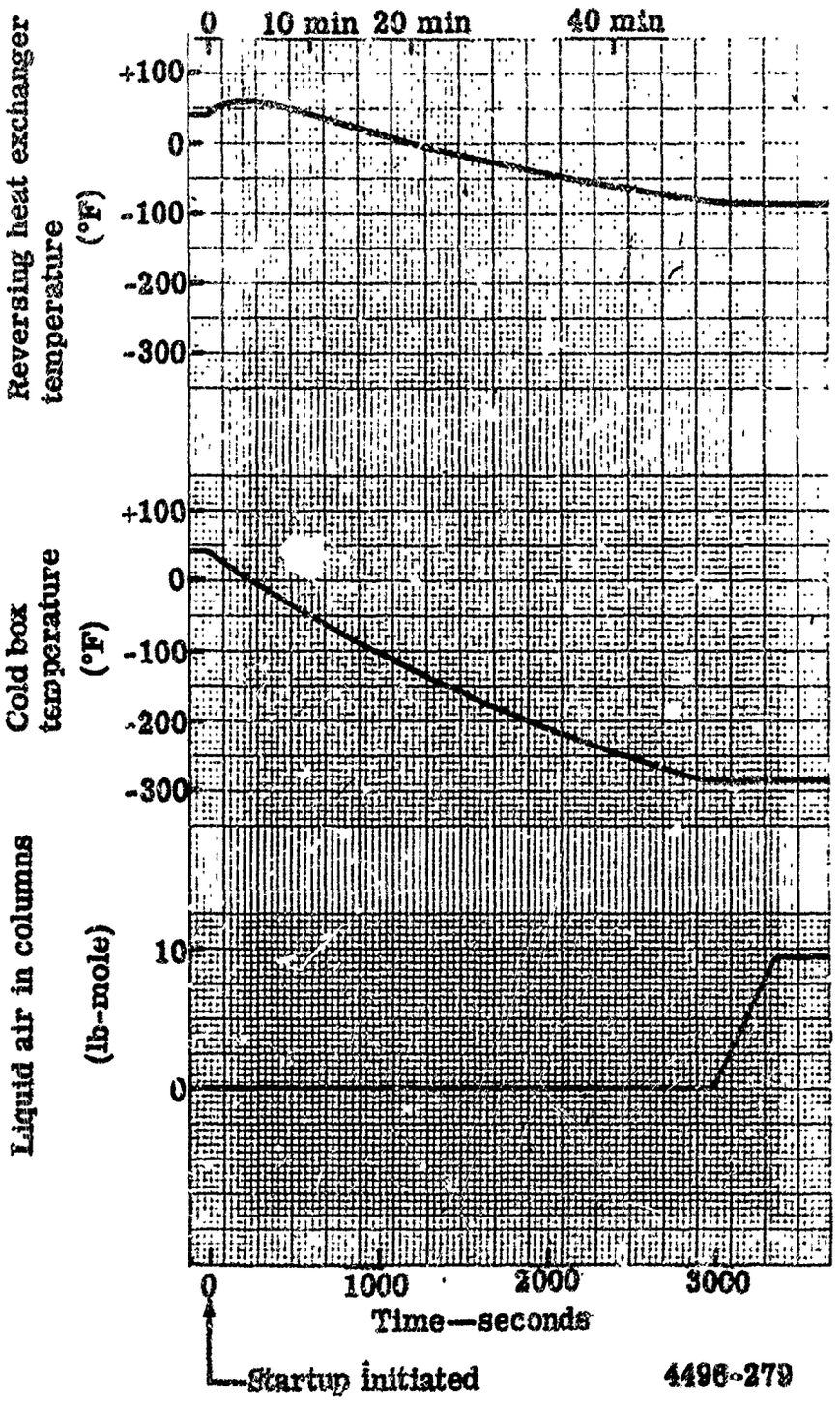


Figure B-37. Nitrogen generation subsystem start-up with full liquid nitrogen flow after 48-hr shutdown—system 11.

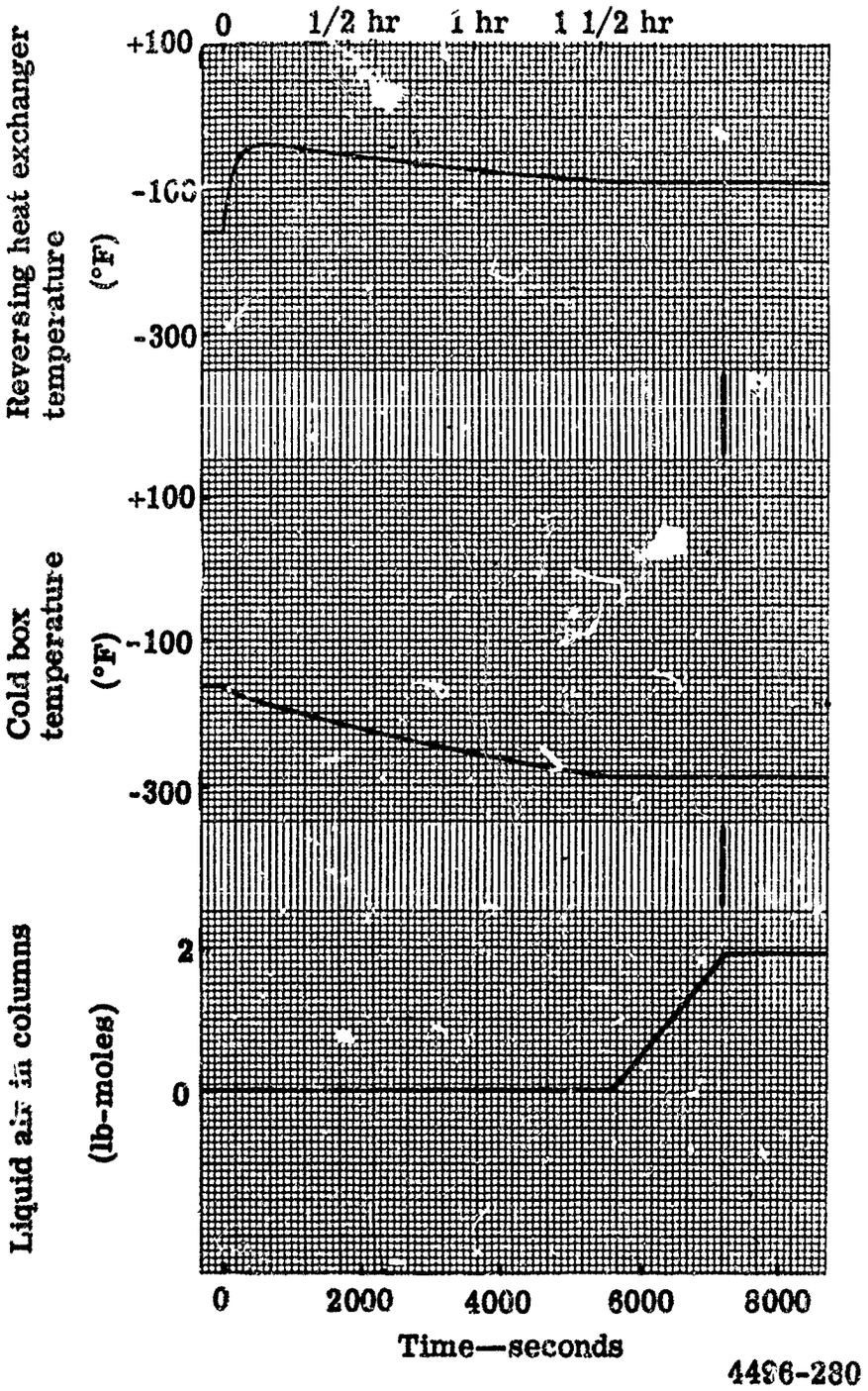


Figure B-28. Nitrogen generation subsystem start-up with 20% liquid nitrogen flow after 8-hr shutdown—systems 1, 2, and 3.

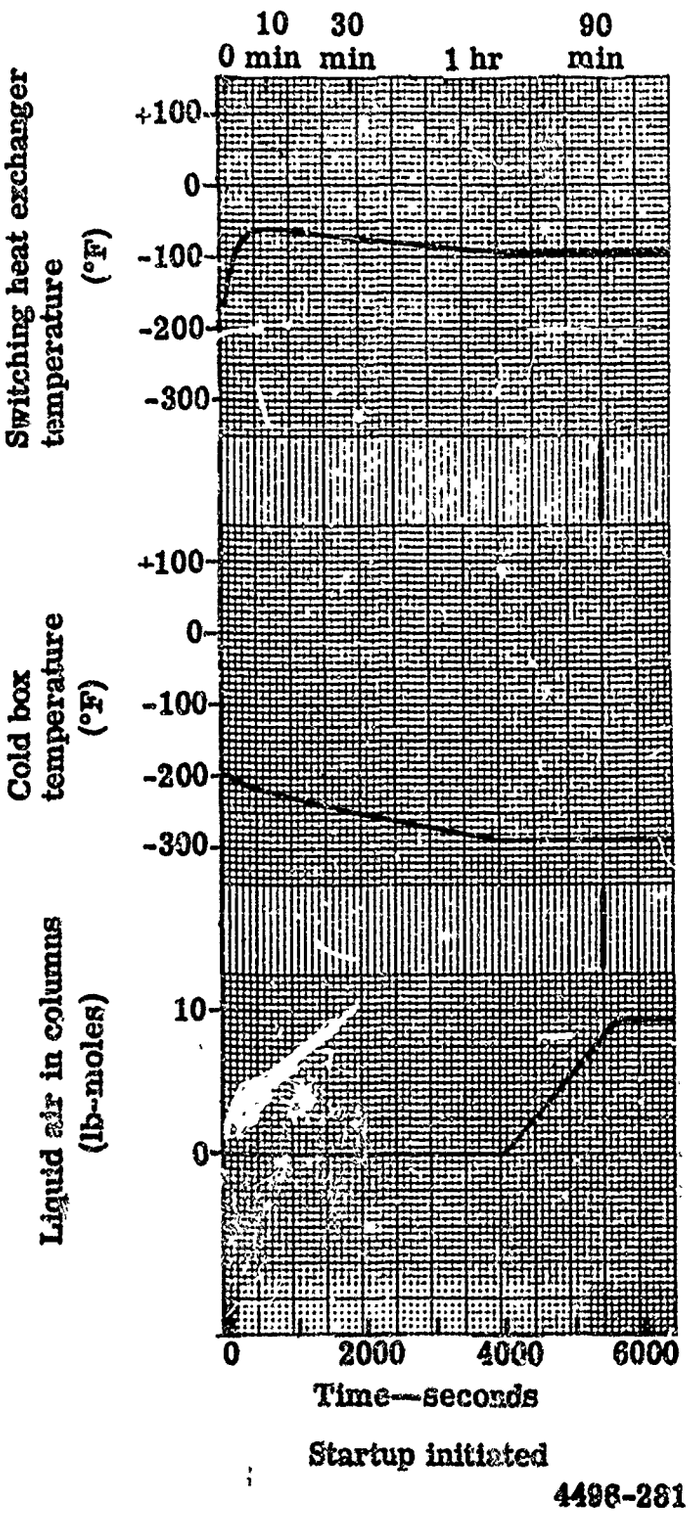


Figure B-29. Nitrogen generation subsystem start-up with 20% liquid nitrogen flow after 8-hr shutdown—system 11.

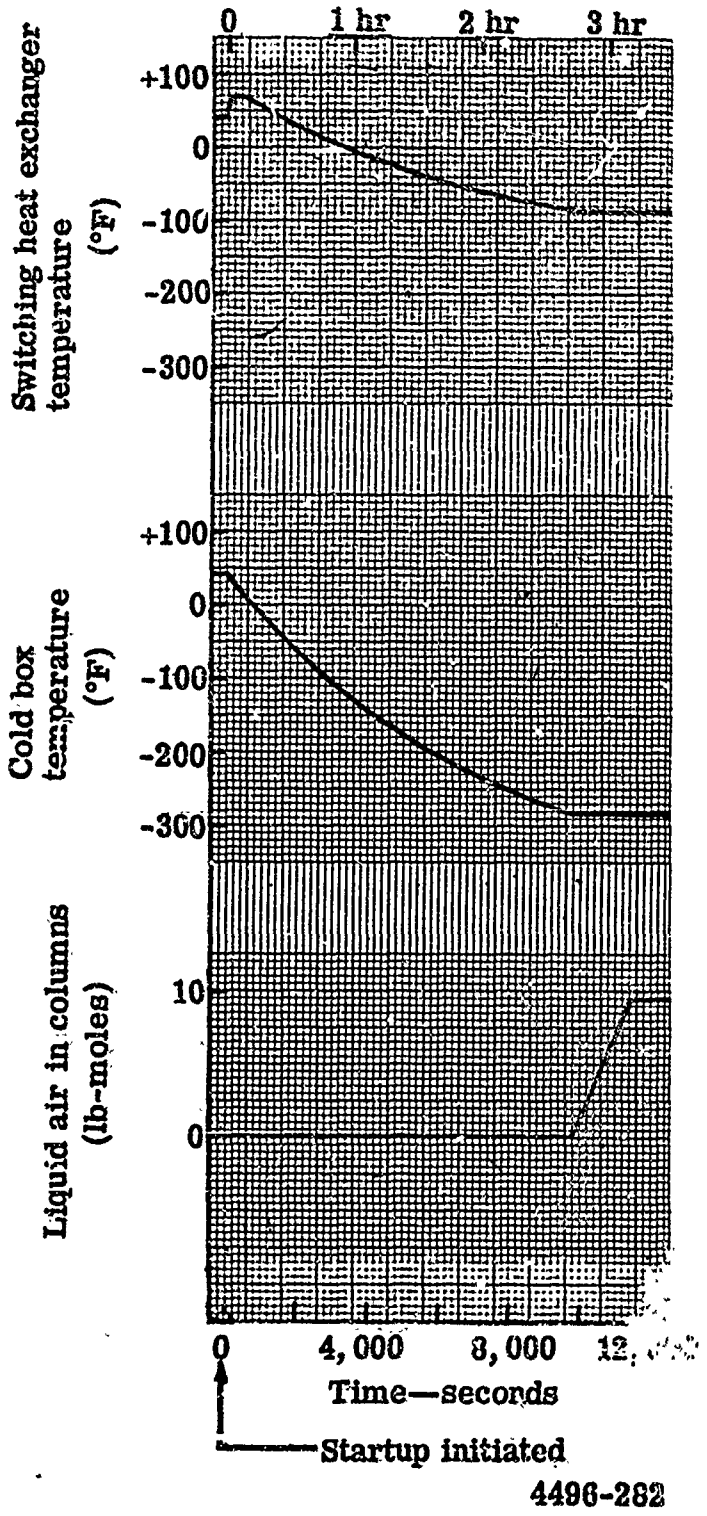


Figure B-30. Nitrogen generation subsystem start-up with 20% nitrogen flow after 48-hr shutdown—system 11.

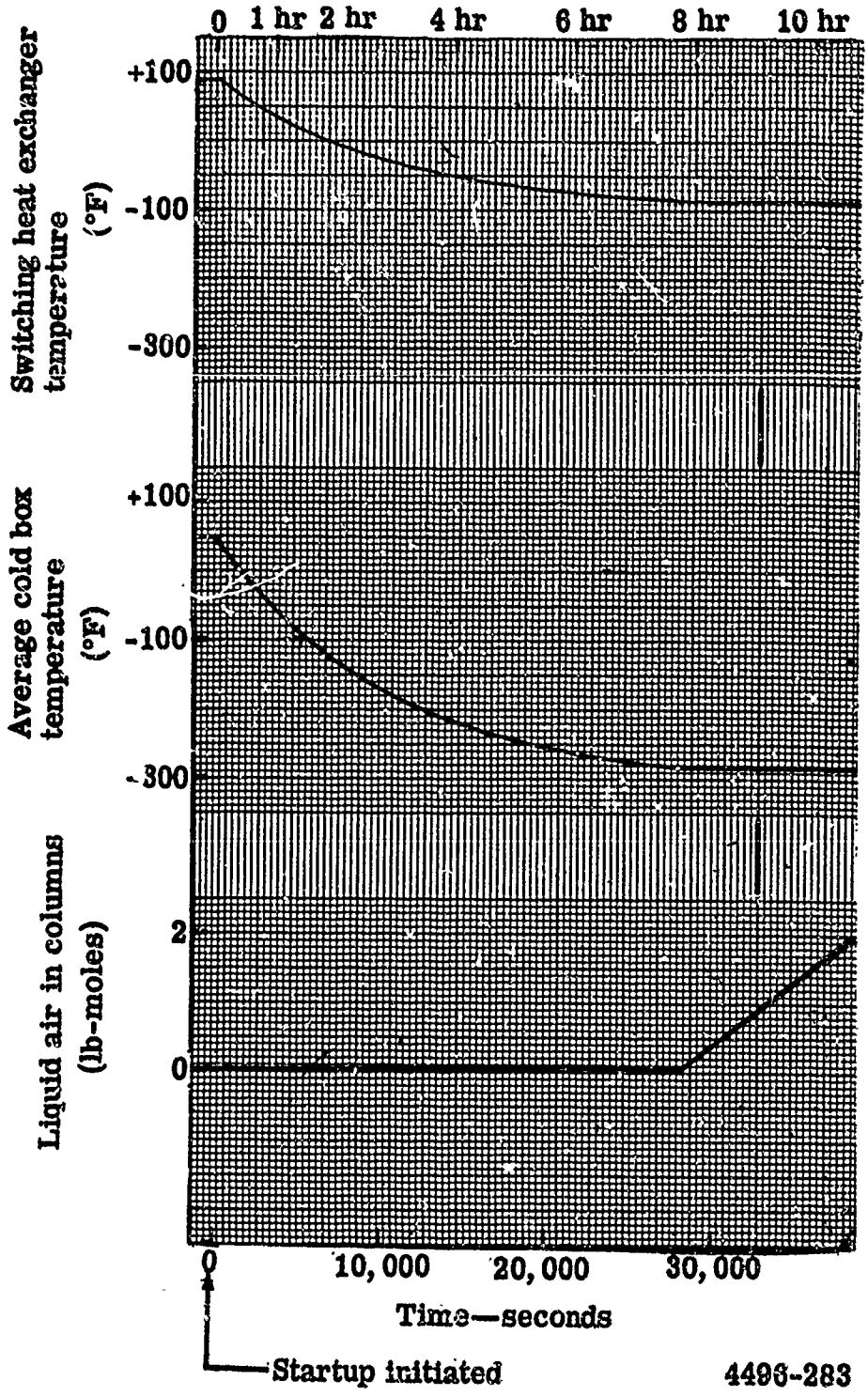


Figure B-31. Nitrogen generation subsystem start-up with no liquid nitrogen flow—systems 1, 2, and 3 (77°F initial subsystem temperature).

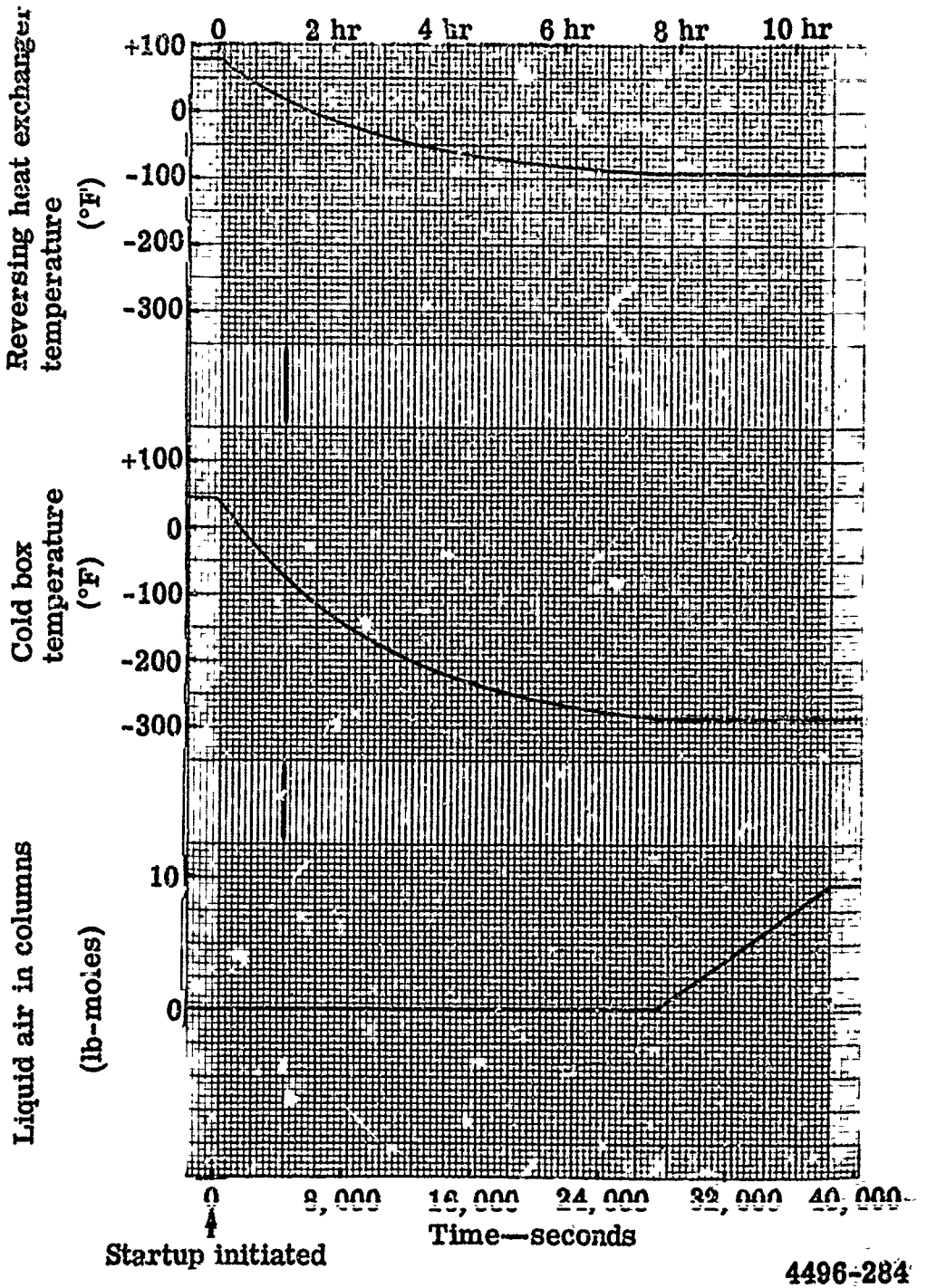


Figure B-32. Nitrogen generation subsystem start-up with no liquid nitrogen flow—system 11 (77°F initial subsystem temperature).

cells, but is also dependent on the temperature of the electrolyte. The maximum voltage of the power conditioning subsystem, as specified, limits the cells to less than full design power when the electrolyte is at ambient temperature. Therefore, only 60% of full hydrogen production is available initially. If a d-c voltage, greater than 619v, were available, additional hydrogen would be available initially and the time to achieve full hydrogen production would be shortened. This voltage limitation does not present any problem to the start-up of the fuel production system however since full hydrogen or oxygen production is not required until all phases of system start-up are completed. See Control Analysis portion of Section III.

Table B-I gives the calculated time to reach full hydrogen production assuming that an average of 20% of the power available to the electrolysis modules during the warm-up process produces heat. In the actual system, the power which results in heat varies from 15 to 25% during start-up.

Table B-I.
Time for the hydrogen production subsystem to reach steady-state operating temperature.

System	Temperature stabilization time (min)
1	32
2	31
3	48
4	25
5	27
6	26
7	27
8	27
9	22
10	24
11	19

AMMONIA SYNTHESIS SUBSYSTEM OPERATING POINT OPTIMIZATION

Since the ammonia fuel product removes hydrogen and nitrogen from the synthesis loop at a fixed ratio, the feed ratio to the ammonia synthesis system must be controlled to maintain the desired hydrogen-to-nitrogen molar ratio in the ammonia synthesis loop. Optimization of the molar ratio in the synthesis gas stream is accomplished by a sampling and comparison technique that biases the nitrogen flow rate slightly to change the ratio within the ammonia synthesis loop at a very slow rate. In the control subsystem specification, this control is connected to the converter temperature rise and synthesis gas flow rate which is an indication of ammonia synthesis rate.

The controls which produce the optimum hydrogen-nitrogen ratio in the ammonia synthesis converter were simulated using a combination of analog and digital equipment. These controls sample a process signal and commit it to analog memory through a track-and-hold amplifier. After an interval of time, the signal in memory is compared with the present process signal. If the difference between the remembered signal and the present signal shows the system to be progressing to a more nearly optimum operating point, i.e., the ammonia synthesis rate has increased, the nitrogen bias remains in the same direction. If the system is shown to be progressing to a less optimum operating point, the bias on the nitrogen flow is reversed so that hydrogen-nitrogen ratio in the synthesis loop returns toward a more optimum point.

Figure B-33 shows the computer diagram for the sampling and optimizing program. One-shot multivibrators 00 and 01 produce the basic timing for the system. Flip flops 1 through 4 produce gate input signals which operate the track-and-hold amplifier and compare its output with the incoming signal at the proper time. Flip flop 5 reverses the direction of the nitrogen bias signal when the system output moves away from optimum.

Although the ammonia synthesis rate does not increase permanently, there is a transient increase in production that is sensed by the controller and the proper correction is made. The analog run (Figure 3.8-12 in the Control Analysis portion of Section III) shows the transient starting at a 3-to-1 hydrogen-nitrogen molar ratio and going to the optimum 2.5-to-1 ratio.

Other variables can be used to give an indication of optimum operating point when a fixed amount of synthesis gas is available. Synthesis loop pressure (Figure 3.8-12 of Section III) decreases as the nitrogen flow rate is biased to achieve an optimum hydrogen-nitrogen ratio. This pressure would increase if the hydrogen-nitrogen synthesis gas molar ratio drops below the optimum point.

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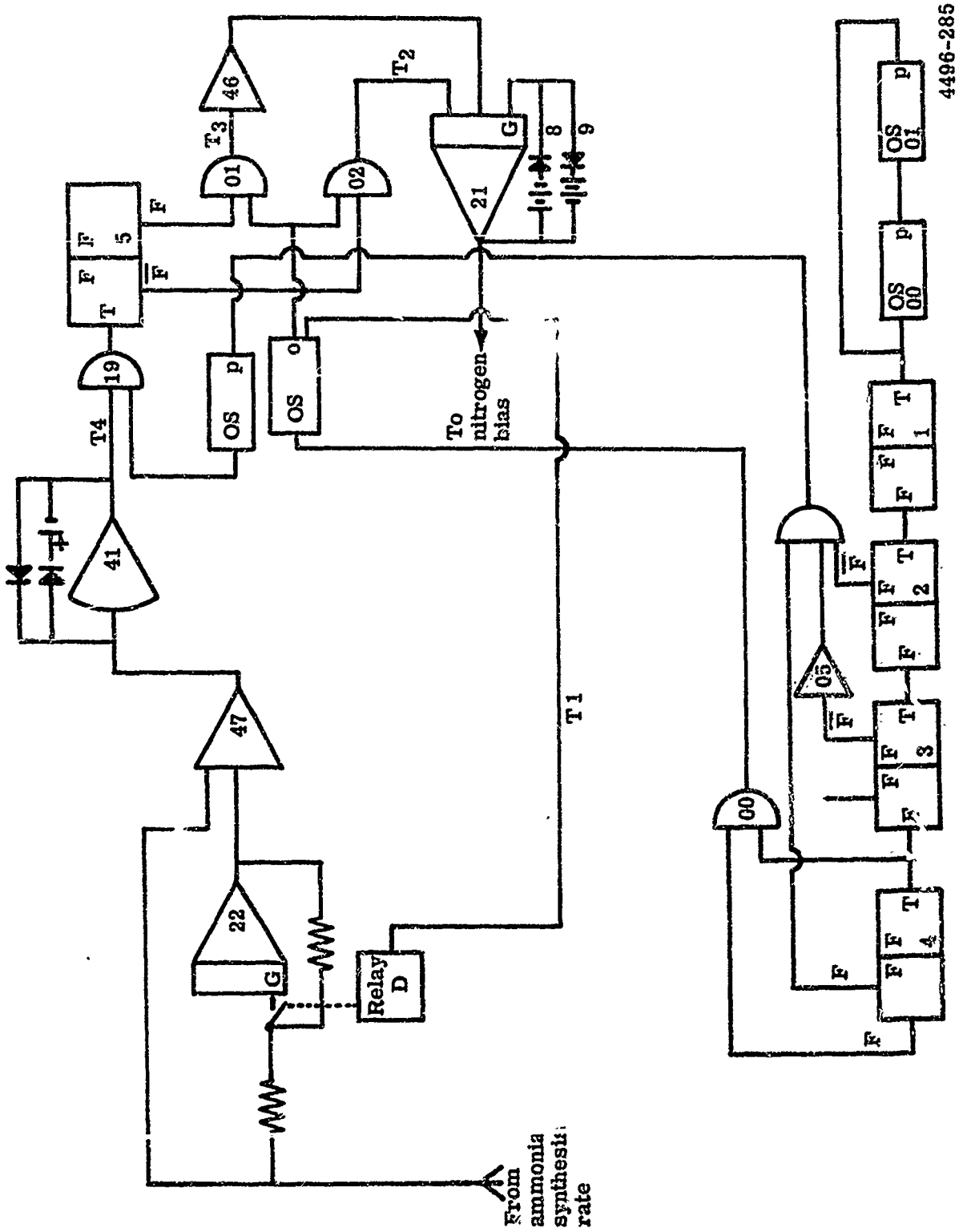


Figure B-33. Ammonia synthesis rate optimization analog program.

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Allison Division of General Motors, Indianapolis, Indiana; Final Report—Ammonia Production Systems Integration; Authors—B. N. Thomson, N. T. Mills, G. E. Hook, R. M. Reinstrom, R. A. Coval, C. L. VanSickle, and D. L. Dimick; Allison EDR 4496. This report consists of 435 pages, including 197 illustrations. 22 February 1966. Contract No. DA-44-009-AMC-1166(X).

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Unclassified
Security Classification

DOCUMENT CONTROL DATA - R&D

(Security classification of title, body of abstract and indexing annotation must be entered when the overall report is classified)

1 ORIGINATING ACTIVITY (Corporate author)
Allison Division of General Motors
P.O. Box 894
Indianapolis, Indiana 46206

2a. REPORT SECURITY CLASSIFICATION
Unclassified

2b GROUP

3 REPORT TITLE
FINAL REPORT
AMMONIA PRODUCTION SYSTEMS INTEGRATION

4 DESCRIPTIVE NOTES (Type of report and inclusive dates)
Final Report (22 June 1965—22 February 1966)

5. AUTHOR(S) (Last name, first name, initial)
Thomson, Bruce N. Reinstrom, Robert M. Dimick, David L.
Mills, Norman T. Coval, Robert A.
Hook, Gerald E. VanSickle, Carl L.

6. REPORT DATE	7a. TOTAL NO. OF PAGES	7b. NO. OF REFS
22 February 1966	435	40

8a. CONTRACT OR GRANT NO.
DA-44-009-AMC-1166 (X)

b. PROJECT NO.

c.

d.

9a. ORIGINATOR'S REPORT NUMBER(S)
EDR 4496

9b. OTHER REPORT NO(S) (Any other numbers that may be assigned this report)

10 AVAILABILITY/LIMITATION NOTICES
U.S. Government agencies may obtain copies of this report directly from the Defense Documentation Center (DDC). Other qualified DDC users should request through Director, USAERDL, Fort Belvoir, Virginia

11. SUPPLEMENTARY NOTES

12. SPONSORING MILITARY ACTIVITY
Research and Development Procurement Office; U.S. Army Engineer Research and Development Laboratories; Fort Belvoir, Virginia 22060

13. ABSTRACT

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DD FORM 1 JAN 64 1473

Unclassified
Security Classification

Unclassified

Security Classification

14. KEY WORDS	LINK A		LINK B		LINK C	
	ROLE	WT	ROLE	WT	ROLE	WT
Ammonia synthesis						
Chemical reactions						
Fuel systems						
Fuels						
Air fractionation						
Purification						
Power equipment						
Electrolysis						
Decomposition						

INSTRUCTIONS

1. ORIGINATING ACTIVITY: Enter the name and address of the contractor, subcontractor, grantee, Department of Defense activity or other organization *(corporate author)* issuing the report.

2a. REPORT SECURITY CLASSIFICATION: Enter the overall security classification of the report. Indicate whether "Restricted Data" is included. Marking is to be in accordance with appropriate security regulations.

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3. REPORT TITLE: Enter the complete report title in all capital letters. Titles in all cases should be unclassified. If a meaningful title cannot be selected without classification, show title classification in all capitals in parenthesis immediately following the title.

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6. REPORT DATE: Enter the date of the report as day, month, year; or month, year. If more than one date appears on the report, use date of publication.

7a. TOTAL NUMBER OF PAGES: The total page count should follow normal pagination procedures, i.e., enter the number of pages containing information.

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Unclassified

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