NASA-CR-145282

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January 20, 1978

(NASA-CR-145282) STUDY OF THE POTENTIAL FOR IMPROVING THE ECONOMICS OF HYDROGEN LIQUEFACTION THROUGH THE USE OF CENTRIFUGAL COMPRESSORS AND THE ADDITION OF A HEAVY WATER PLANT (Union Carbide Corp.) 146 p HC / A07/MF A01

N78-15564
Unclas
G3/44 55571

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NASA CONTRACTOR REPORT 145282

STUDY OF THE POTENTIAL FOR IMPROVING THE ECONOMICS OF HYDROGEN LIQUEFACTION THROUGH THE USE OF CENTRIFUGAL COMPRESSORS AND THE ADDITION OF A HEAVY WATER PLANT

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CONTRACT NAS 1-14698
DECEMBER 1977

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Space Administration
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PREFACE

This study is a continuation of two previous investigations of the use of liquid hydrogen as a jet aircraft fuel. The original study entitled "Survey Study of the Efficiency and Economics of Hydrogen Liquefaction" was carried out under Contract NAS1-13395, April 1975, for the National Aeronautics and Space Administration, Langley Research Center, and was documented in Report NASA CR-132631. That report covered the production of gaseous hydrogen starting with coal. The products of coal gasification were cleaned, processed and subsequently liquefied and stored for use as a jet fuel.

A second study entitled "Study of the Potentials for Improving the Efficiency and Economics of Liquid Hydrogen Produced from Coal" was carried out under Contract NAS1-14120, July 1976, and was documented in Report NASA CR-145077. That study included the investigation of operating the Koppers-Totzek gasifier under pressure, the use of low sulfur compliance coal for power generation, the use of Solvent Refined Coal as a gasifier feedstock, the reduction of leakage losses in the production and liquefaction system, the employment of wet turbines in the liquefaction cycle, the establishment of a thermodynamic grading system for coal gasification and a survey of acid gas removal systems applicable to coal gasification plants.

The potentials for improving the economics of hydrogen liquefaction by the substitution of centrifugal for recycle compressors in the liquefaction process and the addition of a heavy water plant were the subjects of investigation in this study.

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The work presented in this report was carried out by members of the Industrial Gas Process Division, Linde Division, Union Carbide Corporation, Tonawanda, New York, 14150, initially under the direction of Richard L. Shaner, Manager, and subsequently under the direction of Charles R. Baker, Consultant, for the duration of the project. Philip Leckner, Engineer, was also a principal contributor. Mr. Robert D. Witcofski, of the Aeronautical Systems Division of NASA Langley Research Center was technical monitor for the contract.

UNITS OF MEASURE
AND
DOLLAR BASE

All computations performed under this contract were in U.S. Customary engineering units.

In compliance with Form PROC./P-72, all results are presented in the International System of Units (SI) followed, in parentheses, by the U.S. Customary equivalent from which they were converted.

To be consistent with the preceding studies "Survey Study of the Efficiency and Economics of Hydrogen Liquefaction", NASA CR-132631, and "Study of the Potentials for Improving the Efficiency and Economics of Liquid Hydrogen Produced from Coal", NASA CR-145077, all economic values are given in terms of mid-1974 dollars. Adjustment must be made for inflation since mid-1974 to bring costs up to date.

1.0 INTRODUCTION

Liquid hydrogen has been identified as an attractive alternate aviation fuel for the future. Fuel cost is an important parameter in determining the viability of such a fuel for this application.

Studies concerning the production and liquefaction of hydrogen for use as a jet aircraft fuel have recently been completed. This work, done under Contract NAS1-13395 "Survey Study of the Efficiency and Economics of Hydrogen Liquefaction", NASA CR-132631, and Contract NAS1-14120 "Study of the Potentials for Improving the Efficiency and Economics of Liquid Hydrogen Produced from Coal", NASA CR-145077, have shown that it is feasible to produce large quantities of liquid hydrogen starting with coal as the raw material. These studies have also defined equipment requirements for providing the large quantities of liquid hydrogen involved and have shown such plants to be enormous in size and complexity. This applies particularly to the hydrogen compressors required in the liquefaction of hydrogen where processes used to date call for massive, multi-stage reciprocating machines. The capacities required have far exceeded the capacity of even the largest machines available today and thus multiple units have been specified.

The use of compact centrifugal machines for the compression of hydrogen is not feasible because of hydrogen's low density. A high-pressure centrifugal hydrogen compressor would require an inordinate number of impellers as a result of the low pressure head developed by each impeller. Centrifugal compressors are therefore generally uneconomical in the conventional hydrogen liquefaction process.

A novel approach to the liquefaction of hydrogen has been developed which permits the application of standard centrifugal compressors in place of reciprocating machines. Briefly, a second fluid such as propane is added to the hydrogen prior to compression to form a mixture having a molecular weight much greater than that of hydrogen alone so that a standard type of centrifugal compressor can be used. After compression, this mixture is cooled to cryogenic temperature levels where the propane condenses out of the mixture and is separated as a liquid. The low freezing point of propane permits attainment of the low temperatures and essentially complete liquefaction of propane. The liquid propane is then throttled and readmitted to the cold returning hydrogen recycle streams for recovery of latent and sensible heats prior to recompression. The refrigeration section of the process utilizes conventional refrigeration techniques such as work expansion in turbines and Joule-Thomson throttling to generate the low temperatures required for the liquefaction of hydrogen. Ortho-para conversion is carried out in the usual manner.

The advantages of this system are:

1. Use of standard centrifugal compressors with their attendant low cost.
2. Increased reliability of standard centrifugal compressors compared with reciprocating compressors, giving greater on-stream performance.
3. Reduced maintenance.
4. Smaller and simpler compressor foundations.
5. Simpler installation of compression equipment.
6. Reduced space requirements.
7. Reduced amount of nitrogen refrigeration required.

During the liquefaction of hydrogen, a small quantity of deuterium is produced. In a very large liquefaction plant, the amount of deuterium produced is quite sizeable and may warrant recovery and sale as a potential co-product. The deuterium can be separated from the liquid hydrogen by fractional distillation in a column.

Deuterium, when oxidized, forms heavy water. Heavy water is used in certain nuclear reactors as a neutron moderator to reduce the neutron velocity and enhance the probability of neutron collision with uranium nucleii.

It is the purpose of this study to perform analyses of the potential economic benefits of both the substitution of centrifugal compressors for the reciprocating compressors used in the hydrogen liquefaction process and the addition of a heavy water plant to the hydrogen liquefaction cycle.

2.0 OBJECTIVES

The objectives of this study are to:

2.1 Investigate the potential economic advantages of the substitution of of centrifugal compressors for the reciprocating compressors used in the liquefaction of hydrogen and to determine the effect of such a substitution on the thermal efficiency of the liquefaction process.

2.2 Investigate the potential economic advantages of the addition of a heavy water plant to the hydrogen liquefaction cycle.

3.0 SUMMARY OF ACCOMPLISHMENTS

3.1 Substitution of Centrifugal for Reciprocating Compressors.

It has been determined that centrifugal compressors can be substituted for reciprocating compressors in the liquefaction of hydrogen and that such substitution results in both technological improvement and economic gain. The feature which makes feasible the use of centrifugal compressors is the admixture of a high molecular weight component (propane in this study) with the hydrogen prior to compression to increase the molecular weight, and hence the density, of the fluid being compressed. This same feature requires that the liquefaction process be revised to permit separation of the propane from the hydrogen subsequent to compression and remixing as part of the refrigeration process.

Two liquefaction processes have been comparatively evaluated. One process is the conventional scheme using reciprocating compressors for the compression of recycle hydrogen while the other is a novel arrangement adapted for the separation and remixing of the hydrogen-propane recycle stream, using centrifugal compressors to meet recycle compression requirements. It was determined that a 15% propane-85% hydrogen mixture should be used as the recycle fluid. With this composition, the centrifugal compressor process required a unit power for liquefaction of 13.27 kWh/kg (6.21 kWh/lb) compared with a unit power of 13.05 kWh/kg* (5.92 kWh/lb) for the conventional process. Although the centrifugal compressors required approximately 15% more power because of the added propane, there were offsetting power reductions resulting from increased power return from the hydrogen turbines and a lower feedstock pressure, and overall net power requirements are greater by only 4.8%.

Capital investment for the liquefaction complex was reduced 13% through the use of less costly centrifugal compressors. The cost advantage of the centrifugal compressor was partly negated by the 100% increase in size of the heat exchangers used for cooldown of the feed and recycle streams.

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* This unit power is greater than the value of 12.50 kWh/kg (5.67 kWh/lb) previously reported(1). The difference results from the use of manufacturer's data for the specific hydrogen compressors (Table 11) which shows lower efficiencies than the 80% assumed in the earlier study.

Steam turbines were found to be the preferred drivers for the centrifugal recycle compressors, whereas reciprocating compressors were driven by synchronous electric motors. The centrifugal compressor process was therefore powered by a gas fueled steam generating system with steam turbines driving the centrifugal compressors directly and steam turbine-generator sets providing for electrical power needs. The reciprocating compressor process used gas turbine-generator sets to provide electricity for all power requirements. A significant advantage of the steam power supply was the ability to use the heat generated in the water gas shift for preheating boiler feed water. Sufficient heat was available to reduce the net energy input to the power plant section by approximately 6% and consequently decrease the size of the power gasifier. There was no application for this heat in the gas turbine power supply and it was wasted to cooling water.

The steam turbine power system was found to have a considerably lower cost than the gas turbine power system and was principally responsible for the $179 million capital investment reduction of the coal gasification complex. Other differences are minor and result from the slightly smaller power gasifier required to provide fuel gas for the centrifugal compressor process. The feedstock gasifier is identical for both cases.

Total plant investment for the centrifugal compressor process amounts to $2002 million of which $1389 million is for coal gasification and the remainder for the cryogenic equipment. For the conventional process, plant investment totals $2274 million of which $1569 million is for coal gasification. The difference of $272 million represents a 12% investment advantage in favor of centrifugal compressors. For DCF financing, the total unit cost for the centrifugal compressor process ranges from \$0.929/kg (42.13 ¢/lb) to \$1.159/kg (52.58 ¢/lb), depending on the cost of coal. Corresponding costs on a unit energy basis are \$6.54/GJ (\$6.90/MM Btu) to \$8.16/GJ (\$8.61/MM Btu). These costs are 9.9%-8.4% lower than unit costs for the reciprocating compressor process. For utility financing, unit costs range from \$0.693/kg (31.44 ¢/lb) to \$0.920/kg (41.72 ¢/lb) or from \$4.88/GJ (\$5.15/MM Btu) to \$6.48/GJ (\$6.83/MM Btu), down 9.3% - 7.5% from those for the conventional process.

3.2 Addition of a Heavy Water Plant

Deuterium exists primarily as hydrogen deuteride in the liquid hydrogen plant product, and can be effectively and economically recovered by fractional distillation of the hydrogen with minimum product loss. At a liquid hydrogen capacity of 26.25 kg/s (2500 TPD), an annual production of 1.053 Gg (1161 Tons) of heavy water can be recovered which is approximately equal to the total current production by the Canadian heavy water industry and about 20% of the planned capacity after expansion in the early 1980's.

The required capital investment for production of heavy water at this capacity totals \$95.4 million, which includes, in addition to the cryogenic dueterium recovery equipment, allowances for additional recycle compression capability to power the deuterium unit, as well as the enlargement of the power gasifier for the same purpose. Total unit production cost, based on DCF financing, for the heavy water is \$31.64/kg (\$14.35/lb) for coal at \$0.332/GJ (35¢/MM Btu) and increases only a modest amount to \$36.55/kg (\$16.58/lb) for coal at \$0.711/GJ (75¢/MM Btu). For utility financing, the unit cost is 25-30% less. Based on a \$121.25/kg (\$55/lb) market price, the sale of this heavy water would generate a net annual income in the range of \$89 - \$104 million, which is sufficient to defray at least 9.2%, and as much as 17.8%, of the cost of the liquid hydrogen product. It is also sufficient to generate a 1-year payout on the heavy water plant investment.

Although the natural occurrence of deuterium in hydrogen is about 1 part in 7000, it has been found to be somewhat variable. The normal concentration in fresh water has been placed at 148 ppm and the concentration in coal is apparently about the same. Processing in the coal gasifier section will alter this concentration via isotopic exchange between HD and H_2O to yield H_2 and HDO. The deuterium in HDO is lost via condensation in the process water. The loss is most severe in the water gas shift converter where a combination of unfavorable equilibrium temperature and presence of a catalyst promote the exchange toward HDO. Condensation and recycling of the process steam serves to prevent this loss and maintain the concentration of HD in the feedstock hydrogen at 280 ppm (140 ppm as D in H).

4.0 SUBSTITUTION OF RECIPROCATING WITH CENTRIFUGAL COMPRESSORS

4.1 Role of Reciprocating Compressors in Hydrogen Liquefaction

Recently completed studies(1) concerning the production and liquefaction of hydrogen for use as a jet aircraft fuel have identified the equipment necessary to produce the amount of fuel required to meet the expected needs of a major airport in the 1990 decade. The quantity of liquid hydrogen to be produced at such a facility will range up to a daily output of 2.27 Gg (2[illegible]0 Tons). A major investment item in a hydrogen liquefaction plant is the compression machinery which typically amounts to over 40% of the total investment. Although there are several functional applications, the major one is the recompression of the stream of recycle hydrogen which is used to produce the refrigeration for the process region below temperatures of 80°K. This is illustrated in the utility summary given in Table 13 of the Survey Study Report (1) where 56% of the total plant electrical power requirement is consumed by the hydrogen recycle compressors. The process location of these compressors is given in Figure 5 of this report and they are identified as items P-RC and P-SCF on this flow sheet.

To date, compression requirements have been met with massive, multi-stage reciprocating compressors. These compressors are the largest available in today's market place. A typical machine for this purpose would have a balanced-opposed arrangement of 10 cylinders, with a cylinder diameter up to 111.8 cm (44 in.), a stroke of 43.2 cm (17 in.) and overall dimensions of 12.19 m (40 ft.) long by 8.53 m (28 ft.) wide. This compressor would have a power rating of up to 11,180 kW (15,00 HP) so it is obvious that to meet total compression requirements of 635,000 kW for a 26.25 kg/s (2500 TPD) plant, an array of multiple units, numbering at least 57, and installed in parallel, would be required. As shown later in this study, each compressor has a lower than maximum power rating and the total complement of machines is 80.

4.2 Reciprocating vs. Centrifugal Compressors

Hydrogen is quite amenable to compression via reciprocating compressors which function by physically reducing the volume of a fixed quantity of gas admitted to the cylinder. Hydrogen possesses no physical nor thermodynamic attribute which contributes markedly to the difficulty of such a process. The basic principle of compression via centrifugal compressors, however, is quite different and this difference causes difficulty in compressing light, or low density, gases such as hydrogen. A centrifugal compressor functions by imparting kinetic energy to the gas by action of a high-speed rotating impeller. Upon discharge from the impeller, the kinetic energy is converted in the diffuser to static pressure. A low density gas, such as hydrogen, can possess appreciable levels of kinetic energy only at high velocities, and because impeller speeds are normally limited due to physical and material constraints, it follows that a single impeller is capable of producing a much smaller pressure rise for a low density than for a high density gas. A centrifugal compressor in hydrogen service would therefore require a large number of impellers to attain even a moderate discharge pressure level. This subject will be taken up in more detail in paragraph 4.3, entitled Adiabatic Head.

There are a number of reasons which would make centrifugal compressors otherwise attractive for use in hydrogen service.

- Centrifugal compressors are available in larger sizes than reciprocating compressors with respect to both flow and power rating. Machines exist which are capable of handling 612,000 m^3/h (360,000 ICFM) of gaseous flow or as much as 52,200 kw (70,000 HP) of power. This larger capacity for a given machine results in lower capital costs for compression equipment.
- The larger capacity per machine for centrifugal vs. reciprocating compressors reduces the number of machines which must be installed and reduces the complexity of the installation. Piping for gas and cooling water, coolers, power feeder systems and electrical switchgear, and compressor control systems are all simplified.

- Centrifugal compressors are typically more reliable and thus capable of longer periods of continuous operation than reciprocating compressors. This reduces plant outage, cuts maintenance costs and permits reduction or even elimination of spare machines. Minimum periods of 24,000 hours or 3 years of uninterrupted operation are common for centrifugal compressors. For reciprocating machines operating on dry gas at normal pressures, 5,000 hours may be more typical although, because of the larger number of vulnerable components such as valves, piston rings, and packing rings, there is greater probability for failure and there will be greater variability in the period of uninterrupted operation.
- Centrifugal compressors are subject to fewer maintenance problems than reciprocating compressors. For centrifugal compressors, maintenance is primarily directed to bearings and seals. Modern design of journal and thrust bearings is so well advanced that bearing maintenance is seldom a problem, even at high speeds. Low maintenance can be expected for both bearings and seals for well-balanced, well-aligned rotating parts handling clean gas streams.
- For reciprocating compressors, valves, piston rings and packing rings provide for most of the maintenance. Valves are one of the most critical items because of the cyclical nature of their operation, the impact loads during opening and closing, the high gas velocity through them and the wear on the valve seats. The large number of moving parts, where there exist a set of suction and a set of discharge valves for each cylinder, perhaps several valves per set and several valve plates

and springs per valve, also help to make the valves subject to increased maintenance. Piston and packing rings are subject to wear and must be periodically replaced.

- Foundations are simpler for centrifugal than for reciprocating compressors. Two basic requirements must be met. The first requirement is to keep the centrifugal compressor and its driver in perfect alignment and the second is to carry and distribute the weight of the compressor and driver to the soil or other supporting structure. Poured reinforced concrete slabs are generally used with steel sole plates anchored in the concrete. This provides a suitable surface for alignment of the equipment.

  Foundations for reciprocating compressors require more mass than for an equivalent centrifugal compressor. The fundamental requirement of the foundation for a reciprocating compressor is to withstand the forces imposed upon it by the operating machine in addition to the need for weight distribution and equipment alignment. Design of the foundation to withstand all loads and operating forces can often be quite complex.

- Centrifugal compressors differ from reciprocating compressors in their operating speed. Centrifugals are high-speed machines and usually run at speeds measured in thousands of RPM while reciprocating machines are much slower and typically operate at speeds measured in hundreds of RPM. This has an influence on the selection of the driver. The low speed synchronous electric motor is a natural choice for the reciprocating compressor and is widely used. Gas engines are also low speed machines and have enjoyed wide application as drivers for reciprocating compressors, particularly where low cost natural gas or process off gas is available as fuel.

For centrifugal compressor, a high speed driver is normally selected which can consist of a steam turbine, a gas turbine or an electrical induction motor which is connected directly to the compressor. Speed mismatches between compressor and driver can be accommodated by geared speed reducers but these are usually avoided if possible because of the cost and energy loss attendant with their use.

4.3 Adiabatic Head

The problems of compressing hydrogen gas using centrifugal compressors can be best illustrated through the concept of adiabatic head. At this point a general discussion of the fundamentals of centrifugal compression is in order and, although somewhat elementary, is also informative.

Head is a measure of unit energy imparted to the gas stream by the compressor. In customary units, it is expressed as ft lb of energy per lb of throughput or frequently referred to as simply "feet of head". The corresponding SI units would be joules per kg. The use of the term adiabatic head implies that the compression process is adiabatic which may or may not precisely describe the actual process. However, being frequently used, it is chosen here for illustrative purposes to describe how molecular weight influences centrifugal compression.

Adiabatic head may be expressed in equation form.

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$$H_{ad} = ZRT \frac{k}{K-1} \left[Pr^{\frac{k-1}{k}} - 1 \right] \quad (1)$$

Z - Compressibility Factor

R - Gas Constant

T - Inlet Temperature

k - Adiabatic Compression Exponent $= C_p/C_V$

C_p - Specific Heat at Constant Pressure

C_V - Specific Heat at Constant Volume

P_r - Compression ratio $= P_i/P_o$

P_i - Inlet pressure

P_o - Discharge pressure

The role of molecular weight of the gas being compressed enters through the gas constant where

$$R = \frac{8.307}{MW} \; \frac{J}{g,K} \quad \text{or} \quad \frac{1554}{MW} \; \frac{ft\ lb}{lb,R} \tag{2}$$

MW = Molecular weight.

For low molecular weight gases, the head, or unit energy required increases as the molecular weight decreases, all other parameters remaining constant. Note however that total energy required does not change, because of the inverse proportionality between unit energy and molecular weight.

The impeller of a centrifugal compressor is designed to give a specific value for head. The relation is:

$$H_{ad} \text{ (per impeller)} = \frac{\text{Const}}{g_c} V_T^2 \qquad (3)$$

V_T = Impeller tip speed

g_c = gravitational constant

Head thus increases rapidly with tip speed, which is constrained by the material strength of the impeller as it is stressed by centrifugal forces. Maximum head per impeller is typically 29.9 - 35.9 J/g (10,000 - 12,000 ft lb/lb) and may go as high as 44.8 J/g (15,000 ft lb/lb).

With a fixed maximum head imposed on the compressor, the effect of molecular weight is upon the discharge pressure, Equation (1). Discharge pressure is a direct function of molecular weight. Figure 1 illustrates this relationship for the compression of gases of different molecular weights in a centrifugal compressor having an assembly of impellers which develop a total head of 314 J/g (105,000 ft). This could be 7 impellers developing 44.9 J/g (15,000 ft) of head each, 10 impellers developing 31.4 J/g (10,500 ft) of head each, or any other similar combination. It is assumed that all gases have the same value of the adiabatic compression exponent, $k = 1.4$, and that the suction pressure is atmospheric. The discharge pressure decreases from 197 psia for a 30 molecular weight gas to a value as low as 18.9 psia for a molecular weight of 2 (e.g. hydrogen). To develop a discharge pressure of 197 psia for hydrogen would require a total adiabatic head of 4661 J/g (1,560,000 ft) for which as many as 156 impellers, each developing 29.9 J/g (10,000 ft), would have to be provided. Since a compressor body rarely contains more than 10 impellers, we are talking about at least 16 bodies for this particular application. Thus, for the compression of pure hydrogen gas, designers have resorted to reciprocating compressors for all applications except perhaps for low discharge pressures.

The effect of molecular weight on the adiabatic head shown in equations (1) and (2) apply to the effective molecular weight of a gas mixture as well as to the molecular weight of a pure gas. Thus, a light molecular weight gas such as hydrogen (MW = 2.016) can be admixed with a heavier molecular weight gas such as propane (MW = 44.097) to give a mixture having an intermediate molecular weight which is a function of composition. The compositional variation of molecular weight for hydrogen-propane mixtures is presented on Figure 2. In order to achieve a 5-fold increase in molecular weight, for example, propane must be added to bring its concentration to 19.2%. The use of an additive having a higher molecular weight than propane would reduce the amount of additive required to attain the same molecular weight for the mixture.

The effect of propane additive on the discharge pressure for the compression of hydrogen-propane mixtures is shown on Figure 3. This graph assumes a suction pressure of 290 kPa (42 psia) and an adiabatic head of 44,800 J/kg (15,000 ft) per impeller. (The suction pressure level may be recognized as that of the main recycle hydrogen compressor of the hydrogen liquefier.) The adiabatic compression exponent, k, is taken as a function of composition. Thus, while an 8-impeller compressor can develop a discharge pressure of only 384 kPa (55.7 psia) when compressing pure hydrogen, it can discharge at 1765 kPa (256 psia) when compressing a 30-70 propane-hydrogen mixture.

The addition of propane to pure hydrogen streams to allow centrifugal compression is not without disadvantage, however. For a 30% propane concentration, the added propane increases the compressor throughput by 43% and produces a considerable increase in compressor power. There is also the problem of separating the propane from the compressed gas mixture unless one has application for the hydrogen-propane mixture, which situation does not apply to the hydrogen recycle stream of the conventional hydrogen liquefier. However, a hydrogen liquefaction process has been devised (2) using centrifugal compressors to compress a recycle stream of hydrogen plus an additive for the purpose of increasing the effective molecular weight, and then to separate the additive from the hydrogen by cryogenic means in the liquefaction process.

4.4 Hydrogen Liquefaction Process

4.4.1 Process Designation

For the purposes of this report, the hydrogen liquefaction process which uses centrifugal compressors to compress a recycle stream of hydrogen mixed with an additive will be designated Process CRC, which is an acronym for Centrifugal Recycle Compressors. The conventional hydrogen liquefaction process which uses reciprocating compressors to compress a recycle stream of pure hydrogen will be designated Process RRC, which is an acronym for Reciprocating Recycle Compressor.

4.4.2 Requirements of Process

The purpose of the hydrogen recycle stream in a hydrogen liquefaction process is to provide a suitable working fluid for the generation of refrigeration at temperature levels below about 80°K. This hydrogen is:

1. Work expanded through expansion turbines exhausting at temperatures as low as approximately 26°K, and

2. Throttled through a valve to produce liquid hydrogen at temperatures below 21°K.

Since these temperatures are below the triple point of all other materials, except helium, there is a clear necessity for removing, entirely, the admixed component. This can be done by cooling the compressed gas mixture, partially condensing the additive, and separating the condensed phase from the gaseous hydrogen. If an additive component having suitable properties is used and the proper process conditions are imposed, then the condensation/separation can yield a high degree of separation from the hydrogen. This is achieved by using an additive component which has a wide separation between the boiling and freezing points. A wide separation typically produces a very low vapor pressure at the freezing point so that the equilibrium vapor phase concentration will be similarly low. It also provides for a relatively high boiling point so that cooling requirements in the warm section of the system can be kept low.

The process must also provide for the remixing of the added component with the low pressure return recycle hydrogen and for recovery of its refrigeration values. Failure to do this would result in intolerable process inefficiencies.

4.4.3 Propane as an Additive

Although there are many materials which would be a satisfactory additive, propane is well qualified for several reasons and is the material of choice here. Propane has a molecular weight of 44.1 so that excessive amounts are not required to produce a mixture that can readily be compressed in centrifugal compressors. Figure 2 presents the relationship between composition and weight. Propane, with an atmospheric boiling point of 231.1 K and a triple point of 85.5 K, possesses a wide liquid range so that much of the condensing heat load occurs at warm temperature levels and nearly all of the propane will be condensed at temperatures approaching the triple point. For example, a 15% propane-85% hydrogen mixture at a pressure of 2137 kPa (310 psia), which are process conditions of interest as will subsequently be discussed in fuller detail, will start to condense (i.e. its dewpoint) at 257 K and the contained propane will be 94% condensed at a temperature of 200 K. At 95 K the propane concentration in the equilibrium vapor is estimated to be considerably less than 1 ppm. Propane can therefore be almost completely removed by condensation in the temperature range between ambient and liquid nitrogen.

Availability and low cost are also desirable attributes of an additive material so that inevitable process losses do not impose a severe economic burden. Propane possesses both of these attributes.

4.4.4 Process Description - Process RRC

The process for hydrogen liquefaction using reciprocating recycle compressors is conventional as depicted in the process flow sheet, Figure 14, and does not differ from that portrayed in the report of original study (1). The following process description is excerpted from that report.

> "The hydrogen liquefier consists of a nitrogen precooled, expander process in which hydrogen is recycled to provide refrigeration at three temperature levels below 80°K. Two levels of refrigeration are provided by the hydrogen turbines and the third by Joule-Thomson throttling of a portion of the high pressure recycle hydrogen. At and above 80°K, refrigeration is provided by 1.) a stream of cold nitrogen gas which is used to help precool the combined feed and recycle stream and 2.) a stream of

nitrogen liquid which is used for additional cooling as well as for partial ortho-para conversion of the feed stream. The hydrogen feed stream is further converted, continuously, down to the temperature level of the exhaust of the cold turbine, after which it is throttled, passed through a catalytic converter for trimming purposes, and then subcooled in heat exchange with hydrogen boiling at low pressure."

"Cold exhaust streams from the expanders are warmed in counter-current heat exchange with cooling hydrogen streams, combined and finally returned to the suction of the recycle compressor. Vaporized hydrogen from the final subcooler is combined with flash vapor from throttling, warmed in heat exchange with a cooling stream and returned to suction of the subcooling-fluid compressor which, in turn, discharges to the suction of the recycle compressor."

4.4.5 Process Description - Process CRC

The CRC hydrogen liquefaction process will be described with the aid of the process flow diagram, Figure 4. This was constructed for use with the process design computer model and shows only the portion of the hydrogen liquefier for the temperature range between ambient and 80 K.

For lower temperatures, the process is the same as for the conventional hydrogen liquefier, Process RRC, and is shown on Figure 5. Stream numbering at the interface between Figure 4 and Figure 5 is consistent.

The compressed recycle stream, stream number 133, consisting of a mixture of hydrogen and propane, is cooled in heat exchanger X-1A to a temperature level where most of the propane has condensed. After a phase split at separator S-4 the hydrogen, stream 101 , is further cooled in heat exchanger X-1B where most of the residual propane is condensed out and the gas phase contains less than 1 ppm propane. Another phase split is taken at the separator S-5 and the resulting hydrogen gas stream (No. 108) is passed to an adsorber for removal of final traces of propane before entering heat exchanger X-1C. The now-pure hydrogen gas stream number 7 emerging from heat exchanger X-1C is split into two streams, the major stream (stream number 8) being further cooled with boiling liquid nitrogen in heat exchanger XC-2 and the minor stream being fed to heat exchanger X-7 (Figure 5) as stream number 50.

Recycle hydrogen which has been expanded in the low temperature hydrogen turbines returns at a lower pressure as stream number 45. Similarly, stream 55 is a low pressure return stream of hydrogen which was liquefied and used to subcool the liquid hydrogen product (heat exchanger X-6, Figure 5).

The liquid propane which was removed in the phase cut at separator S-4 is throttled and divided into major (stream 105) and minor (stream 103) streams, whose flows are in proportion to the quantity of returning recycle hydrogen, for the purpose of reconstituting each stream at its proper propane content. The remixing of the liquid propane with the returning hydrogen stream before entering heat exchanger X-1A is an important and necessary feature for optimum performance. The hydrogen in the mixture serves to reduce the partial pressure of the propane in the vapor phase and thus reduce the temperature at which the mixture starts to boil. Refrigeration for the condensation of the compressed recycle stream is thereby made available at lower temperatures, which provides an increase in temperature driving force in the heat exchanger and reduces the heat transfer surface requirement.

The liquid propane removed in the phase cut at separator S-5 is likewise split into two streams (numbers 112 and 110) throttled and mixed with returning recycle hydrogen streams. However, since most of the condensation occurred before the recycle stream entered heat exchanger X-[illegible]B, the amounts of liquid remixed with the hydrogen at this level will be quite small.

All low pressure return recycle streams are rewarmed to ambient temperature and are recompressed to the desired recycle pressure in compressors designated P-SCF and P-RC.

Meanwhile, the purified, compressed product stream (No. 1) is fed to the liquefier and cooled as a separate stream in the several heat exchangers. There is no purpose in mixing it with the recycle stream and, in fact, it is even disadvantageous to do so because the increased amount of hydrogen would cause the condensation to occur at lower temperatures as previously described for the boiling, low-pressure return stream. The consequence of this arrangement would be a decrease in the temperature driving force and an increase in the heat transfer surface requirement. The final stage of heat exchange for the product stream is a cooldown using boiling liquid nitrogen and a partial ortho-para conversion. The product stream leaves exchanger-converter XC-2 as stream number 12 for further processing as shown on Figure 5.

Refrigeration to sustain the process at temperature levels higher than about 80°K is provided in part by liquid nitrogen and cold nitrogen gas. This is generated in the Nitrogen Refrigerator, shown in flow seet, Figure 6. Some refrigeration is supplied by the propane in the recycle stream but this is generally inadequate to supply the total requirement. Liquid nitrogen is used in exchanger-converter XC-2 to desuperheat product and recycle hydrogen streams as well as to absorb the heat evolved in the ortho-para conversion. Nitrogen vapor from this operation is combined with a stream of cold, low-pressure nitrogen vapor returning from the hydrogen purifier (stream number 92) and warmed to ambient temperature for refrigeration recovery as is a stream of cold, high-pressure gas from the nitrogen refrigerator (stream number 80). These warmed nitrogen streams are recycled to the nitrogen refrigerator for cooling and liquefaction.

4.5 Case Study

A series of case studies was conducted using a process computer program modeled to represent the hydrogen liquefier, Process CRC, as illustrated by the flow diagram, Figure 4. The purpose of the case studies was to evaluate the effect of various process parameters and to optimize the process. Of particular interest, was the effect of recycle stream composition and pressure on the process power requirements and on the surface requirements for the cooldown heat exchangers, X-1A, X-1B and X-1C.

Results from 11 selected cases are presented in Table 1. The effect of recycle stream composition is shown by comparison of several pairs of cases. Reducing propane content from 30% to 25% resulted in a power reduction of 6.3% (Case No's 10 and 11) while a further reduction from 25% to 15% propane resulted in an additional power reduction of 9.4% (Case No's 3 and 9). A comparison of Case No's 1 and 5 shows that,with a 3.9% power increase, an optimum does not exist at a 20% propane concentration.

The effect of pressure is shown, for a 20% propane content, by Case No's 5-8 where process power decreases continuously with pressure to 2070 kPa (300 psia), the lowest pressure considered. For a 15% propane content, an optimum pressure occurs at a pressure of 2240 kPa (325 psia) although the pressure-power relationship is relatively flat over the 2070-2410 kPa (300-350 psia) pressure range.

The size of the cooldown heat exchanger, as measured by the UA product, also decreases with decreasing propane content and is relatively invariant with recycle pressure. At high propane concentrations, the existence of temperature pinch conditions in the cooldown process is responsible for the great increase in heat transfer surface requirements and even results in an increase in warm temperature approach.

As a result of the case studies the following process parameters were selected for continuing comparative evaluations between Process CRC and Process RRC.

Recycle Stream Composition	- 85% H_2 - 15% propane
Recycle Pressure	- 2140 kPa (310 psia)
Warm Temperature Approach	- 8 K

Although the 2140 kPa recycle pressure was not optimum in the case studies it was near optimum and was selected instead based on adiabatic head considerations. The compression from 290 kPa (42 psia) to 2140 kPa (310 psia) requires an adiabatic head of 690 J/g (230,800 ft). This is already in excess of 43.1 J/g (14,400 ft) per impeller for a 16-impeller configuration or 37.4 J/g (12,500 ft) per impeller for an 18-impeller configuration so that the lower recycle pressure was considered to be a prudent choice.

Stream data for the selected operating parameters for the Process CRC hydrogen liquefier are tabulated in Table 2 in SI units and in Table 3 in corresponding U.S. Customary units. Stream data for the accompanying nitrogen refrigerator for Process CRC are presented in Tables 4 and 5. For Process RRC, stream data which were given in Tables 3 and 4 of a previous publication (1) are presented again for reference purposes in Tables 6 through 9. Stream identification may be made by reference to Figure 4 for Process CRC hydrogen liquefier, to Figure 5 for Process RRC hydrogen liquefier, and to Figure 6 for the nitrogen refrigerator.

4.6 Comparative Process Studies

Comparative studies were made of the hydrogen liquefaction process using centrifugal recycle compressors (Process CRC) and the process using reciprocating recycle compressors (Process RRC) for the purpose of defining a complete liquid hydrogen facility for each process and permitting economic comparisons to be made. In each case, the facility included, in addition to the hydrogen purification and liquefaction units, the hydrogen feedstock coal gasifier section, the power plant coal gasifier section and all auxiliary and supporting equipment.

4.6.1 Compressor Specifications

For each process, a set of specifications was drawn up for the hydrogen liquefier recycle compressors based on the process data contained in Tables 3 and 7. These specifications were then submitted to appropriate compressor manufacturers to obtain performance and cost data. Specific compressors were finally selected through evaluation of the manufacturer's response.

4.6.1.1 Centrifugal Compressors

The specifications for centrifugal compressors for Process CRC are presented in Table 10 together with manufacturer's performance data. There are two compressor functions involved.

1. To compress the flash hydrogen used for subcooling (and then mixed with propane) from near atmospheric pressure to 42 psia, the suction pressure of the main recycle compressor and
2. To compress the main recycle stream of hydrogen and propane from 290 kPa (42 psia) to the 2137 kPa (310 psia) recycle pressure.

Centrifugal compressors can be generally categorized according to two main types of configuration. One is the horizontally split compressor and the other is the vertically split or barrel compressor. The difference is in the way the casing is designed for disassembly. The barrel compressor uses what amounts to a pressure vessel for a casing and is normally used for high pressure applications. Because the casing joint is less extensive, it is also preferred for low molecular weight gases which diffuse readily. A barrel compressor therefore leaks less gas than does a horizontally split compressor when used in hydrogen service. In fact, an API specification (3) defines the conditions for which barrel compressors must be used when compressing hydrogen or hydrogen-containing mixtures. These conditions are shown on Figure 7 which reveals that for the 2140 kPa (310 psia) discharge pressure of the main recycle compressor, the barrel design is mandatory. The flash compressor, which discharges at 290 kPa (42 psia), can use the horizontally split casing.

The centrifugal recycle compressor configuration for Process CRC is shown on Figure 8. This is for a single 2.625 kg/s (250 Tons/day) liquefier module. A 26.25 kg/s (2500 Tons/day) plant would require 10 such units. The main recycle machines would consist of four identical machines, 6-impellers per machine, all on a single shaft and driven by a single steam turbine. This is a 3-stage arrangement because the first stage is a split-flow, parallel arrangement.

The flash compressor consists of 2 machines, 5 impellers per machine, arranged in series on a single shaft and driven by a single steam turbine. Functionally, only one stage of compression is involved in this operation.

The total power required for a 10-module, 26.25 kg/s (2500 Tons/day) complex is 753.00 MW for the recycle gas compression and 31.99 MW for the flash gas compression, for a total of 784.99 MW (1,052,700 HP).

Space requirement for the 10 main recycle compressors and the 10 hydrogen flash compressors for the 26.25 kg/s (2500 TPD) facility amount to a

6.10 m x 21.3m(20 ft. x 70 ft.) area for each main recycle machine and a 5.18 m x 15.2 m (17 ft. x 50 ft.) area for each flash machine. Allowing additional space around each machine in amount equal to twice the space occupied by the machine for clearance purposes gives a total area requirement of approximately 0.65 hectares (1.6 acres). A layout consisting of 5 rows of compressors with 2 compressors per row would require a plot having overall dimensions of 61.0 m x 64.0 m (200 ft. x 210 ft.) for the main recycle machines and 45.7 m x 51.8 m (150 ft. x 170 ft.) for the flash machines.

4.6.1.2 Reciprocating Compressors

The specifications for reciprocating compressors for Process RRC are presented in Table 11 together with manufacturer's performance data. As with the centrifugal compressors, there is a set of main recycle compressors and a set of flash gas compressors, although streams of pure hydrogen rather than hydrogen-propane mixtures are compressed.

These compressors are non-lubricated in a balanced, opposed arrangement, with each compressor having ten cylinders, with diameters up to 1.12 m (44 in.) and a stroke of 0.43 m (17 in.). They are slow speed machines, operating at 300 RPM, compared with the centrifugal compressors which operate at 4500-6000 RPM.

The reciprocating recycle compressor configuration for Process RRC is shown in Figure 9 for a single 2.625 kg/s (250 Tons/day) liquefier module. There is a total of seven reciprocating machines, in parallel, for the compression of the main recycle hydrogen stream. The arrangement is depicted schematically with each stage of the compressor represented by a single cylinder. The true physical arrangement would use multiple cylinders in parallel for each stage, giving a total complement of 10 cylinders for each compressor. The flash gas compressor uses a single machine identical to all the others.

Comparison of the compressor configurations for the two competing processes, shows a greater simplicity of arrangement and manifolding for the centrifugal compressors. There is also greater consolidation of inter and aftercooling heat exchangers. The centrifugal compressors require a total of 4 coolers compared with 22 coolers for the reciprocating compressors.

The total power required for a 10 module, 26.25 kg/s (2500 Tons/day) complex is 649.9 MW for the recycle gas compression and 22.37 MW for the flash gas compression, for a total of 672.27 MW (901,500 HP). This is 14% less power than required by the centrifugal compressors resulting almost entirely from the absence of propane in the recycle gas. Indeed, the main recycle centrifugal compressor is more efficient than its reciprocating ccunterpart although the reverse is true for the smaller flash gas compressor.

Space requirements for the 70 main recycle compressors and the 10 hydrogen flash compressors for the 26.25 kg/s (2500 TPD) facility amount to a 8.69 M x 12.2 M (28.5 ft x 40 ft) area for each machine. Allowing additional clearance space around each compressor in amount equal to twice the space occupied by the compressor gives a total area requirement for the 80 compressors of approximately 2.5 hectares (6.3 acres). A layout consisting of 10 rows of compressors with 8 in each row would fit in a plot 146 M x 174 M (480 ft x 570 ft). The plot size is four times the size of the plot required for the centrifugal compressors.

4.6.2 Compressor Drivers

Reciprocating compressors are slow speed machines and match, very well, the speed of large, synchronous electric motors. Electric motor drive was assumed in previous studies (1) and was also assumed in the present study. The electricity was produced by electrical generators driven by gas turbines which were fueled by the gaseous product fuel from the power plant coal gasifier.

Centrifugal compressors are high speed machines and match, very well, the speed of steam turbines. Usually direct drive can be used without the need for intervening gear box for speed reduction purposes. Furthermore, steam turbines have the capability for variable speed, which can be applied for capacity control of the compressor. Steam turbine drives were assumed for the centrifugal recycle compressors of Process CRC as well as all other centrifugal compressors used in the plant. A single turbine with a 746 MW (100,000 HP) rating is the specified driver for the 4-unit main recycle compressor, Figure 8. The feasibility of a steam turbine mechanical drive of this power rating has been confirmed by a major supplier of steam turbines. The reciprocating hydrogen feed compressor and the forecooler of the nitrogen refrigerator are electric-motor driven and the power for these functions as well as for other electrical requirements of the plant is supplied by steam turbine driven electrical generators.

4.6.3 Steam System

The extensive use of steam turbine drives for Process CRC mandated the installation of a steam generation and supply system to power the plant. For this process, the steam power system supplants the gas turbine driven electrical power system used for Process RRC. The system is represented schematically in Figure 10 which shows a dual pressure level arrangement. The higher pressure steam is used for the large, major power requirements such as the main recycle compressors, the feedstock and fuel gas compressors and the electrical power generation. The lower pressure steam is used in the steam turbine drives for the smaller, remaining centrifugal compressors and also as make up steam for powering the oxygen plant compressors.

The steam power system presented an opportunity for recovery of low level process heat and this advantage was applied to the preheating of boiler feed water. The source of heat is the exothermic heat of reaction at the water gas shift reactor. This heat is normally transferred to cooling water and wasted. The amount of heat involved is 0.544 GJ/s (1.856 x 10^9 Btu/hr) which is sufficient to preheat the boiler feed water 75°K. It constitutes nearly 10% of the total energy input to the steam system and, accordingly, produces a significant economy in the overall energy supply.

4.6.4 Coal Gasifiers

The coal gasification equipment is categorized as either belonging to the Feedstock Gasifier Section or the Power Plant Gasifier Section. The Feedstock Gasifier Section generates the crude hydrogen feedstock, is shown as a block flow diagram in Figure 11, and is identical for both Process RRC and Process CRC.

The Power Plant Gasifier Section generates the fuel gas for heating and power. Because of differences in power requirements for the two competing processes, there will be differences in corresponding Power Plant Gasifiers. This is revealed in the two block flow diagrams where Figure 12 is the diagram of the

Power Plant Gasifier required for Process RRC and Figure 13 is, similarly, for Process CRC. The gasifier size for the latter process is 2-3% smaller than for the former.

4.6.5 Hydrogen Purification and Liquefaction

The crude hydrogen feedstock is treated to remove all impurities which would otherwise freeze out in the subsequent liquefaction process. Because the quantity of feedstock and the extent of purification is the same for the two competing processes, the purifier is the same in each case. The liquefaction process, however, is inherently different, as described previously. One manifestation of the difference is in the heat transfer area requirements of the cooldown heat exchanger. This is identified as heat exchangers X-1 and X-8 on Figure 5 for Process RRC and X-1 (subdivided into sections A, B and C) on Figure 4 for Process CRC. The increase in flow of the recycle stream plus the latent heat of condensation and evaporation of the propane component place a much heavier heat load on this exchanger for Process CRC, specifically, about 2.3 times as great. However, the contortion of the cooling curves (T vs ΔH) caused by the propane phase change produces a larger mean temperature difference, so that the greater heat load is partially compensated. The final design calls for 124,000m ^{2} (1,330,000 sq.ft.) for Process CRC which is twice the area requirement of 60,900 m^2 (655,000 sq.ft.) for Process RRC.

The cooldown and condensation of the hydrogen-propane mixture is a partial condensation process in which the propane must diffuse through a hydrogen gas film before it can reach the heat transfer surface and condense. A mass transfer resistance is thus imposed upon the heat transfer process and requires additional heat transfer surface. This was accounted for in the design of the cooldown heat exchanger, using the method of Colburn and Hougen (4). Block flow diagrams of the cryogenic equipment are shown in Figures 14 and 15 for Processes RRC and CRC, respectively.

Power requirements for the purification-liquefaction section for both processes are listed in Tables 12 and 13, although the centrifugal recycle compressors, including the flash compressors, require about 17% more power than do the reciprocating compressors, the overall power required is only 5% greater because of other gains resulting from the elimination of the hydrogen booster compressor, and the greater return from the hydrogen turbines. More power is returned by the turbines because they are required to produce more refrigeration to overcome the increased warm end temperature approach losses caused by the recirculation of additional recycle hydrogen through the turbines. Additional hydrogen must be recirculated to counterbalance the effect of the reduced expansion pressure ratio across the turbines. Tables 12 and 13 also reflect the losses due to inefficiencies in the electric motor drivers, which are sparingly applied in Process CRC. Table 13 lists the individual power as either electrical or mechanical according to whether the driver is electric motor or stream turbine. Losses due to inefficiency of the steam turbines appear as part of the steam requirements shown on Figure 10.

4.7 Economic Comparisons

4.7.1 Methods

Complete economic analyses were conducted on the two competing processes based on both a Discounted Case Flow (DCF) basis, which is representative of industrial financing, and a Utility financing basis. These financing methods have been previously described (1) and the financing rules for each method are enumerated in Table 14 together with shortcut equations for determining unit costs via each method. The basis for the economic evaluations is given in Table 15.

4.7.2 Gasification Economics

Capital investment requirements for the 26.25 kg/s (2500 TPD) coal gasification complex for each competing process are compared in Table 16. Included are all operating areas as listed and all necessary auxiliary and supporting equipment for a completely installed grass roots facility except for the acquisition of land. Primarily because of a cost reduction in the power and steam generation area and also to a lesser extent because of the slightly smaller power gasifier, the capital investment in the coal gasification section which supports Process CRC is less by $179,400,000 or 11.4%. The steam turbine drive including the steam generation system has shown to be a lower investment arrangement than the electric motor drive with the gas turbine/generator system.

Annual operating costs for Processes RRC and CRC are compared in Table 17. The cost of coal is the major item in the overall cost and, as in the previous study (1), the price of coal is varied over a range from $0.332/GJ (35¢/MM Btu) to $0.7113/GJ (75¢/MM Btu). Operating costs vary from 4 to 5 1/2% lower for the gasification section which supports Process CRC, solely as a result of the smaller power plant gasifier.

Unit gasification costs on a DCF basis are presented in Table 18 and on a Utility Financing basis in Table 19. For Process CRC, unit costs vary from $0.5663/kg (25.69¢/lb) to $0.9701/kg (44.00 ¢/lb) depending on the financing method and the price of coal. These costs are 6.4 to 8.9% less than the unit gasification costs for Process RRC.

4.7.3 Liquefaction Economics

Capital investment requirements for the 26.25 kg/s (2500 TPD) liquefaction complex for the two processes are compared in Table 20. The liquefaction complex includes the hydrogen liquefier, the cryogenic purifier, the nitrogen refrigerator, the hydrogen feed compressor and all on-site ancillary and supporting equipment as an installed grass roots facility. Process CRC shows a $77,200,000 advantage in total plant investment and a $93,100,000 advantage in total capital requirement. Investment differences result mostly from the lower cost of centrifugal compressors compared with reciprocating compressors, which accounts for about 60% of the difference. Decreased cost of associated equipment result from simpler foundations and buildings, reduction in power wiring and electrical switchgear, and elimination of the hydrogen feedstock booster compressor. An additional $14 million is required for Process CRC, however, to provide for the extra cooldown heat exchangers in the hydrogen liquefier.

Annual operating costs are compared in Table 21. The only differences here lie in the lowered cost of maintenance supplies and reduced local taxes and insurance, both being investment related. Table 21 does not contain entries for the cost of feedstock or liquefaction energy. Both are derived from coal and consequently these costs are borne by the feedstock and power gasifiers and have not been allocated to the liquefaction process. Similarly, in Table 22, feedstock and liquefaction energy are not included in the unit liquefaction cost.

Unit liquefaction costs are presented in Table 22 for both DCF and Utility financing. For both financing methods, the unit liquefaction cost is about 14% lower for Process CRC.

4.7.4 Overall Economics

The total investment for the overall facility is the sum of the investments for the gasification and liquefaction complexes and is summarized in Figure 16. The total of $1,886,000,000 for Process CRC is 12% less than the total plant investment for Process RRC.

Table 23 and Figure 17 present the total unit cost for generation and liquefaction of hydrogen for the two competing processes and for the two financing methods. Based on DCF financing, Process CRC produces liquid hydrogen which is approximately \$0.11/kg (5¢/lb) lower in cost, varying with the cost of coal. For utility financing, the advantage still lies with Process CRC to the extent of approximately \$0.073/kg (3.3¢/lb). For either type of financing, the reduction in cost on a percentage basis is nearly the same, varying from 7.5% to 10%, primarily as a function of the cost of coal.

4.7.5 Reliability Economics

One of the favorable attributes of a centrifugal compressor compared with its reciprocating counterpart is greater reliability. Earlier discussion noted that specifications calling for three years of continuous, uninterrupted operation are now quite common. This reliability is seldom achieved with reciprocating machines. The economic impact of this greater reliability has not been included in the economic analyses presented thus far. The reason for this is not neglect but rather the uncertainty in ascribing a typical reliability or on-stream factor to the reciprocating compressor. The problem was resolved by conducting a separate economic analysis for reliability which would permit the evaluation to be made as a function of reliability.

The procedure which was adopted assumed that the centrifugal compressors were completely reliable and that the reciprocating compressors were effectively made to be completely reliable by the installation of spares. A specific situation was selected wherein one spare recycle compressor was arbitrarily provided for a pair of 2.625 kg/s (250 TPD) liquefier modules. For the 10-module plant, there was a total of five spare compressors backing up 70 operating compressors. The five spare compressors out of a total of 70 provides for an outage if 7.1%, which becomes the basis for the reliability economics.

The cost of providing this reliability is shown in Table 24 and amounts to an additional plant investment of $13,269,000. The corresponding incremental unit liquefaction cost amounts to $0.0039/kg (0.18¢/lb) based on DCF financing and $0.0024/kg (0.11¢/lb) based on utility financing. These costs apply for a 92.9% reliability; a direct linear relationship between cost and reliability may be assumed for other reliability levels.

4.8 Thermal Efficiency

Thermal efficiencies for Process RRC and Process CRC are presented in Tables 25 and 26, respectively. The overall facility for each process is subdivided into four major sections and the thermal efficiency of each section is presented individually to more clearly illustrate where major energy losses occur. Process CRC consumes less coal than the other process (495.4 kg/s vs. 502.7 kg/s) and, because the liquid H_2 product output is identical for both, should have a slightly higher overall thermal efficiency; its tabulated value of 25.92% (HHV basis) is indeed slightly better than the 25.54% for Process RRC.

There are some thermal efficiency differences between the two processes for the individual sections. The feedstock gasifier shows a considerably better efficiency for Process CRC which is attributed to the recovery of process heat from the shift converter. Only minor efficiency differences exist for the power plant gasifier and the hydrogen liquefier sections. The thermal efficiency for energy conversion is somewhat greater for Process RRC resulting from the use of gas turbines which have a higher efficiency than the steam turbines. Process CRC also has an additional energy input requirement for the purpose of driving the boiler feed water pumps of the steam generating system which contributes slightly to the decrease in efficiency.

Thermal efficiencies are expressed on both lower and higher heating value bases and the heating values of the several process streams upon which thermal efficiencies depend are presented in Table 27. Distribution of process energy is depicted graphically in Figure 18.

5.0 ADDITION OF HEAVY WATER PLANT

5.1 Background Survey

Early in the development of nuclear power sources where heavy water was used as a neutron moderator, the distillation of liquid hydrogen was recognized (5) as the most attractive of several methods available for recovery of deuterium. The attractiveness exists, however, only for situations where liquid hydrogen production is the principle activity and the deuterium recovery unit is parasitic to the liquid hydrogen plant. Otherwise, at a natural occurrence of one part in about 7000 parts of ordinary hydrogen, the deuterium is too dilute to support, alone, the cost of the liquefaction plant. With the installation of massive production facilities for the supply of 26.25 kg/s (2500 TPD) of liquid hydrogen to meet commercial aviation needs, there is a potential source of deuterium to produce up to 1.12 Gg (1230 Tons) per year of heavy water from each facility.

The recovery of deuterium by distillation of heavy water has been described by several authors (5, 6, 7, 8, 9). Timmerhaus et al (10) have reported on the operation of a hydrogen distillation pilot plant at the NBS Cryogenic Engineering Laboratory in Boulder, CO. Commercially operating plants in Europe and Asia have also been described (8, 11, 12). Sufficient data, information and experience exist to permit the design of a deuterium recovery facility with a reasonable degree of confidence.

5.2 Natural Occurrence of Deuterium

Deuterium occurs in nature in the ratio of one atom of deuterium to approximately 7000 atoms of hydrogen, which translates to a concentration of 143 ppm. It commonly exists in combination with hydrogen as HD, or hydrogen deuteride. Because the hydrogen molecule is diatomic, the corresponding concentration of HD in hydrogen molecules is twice this value, or 286 ppm.

The concentration just cited is completely general. In practice, it is found to vary somewhat from one material to another or from one geographical location to another. For a valid evaluation of deuterium recovery from liquid hydrogen, the deuterium content of the hydrogen must be known which, in turn, depends on the deuterium content of the coal and water from which the hydrogen is generated.

The deuterium content of various natural waters is given by Kirshenbaum (13) where the concentration was found to vary over a range from 139 ppm to 153 ppm depending upon the source and type of the water. Brines and salt water tend to have slightly higher deuterium contents than fresh water. The low values are attributed to weather factors which cause isotopic fractionation and depletion of deuterium. The deuterium content of most of the waters analyzed fell within the range of 148 ± 2 ppm.

A search of the literature revealed just a single value for the deuterium content of coal. The National Bureau of Standards reported (14) that water obtained by the combustion of anthracite coal had a deuterium content that differs no more than 10 ppm from that of normal fresh water.

5.3 Deuterium Process Losses

Water and coal appear to have about the same deuterium content and a rounded value of 150 ppm (300 ppm of HD) was assumed for this evaluation. The exact value is not of great importance because, during subsequent processing, isotopic exchange occurs according to the following reaction. This can result in a loss of deuterium by transfer of D atoms to HDO and subsequent rejection in the process condensate,

$$HD + H_2O = H_2 + HDO$$

The equilibrium constant for this reaction is presented in Figure 19 after the data of Cerrai, et al (15). The exchange of D atoms from HD to HDO increases with decreasing temperature although Cerrai indicates that below about 600°C the uncatalyzed reaction is slow.

Equilibrium calculations based on process conditions encountered in the coal gasification process show only a 0.3% transfer to HDO at the high outlet temperature of the Koppers-Totzek gasifier. At the water gas shift converter, however, the temperature is lower, exchange is much greater, and an exit equilibrium concentration of 232 ppm HD is calculated, which is down from 300 ppm in the feed. The exit temperature of this process step is about 650°C and the reaction rate might be expected to be slow except that the shift catalyst apparently also catalyzes the deuterium exchange reaction. This is verified by a reported (16) reduction of HD concentration in commercially operating hydrogen plants where concentrations of 107-128 ppm deuterium (214-256 ppm as HD) were measured.

The loss of deuterium at the shift converter can be prevented by recycling the condensed process steam. To accomplish this, the shift converter effluent would be cooled and the condensed process water, which is enriched in HDO, would be added to the required amount of makeup water, reconverted to process steam and returned to the water gas shift converter.

For the purpose of determining the location and extent of deuterium losses in the subject coal gasification process, a deuterium balance was made on the entire feedstock gasifier, assuming that process steam is recycled at the shift converter. The balance is summarized on Figure 20. The principal loss in deuterium occurs after the water quench of the process gas following the gasifiers, where the concentration is down to 277 ppm from 300 ppm in the feed. Loss is minimized by recycling process water but cannot be entirely avoided because of leakage, blowdown and cooling tower drift. A gain is actually encountered in the water gas shift process as a result of the recycling, and the product hydrogen leaves the gasification section having an HD concentration of 282 ppm. Based on this result, the concentration of HD in the hydrogen feed to the liquefier was assumed to be 280 ppm.

5.4 Equilibrium Data

The substantial difference in vapor pressure between hydrogen, hydrogen deuteride, and deuterium makes separation possible by means of fractional distillation. Vapor pressures are given by Woolley and others (17) and may be expressed analytically by the following equations.

$$\text{For e - } H_2\text{:} \quad \log_{10} P = 3.76882 - 44.3450/T + 0.02093\,T \qquad (4)$$

$$\text{For HD:} \quad \log_{10} P = 4.17454 - 55.2495/T + 0.01479\,T \qquad (5)$$

For n - D_2: $\log_{10} P = 3.8561 - 58.4619/T + 0.02671\ T$ (6)

where P = vapor pressure, kPa

T = temperature, °K

For ideal liquids, the relative volatility, α^0_{1-2}, is expressed as the ratio of the vapor pressures. The ideal relative volatility for a 1-2 binary is, therefore, given by the following relation.

$$\log_{10} \alpha^0_{12} = \log \frac{P_1}{P_2} = (A_1 - A_2) - (B_1 - B_2)/T + (C_1 - C_2)T \quad (7)$$

As an example, substitution of numerical values for the H_2 - HD binary, produces the following relation.

$$\log_{10} \alpha^0_{H_2 - HD} = -0.40572 + 10.9045/T + 0.00614\ T \quad (8)$$

Liquid phase non-ideality is accounted for by the introduction of liquid phase activity coefficients.

$$\alpha_{12} = \frac{\gamma_1}{\gamma_2} \alpha^0_{12} \quad (9)$$

where α_{12} = actual relative volatility of component 1 with respect to component 2

γ_1 = activity coefficient of component 1

γ_2 = activity coefficient of component 2

α^0_{12} = ideal relative volatility of component 1 with respect to component 2

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Activity coefficients for both the H_2 - HD binary and the H_2 - D_2 binary, taken from Newman (18), are presented in Table 28. Deviations from the ideal liquid assumption are less than 5% for the H_2 - HD system and are minor but not negligible. Actual relative volatilities for this binary are shown on Figure 21 as a function of liquid composition for temperatures of interest. The principal separation, which occurs in H_2-rich mixture at temperatures of approximately 24°K, is made at a relative volatility of approximately 1.5. Enrichments of about 50% per theoretical tray are thus obtained.

The H_2 - D_2 system is somewhat less ideal than the H_2 - HD system, where deviations up to 11% occur.

5.5 Deuterium Recovery Process

The process for production of heavy water, D_2O, via the fractional distillation of liquid hydrogen is represented by the flow diagram, Figure 22. The initial step is a concentration of HD from 280 ppm in the liquid product from hydrogen liquefier to 6.8% in the principal distillation column. The more volatile H_2 is withdrawn as an overhead product containing 11 ppm HD while the enriched HD product is removed from the bottom of the column and fed to a secondary distillation column for further enrichment. The overhead product from the secondary column is returned to the bottom of the primary column as a secondary feed. Only 4% of the HD in the feed to the primary column is lost in the overhead product.

The secondary column is actually a double column arrangement with the two columns being connected thermally. The reboiler requirements of the upper, or low pressure, column are matched and met by the reflux condensation requirements of the lower, or high pressure, column in a manner similar to the double column arrangement used in cryogenic air separation plants. The upper column serves to further concentrate the HD, and an enriched product stream is withdrawn at the reboiler, warmed to ambient temperature, compressed, and passed through a catalytic reactor which converts the hydrogen deuteride to deuterium via the following reaction.

$$2HD = D_2 + H_2$$

The H_2, HD, D_2 mixture leaving the reactor is cooled and returned, as feed, to an intermediate tray of the lower column, where the deuterium is concentrated and removed as bottoms product from the reboiler. The more volatile H_2 and HD leave as overhead product and are transferred, as feed, to an intermediate point in the upper column for further separation.

The D_2 product, 99.8% pure, is reacted with oxygen in a burner to form heavy water, which is cooled and condensed.

$$D_2 + 1/2\ O_2 = D_2O$$

The D_2O production rate from a single 2.625 kg/s (250 TPD) hydrogen liquefaction module is 12.7 kg/hr (27.9 lb/hr) or 105,000 kg (116 tons) per year.

Several supporting streams are required for the operation of the deuterium recovery system. Cold hydrogen exhaust from turbine E-2 is transferred from the hydrogen liquefier to the primary distillation column, where it is condensed to drive the reboiler. It is then throttled into the overhead condenser where it is evaporated against condensing reflux and returned to the hydrogen liquefier as the low temperature recycle stream. The pressure and flow requirements of this stream are significantly different from those of the main recycle stream of a hydrogen liquefier which does not possess a deuterium facility. Flows are considerably higher and the temperature requirement at the condenser establishes a lower return pressure. Recompression of the additional portion required by the primary column is to a pressure intermediate between the suction and discharge pressures of the main recycle compressor. The recycle compressor configuration for a plant having deuterium recovery capability is therefore quite different from one that does not, as is the heat exchanger configuration of the hydrogen liquefier. The compressor arrangement is illustrated by the block flow diagram of Figure 23. The main recycle stream returns at a pressure of 165 kPa (24 psia) rather than at 290 kPa (42 psia) so extra compression equipment is required to restore it to its normal pressure. Figure 23 shows flash recompression to 165 kPa (24 psia)where it joins the main recycle stream and the combined stream is compressed to 290 kPa (42 psia). The portion of the recycle stream which originates from turbine E-1 returns at the normal 290 kPa (42 psia) pressure and a portion of this is diverted and compressed to 393 kPa (57 psia) for subsequent cooling and recycling to the primary HD column.

In addition to providing a stream to drive the primary distillation column, a stream must be provided to drive the lower reboiler of the secondary double column. The pressure of this stream must be greater than that of the E-2 turbine exhaust and it is, consequently, withdrawn from the flash hydrogen circuit of the liquefier after it emerges from heat exchanger X-7 The condensed stream, after leaving the reboiler, is transferred to the condenser of the primary column where it assists in reflux generation.

Addition of a deuterium recovery facility is therefore, not simply a matter of attaching the unit to the product line of an existing hydrogen liquefier because of the alterations required to the liquefier itself. Capacity enlargement of the power gasification section will also be required to provide the energy requirements of the deuterium unit. Although the modification of an existing liquefier would be possible, the preferred procedure would provide for deuterium recovery in the original design in order to allow for integration and optimization benefits.

5.6 Economics of Deuterium Recovery

For the recovery of deuterium from the liquid hydrogen plant, cost additions are encountered in the following three areas.

1. Deuterium separation equipment.
2. Recycle compression equipment to power the deuterium separation.
3. Power gasifier to provide energy for the additional recycle compression

Plant investment required for the deuterium separation units is given in Table 29 which includes investment in additional recycle compression equipment. Costs are based on installation of ten separate deuterium recovery units, one for each of the ten hydrogen liquefier modules. Each unit is completely installed and connected to its host liquefier. The $63,750,000 investment amounts to a 12.1% increase for the total hydrogen liquefaction complex, which includes all cryogenic equipment and facilities.

To provide the additional 28.5 m^3/s (3.620 MM CFH) of fuel gas for the expanded recycle compression requirement, the power gasification section is enlarged by 4.7%. Table 30 presents a summary of the incremental investment required for this expansion. Total plant investment for both deuterium recovery and power gasification equipment amounts to an additional $95,410,000, a 4.6% increase.

The power requirement for the recovery of deuterium is shown in Table 31, as the difference in recycle compression requirements for liquefiers operating with and without deuterium recovery. For the 10 deuterium recovery modules producing 1.05 Gg (1161 Tons) of heavy water annually, this difference amounts to 110,500 kW. For this power consumption rate, the net unit power amounts to 873 Wh/g (396 kWh/lb) of D_2O. This may be compared with the power consumption for the well-known GS process, widely used by the Canadian heavy water industry, reported (19) to amount to 700 Weh/g (318 kWeh/lb) of electrical energy plus an additional 23 GJ/kg (2900 kWth/lb) of steam thermal energy. The thermal energy content of the additional coal fired to the power gasifier to provide the 110,500 kW of additional electrical energy for the subject facility totals 11.7 GJ/kg (1475 kW_th/lb).

The economics of providing the electrical energy required to operate the deuterium separation facility was handled by treating the incremental fuel gas used to generate the electrical energy as a utility at a chargeable rate. A unit fuel gas cost was therefore computed for the three assumed coal prices and for both DCF and Utility financing methods. The alternative procedure of determining the investment and operating costs of the expanded power gasifier section, although generating the same results, would have tended to obscure the incremental fuel gas cost. Unit fuel gas costs, based on the incremental investment of Table 30 and the incremental operating costs of Table 32, are presented in Table 33, and range from \$1.12 to \$2.02 per GJ (\$1.19-2.13 per MM Btu).

Finally, the total unit cost of producing heavy water as a function of the cost of coal is presented in Table 34 for DCF financing, and in Table 35 for Utility financing. These tables also have consolidated into them the elements of operating cost. Total unit costs are also depicted graphically in Figure 24. Depending, again, upon the cost of coal and the financing employed, they vary from a low of about \$22.50/kg (\$10.20/lb) to a high of \$36.50/kg (\$16.60/lb).

For the purpose of illustrating the profitability of heavy water production, an economic summary was prepared and is shown in Table 36. The annual production of 1.053 Gg (1161 Tons) will require a \$95.4 million plant investment, but a \$121.50/kg (\$55/lb) selling price will generate over \$127 million of gross income. Total annual expenditures will vary from \$23.64 million to \$38.50 million leaving a net income in excess of expenditures ranging from \$89.2 million to \$104.1 million. Expenditures, as defined here, already include a 12% discounted rate of return on investment for DCF financing or a 15% rate of return on equity plus a 9% interest rate on debt for Utility financing, so that this net income represents return in addition to the required return. It is sufficient to produce an additional rate of return (not discounted) on investment of as much as 109% or, alternatively, will yield a payout of approximately one year. It may also be employed to help defray the cost of producing the principal plant product, liquid hydrogen and, if so used, would result in a cost reduction ranging from 9% to nearly 18%.

Recovery of deuterium from large liquid hydrogen installations is seen to be highly profitable but such profitability is quite dependent upon maintaining the assumed market price of \$125.25/kg (\$55/lb) under the assault of producing such large quantities of heavy water. The sensitivity of net revenue to variations in the price of heavy water as well as the cost of coal is shown in Figure 25. Although it is beyond the scope of this work to attempt to project future heavy water demand, the annual output of 1.053 Gg (1161 Tons) from a single 26.25 kg/s (2500 TPD) liquid hydrogen facility is about equal to the total current output of the Canadian heavy water industry and about 20% of the total planned capacity for the early 1980's. Although the exercise of caution is advised, severe market price erosion must occur before profitability completely disappears.

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TABLE 1

CASE STUDY

H_2 LIQUEFIER - PROCESS CRC

CASE NO.	1	2	3	4	5	6	7	8	9	10	11
Composition-% Propane	1.	15	15	15	20	20	20	20	25	25	30
Recycle Flow-Nm^3/s	253.7	248.2	242.1	218.4	269.7	236.1	216.7	209.5	276.8	216.4	232.1
Recycle Pressure-kPa	2070	2140	2240	2410	2070	2760	3450	3800	2240	4140	4140
Warm Temp. Approach-K	8	8	8	8	8	8	8	9	8	9	12
Power Requirements-MW											
Recycle compressor	70.40	70.09	70.09	70.44	74.46	75.45	76.73	75.70	80.46	82.21	80.30
Flash compressor	3.23	2.93	2.63	2.16	3.41	2.27	2.25	2.63	2.96	2.36	2.00
N_2 Refrigerator	28.34	27.76	27.84	28.61	28.07	28.31	28.68	29.67	27.53	28.39	29.17
Feed compressor	2.74	2.74	2.74	2.74	2.74	2.74	2.74	2.74	2.74	2.74	2.74
Turbine returns	-3.85	-3.82	-3.78	-3.73	-3.86	-3.63	-3.45	-3.36	-3.79	-3.29	-3.29
Net Power	100.86	99.69	99.53	100.21	104.82	105.14	106.96	107.38	109.90	112.41	119.93
Heat Exchanger Size											
UA-MJ/s, K	10.9	10.7	10.6	10.4	12.3	11.8	12.3	11.9	14.2	27.2	48.1

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TABLE 2
H2 LIQUEFIER - PROCESS CRC
NASA CONTRACT NAS1-14098

STREAM DATA FOR PURE NITROGEN STREAMS
SI UNITS

STREAM NUMBER	080	081	082	083	084	085
FLOW, CMH(STP)	92101.1	92101.1	92101.1	24680.7	24680.8	19426.6
FLOW, GM.MOLE/SEC	1141.44	1141.44	1141.44	305.876	305.876	240.760
PRESSURE, KILOPASCAL	620.528	620.528	579.160	136.86	151.685	151.685
TEMPERATURE, DEG K	97.0000	300.023	299.942	99.9000	81.0186	81.0186
ENTHALPY, JOULES/GM.MOLE	6693.07	12909.5	12909.5	2191.67	2191.67	1032.51
ENTROPY, JOULES/GM.MOLE-DEG C	73.4695	108.629	109.202	25.5956	28.6944	14.3890
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	0.78712	SAT. LIQ
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	087	089	090	091	092	093
FLOW, CMH(STP)	92101.1	19426.6	5254.15	24680.8	5833.16	5833.16
FLOW, GM.MOLE/SEC	1141.44	240.760	65.1164	305.876	72.2923	72.2923
PRESSURE, KILOPASCAL	620.528	151.685	151.685	151.685	151.685	151.685
TEMPERATURE, DEG K	133.948	81.0186	81.0186	81.0186	90.6000	90.6000
ENTHALPY, JOULES/GM.MOLE	7935.42	6477.53	6477.53	6477.53	6772.96	6772.96
ENTROPY, JOULES/GM.MOLE-DEG C	84.3674	81.5870	81.5870	81.5870	85.0341	85.0341
LIQUID FRACTION (MOLE L/F)	S.H. VAP	SAT. VAP	SAT. VAP	SAT. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	094	095	096	097
FLOW, CMH(STP)	30513.9	30513.9	30513.9	30513.9
FLOW, GM.MOLE/SEC	378.168	378.168	378.168	378.168
PRESSURE, KILOPASCAL	151.685	151.685	103.421	151.685
TEMPERATURE, DEG K	82.8386	300.023	299.927	133.948
ENTHALPY, JOULES/GM.MOLE	6514.00	12936.5	12936.5	8074.22
ENTROPY, JOULES/GM.MOLE-DEG C	82.2765	120.417	123.598	96.7834
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:				
NITROGEN	1.00000	1.00000	1.00000	1.00000

TABLE 2
H2 LIQUEFIER - PROCESS CRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR HYDROGEN-PROPANE MIXTURES
SI UNITS

STREAM NUMBER	100	101	102	103	104	105
FLOW, CMH(STP)	829236.	704199.	125037.	9923.35	9923.35	115114.
FLOW, GM.MOLE/SEC	10277.0	8727.36	1549.62	122.983	122.983	1426.64
PRESSURE, KILOPASCAL	2137.38	2137.38	2137.38	2137.38	108.248	2137.38
TEMPERATURE, DEG K	140.000	140.000	140.000	140.000	142.157	140.000
ENTHALPY, JOULES/GM.MOLE	1509.51	4125.14	-13221.5	-13221.5	-13222.1	-13221.5
ENTROPY, JOULES/GM.MOLE-DEG C	166.625	181.952	80.3088	80.3088	81.3759	80.3088
LIQUID FRACTION (MOLE L/F)	0.15079	SAT. VAP	SAT. LIQ	SAT. LIQ	0.99476	SAT. LIQ
COMPOSITION, MOLE FRACTION:						
HYDROGEN	0.85000	0.99994	0.00552	0.00552	0.00552	0.00552
PROPANE	0.15000	55. PPM	0.99448	0.99448	0.99448	0.99448

STREAM NUMBER	106	107	108	109	110	111
FLOW, CMH(STP)	115114.	704199.	704159.	38.9031	3.08748	3.08748
FLOW, GM.MOLE/SEC	1426.64	8727.35	8726.87	0.48214	0.03826	0.03826
PRESSURE, KILOPASCAL	324.053	2137.38	2137.38	2137.38	2137.38	108.248
TEMPERATURE, DEG K	143.260	95.0000	95.0000	95.0000	95.0000	97.7141
ENTHALPY, JOULES/GM.MOLE	-13150.9	3015.59	3016.61	-15458.2	-15458.2	-15459.5
ENTROPY, JOULES/GM.MOLE-DEG C	81.7267	172.402	172.408	61.2762	61.2762	62.6759
LIQUID FRACTION (MOLE L/F)	0.99536	0.00006	SAT. VAP	SAT. LIQ	SAT. LIQ	0.99801
COMPOSITION, MOLE FRACTION:						
HYDROGEN	0.00552	0.99994	1.00000	0.00211	0.00211	0.00211
PROPANE	0.99448	55. PPM	0.884917E-08	0.99789	0.99789	0.99789

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STREAM NUMBER	112	113	117	118	119	120
FLOW, CMH(STP)	35.8156	35.8156	648276.	648312.	648312.	763426.
FLOW, GM.MOLE/SEC	0.44387	0.44387	8034.29	8034.73	8034.73	9461.37
PRESSURE, KILOPASCAL	2137.38	324.053	324.053	324.053	324.053	324.053
TEMPERATURE, DEG K	95.0000	97.4451	88.9208	88.9208	133.948	136.491
ENTHALPY, JOULES/GM.MOLE	-15458.2	-15458.5	2921.77	2920.74	3971.16	1389.53
ENTROPY, JOULES/GM.MOLE-DEG C	61.2762	62.5230	187.423	187.416	196.957	179.597
LIQUID FRACTION (MOLE L/F)	SAT. LIQ	0.99824	S.H. VAP	0.00006	S.H. VAP	0.14998
COMPOSITION, MOLE FRACTION:						
HYDROGEN	0.00211	0.00211	1.00000	0.99994	0.99994	0.85000
PROPANE	0.99789	0.99789	0.0	55. PPM	55. PPM	0.15000

TABLE 2
H2 LIQUEFIER - PROCESS CHC
NASA CONTRACT NAS1-14698

STREAM DATA FOR HYDROGEN-PROPANE MIXTURES
SI UNITS

STREAM NUMBER	121	122	123	124	125	126
FLOW, CMH(STP)	55884.6	55887.6	55887.6	65810.9	65810.9	65810.9
FLOW, GM.MOLE/SEC	692.595	692.633	692.633	815.616	815.616	815.616
PRESSURE, KILOPASCAL	108.248	108.248	108.248	108.248	108.248	103.421
TEMPERATURE, DEG K	88.9208	88.9208	133.948	136.037	300.023	300.020
ENTHALPY, JOULES/GM.MOLE	2929.25	2928.21	3971.16	1378.77	9453.85	9453.84
ENTROPY, JOULES/GM.MOLE-DEG C	196.660	196.652	206.123	187.329	228.115	228.494
LIQUID FRACTION (MOLE L/F)	S.H. VAP	0.00006	S.H. VAP	0.14967	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
HYDROGEN	1.00000	0.99994	0.99994	0.85000	0.85000	0.85000
PROPANE	0.0	55. PPM	55. PPM	0.15000	0.15000	0.15000

STREAM NUMBER	127	128	129	130	131	132
FLOW, CMH(STP)	0.00623	65810.9	65810.9	763426.	763426.	829237.
FLOW, GM.MOLE/SEC	0.00008	815.616	815.616	9461.37	9461.37	10277.0
PRESSURE, KILOPASCAL	103.421	103.421	289.580	324.053	289.580	289.580
TEMPERATURE, DEG K	308.000	300.020	308.000	300.023	300.005	300.647
ENTHALPY, JOULES/GM.MOLE	0.0	9453.84	9741.83	9449.77	9449.77	9472.95
ENTROPY, JOULES/GM.MOLE-DEG C	0.0	228.494	220.862	218.962	219.901	219.978
LIQUID FRACTION (MOLE L/F)	SAT. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
HYDROGEN	0.0	0.85000	0.85000	0.85000	0.85000	0.85000
PROPANE	1.00000	0.15000	0.15000	0.15000	0.15000	0.15000

STREAM NUMBER	133
FLOW, CMH(STP)	829237.
FLOW, GM.MOLE/SEC	10277.0
PRESSURE, KILOPASCAL	2137.38
TEMPERATURE, DEG K	308.000
ENTHALPY, JOULES/GM.MOLE	9714.73
ENTROPY, JOULES/GM.MOLE-DEG C	203.919
LIQUID FRACTION (MOLE L/F)	S.H. VAP
COMPOSITION, MOLE FRACTION:	
HYDROGEN	0.85000
PROPANE	0.15000

TABLE 2
H2 LIQUEFIER - PROCESS CRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR PURE HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	001	002	003	004	005	006
FLOW, CMH(STP)	105140.	105140.	105140.	105140.	105140.	704159.
FLOW, GM.MOLE/SEC	1303.03	1303.03	1303.03	1303.03	1303.03	8726.86
PRESSURE, KILOPASCAL	2378.69	2378.69	2378.69	2378.69	2378.69	2137.38
TEMPERATURE, DEG K	308.000	140.000	95.0000	85.5000	308.000	95.0000
ENTHALPY, JOULES/GM.MOLE	8758.06	4122.85	2987.54	2759.73	0.0	2993.76
ENTROPY, JOULES/GM.MOLE-DEG C	115.841	94.4402	85.5942	83.0051	0.0	86.4368
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	SAT. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	007	008	009	010	011	012
FLOW, CMH(STP)	704159.	105140.	105140.	105140.	105140.	105140.
FLOW, GM.MOLE/SEC	8726.86	1303.03	1303.03	1303.03	1303.03	1303.03
PRESSURE, KILOPASCAL	2137.38	2137.38	2378.69	2368.35	2368.35	2282.17
TEMPERATURE, DEG K	85.5000	85.5000	308.000	85.4846	81.2186	81.0812
ENTHALPY, JOULES/GM.MOLE	2768.39	2759.73	0.0	2759.74	2374.88	2374.88
ENTROPY, JOULES/GM.MOLE-DEG C	83.8986	83.0051	0.0	83.0391	79.2558	79.5386
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	SAT. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.54509	0.54509
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.45491	0.45491

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STREAM NUMBER	013	014	015	016	017	018
FLOW, CMH(STP)	105140.	105140.	105140.	105140.	105140.	105140.
FLOW, GM.MOLE/SEC	1303.03	1303.03	1303.03	1303.03	1303.03	1303.03
PRESSURE, KILOPASCAL	2282.17	2230.45	2230.45	2178.74	930.792	930.792
TEMPERATURE, DEG K	50.0511	49.8683	30.0580	30.0635	29.7109	30.2877
ENTHALPY, JOULES/GM.MOLE	1227.90	1227.90	-160.586	-160.583	-160.584	-160.586
ENTROPY, JOULES/GM.MOLE-DEG C	61.8932	62.0314	27.2779	27.3343	28.7857	28.8357
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.C. LIQ	S.C. LIQ	S.C. LIQ	S.C. LIQ
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.30835	0.30835	0.05605	0.05605	0.05605	0.03739
PARA -HYDROGEN	0.69165	0.69165	0.94395	0.94395	0.94395	0.96261

TABLE 2
H2 LIQUEFIER - PROCESS CRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR PURE HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	019	020	021	022	023	024
FLOW, CMH(STP)	105140.	670779.	670779.	670779.	292739.	292739.
FLOW, GM.MOLE/SEC	1303.03	8313.17	8313.17	8313.17	3628.00	3628.00
PRESSURE, KILOPASCAL	930.792	2127.03	2127.03	2109.80	2109.80	344.738
TEMPERATURE, DEG K	20.5700	85.1183	81.2186	81.1903	81.1903	48.0511
ENTHALPY, JOULES/GM.MOLE	-441.045	2759.73	2666.59	2666.59	2666.59	2023.01
ENTROPY, JOULES/GM.MOLE-DEG C	18.0093	83.8305	82.7345	82.7911	82.7911	86.6340
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.03739	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.96261	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	025	026	027	028	029	030
FLOW, CMH(STP)	378041.	378041.	355538.	355538.	355538.	22502.8
FLOW, GM.MOLE/SEC	4685.18	4685.18	4406.29	4406.29	4406.29	278.884
PRESSURE, KILOPASCAL	2109.80	2109.80	2109.80	2109.80	358.527	2109.80
TEMPERATURE, DEG K	81.1903	50.4637	50.4637	50.4637	28.0580	50.4637
ENTHALPY, JOULES/GM.MOLE	2666.59	1883.61	1883.61	1883.61	1546.53	1883.61
ENTROPY, JOULES/GM.MOLE-DEG C	82.7911	70.5660	70.5660	70.5659	73.7028	70.5660
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	031	032	033	034	035	036
FLOW, CMH(STP)	22502.8	22502.8	47717.6	47717.6	459683.	459683.
FLOW, GM.MOLE/SEC	278.884	278.884	591.380	591.380	5697.00	5697.00
PRESSURE, KILOPASCAL	2109.80	111.695	358.527	358.527	344.738	344.738
TEMPERATURE, DEG K	31.5560	20.5200	28.0580	48.1209	48.0511	77.6903
ENTHALPY, JOULES/GM.MOLE	880.042	880.042	1546.53	2023.01	2023.01	2668.43
ENTROPY, JOULES/GM.MOLE-DEG C	44.7939	50.7501	73.7028	86.4473	86.6340	97.6319
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	0.63198	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

TABLE 2
H2 LIQUEFIER - PROCESS CRC
NASA CONTRACT NAS1-14098

STREAM DATA FOR PURE HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	037	038	039	040	041	042
FLOW, CMH(STP)	459683.	307820.	307820.	307820.	119227.	188593.
FLOW, GM.MOLE/SEC	5697.00	3814.92	3814.92	3814.92	1477.62	2337.29
PRESSURE, KILOPASCAL	324.053	358.527	358.527	344.738	344.738	344.738
TEMPERATURE, DEG K	77.6438	28.0580	48.1209	48.0511	48.0511	48.0511
ENTHALPY, JOULES/GM.MOLE	2668.43	1546.53	2023.01	2023.01	2023.01	2023.01
ENTROPY, JOULES/GM.MOLE-DEG C	98.1163	73.7028	86.4473	86.6340	86.6340	86.6340
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	043	044	045	046	047	048
FLOW, CMH(STP)	188593.	188593.	648276.	648276.	105140.	105140.
FLOW, GM.MOLE/SEC	2337.29	2337.29	8034.29	8034.29	1303.03	1303.03
PRESSURE, KILOPASCAL	344.738	324.053	324.053	324.053	2378.69	2378.69
TEMPERATURE, DEG K	77.5812	77.5345	77.6120	88.9208	308.000	308.000
ENTHALPY, JOULES/GM.MOLE	2666.08	2666.08	2667.75	2916.05	0.0	0.0
ENTROPY, JOULES/GM.MOLE-DEG C	97.5984	98.0832	98.1066	101.268	0.0	0.0
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	SAT. VAP	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	049	050	051	052	053	054
FLOW, CMH(STP)	105140.	33381.8	33381.8	33381.8	55884.6	55884.6
FLOW, GM.MOLE/SEC	1303.03	413.711	413.711	413.711	692.595	692.595
PRESSURE, KILOPASCAL	2137.38	2127.03	2127.03	111.695	111.695	111.695
TEMPERATURE, DEG K	85.5000	85.1183	32.8924	20.5200	20.5200	20.5200
ENTHALPY, JOULES/GM.MOLE	2759.73	2759.73	931.807	931.806	910.962	1444.11
ENTROPY, JOULES/GM.MOLE-DEG C	83.0051	83.8305	46.4129	53.2786	52.2604	78.3027
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.C. LIQ	0.57399	0.59734	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

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TABLE 2
H2 LIQUEFIER - PROCESS CRC
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STREAM DATA FOR PURE HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	055	056	057	058	059	060
FLOW, CMH(STP)	55884.6	55884.6	55884.6	105140.	105140.	105140.
FLOW, GM.MOLE/SEC	692.595	692.595	692.595	1303.03	1303.03	1303.03
PRESSURE, KILOPASCAL	111.695	108.248	108.[illegible]48	2378.69	2378.69	2378.69
TEMPERATURE, DEG K	71.0183	71.0[illegible]	38.9208	308.000	308.000	308.000
ENTHALPY, JOULES/GM.MOLE	2535.99	2536.[illegible]	2924.15	0.0	0.0	0.0
ENTROPY, JOULES/GM.MOLE-DEG C	105.232	105.431	110.590	0.0	0.0	0.0
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	SAT. VAP	SAT. VAP	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	061	062	063	064	065	066
FLOW, CMH(STP)	105140.	[illegible]382.1	33382.1	22502.5	105140.	2571.59
FLOW, GM.MOLE/SEC	1303.03	4[illegible].715	413.715	278.880	1303.03	31.8705
PRESSURE, KILOPASCAL	2378.69	111.695	111.695	111.695	101.353	101.353
TEMPERATURE, DEG K	308.000	20.5200	20.5200	20.5200	20.2337	20.2337
ENTHALPY, JOULES/GM.MOLE	0.0	551.573	1444.11	1444.11	-441.045	432.929
ENTROPY, JOULES/GM.MOLE-DEG C	0.0	34.7056	78.3027	78.3028	19.1478	62.3552
LIQUID FRACTION (MOLE L/F)	SAT. VAP	SAT. LIQ	SAT. VAP	SAT. VAP	0.97554	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.03739	0.03739
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.96261	0.96261

STREAM NUMBER	067
FLOW, CMH(STP)	92568.
FLOW, GM.MOLE/SEC	1271.16
PRESSURE, KILOPASCAL	101.353
TEMPERATURE, DEG K	20.2337
ENTHALPY, JOULES/GM.MOLE	-462.958
ENTROPY, JOULES/GM.MOLE-DEG C	18.0645
LIQUID FRACTION (MOLE L/F)	SAT. LIQ
COMPOSITION, MOLE FRACTION:	
ORTHO-HYDROGEN	0.03739
PARA -HYDROGEN	0.96261

TABLE
H2 LIQUEFIER - PROCESS CRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR PURE NITROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	080	081	082	083	084	085
FLOW, CFH(NTP)	3503942.	3503943.	3503943.	938967.	938967.	739075.
FLOW, LB.MOLE/HR	9059.03	9059.04	9059.04	2427.59	2427.59	1910.79
PRESSURE, PSIA	90.0000	90.0000	84.0000	600.000	22.0000	22.0000
TEMPERATURE, DEG K	97.0300	300.023	299.942	99.9000	81.0186	81.0186
ENTHALPY, BTU/LB.MOLE	2879.48	5553.91	5553.91	942.897	942.896	444.204
ENTROPY, BTU/LB.MOLE-DEG K	31.6080	46.7341	46.9805	11.0117	12.3449	6.19041
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	0.78712	SAT. LIQ
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	087	089	090	091	092	093
FLOW, CFH(NTP)	3503943.	739076.	199892.	938967.	221920.	221920.
FLOW, LB.MOLE/HR	9059.04	1910.79	516.797	2427.59	573.748	573.748
PRESSURE, PSIA	90.0000	22.0000	22.0000	22.0000	22.0000	22.0000
TEMPERATURE, DEG K	133.948	81.0186	81.0186	81.0186	90.6000	90.6000
ENTHALPY, BTU/LB.MOLE	3413.97	2786.75	2786.75	2786.75	2913.85	2913.85
ENTROPY, BTU/LB.MOLE-DEG K	36.2964	35.1003	35.1003	35.1003	36.5832	36.5832
LIQUID FRACTION (MOLE L/F)	S.H. VAP	SAT. VAP	SAT. VAP	SAT. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	094	095	096	097
FLOW, CFH(NTP)	1160887.	1160887.	1160887.	1160887.
FLOW, LB.MOLE/HR	3001.34	3001.34	3001.34	3001.34
PRESSURE, PSIA	22.0000	22.0000	15.0000	22.0000
TEMPERATURE, DEG K	82.8386	300.023	299.927	133.948
ENTHALPY, BTU/LB.MOLE	2411.05	5565.52	5565.52	3473.68
ENTROPY, BTU/LB.MOLE-DEG K	35.3969	51.8055	53.1742	41.6380
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:				
NITROGEN	1.00000	1.00000	1.00000	1.00000

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TABLE 1
H2 LIQUEFIER - PROCESS CRC
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STREAM DATA FOR HYDROGEN-PROPANE MIXTURES
US CUSTOMARY UNITS

STREAM NUMBER	100	101	102	103	104	105
FLOW, CFH(NTP)	31547884.	26790912.	4756973.	377529.	377529.	4379441.
FLOW, LB.MOLE/HR	81563.4	69264.8	12298.6	976.057	976.057	11322.5
PRESSURE, PSIA	310.000	310.000	310.000	310.000	15.7000	310.000
TEMPERATURE, DEG K	140.000	140.000	140.000	140.000	142.157	140.000
ENTHALPY, BTU/LB.MOLE	649.421	1774.71	-5688.13	-5688.13	-5688.41	-5688.13
ENTROPY, BTU/LB.MOLE-DEG K	71.6853	78.2790	34.5503	34.5503	35.0094	34.5503
LIQUID FRACTION (MOLE L/F)	0.15079	SAT. VAP	SAT. LIQ	SAT. LIQ	0.99476	SAT. LIQ
COMPOSITION, MOLE FRACTION:						
HYDROGEN	0.85000	0.99994	0.00552	0.00552	.00552	0.00552
PROPANE	0.15000	55. PPM	0.99448	0.99448	0.99448	0.99448

STREAM NUMBER	106	107	108	109	110	111
FLOW, CFH(NTP)	4379441.	26790896.	26789408.	1480.05	117.462	117.462
FLOW, LB.MOLE/HR	11322.5	69264.7	69260.9	3.82650	0.30368	0.30368
PRESSURE, PSIA	47.0000	310.000	310.000	310.000	310.000	15.7000
TEMPERATURE, DEG K	143.260	95.0000	95.0000	95.0000	95.0000	97.7141
ENTHALPY, BTU/LB.MOLE	-5657.77	1297.36	1297.80	-6650.41	-6650.41	-6650.97
ENTROPY, BTU/LB.MOLE-DEG K	35.1603	74.1704	74.1731	26.3622	26.3622	26.9644
LIQUID FRACTION (MOLE L/F)	0.99536	0.00006	SAT. VAP	SAT. LIQ	SAT. LIQ	0.99801
COMPOSITION, MOLE FRACTION:						
HYDROGEN	0.00552	0.99994	1.00000	0.00211	0.00211	0.00211
PROPANE	0.99448	55. PPM	0.884917E-08	0.99789	0.99789	0.99789

STREAM NUMBER	112	113	117	118	119	120
FLOW, CFH(NTP)	1362.59	1362.59	44663360.	24664720.	24664720.	29044160.
FLOW, LB.MOLE/HR	3.52281	3.52281	63764.2	63767.8	63767.8	75090.3
PRESSURE, PSIA	310.000	47.0000	47.0000	47.0000	47.0000	47.0000
TEMPERATURE, DEG K	95.0000	97.4451	88.9208	88.9208	133.948	136.491
ENTHALPY, BTU/LB.MOLE	-6650.41	-6650.54	1257.00	1256.56	1708.47	597.801
ENTROPY, BTU/LB.MOLE-DEG K	26.3622	26.8986	80.6330	80.6300	84.7346	77.2660
LIQUID FRACTION (MOLE L/F)	SAT. LIQ	0.99824	S.H. VAP	0.00006	S.H. VAP	0.14998
COMPOSITION, MOLE FRACTION:						
HYDROGEN	0.00211	0.00211	1.00000	0.99994	0.99994	0.85000
PROPANE	0.99789	0.99789	0.0	55. PPM	55. PPM	0.15000

TABLE 3
H2 LIQUEFIER - PROCESS CRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR HYDROGEN-PROPANE MIXTURES
US CUSTOMARY UNITS

STREAM NUMBER	121	122	123	124	125	126
FLOW, CFH(NTP)	2126101.	2126218.	2126218.	2503747.	2503747.	2503747.
FLOW, LB.MOLE/HR	5496.79	5497.09	5497.09	6473.14	6473.14	6473.14
PRESSURE, PSIA	15.7000	15.7000	15.7000	15.7000	15.7000	15.0000
TEMPERATURE, DEG K	88.9208	88.9208	133.948	136.037	300.023	300.020
ENTHALPY, BTU/LB.MOLE	1260.22	1259.77	1708.47	593.175	4067.22	4067.22
ENTROPY, BTU/LB.MOLE-DEG K	84.6068	84.6035	88.6781	80.5925	98.1391	98.3024
LIQUID FRACTION (MOLE L/F)	S.H. VAP	0.00006	S.H. VAP	0.14967	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
HYDROGEN	1.00000	0.99994	0.99994	0.85000	0.85000	0.85000
PROPANE	0.0	55. PPM	55. PPM	0.15000	0.15000	0.15000

STREAM NUMBER	127	128	129	130	131	132
FLOW, CFH(NTP)	0.23706	2503747.	2503747.	29044160.	29044160.	31547904.
FLOW, LB.MOLE/HR	0.00061	6473.14	6473.14	75090.3	75090.3	81563.4
PRESSURE, PSIA	15.0000	15.0000	42.0000	47.0000	42.0000	42.0000
TEMPERATURE, DEG K	308.000	300.020	308.000	300.023	300.005	300.647
ENTHALPY, BTU/LB.MOLE	0.0	4067.22	4191.12	4065.47	4065.47	4075.44
ENTROPY, BTU/LB.MOLE-DEG K	0.0	98.3024	95.0188	94.2013	94.6054	94.6386
LIQUID FRACTION (MOLE L/F)	SAT. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
HYDROGEN	0.0	0.85000	0.85000	0.85000	0.85000	0.85000
PROPANE	1.00000	0.15000	0.15000	0.15000	0.15000	0.15000

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STREAM NUMBER	133
FLOW, CFH(NTP)	31547904.
FLOW, LB.MOLE/HR	81563.4
PRESSURE, PSIA	310.000
TEMPERATURE, DEG K	308.000
ENTHALPY, BTU/LB.MOLE	4179.46
ENTROPY, BTU/LB.MOLE-DEG K	87.7296
LIQUID FRACTION (MOLE L/F)	S.H. VAP
COMPOSITION, MOLE FRACTION:	
HYDROGEN	0.85000
PROPANE	0.15000

TABLE 3
H2 LIQUEFIER - PROCESS CRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR PURE HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	001	002	003	004	005	006
FLOW, CFM(NTP)	4000000.	4000000.	4000000.	4000000.	4000000.	26789392.
FLOW, LB.MOLE/HR	10341.5	10341.5	10341.5	10341.5	10341.5	69260.8
PRESSURE, PSIA	345.000	345.000	345.000	345.000	345.000	310.000
TEMPERATURE, DEG K	308.000	140.000	95.0000	85.5000	308.000	95.0000
ENTHALPY, BTU/LB.MOLE	3767.88	1773.73	1285.29	1187.29	0.0	1287.97
ENTROPY, BTU/LB.MOLE-DEG K	49.8371	40.6299	36.8242	35.7103	0.0	37.1867
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	SAT. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	007	008	009	010	011	012
FLOW, CFM(NTP)	26789392.	4000000.	4000000.	4000000.	4000000.	4000000.
FLOW, LB.MOLE/HR	69260.8	10341.5	10341.5	10341.5	10341.5	10341.5
PRESSURE, PSIA	310.000	310.000	345.000	343.500	343.500	331.000
TEMPERATURE, DEG K	85.5000	85.5000	308.000	85.4846	81.2186	81.0812
ENTHALPY, BTU/LB.MOLE	1191.91	1187.29	0.0	1187.29	1021.72	1021.72
ENTROPY, BTU/LB.MOLE-DEG K	36.0947	35.7103	0.0	35.7250	34.0973	34.2190
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	SAT. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.54509	0.54509
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.45491	0.45491

STREAM NUMBER	013	014	015	016	017	018
FLOW, CFM(NTP)	4000000.	4000000.	4000000.	4000000.	4000000.	4000000.
FLOW, LB.MOLE/HR	10341.5	10341.5	10341.5	10341.5	10341.5	10341.5
PRESSURE, PSIA	331.000	323.500	323.500	316.000	135.000	135.000
TEMPERATURE, DEG K	50.0511	49.8683	30.0580	30.0635	29.7109	30.2877
ENTHALPY, BTU/LB.MOLE	528.266	528.266	-69.0872	-69.0860	-69.0864	-69.0869
ENTROPY, BTU/LB.MOLE-DEG K	26.6276	26.6871	11.7355	11.7597	12.3841	12.4057
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.C. LIQ	S.C. LIQ	S.C. LIQ	S.C. LIQ
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.30835	0.30835	0.05605	0.05605	0.05605	0.03739
PARA -HYDROGEN	0.69165	0.69165	0.94395	0.94395	0.94395	0.96261

TABLE 3
H_2 LIQUEFIER - PROCESS CRC
NASA CONTRACT NAS1-14098

STREAM DATA FOR PURE HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	019	020	021	022	023	024
FLOW, CFM(NTP)	4000000.	25519456.	25519456.	25519456.	11137101.	11137101.
FLOW, LB.MOLE/HR	10341.5	65977.6	65977.6	65977.6	28793.7	28793.7
PRESSURE, PSIA	135.000	308.500	308.500	306.000	306.000	50.0000
TEMPERATURE, DEG K	20.5700	85.1183	81.2186	81.1903	81.1903	48.0511
ENTHALPY, BTU/LB.MOLE	-189.746	1187.29	1147.21	1147.21	1147.21	870.336
ENTROPY, BTU/LB.MOLE-DEG K	7.74794	36.0654	35.5939	35.6183	35.6183	37.2716
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.03739	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.96261	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	025	026	027	028	029	030
FLOW, CFM(NTP)	14382380.	14382380.	13526272.	13526272.	13526272.	856108.
FLOW, LB.MOLE/HR	37184.0	37184.0	34970.6	34970.6	34970.6	2213.37
PRESSURE, PSIA	306.000	306.000	306.000	306.000	52.0000	306.000
TEMPERATURE, DEG K	81.1903	50.4637	50.4637	50.4637	28.0580	50.4637
ENTHALPY, BTU/LB.MOLE	1147.21	810.363	810.363	810.363	665.344	810.363
ENTROPY, BTU/LB.MOLE-DEG K	35.6183	30.3588	30.3588	30.3588	31.7083	30.3588
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

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STREAM NUMBER	031	032	033	034	035	036
FLOW, CFM(NTP)	856108.	856108.	1815395.	1815395.	17488432.	17488432.
FLOW, LB.MOLE/HR	2213.37	2213.37	4693.49	4693.49	45214.3	45214.3
PRESSURE, PSIA	306.000	16.2000	52.0000	52.0000	50.0000	50.0000
TEMPERATURE, DEG K	31.5580	20.5200	28.0580	48.1209	48.0511	77.6903
ENTHALPY, BTU/LB.MOLE	378.610	378.610	665.344	870.336	870.336	1148.01
ENTROPY, BTU/LB.MOLE-DEG K	19.2711	21.8336	31.7083	37.1913	37.2716	42.0030
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	0.63198	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

TABLE 7
H2 LIQUEFIER - PROCESS CRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR PURE HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	037	038	039	040	041	042
FLOW, CFH(NTP)	17488432.	11710877.	11710877.	11710877.	4535949.	7174928.
FLOW, LB.MOLE/HR	45214.3	30277.1	30277.1	30277.1	11727.2	18549.9
PRESSURE, PSIA	47.0000	52.0000	52.0000	50.0000	50.0000	50.0000
TEMPERATURE, DEG K	77.6438	28.0580	48.1209	48.0511	48.0511	48.0511
ENTHALPY, BTU/LB.MOLE	1148.01	665.344	870.336	870.336	870.336	870.336
ENTROPY, BTU/LB.MOLE-DEG K	42.2115	31.7083	37.1913	37.2716	37.2716	37.2716
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	043	044	045	046	047	048
FLOW, CFH(NTP)	7174928.	7174928.	24663360.	24663360.	4000000.	4000000.
FLOW, LB.MOLE/HR	18549.9	18549.9	63764.2	63764.2	10341.5	10341.5
PRESSURE, PSIA	50.0000	47.0000	47.0000	47.0000	345.000	345.000
TEMPERATURE, DEG K	77.5812	77.5345	77.6120	88.9208	308.000	308.000
ENTHALPY, BTU/LB.MOLE	1147.00	1147.00	1147.71	1254.54	0.0	0.0
ENTROPY, BTU/LB.MOLE-DEG K	41.9887	42.1972	42.2073	43.5672	0.0	0.0
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	SAT. VAP	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	049	050	051	052	053	054
FLOW, CFH(NTP)	4000000.	1269993.	1269993.	1269993.	2126101.	2126101.
FLOW, LB.MOLE/HR	10341.5	3283.42	3283.42	3283.42	5496.79	5496.79
PRESSURE, PSIA	310.000	308.500	308.500	16.2000	16.2000	16.2000
TEMPERATURE, DEG K	85.5000	85.1183	32.8924	20.5200	20.5200	20.5200
ENTHALPY, BTU/LB.MOLE	1187.29	1187.29	400.881	400.880	391.913	621.284
ENTROPY, BTU/LB.MOLE-DEG K	35.7103	36.0654	19.9677	22.9215	22.4834	33.6873
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.C. LIQ	0.57399	0.59734	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

TABLE 3
H2 LIQUEFIER - PROCESS CRC
NASA CONTRACT NAS1-14098

STREAM DATA FOR PURE HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	055	056	057	058	059	060
FLOW, CFM(NTP)	2126101.	2126101.	2126101.	4000000.	4000000.	4000000.
FLOW, LB.MOLE/HR	5496.79	5496.79	5496.79	10341.5	10341.5	10341.5
PRESSURE, PSIA	16.2000	15.7000	15.7000	345.000	345.000	345.000
TEMPERATURE, DEG K	71.0183	71.0183	88.9208	308.000	308.000	308.000
ENTHALPY, BTU/LB.MOLE	1091.03	1091.11	1258.02	0.0	0.0	0.0
ENTROPY, BTU/LB.MOLE-DEG K	45.2729	45.3582	47.5777	0.0	0.0	0.0
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	SAT. VAP	SAT. VAP	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	061	062	063	064	065	066
FLOW, CFM(NTP)	4000000.	1270005.	1270005.	856095.	4000000.	97834.8
FLOW, LB.MOLE/HR	10341.5	3283.45	3283.45	2213.33	10341.5	252.941
PRESSURE, PSIA	345.000	16.2000	16.2000	16.2000	14.7000	14.7000
TEMPERATURE, DEG K	308.000	20.5200	20.5200	20.5200	20.2337	20.2337
ENTHALPY, BTU/LB.MOLE	0.0	237.297	621.284	621.283	-189.746	186.254
ENTROPY, BTU/LB.MOLE-DEG K	0.0	14.9310	33.6873	33.6873	8.23776	26.8264
LIQUID FRACTION (MOLE L/F)	SAT. VAP	SAT. LIQ	SAT. VAP	SAT. VAP	0.97554	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.03739	0.03739
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.96261	0.96261

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STREAM NUMBER	067
FLOW, CFM(NTP)	3902165.
FLOW, LB.MOLE/HR	10088.6
PRESSURE, PSIA	14.7000
TEMPERATURE, DEG K	20.2337
ENTHALPY, BTU/LB.MOLE	-199.173
ENTROPY, BTU/LB.MOLE-DEG K	7.77171
LIQUID FRACTION (MOLE L/F)	SAT. LIQ
COMPOSITION, MOLE FRACTION:	
ORTHO-HYDROGEN	0.03739
PARA -HYDROGEN	0.96261

TABLE 4
N2 REFRIGERATOR - PROCESS CRC
NASA CONTRACT NAS1-14098

STREAM DATA FOR N2 REFRIGERATOR STREAMS
SI UNITS

STREAM NUMBER	001	002	003	004	005	006
FLOW, CMH(STP)	369232.	369232.	337169.	136814.	136814.	200354.
FLOW, GM.MOLE/SEC	4576.01	4576.01	4178.64	1695.58	1695.58	2483.06
PRESSURE, KILOPASCAL	2817.51	3358.02	4136.86	4136.86	606.739	4136.86
TEMPERATURE, DEG K	308.000	308.000	308.000	235.000	147.536	235.000
ENTHALPY, JOULES/GM.MOLE	13027.9	13000.3	12961.1	10650.9	8364.43	10650.9
ENTROPY, JOULES/GM.MOLE-DEG C	96.4698	94.9251	93.0685	84.4956	87.5968	84.4956
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	007	008	009	010	011	012
FLOW, CMH(STP)	200354.	245068.	245068.	32063.8	32063.8	32063.8
FLOW, GM.MOLE/SEC	2483.06	3037.20	3037.20	397.376	397.376	397.376
PRESSURE, KILOPASCAL	4136.86	606.739	606.739	4136.86	4136.86	4136.86
TEMPERATURE, DEG K	162.416	148.446	232.000	235.000	162.416	140.000
ENTHALPY, JOULES/GM.MOLE	8000.39	8392.60	10906.3	10650.9	8000.39	6763.73
ENTROPY, JOULES/GM.MOLE-DEG C	70.8873	87.7872	101.241	84.4956	70.8873	62.6387
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	013	014	015	016	017	018
FLOW, CMH(STP)	32063.8	108253.	108253.	108253.	32063.8	32063.8
FLOW, GM.MOLE/SEC	397.376	1341.62	1341.62	1341.62	397.376	397.376
PRESSURE, KILOPASCAL	4136.86	620.528	620.528	606.739	2757.90	2757.90
TEMPERATURE, DEG K	99.0000	97.0000	138.000	149.597	100.163	100.163
ENTHALPY, JOULES/GM.MOLE	2137.69	6693.07	8063.26	8428.20	2196.66	2196.66
ENTROPY, JOULES/GM.MOLE-DEG C	25.0528	73.4695	85.3076	88.0261	26.1933	26.1933
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	SAT. LIQ
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

TABLE 4
N2 REFRIGERATOR - PROCESS CRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR N2 REFRIGERATOR STREAMS
SI UNITS

STREAM NUMBER	019	020	021	022	023	025
FLOW, CMH(STP)	0.0	92101.1	92101.1	200354.	0.0	337169.
FLOW, GM.MOLE/SEC	0.0	1141.44	1141.44	2483.06	0.0	4178.64
PRESSURE, KILOPASCAL	2757.90	620.528	620.528	620.528	620.528	4136.86
TEMPERATURE, DEG K	100.163	97.0000	97.0000	97.0000	63.1365	235.000
ENTHALPY, JOULES/GM.MOLE	0.0	6693.07	6693.07	6693.07	-0.00908	10650.9
ENTROPY, JOULES/GM.MOLE-DEG C	0.0	73.4695	73.4695	73.4695	-0.24987	84.4957
LIQUID FRACTION (MOLE L/F)	SAT. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	026	027	028	029	030	031
FLOW, CMH(STP)	337169.	32063.8	32063.8	245068.	369232.	369232.
FLOW, GM.MOLE/SEC	4178.64	397.376	397.376	3037.20	4576.01	4576.01
PRESSURE, KILOPASCAL	4136.86	4136.86	4136.86	592.949	4136.86	579.160
TEMPERATURE, DEG K	263.586	308.000	263.586	303.000	308.000	302.665
ENTHALPY, JOULES/GM.MOLE	11573.4	12961.1	11573.4	12998.5	12961.1	12989.4
ENTROPY, JOULES/GM.MOLE-DEG C	88.2011	93.0685	88.2011	109.301	93.0685	109.467
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

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STREAM NUMBER	032	033	034	035
FLOW, CMH(STP)	92101.1	277131.	32063.8	32063.8
FLOW, GM.MOLE/SEC	1141.44	3434.58	397.376	397.376
PRESSURE, KILOPASCAL	579.160	592.949	592.949	101.353
TEMPERATURE, DEG K	300.000	303.578	308.000	300.000
ENTHALPY, JOULES/GM.MOLE	12911.2	13015.4	13145.2	12938.7
ENTROPY, JOULES/GM.MOLE-DEG C	109.207	109.357	109.781	123.773
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:				
NITROGEN	1.00000	1.00000	1.00000	1.00000

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                                              TABLE 5
                                  N2 REFRIGERATOR - PROCESS CRC
                                   NASA CONTRACT NAS1-14698

                                       STREAM DATA FOR N2 REFRIGERATOR STREAMS
                                               US CUSTOMARY UNITS

STREAM NO., N-(STYP/KTYP/HTYP)       001 (01/03/01)  002 (01/03/01)  003 (01/03/01)  004 (01/03/01)  005 (01/03/01)  006 (01/03/01)

FLOW, CFH(NTP)                           14047270.       14047270.       12827420.        5205033.        5205033.        7622387.
FLOW, LB.MOLE/HR                           36317.6         36317.6         33163.8         13457.0         13457.0         19706.8
PRESSURE, PSIA                             408.646         487.039         600.000         600.000         88.0000         600.000
TEMPERATURE, DEG K                         308.000         308.000         308.000         235.000         147.536         235.000
ENTHALPY, BTU/LB.MOLE                      5604.84         5592.95         5576.10         4582.21         3598.54         4582.21
ENTROPY, BTU/LB.MOLE-DEG K                 41.5031         40.8385         40.0398         36.3516         37.6858         36.3516
LIQUID FRACTION (MOLE  L/F)               S.H. VAP        S.H. VAP        S.H. VAP        S.H. VAP        S.H. VAP        S.H. VAP
COMPOSITION, MOLE FRACTION:
  NITROGEN                                 1.00000         1.00000         1.00000         1.00000         1.00000         1.00000


STREAM NO., N-(STYP/KTYP/HTYP)       007 (01/03/01)  008 (01/03/01)  009 (01/03/01)  010 (01/03/01)  011 (01/03/01)  012 (01/03/01)

FLOW, CFH(NTP)                            7622387.        9323478.        9323478.        1219850.        1219850.        1219850.
FLOW, LB.MOLE/HR                           19706.8         24104.8         24104.8         3153.78         3153.78         3153.78
PRESSURE, PSIA                             600.000         88.0000         88.0000         600.000         600.000         600.000
TEMPERATURE, DEG K                         162.416         148.446         232.000         235.000         162.416         140.000
ENTHALPY, BTU/LB.MOLE                      3441.92         3610.65         4692.08         4582.21         3441.92         2909.88
ENTROPY, BTU/LB.MOLE-DEG K                 30.4970         37.7677         43.5560         36.3516         30.4970         26.9483
LIQUID FRACTION (MOLE  L/F)               S.H. VAP        S.H. VAP        S.H. VAP        S.H. VAP        S.H. VAP        S.H. VAP
COMPOSITION, MOLE FRACTION:
  NITROGEN                                 1.00000         1.00000         1.00000         1.00000         1.00000         1.00000


STREAM NO., N-(STYP/KTYP/HTYP)       013 (01/03/01)  014 (01/03/01)  015 (01/03/01)  016 (01/03/01)  017 (01/03/01)  018 (01/03/01)

FLOW, CFH(NTP)                            1219850.        4118445.        4118445.        4118445.        1219850.        1219850.
FLOW, LB.MOLE/HR                           3153.78         10647.8         10647.8         10647.8         3153.78         3153.78
PRESSURE, PSIA                             600.000         90.0000         90.0000         88.0000         400.000         400.000
TEMPERATURE, DEG K                         99.0000         97.0000         138.000         149.597         100.163         100.163
ENTHALPY, BTU/LB.MOLE                      919.675         2879.48         3468.97         3625.97         945.044         945.044
ENTROPY, BTU/LB.MOLE-DEG K                 10.7782         31.6080         36.7009         37.8705         11.2689         11.2689
LIQUID FRACTION (MOLE  L/F)               S.C. LIQ        S.H. VAP        S.H. VAP        S.H. VAP        S.C. LIQ        SAT. LIQ
COMPOSITION, MOLE FRACTION:
  NITROGEN                                 1.00000         1.00000         1.00000         1.00000         1.00000         1.00000
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TABLE 5
N2 REFRIGERATOR - PROCESS CRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR N2 REFRIGERATOR STREAMS
US CUSTOMARY UNITS

STREAM NO., N-(STYP/KTYP/HTYP)	019 (01/03/01)	020 (01/03/01)	021 (01/03/01)	022 (01/03/01)	023 (01/03/01)	025 (01/03/01)
FLOW, CFH(NTP)	0.0	3503942.	3503942.	7622387.	0.0	12827420.
FLOW, LB.MOLE/HR	0.0	9059.03	9059.03	19706.8	0.0	33163.8
PRESSURE, PSIA	400.000	90.0000	90.0000	90.0000	90.0000	600.000
TEMPERATURE, DEG K	100.163	97.0000	97.0000	97.0000	63.1365	235.000
ENTHALPY, BTU/LB.MOLE	0.0	2879.48	2879.48	2879.48	-0.0039	4582.20
ENTROPY, BTU/LB.MOLE-DEG K	0.0	31.6060	31.6080	31.6080	-0.10750	36.3516
LIQUID FRACTION (MOLE L/F)	SAT. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NO., N-(STYP/KTYP/HTYP)	026 (01/03/01)	027 (01/03/01)	028 (01/03/01)	029 (01/03/01)	030 (01/03/01)	031 (01/03/01)
FLOW, CFH(NTP)	12827420.	1219850.	1219850.	9323478.	14047270.	14047270.
FLOW, LB.MOLE/HR	33163.8	3153.78	3153.78	24104.8	36317.6	36317.6
PRESSURE, PSIA	600.000	600.000	600.000	86.0000	600.000	84.0000
TEMPERATURE, DEG K	263.586	308.000	263.586	303.000	308.000	302.665
ENTHALPY, BTU/LB.MOLE	4979.08	5576.10	4979.08	5592.18	5576.10	5588.29
ENTROPY, BTU/LB.MOLE-DEG K	37.9458	40.0398	37.9458	47.0233	40.0398	47.0945
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

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STREAM NO., N-(STYP/KTYP/HTYP)	[illegible]2 (01/03/01)	033 (01/03/01)	034 (01/03/01)	035 (01/03/01)
FLOW, CFH(NTP)	3503942.	10543328.	1219850.	1219850.
FLOW, LB.MOLE/HR	9059.03	27258.5	3153.78	3153.78
PRESSURE, PSIA	84.0000	86.0000	86.0000	14.7000
TEMPERATURE, DEG K	300.000	303.578	308.000	300.000
ENTHALPY, BTU/LB.MOLE	5554.64	5599.48	5655.30	5566.48
ENTROPY, BTU/LB.MOLE-DEG K	46.9828	47.0475	47.2300	53.2496
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:				
NITROGEN	1.00000	1.00000	1.00000	1.00000

TABLE 6
H2 LIQUEFIER - PROCESS RRC
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STREAM DATA FOR PURE NITROGEN STREAMS
SI UNITS

STREAM NUMBER	080	081	082	083	084	085
FLOW, CMH(STP)	93920.1	93920.1	93920.1	22229.6	22229.6	17497.3
FLOW, GM.MOLE/SEC	1163.98	1163.98	1163.98	275.498	275.498	216.849
PRESSURE, KILOPASCAL	620.528	620.528	579.160	4136.86	151.685	151.685
TEMPERATURE, DEG K	97.0000	300.000	299.919	99.9000	81.0186	81.0186
ENTHALPY, JOULES/GM.MOLE	6693.07	12908.8	12908.8	2191.67	2191.67	1032.51
ENTROPY, JOULES/GM.MOLE-DEG C	73.4695	108.626	109.199	25.5956	28.6944	14.3890
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	0.78712	SAT. LIQ
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	086	087	088	089	090	091
FLOW, CMH(STP)	8079.84	8079.84	9417.41	9417.41	4732.34	22229.6
FLOW, GM.MOLE/SEC	100.136	100.136	116.713	116.713	58.6494	275.498
PRESSURE, KILOPASCAL	151.685	151.685	151.685	151.685	151.685	151.685
TEMPERATURE, DEG K	81.0186	81.0186	81.0186	81.0186	81.0186	81.0186
ENTHALPY, JOULES/GM.MOLE	1032.51	6477.53	1032.51	6477.53	6477.53	6477.53
ENTROPY, JOULES/GM.MOLE-DEG C	14.3890	81.5870	14.3890	81.5870	81.5870	81.5870
LIQUID FRACTION (MOLE L/F)	SAT. LIQ	SAT. VAP	SAT. LIQ	SAT. VAP	SAT. VAP	SAT. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	092	093	094	095	096
FLOW, CMH(STP)	5833.16	5833.16	28062.8	28062.8	28062.8
FLOW, GM.MOLE/SEC	72.2923	72.2923	347.791	347.791	347.791
PRESSURE, KILOPASCAL	151.685	151.685	151.685	151.685	103.421
TEMPERATURE, DEG K	90.6000	90.6000	82.9976	300.000	299.904
ENTHALPY, JOULES/GM.MOLE	6772.96	6772.96	6538.93	12935.8	12935.8
ENTROPY, JOULES/GM.MOLE-DEG C	85.0341	85.0341	82.3359	120.414	123.596
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:					
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000

TABLE 6
H2 LIQUEFIER - PROCESS RRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR PURE HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	001	002	003	004	005	006
FLOW, CMH(STP)	105140.	72865.8	32274.2	528831.	528831.	45335.5
FLOW, GM.MOLE/SEC	1303.03	903.049	399.984	6553.96	6553.96	561.857
PRESSURE, KILOPASCAL	4136.86	4136.86	4136.86	4136.86	4136.86	4136.86
TEMPERATURE, DEG K	308.000	308.000	308.000	308.000	85.5000	85.5000
ENTHALPY, JOULES/GM.MOLE	8779.20	8779.20	8779.20	8779.20	2697.04	2697.04
ENTROPY, JOULES/GM.MOLE-DEG C	112.301	112.301	112.301	112.301	77.6278	77.6278
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	007	008	009	010	011	012
FLOW, CMH(STP)	574166.	105140.	574166.	105140.	105140.	105140.
FLOW, GM.MOLE/SEC	7115.82	1303.03	7115.82	1303.03	1303.03	1303.03
PRESSURE, KILOPASCAL	4136.86	[illegible].86	4126.51	4126.51	4126.51	4040.33
TEMPERATURE, DEG K	85.5000	308.000	85.4874	85.4874	81.2186	81.1027
ENTHALPY, JOULES/GM.MOLE	2697.04	[illegible]	2697.04	2697.04	2305.57	2305.57
ENTROPY, JOULES/GM.MOLE-DEG C	77.6278	[illegible]	77.6465	77.6465	73.7822	73.9426
LIQUID FRACTION (MOLE L/F)	S.H. VAP	[illegible]	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	[illegible]	0.75000	0.75000	0.54509	0.54509
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.45491	0.45491

STREAM NUMBER	013	014	015	016	017	01[illegible]
FLOW, CMH(STP)	105140.	105140.	105140.	105140.	105140.	105140.
FLOW, GM.MOLE/SEC	1303.03	1303.03	1303.03	1303.03	1303.03	1303.03
PRESSURE, KILOPASCAL	4040.33	3988.62	3988.62	3936.91	930.792	930.792
TEMPERATURE, DEG K	41.1734	41.1034	28.0000	28.0281	28.6723	29.0895
ENTHALPY, JOULES/GM.MOLE	471.091	471.088	-220.310	-220.311	-220.315	-220.316
ENTROPY, JOULES/GM.MOLE-DEG C	42.2678	42.3413	23.3969	23.4526	26.8947	26.8965
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	S.C. LIQ	S.C. LIQ	S.C. LIQ	S.C. LIQ	S.C. LIQ
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.20596	0.20596	0.04132	0.04132	0.04132	0.03029
PARA -HYDROGEN	0.79404	0.79404	0.95868	0.95868	0.95868	0.96971

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TABLE 6
H2 LIQUEFIER - PROCESS RRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR PURE HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	019	020	021	022	023	024
FLOW, CMH(STP)	105140.	469026.	469026.	469026.	177463.	177463.
FLOW, GM.MOLE/SEC	1303.03	5812.79	5812.79	5812.79	2199.35	2199.35
PRESSURE, KILOPASCAL	930.792	4126.51	4126.51	4109.27	4109.27	344.738
TEMPERATURE, DEG K	20.5700	85.4874	81.2186	81.1955	81.1955	39.1734
ENTHALPY, JOULES/GM.MOLE	-451.097	2697.04	2587.71	2587.72	2587.72	1823.71
ENTROPY, JOULES/GM.MOLE-DEG C	17.6819	77.6465	76.3563	76.3869	76.3869	82.0290
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION MOLE FRACTION:						
ORTHO-HYDROGEN	0.03029	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.96971	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	025	026	027	028	029	030
FLOW, CMH(STP)	291564.	291564.	274850.	274850.	274850.	16714.2
FLOW, GM.MOLE/SEC	3613.44	3613.44	3406.30	3406.30	3406.30	207.144
PRESSURE, KILOPASCAL	4109.27	4109.27	4109.27	4109.27	358.527	4109.27
TEMPERATURE, DEG K	81.1955	58.3915	58.3915	58.3915	26.0000	58.3915
ENTHALPY, JOULES/GM.MOLE	2587.72	1944.92	1944.92	1944.92	1487.88	1944.92
ENTROPY, JOULES/GM.MOLE-DEG C	76.3869	66.9931	66.9931	66.9930	71.6835	66.9931
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	031	032	033	034	035	036
FLOW, CMH(STP)	16714.2	16714.2	56348.3	56348.3	222655.	222655.
FLOW, GM.MOLE/SEC	207.144	207.144	698.343	698.343	2759.43	2759.43
PRESSURE, KILOPASCAL	4109.27	111.695	358.527	358.527	344.738	344.738
TEMPERATURE, DEG K	29.5000	20.5200	26.0000	39.2700	39.1734	77.6955
ENTHALPY, JOULES/GM.MOLE	821.933	821.933	1487.88	1823.71	1823.71	2668.54
ENTROPY, JOULES/GM.MOLE-DEG C	40.8006	47.9117	71.6835	81.8447	82.0290	97.6332
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	0.69709	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

TABLE 6
H2 LIQUEFIER - PROCESS RHC
NASA CONTRACT NAS1-14698

STREAM DATA FOR PURE HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	037	038	039	040	041	042
FLOW, CMH(STP)	222655.	218501.	218501.	218501.	-11156.4	229658.
FLOW, GM.MOLE/SEC	2759.43	2707.96	2707.96	2707.96	-138.265	2846.22
PRESSURE, KILOPASCAL	324.053	358.527	358.527	344.738	344.738	344.738
TEMPERATURE, DEG K	77.5489	26.0000	39.2700	39.1734	39.1734	39.1734
ENTHALPY, JOULES/GM.MOLE	2666.54	1487.88	1823.71	1823.71	1823.71	1823.71
ENTROPY, JOULES/GM.MOLE-DEG C	98.1178	71.6835	81.8447	82.0290	82.0290	82.0290
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	043	044	045	046	047	048
FLOW, CMH(STP)	229658.	229658.	452312.	452312.	72865.6	45335.5
FLOW, GM.MOLE/SEC	2846.22	2846.22	5605.65	5605.65	903.049	561.857
PRESSURE, KILOPASCAL	344.738	324.053	324.053	324.053	4136.86	4136.86
TEMPERATURE, DEG K	77.6027	77.5561	77.6018	300.000	85.5000	85.5000
ENTHALPY, JOULES/GM.MOLE	2666.54	2666.55	2667.53	8505.97	2697.04	2697.04
ENTROPY, JOULES/GM.MOLE-DEG C	97.6050	98.0897	98.1036	132.491	77.6278	77.6278
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

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STREAM NUMBER	049	050	[illegible]	052	053	054
FLOW, CMH(STP)	27530.2	27530.2	27530.2	27530.2	44244.4	44244.4
FLOW, GM.MOLE/SEC	341.191	341.191	341.191	341.191	548.334	548.334
PRESSURE, KILOPASCAL	4136.86	41[illegible]	4126.51	111.695	111.695	111.695
TEMPERATURE, DEG K	85.5000	85.4874	33.2954	20.5200	20.5200	20.5200
ENTHALPY, JOULES/GM.MOLE	2697.04	2697.0[illegible]	929.315	929.315	888.749	1444.11
ENTROPY, JOULES/GM.MOLE-DEG C	77.6278	77.64[illegible]	44.2055	53.1569	51.1754	78.3027
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.C. LIQ	0.57678	0.62223	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

TABLE 6
H2 LIQUEFIER - PROCESS RRC
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STREAM DATA FOR PURE HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	055	056	057	058	059	060
FLOW, CMH(STP)	44244.4	44244.4	44244.4	44244.4	44244.4	496556.
FLOW, GM.MOLE/SEC	548.334	548.334	548.334	548.334	548.334	6153.98
PRESSURE, KILOPASCAL	111.695	108.248	108.248	103.421	289.580	289.580
TEMPERATURE, DEG K	71.3873	71.3782	300.000	300.002	308.000	300.728
ENTHALPY, JOULES/GM.MOLE	2544.05	2544.04	8503.41	8503.40	8732.93	8526.20
ENTROPY, JOULES/GM.MOLE-DEG C	105.345	105.540	141.616	142.020	133.767	133.184
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	061	062	063	064	065	066
FLOW, CMH(STP)	496556.	27530.1	27530.1	16714.3	105140.	2571.59
FLOW, GM.MOLE/SEC	6153.98	341.189	341.189	207.146	1303.03	31.8706
PRESSURE, KILOPASCAL	4136.86	111.695	111.695	111.695	101.353	101.353
TEMPERATURE, DEG K	308.000	20.5200	20.5200	20.5200	20.2337	20.2337
ENTHALPY, JOULES/GM.MOLE	8779.20	551.573	1444.11	1444.11	-451.097	422.877
ENTROPY, JOULES/GM.MOLE-DEG C	112.301	34.7056	78.3027	78.3028	18.8205	62.0279
LIQUID FRACTION (MOLE L/F)	S.H. VAP	SAT. LIQ	SAT. VAP	SAT. VAP	0.97554	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.03029	0.03029
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.96971	0.96971

STREAM NUMBER	067
FLOW, CMH(STP)	102568.
FLOW, GM.MOLE/SEC	1271.16
PRESSURE, KILOPASCAL	101.353
TEMPERATURE, DEG K	20.2337
ENTHALPY, JOULES/GM.MOLE	-473.010
ENTROPY, JOULES/GM.MOLE-DEG C	17.7372
LIQUID FRACTION (MOLE L/F)	SAT. LIQ
COMPOSITION, MOLE FRACTION:	
ORTHO-HYDROGEN	0.03029
PARA -HYDROGEN	0.96971

TABLE 7
H2 LIQUEFIER - PROCESS RRC
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STREAM DATA FOR PURE NITROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	080	081	082	083	084	085
FLOW, CFM(NTP)	3573146.	3573146.	3573146.	845714.	845714.	665675.
FLOW, LB.MOLE/HR	9237.95	9237.95	9237.95	2186.50	2186.50	1721.02
PRESSURE, PSIA	90.0000	90.0000	84.0000	600.000	22.0000	22.0000
TEMPERATURE, DEG K	97.0000	300.000	299.919	99.9000	81.0186	81.0186
ENTHALPY, BTU/LB.MOLE	2879.48	5553.63	5553.63	942.897	942.896	444.204
ENTROPY, BTU/LB.MOLE-DEG K	31.6080	46.7331	46.9795	11.0117	12.3449	6.19041
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	0.78712	SAT. LIQ
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	086	087	088	089	090	091
FLOW, CFM(NTP)	307394.	307394.	358281.	358281.	180040.	845714.
FLOW, LB.MOLE/HR	794.731	794.731	926.293	926.293	465.471	2186.50
PRESSURE, PSIA	22.0000	22.0000	22.0000	22.0000	22.0000	22.0000
TEMPERATURE, DEG K	81.0186	81.0186	81.0186	81.0186	81.0186	81.0186
ENTHALPY, BTU/LB.MOLE	444.204	2786.75	444.204	2786.75	2786.75	2786.75
ENTROPY, BTU/LB.MOLE-DEG K	6.19041	35.1003	6.19041	35.1003	35.1003	35.1003
LIQUID FRACTION (MOLE L/F)	SAT. LIQ	SAT. VAP	SAT. LIQ	SAT. VAP	SAT. VAP	SAT. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

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STREAM NUMBER	092	093	094	095	096
FLOW, CFM(NTP)	221920.	221920.	1067634.	1067634.	1067634.
FLOW, LB.MOLE/HR	573.748	573.748	2760.24	2760.24	2760.24
PRESSURE, PSIA	22.0000	22.0000	22.0000	22.0000	15.0000
TEMPERATURE, DEG K	90.6000	90.6000	82.9976	300.000	299.904
ENTHALPY, BTU/LB.MOLE	2913.85	2913.85	2813.17	5565.23	5565.23
ENTROPY, BTU/LB.MOLE-DEG K	36.5832	36.5832	35.4224	51.8045	53.1732
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:					
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000

TABLE 7
H2 LIQUEFIER - PROCESS RRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR PURE HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	001	002	003	004	005	006
FLOW, CFH(NTP)	4000000.	2772144.	1227856.	20119104.	20119104.	1724769.
FLOW, LB.MOLE/HR	10341.5	7167.05	3174.48	52015.6	52015.6	4459.19
PRESSURE, PSIA	600.000	600.000	600.000	600.000	600.000	600.000
TEMPERATURE, DEG K	308.000	308.000	308.000	308.000	85.5000	85.5000
ENTHALPY, BTU/LB.MOLE	3776.98	3776.98	3776.98	3776.98	1160.32	1160.32
ENTROPY, BTU/LB.MOLE-DEG K	48.3138	48.3138	48.3138	48.3138	33.3969	33.3969
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	007	008	009	010	011	012
FLOW, CFH(NTP)	21843872.	4000000.	21843872.	4000000.	4000000.	4000000.
FLOW, LB.MOLE/HR	56474.8	10341.5	56474.8	10341.5	10341.5	10341.5
PRESSURE, PSIA	600.000	600.000	598.500	598.500	598.500	586.000
TEMPERATURE, DEG K	85.5000	308.000	85.4874	85.4874	81.2186	81.1027
ENTHALPY, BTU/LB.MOLE	1160.32	0.0	1160.32	1160.32	991.899	991.898
ENTROPY, BTU/LB.MOLE-DEG K	33.3969	0.0	33.4050	33.4050	31.7425	31.8115
LIQUID FRACTION (MOLE L/F)	S.H. VAP	SAT. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.54509	0.54509
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.45491	0.45491

STREAM NUMBER	013	014	015	016	017	018
FLOW, CFH(NTP)	4000000.	4000000.	4000000.	4000000.	4000000.	4000000.
FLOW, LB.MOLE/HR	10341.5	10341.5	10341.5	10341.5	10341.5	10341.5
PRESSURE, PSIA	586.000	578.500	578.500	571.000	135.000	135.000
TEMPERATURE, DEG K	41.1734	41.1034	28.0000	28.0281	28.6723	29.0895
ENTHALPY, BTU/LB.MOLE	202.672	202.671	-94.7814	-94.7821	-94.7836	-94.7838
ENTROPY, BTU/LB.MOLE-DEG K	18.1844	18.2160	10.0658	10.0897	11.5706	11.5714
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	S.C. LIQ	S.C. LIQ	S.C. LIQ	S.C. LIQ	S.C. LIQ
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.20596	0.20596	0.04132	0.04132	0.04132	0.03029
PARA -HYDROGEN	0.79404	0.79404	0.95868	0.95868	0.95868	0.96971

TABLE 7
H2 LIQUEFIER - PROCESS RRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR PURE HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	019	020	021	022	023	024
FLOW, CFH(NTP)	4000000.	17843872.	17843872.	17843872.	6751485.	6751485.
FLOW, LB.MOLE/HR	10341.5	46133.3	46133.3	46133.3	17455.2	17455.2
PRESSURE, PSIA	135.000	598.500	598.500	596.000	596.000	50.0000
TEMPERATURE, DEG K	20.5700	85.4874	81.2186	81.1955	81.1955	39.1734
ENTHALPY, BTU/LB.MOLE	-194.070	1160.32	1113.28	1113.28	1113.28	784.596
ENTROPY, BTU/LB.MOLE-DEG K	7.60710	33.4050	32.8499	32.8631	32.8631	35.2904
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.03029	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.96971	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	025	026	027	028	029	030
FLOW, CFH(NTP)	11092402.	11092402.	10456521.	10456521.	10456521.	635882.
FLOW, LB.MOLE/HR	28678.1	28678.1	27034.1	27034.1	27034.1	1644.00
PRESSURE, PSIA	596.000	596.000	596.000	596.000	52.0000	596.000
TEMPERATURE, DEG K	81.1955	58.3915	58.3915	58.3915	26.0000	58.3915
ENTHALPY, BTU/LB.MOLE	1113.28	836.741	836.741	836.740	640.115	836.741
ENTROPY, BTU/LB.MOLE-DEG K	32.8631	28.8217	28.8217	28.8216	30.8396	28.8217
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

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STREAM NUMBER	031	032	033	034	035	036
FLOW, CFH(NTP)	635882.	635882.	2143745.	2143745.	8470790.	8470790.
FLOW, LB.MOLE/HR	1644.00	1644.00	5542.40	5542.40	21900.2	21900.2
PRESSURE, PSIA	596.000	16.2000	52.0000	52.0000	50.0000	50.0000
TEMPERATURE, DEG K	29.5000	20.5200	26.0000	39.2700	39.1734	77.6955
ENTHALPY, BTU/LB.MOLE	353.611	353.611	640.115	784.594	784.596	1148.06
ENTROPY, BTU/LB.MOLE-DEG K	17.5532	20.6125	30.8396	35.2111	35.2904	42.0036
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	0.69709	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

TABLE 7
H2 LIQUEFIER - PROCESS RRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR PURE HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	037	038	039	040	041	042
FLOW, CFM(NTP)	8470790.	8312776.	8312776.	8312776.	-426440.	8737216.
FLOW, LB.MOLE/HR	21900.2	21491.7	21491.7	21491.7	-1097.34	22589.1
PRESSURE, PSIA	47.0000	52.0000	52.0000	50.0000	50.0000	50.0000
TEMPERATURE, DEG K	77.6489	26.0000	39.2700	39.1734	39.1734	39.1734
ENTHALPY, BTU/LB.MOLE	1148.06	640.115	784.594	784.596	784.596	784.596
ENTROPY, BTU/LB.MOLE-DEG K	42.2121	30.8396	35.2111	35.2904	35.2904	35.2904
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	043	044	045	046	047	048
FLOW, CFM(NTP)	8737216.	8737216.	17208000.	17208000.	2772144.	1724769.
FLOW, LB.MOLE/HR	22589.1	22589.1	44489.3	44489.3	7167.05	4459.19
PRESSURE, PSIA	50.0000	47.0000	47.0000	47.0000	600.000	600.000
TEMPERATURE, DEG K	77.6027	77.5561	77.6018	300.000	85.5000	85.5000
ENTHALPY, BTU/LB.MOLE	1147.20	1147.20	1147.62	3659.43	1160.32	1160.32
ENTROPY, BTU/LB.MOLE-DEG K	41.9915	42.2000	42.2060	57.0000	33.3969	33.3969
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	049	050	051	052	053	054
FLOW, CFM(NTP)	1047375.	1047375.	1047375.	1047375.	1683256.	1683256.
FLOW, LB.MOLE/HR	2707.87	2707.87	2707.87	2707.87	4351.86	4351.86
PRESSURE, PSIA	600.000	598.500	598.500	16.2000	16.2000	16.2000
TEMPERATURE, DEG K	85.5000	85.4874	33.2954	20.5200	20.5200	20.5200
ENTHALPY, BTU/LB.MOLE	1160.32	1160.32	399.809	399.809	382.356	621.284
ENTROPY, BTU/LB.MOLE-DEG K	33.3969	33.4050	19.0180	22.8691	22.0166	33.6873
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.C. LIQ	0.57678	0.62223	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

TABLE 7
H2 LIQUEFIER - PROCESS RRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR PURE HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	055	056	057	058	059	060
FLOW, CFH(NTP)	1683256.	1683256.	1683256.	1683256.	1683256.	18891248.
FLOW, LB.MOLE/HR	4351.86	4351.86	4351.86	4351.86	4351.86	48841.1
PRESSURE, PSIA	16.2000	15.7000	15.7000	15.0000	42.0000	42.0000
TEMPERATURE, DEG K	71.3873	71.3782	300.000	300.002	308.000	300.728
ENTHALPY, BTU/LB.MOLE	1094.50	1094.49	3658.32	3658.32	3757.07	3668.13
ENTROPY, BTU/LB.MOLE-DEG K	45.3214	45.4054	60.9259	61.0995	57.5489	57.2980
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.75000	0.75000
PARA - HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	061	062	063	064	065	066
FLOW, CFH(NTP)	18891248.	1047368.	1047369.	635888.	4000000.	97835.1
FLOW, LB.MOLE/HR	48841.1	2707.85	2707.85	1644.01	10341.5	252.941
PRESSURE, PSIA	600.000	16.2000	16.2000	16.2000	14.7000	14.7000
TEMPERATURE, DEG K	308.000	20.5200	20.5200	20.5200	20.2337	20.2337
ENTHALPY, BTU/LB.MOLE	3776.98	237.297	621.284	621.283	-194.070	181.930
ENTROPY, BTU/LB.MOLE-DEG K	48.3138	14.9310	33.6873	33.6873	8.09691	26.6855
LIQUID FRACTION (MOLE L/F)	S.H. VAP	SAT. LIQ	SAT. VAP	SAT. VAP	0.97554	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.03029	0.03029
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.96971	0.96971

STREAM NUMBER	067
FLOW, CFH(NTP)	3902164.
FLOW, LB.MOLE/HR	10088.6
PRESSURE, PSIA	14.7000
TEMPERATURE, DEG K	20.2337
ENTHALPY, BTU/LB.MOLE	-203.498
ENTROPY, BTU/LB.MOLE-DEG K	7.63086
LIQUID FRACTION (MOLE L/F)	SAT. LIQ
COMPOSITION, MOLE FRACTION:	
ORTHO-HYDROGEN	0.03029
PARA -HYDROGEN	0.96971

TABLE 8
N2 REFRIGERATOR - PROCESS RRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR N2 REFRIGERATOR STREAMS
SI UNITS

STREAM NUMBER	001	002	003	004	005	006
FLOW, CMH(STP)	364219.	364219.	334050.	137307.	137307.	196743.
FLOW, GM.MOLE/SEC	4513.89	4513.89	4139.98	1701.69	1701.69	2438.29
PRESSURE, KILOPASCAL	2809.74	3346.24	4136.86	4136.86	606.739	4136.86
TEMPERATURE, DEG K	308.000	308.000	308.000	235.000	147.536	235.000
ENTHALPY, JOULES/GM.MOLE	13028.3	13000.8	12961.1	10650.9	8364.43	10650.9
ENTROPY, JOULES/GM.MOLE-DEG C	96.4940	94.9562	93.0685	84.4956	87.5968	84.4956
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	007	008	009	010	011	012
FLOW, CMH(STP)	196743.	239166.	239166.	30169.9	30169.9	30169.9
FLOW, GM.MOLE/SEC	2438.29	2964.06	2964.06	373.905	373.905	373.905
PRESSURE, KILOPASCAL	4136.86	606.739	606.739	4136.86	4136.86	4136.86
TEMPERATURE, DEG K	162.416	148.413	232.000	235.000	162.416	140.000
ENTHALPY, JOULES/GM.MOLE	8000.39	8391.59	10706.3	10650.9	8000.39	6763.73
ENTROPY, JOULES/GM.MOLE-DEG C	70.8873	87.7804	101.241	84.4956	70.8873	62.6387
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	013	014	015	016	017	018
FLOW, CMH(STP)	30169.9	101859.	101859.	101859.	30169.9	30169.9
FLOW, GM.MOLE/SEC	373.905	1262.38	1262.38	1262.38	373.905	373.905
PRESSURE, KILOPASCAL	4136.86	620.528	620.528	606.739	2757.90	2757.90
TEMPERATURE, DEG K	99.0000	97.0000	138.000	149.597	100.223	100.223
ENTHALPY, JOULES/GM.MOLE	2137.69	6693.07	8063.26	8428.20	2200.35	2200.35
ENTROPY, JOULES/GM.MOLE-DEG C	25.0528	73.4695	85.3076	88.0261	26.2303	26.2303
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	SAT. LIQ
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

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TABLE 8
N2 REFRIGERATOR - PROCESS RRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR N2 REFRIGERATOR STREAMS
SI UNITS

STREAM NUMBER	019	020	021	022	023	025
FLOW, CMH(STP)	0.0	94883.3	94883.3	196743.	0.0	334050.
FLOW, GM.MOLE/SEC	0.0	1175.92	1175.92	2438.29	0.0	4139.98
PRESSURE, KILOPASCAL	2757.90	620.528	620.528	620.528	620.528	4136.86
TEMPERATURE, DEG K	100.223	97.0000	97.0000	97.0000	63.1366	235.000
ENTHALPY, JOULES/GM.MOLE	0.0	6693.07	6693.07	6693.07	-0.00908	10650.9
ENTROPY, JOULES/GM.MOLE-DEG C	0.0	73.4695	73.4695	73.4695	-0.25012	84.4957
LIQUID FRACTION (MOLE L/F)	SAT. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	026	027	028	029	030	031
FLOW, CMH(STP)	334050.	30169.9	30169.9	239166.	364219.	364219.
FLOW, GM.MOLE/SEC	4139.98	373.905	373.905	2964.06	4513.89	4513.89
PRESSURE, KILOPASCAL	4136.86	4136.86	4136.86	592.949	4136.86	579.160
TEMPERATURE, DEG K	264.051	308.000	264.051	303.000	308.000	302.613
ENTHALPY, JOULES/GM.MOLE	11588.1	12961.1	11588.1	12998.5	12961.1	12987.9
ENTROPY, JOULES/GM.MOLE-DEG C	88.2572	93.0685	88.2572	109.301	93.0685	109.461
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NUMBER	032	033	034	035
FLOW, CMH(STP)	94883.3	269336.	30169.9	30169.9
FLOW, GM.MOLE/SEC	1175.92	3337.97	373.905	373.905
PRESSURE, KILOPASCAL	579.160	592.949	592.949	101.353
TEMPERATURE, DEG K	300.000	303.560	308.000	300.000
ENTHALPY, JOULES/GM.MOLE	12911.2	13014.9	13145.2	12938.7
ENTROPY, JOULES/GM.MOLE-DEG C	109.207	109.355	109.781	123.773
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:				
NITROGEN	1.00000	1.00000	1.00000	1.00000

TABLE 9
N2 REFRIGERATOR - PROCESS RRC
NASA CONTRACT NAS1-14698

STREAM DATA FOR N2 REFRIGERATOR STREAMS
US CUSTOMARY UNITS

STREAM NO., N-(STYP/KTYP/HTYP)	001 (01/03/01)	002 (01/03/01)	003 (01/03/01)	004 (01/03/01)	005 (01/03/01)	006 (01/03/01)
FLOW, CFH(NTP)	13856555.	13856555.	12708755.	5223775.	5223775.	7484980.
FLOW, LB.MOLE/HR	35824.5	35824.5	32857.0	13505.5	13505.5	19351.5
PRESSURE, PSIA	407.519	485.331	600.000	600.000	88.0000	600.000
TEMPERATURE, DEG K	308.000	308.000	308.000	235.000	147.536	235.000
ENTHALPY, BTU/LB.MOLE	5505.02	5593.21	5576.10	4582.21	3598.54	4582.21
ENTROPY, BTU/LB.MOLE-DEG K	41.5135	40.8519	40.0398	36.3516	37.6858	36.3516
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

STREAM NO., N-(STYP/KTYP/HTYP)	007 (01/03/01)	008 (01/03/01)	009 (01/03/01)	010 (01/03/01)	011 (01/03/01)	012 (01/03/01)
FLOW, CFH(NTP)	7484980.	9098965.	9098965.	1147800.	1147800.	1147800.
FLOW, LB.MOLE/HR	19351.5	23524.3	23524.3	2967.50	2967.50	2967.50
PRESSURE, PSIA	600.000	88.0000	88.0000	600.000	600.000	600.000
TEMPERATURE, DEG K	162.416	148.413	232.000	235.000	162.416	140.000
ENTHALPY, BTU/LB.MOLE	3441.92	3610.22	4692.08	4582.21	3441.92	2909.88
ENTROPY, BTU/LB.MOLE-DEG K	30.4970	37.7648	43.5560	36.3516	30.4970	26.9483
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

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STREAM NO., N-(STYP/KTYP/HTYP)	013 (01/03/01)	014 (01/03/01)	015 (01/03/01)	016 (01/03/01)	017 (01/03/01)	018 (01/03/01)
FLOW, CFH(NTP)	1147800.	3875190.	3875190.	3875190.	1147800.	1147800.
FLOW, LB.MOLE/HR	2967.50	10018.9	10018.9	10018.9	2967.50	2967.50
PRESSURE, PSIA	600.000	90.0000	90.0000	88.0000	400.000	400.000
TEMPERATURE, DEG K	99.0000	97.0000	138.000	149.597	100.223	100.223
ENTHALPY, BTU/LB.MOLE	919.675	2879.48	3468.97	3625.97	946.631	946.631
ENTROPY, BTU/LB.MOLE-DEG K	10.7782	31.6080	36.7009	37.8705	11.2848	11.2848
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	SAT. LIQ
COMPOSITION, MOLE FRACTION:						
NITROGEN	1.00000	1.00000	1.00000	1.00000	1.00000	1.00000

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                                        TABLE 9
                              N2 REFRIGERATOR - PROCESS RRC
                                NASA CONTRACT NAS1-14698

                                            STREAM DATA FOR N2 REFRIGERATOR STREAMS
                                                    US CUSTOMARY UNITS

 STREAM NO., N-(STYP/KTYP/HTYP)      019 (01/03/01)  020 (01/03/01)  021 (01/03/01)  022 (01/03/01)  023 (01/03/01)  025 (01/03/01)

 FLOW, CFM(NTP)                                0.0        3609790.        3609790.        7484980.             0.0       12708755.
 FLOW, LB.MOLE/HR                              0.0         9332.69         9332.69         19351.5             0.0         32857.0
 PRESSURE, PSIA                            400.000         90.0000         90.0000         90.0000         90.0000         600.000
 TEMPERATURE, DEG K                        100.223         97.0000         97.0000         97.0000         63.1366         235.000
 ENTHALPY, BTU/LB.MOLE                         0.0         2879.48         2879.48         2879.48        -0.00391         4582.20
 ENTROPY, BTU/LB.MOLE-DEG K                    0.0         31.6080         31.6080         31.6080        -0.10760         36.3516
 LIQUID FRACTION (MOLE  L/F)              SAT. VAP        S.H. VAP        S.H. VAP        S.H. VAP        S.C. LIQ        S.H. VAP
 COMPOSITION, MOLE FRACTION:
   NITROGEN                                1.00000         1.00000         1.00000         1.00000         1.00000         1.00000


 STREAM NO., N-(STYP/KTYP/HTYP)      026 (01/03/01)  027 (01/03/01)  028 (01/03/01)  029 (01/03/01)  030 (01/03/01)  031 (01/03/01)

 FLOW, CFM(NTP)                          12708755.        1147800.        1147800.        9098965.       13856555.       13856555.
 FLOW, LB.MOLE/HR                          32857.0         2967.50         2967.50         23524.3         35824.5         35824.5
 PRESSURE, PSIA                            600.000         600.000         600.000         86.0000         600.000         84.0000
 TEMPERATURE, DEG K                        264.051         308.000         264.051         303.000         308.000         302.613
 ENTHALPY, BTU/LB.MOLE                     4985.43         5576.10         4985.43         5592.18         5576.10         5587.63
 ENTROPY, BTU/LB.MOLE-DEG K                37.9699         40.0398         37.9699         47.0233         40.0398         47.0923
 LIQUID FRACTION (MOLE  L/F)              S.H. VAP        S.H. VAP        S.H. VAP        S.H. VAP        S.H. VAP        S.H. VAP
 COMPOSITION, MOLE FRACTION:
   NITROGEN                                1.00000         1.00000         1.00000         1.00000         1.00000         1.00000


 STREAM NO., N-(STYP/KTYP/HTYP)      032 (01/03/01)  033 (01/03/01)  034 (01/03/01)  035 (01/03/01)

 FLOW, CFM(NTP)                           3609790.       10246765.        1147800.        1147800.
 FLOW, LB.MOLE/HR                          9332.69         26491.8         2967.50         2967.50
 PRESSURE, PSIA                            84.0000         86.0000         86.0000         14.7000
 TEMPERATURE, DEG K                        300.000         303.560         308.000         300.000
 ENTHALPY, BTU/LB.MOLE                     5554.64         5599.25         5655.30         5566.48
 ENTROPY, BTU/LB.MOLE-DEG K                46.9828         47.0467         47.2300         53.2496
 LIQUID FRACTION (MOLE  L/F)              S.H. VAP        S.H. VAP        S.H. VAP        S.H. VAP
 COMPOSITION, MOLE FRACTION:
   NITROGEN                                1.00000         1.00000         1.00000         1.00000
```

TABLE 10

SPECIFICATIONS

CENTRIFUGAL HYDROGEN RECYCLE COMPRESSORS

	UNITS		MAIN RECYCLE		FLASH	
	SI	Customary	SI	Customary	SI	Customary
Flow	m^3/s	MCFH-NTP	2481	315,480	197.0	25,040
Suction Pressure	kPa	psia	289.6	42	103.4	15
Discharge Pressure	kPa	psia	2137	310	289.6	42
Suction Temperature	K	F	300	80	300	80
Cooling Water Temperature	K	F	300	80	300	80
Power	MWm	BHP	753.00	1,009,800	31.99	42,900
Adiabatic Efficiency	%	%	75.54	75.54	73.2	73.2
Fluid			85% H_2 - 15% propane		85% H_2 - 15% Propane	
No. of Compressors			40		20	
Stages per Compressor			3		1	
Type			Barrel		Horizontally Split	
No. Impellers/Compressor			6		5	
Speed		RPM	4450		6000	

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TABLE 11

SPECIFICATIONS
RECIPROCATING HYDROGEN RECYCLE COMPRESSORS

	UNITS		MAIN RECYCLE		FLASH	
	SI	CUSTOMARY	SI	CUSTOMARY	SI	CUSTOMARY
Flow	m^3/s	MCFH-NTP	152.6	19,400	13.37	1,700
Suction Pressure	kPa	psia	289.6	42	103.4	15
Discharge Pressure	kPa	psia	4137	600	289.6	42
Suction Temperature	K	F	301	82	301	82
Cooling Water Temperature	K	F	303	86	303	86
Power	MWm	BHP	649.90	871,500	22,37	30,000
Adiabatic Efficiency	%	%	74.24	74.24	77.9	77.9
Fluid			Pure Hydrogen		Pure Hydrogen	
No. of Compressors			70		10	
Stages per Compressor			3		1	
Type			Non-Lube		Non-Lube	
No. of Cylinders/Compressor			10		10	
Arrangement			Balanced opposed		Balanced opposed	
Speed		RPM	300		300	

TABLE 12

POWER SUMMARY

H_2 LIQUEFIER USING RECIPROCATING RECYCLE COMPRESSORS - PROCESS RRC

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26.25 kg/s (2500 TPD) LIQUID H_2

PRODUCTION	BHP	MW_e
H_2 Recycle Compressors	871,500	660.75
H_2 Flash Compressors	30,000	22.75
Nitrogen Recycle Compressors	340,920	262.20
Forecoolers	42,130	33.10
Air Compressors, N_2 Plant	20,000	15.95
Purifier Heat Pump Compressors	42,550	33.43
Hydrogen Feed/Booster Compressors	72,440	56.30
Nitrogen Feed Compressors	34,000	26.71
Hydrogen Driers	1,050	7.26
Pumps	2,040	1.81
Subtotal	1,456,630	1,120.31
Hydrogen Turbine Return	-44,500	-31.55
Net Subtotal	1,412,130	1,088.76
PRODUCTION AUXILIARIES		
Cooling Tower & Water Supply	58,000	49.52
Plant Air Compressor & Drier	4,750	3.14
Purge Blower & Thaw Heater	4,750	13.43
Miscellaneous	950	17.86
Subtotal	68,450	83.95
Process Contingency	74,030	58.63
Subtotal	1,554,610	1,231.34
PLANT AUXILIARIES		
Road & Exterior Lighting		0.50
Building Lighting, Heating, Air Conditioning		1.30
Cranes		0.40
		2.20
TOTAL POWER - ELECTRICAL		1233.54

TABLE 13

POWER SUMMARY

H_2 LIQUEFIER USING CENTRIFUGAL RECYCLE COMPRESSORS - PROCESS CRC

26.25 kg/s (2500 TPD) LIQUID H_2

PRODUCTION	BHP	MW_e	MWm
H_2 Recycle Compressors	1,009,800		753.01
H_2 Flash Compressors	42,900		31.99
Nitrogen Recycle Compressors	345,320		258.25
Forecoolers	34,740	27.29	
Air Compressors, N_2 Plant	20,000		14.91
Purifier Heat Pump Compressors	42,550		31.73
Hydrogen Feed Compressors	36,760	28.57	
Nitrogen Feed Compressors	36,090		26.92
Hydrogen Driers	1,050	7.26	
Pumps	2,040	1.81	
Subtotal	1,572,250	64.93	1,116.81
Hydrogen Turbine Return	- 51,230	-36.29	
Net Subtotal	1,521,020	28.64	1,116.81
PRODUCTION AUXILIARIES			
Cooling Tower & Water Supply	58,000	49.52	
Plant Air Compressor and Driers	4,750	3.14	
Purge Blowers and Thaw Heaters	4,750	13.43	
Miscellaneous	950	17.86	
Subtotal	68,450	83.95	
Process Contingency		5.68	55.79
Subtotal		118.27	1,172.60
PLANT AUXILIARIES			
Road and Exterior Lighting		0.500	
Building Lighting, Heating Air Conditioning		1.300	
Cranes		0.400	
		2.200	
TOTAL POWER - ELECTRICAL		120.47	
MECHANICAL			1,172.60

TABLE 14
SUMMARY OF FINANCING RULES
DCF AND UTILITY FINANCING

DISCOUNTED CASH FLOW (DCF) FINANCING

25 Year Project Life
16 Year Sum-of-the-Years Digits Depreciation
100% Equity Capital
12% Discounted Rate of Return
48% Federal Income Tax Rate
No Escalation

$$\text{Unit Cost} = \frac{aN + 0.2353\ I + 0.1275\ S + 0.2308\ W}{G}$$

UTILITY FINANCING

20 Year Project Life
5% Straight Line Depreciation on Total Capital Investment
48% Federal Income Tax Rate
3:1 Debt to Equity
9% Interest Rate on Debt
15% Rate of Return on Equity
No Escalation

$$\text{Unit Cost} = \frac{aN + 0.05(C-W) + 0.005\ (P + \frac{48}{52}\ [1-d]\ r)(C-W)}{G}$$

NOMENCLATURE

a = Escalation Factor = 1.00 for No Escalation
N = Total Net Annual Operating Cost, \$MM
I = Total Plant Investment, \$MM
S = Startup Costs, \$MM
C = Total Capital Requirement, \$MM
W = Working Capital, \$MM

G = Annual Liquid H_2 Production, Mg/yr
d = Fraction Debt
i = Interest Rate on Debt, %
r = Return on Equity, %
p = Return on Rate Base
$p = d(i) + (1-d)r$
Unit Cost in \$/kg

TABLE 15

BASIS FOR ECONOMIC EVALUATIONS

1. 1974 dollars
2. CE Plant Cost Index (20) for escalation
3. Plant capacity = 26.25 kg/s (2500 TPD) LH_2
4. Interest during construction at 9% for 1.875 yrs.
5. Startup costs at 20% of total gross operating cost for coal gasifiers
6. Startup costs at 2.75% of total plant investment for H_2 liquefier.
7. Working capital at 0.9% of total plant investment plus raw materials (coal) inventory of 60 days for coal gasifiers.
8. Working capital at 0.9% of total plant investment plus net receivables at 1/24 of annual liquid hydrogen revenue at $2.85/Gj ($3/MM Btu) for H_2 liquefier.
9. Makeup water at 7.93 ¢/m^3 (30¢/M gal)
10. Operating labor at $6.00/hr.
11. Maintenance labor at 1.5% of total plant investment
12. Administration and overhead at 60% of total labor.
13. Operating supplies at 30% of operating labor
14. Maintenance supplies at 1.5% of total plant investment.
15. Local taxes and insurance at 2.7% of total plant investment.
16. Sale of by-product sulfur at 2.66¢/kg ($27/long ton)
17. 8322 operating hours per year (95% on stream)
18. Transmission of feedstock and energy from gasifiers to liquefaction complex not included.
19. Land acquisition not included.

TABLE 16

CAPITAL INVESTMENT

COAL GASIFICATION COMPLEX

26.25 kg/s (2500 TPD) LIQUID H_2

$1,000's

PROCESS	PROCESS	
	RRC	CRC
H_2 and Fuel Gas Production		
Coal Preparation and Water Gas Shift	416,000	410,300
Raw Gas Compression	131,800	129,900
H_2 and Fuel Gas Purification	111,100	110,400
Sulfur and CO_2 Removal		
O_2 Plant and Compression	314,100	309,200
Power and Steam Generation	287,700	178,700
Electrical Substation and Switchgear	52,800	19,800
Water Treatment and Cooling	29,000	28,300
General Facility, Roads, Buildings, Etc.	21,700	21,600
Subtotal, Plant Investment	1,364,200	1,208,200
Project Contingency at 15%	204,600	181,200
TOTAL PLANT INVESTMENT	1,568,800	1,389,400

TABLE 17

ANNUAL OPERATING COST

COAL GASIFICATION COMPLEX

26.25 kg/s (2500 TPD) LIQUID H_2

$1,000's

		PROCESS	
PROCESS		RRC	CRC
Coal Consumption	kg/s (TPD)	20.946 (1994.9)	20.582 (1960.2)
Coal @ $0.7113/GJ	($0.75/MM Btu)	323,730	319,024
$0.4742/GJ	($0.50/MM Btu)	215,820	212.683
$0.3320/GJ	($0.35/MM Btu)	151,074	148,878
Catalysts and Chemicals		1,213	1,194
Process Water		1,039	1,021
LABOR -	Process	4,494	4,494
	Maintenance	23,533	20,840
	Supervision	1,332	1,332
Administration and	Overhead	17,509	15,999
Supplies			
	Operating	1,348	1,348
	Maintenance	23,533	20,840
Local Taxes and Insurance		42,359	37,512
TOTAL GROSS OPERATING COST		$440,090	$423,603
		332,180	317,262
		267,434	253,457
SALE OF BY-PRODUCT SULFUR		14,164	14,044
TOTAL NET OPERATING COST		$425,926	$409,559
		318,016	303,218
		253,270	239,413

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TABLE 18
UNIT GASIFICATION COST
2500 TPD LIQUID H_2
DCF FINANCING

I = Total Plant Investment, $MM
S = Startup Costs, $MM
W = Working Capital, $MM
N = Total Net Annual Operating Cost, $MM
a = Escalation Factor = 1.00
G = Annual Liquid H_2 Production 786.43 Mg/yr(1733.8 MM lb/yr)

POWER GASIFICATION

	PROCESS RRC			PROCESS CRC		
Coal Cost $/Gj ($/MM Btu)	0.3320 (0.35	0.4742 (0.50)	0.7113 (0.75)	0.3320 (0.35)	0.4742 (0.50)	0.7113 (0.75)
I	1036.44	1036.44	1036.44	857.00	857.00	857.00
S	33.82	41.77	55.04	31.03	38.79	51.74
W	25.38	32.27	43.75	23.39	30.12	41.32
N	160.37	200.17	266.49	146.51	185.37	250.12
Unit Cost $/kg (¢/lb)	0.5269 (23.90)	0.5809 (26.35)	0.6707 (30.42)	0.4545 (20.62)	0.5072 (23.01)	0.5949 (26.99)

FEEDSTOCK GASIFICATION

Coal Cost $/GJ ($/MM Btu)	0.3320 (0.35)	0.4742 (0.50)	0.7113 (0.75)
I	532.40	532.40	532.40
S	19.67	24.67	32.98
W	14.87	19.18	26.38
N	92.90	117.85	159.44
Unit Cost $/kg (¢/lb)	0.2851 (12.93)	0.3188 (14.46)	0.3752 (17.02)

NOTE: Feedstock gasifier is identical for Processes RRC and CRC.

TABLE 19

GASIFICATION COST

UTILITY FINANCING

26.25 kg/s (2500 TPD) LIQUID H_2

C = Total Capital Requirement, $MM

W = Working Capital, $MM

N = Total Net Annual Operating Cost, $MM

G = Annual Liquid H_2 Production, 786.43 Mg/yr (1733.8 MM lb/yr)

d = Fraction debt = 0.75

r = Return on Equity = 15%

i = Interest Rate on debt = 9%

p = Return on Rate Base

a = Escalation Factor = 1.00

$$p = d(i) + (1 - d)r$$

$$p = 10.5$$

POWER GASIFICATION

	PROCESS RRC			PROCESS CRC		
Coal Cost, $/Gj	0.3320	0.4742	0.7113	0.3320	0.4742	0.7113
($/MMBtu)	(0.35)	(0.50)	(0.75)	(0.35)	(0.50)	(0.75)
I	1036.44	1036.44	1036.44	857.00	857.00	857.00
S	33.82	41.77	55.04	31.03	38.79	51.74
W	25.38	32.27	43.75	23.39	30.12	41.32
IDC	174.90	174.90	174.90	144.62	144.62	144.62
C	1270.54	1285.38	1310.13	1056.04	1070.53	1094.68
N	160.37	200.17	266.49	146.51	185.37	250.12
Unit Cost $/kg	0.3981	0.4513	0.5395	0.3478	0.3994	0.4857
(¢/lb)	(18.06)	(20.46)	(24.47)	(15.77)	(18.12)	(22.04)

TABLE 19 - Cont'd

GASIFICATION COST

UTILITY FINANCING

2500 TPD LIQUID H_2

FEEDSTOCK GASIFICATION

Coal Cost $/Gj	0.3320	0.4742	0.7113
($/MM Btu)	(0.35)	(0.50)	(0.75)
I	532.40	532.40	532.40
S	19.67	24.67	32.98
W	14.87	19.18	26.38
IDC	89.84	89.84	89.84
C	656.78	666.09	681.60
N	92.90	117.85	159.44
Unit Cost $/kg	0.2185	0.2518	0.3073
(¢/lb)	(9.91)	(11.42)	(13.94)

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TABLE 20

CAPITAL INVESTMENT

LIQUEFACTION COMPLEX

26.25 kg/s (2500 TPD) LIQUID H_2

$1,000's

	PROCESS	
	RRC	CRC
TOTAL PLANT INVESTMENT	$574,100	$496,900
INTEREST DURING CONSTRUCTION	96,900	83,800
STARTUP COSTS	15,800	13,700
WORKING CAPITAL	18,000	17,300
TOTAL CAPITAL REQUIREMENT	$704,800	$611,700

TABLE 21

ANNUAL OPERATING COST

LIQUEFACTION COMPLEX

26.25 kg/s (2500 TPD) LIQUID H_2

	PROCESS	
	RRC	CRC
RAW MATERIALS		
Feedstock - From Coal Gasifier	--	--
CHEMICALS AND ADSORBERS		
Sulfuric Acid	$832,000	$832,000
Dessicants and Adsorbents	301,500	301,500
UTILITIES		
Makeup Water	2,247,000	2,247,000
Electricity - From Power Gasifier	--	--
LABOR		
Operating	1,797,600	1,797,600
Supervision	230,800	230,800
ADMINISTRATION AND OVERHEAD	1,217,000	1,217,000
SUPPLIES		
Operating	540,000	540,000
Maintenance	8,611,000	5,474,000
LOCAL TAXES AND INSURANCE	15,500,000	13,416,000
TOTAL OPERATING COST	$31,276,900	$26,055,900

TABLE 22

UNIT LIQUEFACTION COST

26.25 kg/s (2500 TPD) LIQUID H_2

$1,000,000's

		PROCESS	
		RRC	CRC
I	- TOTAL PLANT INVESTMENT	574.1	496.9
S	- STARTUP COSTS	15.8	13.7
W	- WORKING CAPITAL	18.0	17.3
C	- TOTAL CAPITAL REQUIREMENT	704.8	611.7
N	- TOTAL NET ANNUAL OPERATING COST	31.28	26.06
G	- ANNUAL LIQUID H_2 PRODUCTION, Mg/yr (MM lb/yr)	786.43 (1733.8)	786.43 (1733.8)
a	- ESCALATION FACTOR	1.00	1.00

DCF FINANCING

		RRC	CRC
UNIT COST	$/kg	0.2194	0.1890
	(¢/lb)	(9.95)	(8.58)

UTILITY FINANCING

		RRC	CRC
UNIT COST	$/kg	0.1476	0.1268
	(¢/lb)	(6.69)	(5.75)

NOTE: Does not include feedstock and liquefaction energy costs.
See text, Section 4.7.3

TABLE 23

TOTAL UNIT COST OF LIQUID H_2

	PROCESS RRC			PROCESS CRC		
Coal Cost, $/Gj	0.3320	0.4742	0.7113	0.3320	0.4742	0.7113
($/MM Btu)	(0.35)	(0.50)	(0.75)	(0.35)	(0.56)	(0.75)
DCF FINANCING						
Gasification						
Feedstock	0.2851	0.3188	0.3752	0.2851	0.3188	0.3752
Power	0.5269	0.5809	0.6707	0.4545	0.5072	0.5949
Liquefaction	0.2194	0.2194	0.2194	0.1890	0.1890	0.1890
Total, $/kg	1.0314	1.1191	1.2653	0.9286	1.0150	1.1591
(¢/lb)	(46.78)	(50.76)	(57.39)	(42.13)	(46.05)	(52.58)
$/GJ	7.26	7.88	8.91	6.54	7.15	8.16
($/MM Btu)	(7.66)	(8.31)	(9.39)	(6.90)	(7.54)	(8.61)
UTILITY FINANCING						
Gasification						
Feedstock	0.2185	0.2518	0.3073	0.2185	0.2518	0.3073
Power	0.3981	0.4513	0.5395	0.3478	0.3994	0.4857
Liquefaction	0.1476	0.1476	0.1476	0.1268	0.1268	0.1268
Total, $/kg	0.7642	0.8507	0.9944	0.6931	0.7780	0.9198
(¢/lb)	(34.66)	(38.59)	(45.10)	(31.44)	(35.29)	(41.72)
$/GJ	5.38	5.99	7.00	4.88	5.48	6.48
($ MM Btu)	(5.67)	(6.32)	(7.38)	(5.15)	(5.78)	(6.83)

NOTE: Feedstock and liquefaction energy costs included under Feedstock and Power Gasification

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TABLE 24

RECIPROCATING RECYCLE COMPRESSORS

COST OF SPARES

No. of compressors required - Main Recycle	70
No. of spare compressors	5
H_2 Liquefaction complex, no spares	
Total plant investment (Table 20)	$574,100,000
Additional Investment for spares	13,269,000
Total Plant Investment, with Spares	$587,369,000
Interest during construction	99,119,000
Startup	16,153,000
Working Capital	18,117,000
Total Capital Requirement	$720,758,000

UNIT COST OF H_2 LIQUEFACTION

			FINANCING			
	UNITS		DCF		UTILITY	
	SI	Customary	SI	Customary	SI	Customary
With Spares	$/kg	¢/lb	0.2233	10.13	0.1499	6.80
Without Spares	$/kg	¢/lb	0.2194	9.95	0.1476	6.69
Cost of 5 Spares	$/kg	¢/lb	0.0039	0.18	0.0024	0.11

TABLE 25

THERMAL EFFICIENCY FOR PRODUCING LIQUID H_2 USING RECIPROCATING RECYCLE COMPRESSORS - PROCESS RRC

			ENERGY - GJ/s	
FEEDSTOCK GASIFIER		Quantity	HHV	LHV
Input:	Coal, kg/s	193.8	5.631	5.405
	Mechanical Energy, kWm	241,300	0.241	0.241
	Electrical Energy, kW_e	67,800	0.068	0.068
	Fuel Gas, g mole/s	5,32[illegible]	1.397	1.316
	Total In		7.337	7.030
Output:	H_2 Feedstock, g moles/s	15,540	4.302	3.646
POWER GASIFIER				
Input:	Coal, kg/s	308.9	8.977	8.618
	Mechanical Energy, kWm	304,200	0.304	0.304
	Electrical Energy, kW_e	60,800	0.061	0.061
	Total In		9.342	8.983
Output:	Fuel Gas, g mole/s	25,680	6.741	6.348
H_2 LIQUEFIER				
Input:	H_2 Feedstock, g mole/s	15,540	4.302	3.646
	Electrical Energy, kW_e	1,233,540	1.234	1.234
	Total In		5.536	4.880
Output:	Liquid H_2, g mole/s	13,030	3.731	3.151
	Tail Gas, g mole/s	871	0.164	0.147
	Total Out		3.895	3.298
ENERGY CONVERSION				
Input:	Fuel Gas, g mole/s	20,356	5.344	5.032
Output:	Mechanical Energy, kWm	545,550	0.546	0.546
	Electrical Energy, kW_e	1,362,140	1.362	1.362
	Total Out		1.908	1.908
THERMAL EFFICIENCY - %				
Feedstock Gasifier			58.63	51.86
Power Gasifier			72.16	70.67
H_2 Liquefier			70.36	67.58
Energy Conversion			35.70	37.92
Overall			25.54	22.47

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TABLE 26

THERMAL EFFICIENCY FOR PRODUCING LIQUID H_2 USING CENTRIFUGAL RECYCLE COMPRESSORS - PROCESS CRC

			ENERGY - GJ/s	
		Quantity	HHV	LHV
FEEDSTOCK GASIFIER				
Input:	Coal, kg/s	193.8	5.631	5.405
	Mechanical Energy, kWm	241,300	0.241	0.241
	Electrical Energy, kW_e	67,800	0.068	0.068
	Fuel Gas, g mole/s	5,320	1.397	1.316
	Total In		7.337	7.030
Output:	H_2 Feedstock, g mole/s	15,540	4.302	3.646
	Process Heat		0.544	0.544
	Total Out		4.846	4.190
POWER GASIFIER				
Input:	Coal, kg/s	301.6	8.764	8.414
	Mechanical Energy, kWm	297,300	0.297	0.297
	Electrical Energy, kW_e	59,340	0.059	0.059
	Total In		9.120	8.770
Output:	Fuel Gas, g mole/s	25,092	6.588	6.204
H_2 LIQUEFIER				
Input:	H_2 Feedstock, g mole/s	15,540	4.302	3.646
	Mechanical Energy, kWm	1,134,900	1.135	1.135
	Electrical Energy, kW_e	119,210	0.119	0.119
	Total In		5.556	4.900
Output:	Liquid H_2, g mole/s	13,030	3.731	3.151
	Tail Gas, g mole/s	871	0.164	0.147
	Total Out		3.895	3.298

TABLE 26 (Cont'd)

ENERGY CONVERSION				
Input:	Fuel Gas, g mole/s	19,760	5.188	4.886
	Process Heat		0.544	0.544
	Total In		5.732	5.430
Output:	Mechanical Energy, kWm	1,673,000	1.673	1.673
	Electrical Energy, kW_e	246,400	0.246	0.246
	Total Out		1.919	1.919
THERMAL EFFICIENCY - %				
Feedstock Gasifier			66.05	59.60
Power Gasifier			72.24	70.74
H_2 Liquefier			70.10	67.31
Energy Conversion			33.48	35.34
	Overall		25.92	22.80

TABLE 27

HEATING VALUES OF PROCESS STREAMS

	Units		Heating Value			
	SI	Customary	SI		CUSTOMARY	
			HHV	LHV	HHV	LHV
Coal	kJ/g	Btu/lb	29.06	27.89	12,500	12,000
Fuel Gas	kJ/g mole	Btu/SCF	263	247	292	275
Crude H_2	kJ/g mole	Btu/SCF	277	235	308	261
Product H_2	kJ/g mole	Btu/SCF	286	242	318.5	269
Tail gas	kJ/g mole	Btu/SCF	188	168	209	187

TABLE 28

LIQUID PHASE ACTIVITY COEFFICIENTS*

FOR H_2-HD and H_2-D_2 SYSTEMS

% H_2 in Liquid	H_2-HD System		H_2-D_2 System	
	$\log_{10}\gamma_{H_2}$	$\log_{10}\gamma_{HD}$	$\log_{10}\gamma_{H_2}$	$\log_{10}\gamma_{D_2}$
0	0.0145	0	0.038	0
20	0.0100	0.0010	0.025	0.0017
40	0.0058	0.0034	0.014	0.0064
60	0.0024	0.0073	0.006	0.0160
80	0.0006	0.0130	0.0015	0.029
100	0	0.0198	0	0.045

* Data of Newman (18)

TABLE 29

INVESTMENT SUMMARY

DEUTERIUM RECOVERY FACILITY

For 10-Module 26.25 kg/s (2500 TPD) H_2 Liquefier

Total Plant Investment	$63,750,000
Interest During Construction	10,758,000
Startup Costs	1,753,000
Working Capital	4,895,000
TOTAL CAPITAL REQUIREMENT	$81,156,000

TABLE 30

INCREMENTAL INVESTMENT
POWER GASIFIER SECTION
FOR DEUTERIUM RECOVERY

BASIS: Generation of 28.5 m^3/s (3.620 MM CFH) Additional Fuel Gas
Requiring 14.2 kg/s (1350 TPD) Additional Coal

Incremental Investment		$31,660,000
Interest During Construction		5,343,000
Startup Costs	(a)	1,231,000
	(b)	1,596,000
	(c)	2,204,000
Working Capital	(a)	965,000
	(b)	1,257,000
	(c)	1,743,000
Total Incremental Capital	(a)	39,199,000
	(b)	39,856,000
	(c)	40,950,000

COST OF COAL:		$/GJ	($/MM Btu)
	(a)	0.3320	(0.35)
	(b)	0.4742	(0.50)
	(c)	0.7113	(0.75)

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TABLE 31

POWER REQUIREMENTS

DEUTERIUM RECOVERY FACILITY

BASIS: 10-Module 26.25 kg/s (2500 TPD) H_2 Liquefier

HYDROGEN LIQUEFIER POWER REQUIRED, kW	DEUTERIUM RECOVERY WITH	DEUTERIUM RECOVERY WITHOUT
Flash Compressor	10,260	215,700
Main Recycle		
165-290 kPa	105,290	N.A.
290-393 kPa	8,930	N.A.
290-4137 kPa	594,870	587,280
HD Compressor	20	N.A.
TOTAL	719,370	608,850
Difference, kW	110,500	

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TABLE 32

INCREMENTAL ANNUAL OPERATING COST

POWER GASIFIER SECTION

FOR DEUTERIUM RECOVERY

Incremental Coal Consumption: 1.42 kg/s (135 TPD)

Coal:	(a)	$913,600
	(b)	609,100
	(c)	426,400
Catalysts & Chemicals		2,000
Process Water		3,000
Labor, Maintenance		47,800
Supplies		51,600
Local Taxes and Insurance		86,000
Subtotal	(a)	$1,104,000
	(b)	799,500
	(c)	616,800
By-Product Sulfur (Credit)		40,300
Net Operating Cost	(a)	$1,063,700
	(b)	759,200
	(c)	$ 576,500

Coal Cost, $/GJ ($/MM Btu):	(a)	0.3320	(0.35)
	(b)	0.4742	(0.50)
	(c)	0.7113	(0.75)

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TABLE 33

INCREMENTAL COST OF FUEL GAS

POWER GASIFIER SECTION

FOR DEUTERIUM RECOVERY

BASIS: 28.5 m^3/s (3.62 x 10^6 cfh) Incremental Fuel Gas Production

Cost of Coal:			
$/Gj	0.3320	0.4742	0.7113
($/MM Btu)	(0.35)	(0.50)	(0.75)
I - $MM	31.660	31.660	31.660
S - $MM	1.231	1.596	2.204
W - $MM	0.965	1.[illegible]7	1.743
C - $MM	39.199	39.856	40.950
N - $MM	5.750	7.575	10.618

Unit Cost				
DCF Financing:	$/GJ	1.464	1.673	2.022
	($/MM Btu)	(1.544)	(1.764)	(2.132)
Utility Financing:	$/GJ	1.128	1.333	1.676
	($/MM Btu)	(1.189)	(1.406)	(1.768)

Where

I - Total Incremental Plant Investment

S - Incremental Startup Costs

W - Incremental Working Capital

C - Total Incremental Capital Requirement

N - Net Incremental Operating Cost

G - Annual Incremental Fuel Heating Value,
9273 TJ (8.795 x 10^{12} Btu)

Unit Costs computed by equations in Table 14.

TABLE 34

UNIT COST OF HEAVY WATER

DCF FINANCING

Operating Cost	$MM		
Cost of Coal: $/GJ	0.3320	0.4742	0.7113
($/MM Btu)	(0.35)	(0.50)	(0.75)
Fuel Gas	13.58	15.51	18.75
Loss of LH_2 Product	0.10	0.11	0.12
Oxygen for D_2 Combustion	0.05	0.05	0.05
Labor	0.30	0.30	0.30
Administration & Overhead	0.18	0.18	0.18
Supplies, Operating	0.08	0.08	0.08
Maintenance	0.96	0.96	0.96
Local Taxes & Insurance	1.72	1.72	1.72
Total	16.97	18.91	22.16
Total Plant Investment	63.75	63.75	63.75
Startup Cost	1.75	1.75	1.75
Working Capital	4.89	4.89	4.89
Unit Cost of D_2O, $/kg	31.64	33.47	36.55
($/lb)	(14.35)	(15.18)	(16.58

G = Annual Production of Heavy Water
= 1.053 Gg (1161 Tons)

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TABLE 35

UNIT COST OF HEAVY WATER

UTILITY FINANCING

Operating Cost	$MM		
Cost of Coal: $/GJ	0.3320	0.4742	0.7113
($/MM Btu)	(0.35)	(0.50)	(0.75)
Fuel Gas	10.45	12.36	15.55
Loss of LH_2 Product	0.08	0.09	0.10
Oxygen for D_2 Combustion	0.05	0.05	0.05
Labor	0.30	0.30	0.30
Administration & Overhead	0.18	0.18	0.18
Supplies, Operating	0.08	0.08	0.08
Maintenance	0.96	0.96	0.96
Local Taxes & Insurance	1.72	1.72	1.72
Total	13.82	15.74	18.94
Total Capital Requirement	81.15	81.15	81.15
Working Capital	4.89	4.89	4.89
Unit Cost of D_2O, $/kg	22.44	24.27	27.30
($/lb)	(10.18)	(11.01)	(12.38)

G = Annual Production of Heavy Water
= 1.053 Gg (1161 Tons)

TABLE 36

ECONOMIC SUMMARY

IMPACT OF HEAVY WATER PRODUCTION

BASIS: 1.053 Gg (1161 T) D_2O Annually - From 26.25 kg/s (2500 TPD) LH_2

		TOTAL PLANT INVESTMENT $ MILLIONS	
Power Gasifier (Incremental)		31.660	
Deuterium Recovery & H_2 Liquefier		63.750	
	Total	95.410	
Cost of Coal: $/GJ	0.3320	0.4742	0.7113
($/MM Btu)	(0.35)	(0.50)	(0.75)
Income @ $121.25/kg ($55/lb) D_2O	127.70	127.70	127.70
Annual Cost of D_2O, $MM			
DCF Financing	33.32	35.25	38.50
Utility Financing	23.64	25.56	28.74
Annual Cost of LH_2, $MM			
DCF Financing	787.1	855.1	967.8
Utility Financing	584.3	651.2	762.5
Income less Cost, $MM			
DCF Financing	94.38	92.45	89.20
Utility Financing	104.06	102.14	98.96
Income less Cost, as % of LH_2 Cost			
DCF Financing	12.0	10.8	9.2
Utility Financing	17.8	15.7	13.0

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EFFECT OF MOLECULAR WEIGHT ON DISCHARGE PRESSURE – CENTRIFUGAL COMPRESSOR

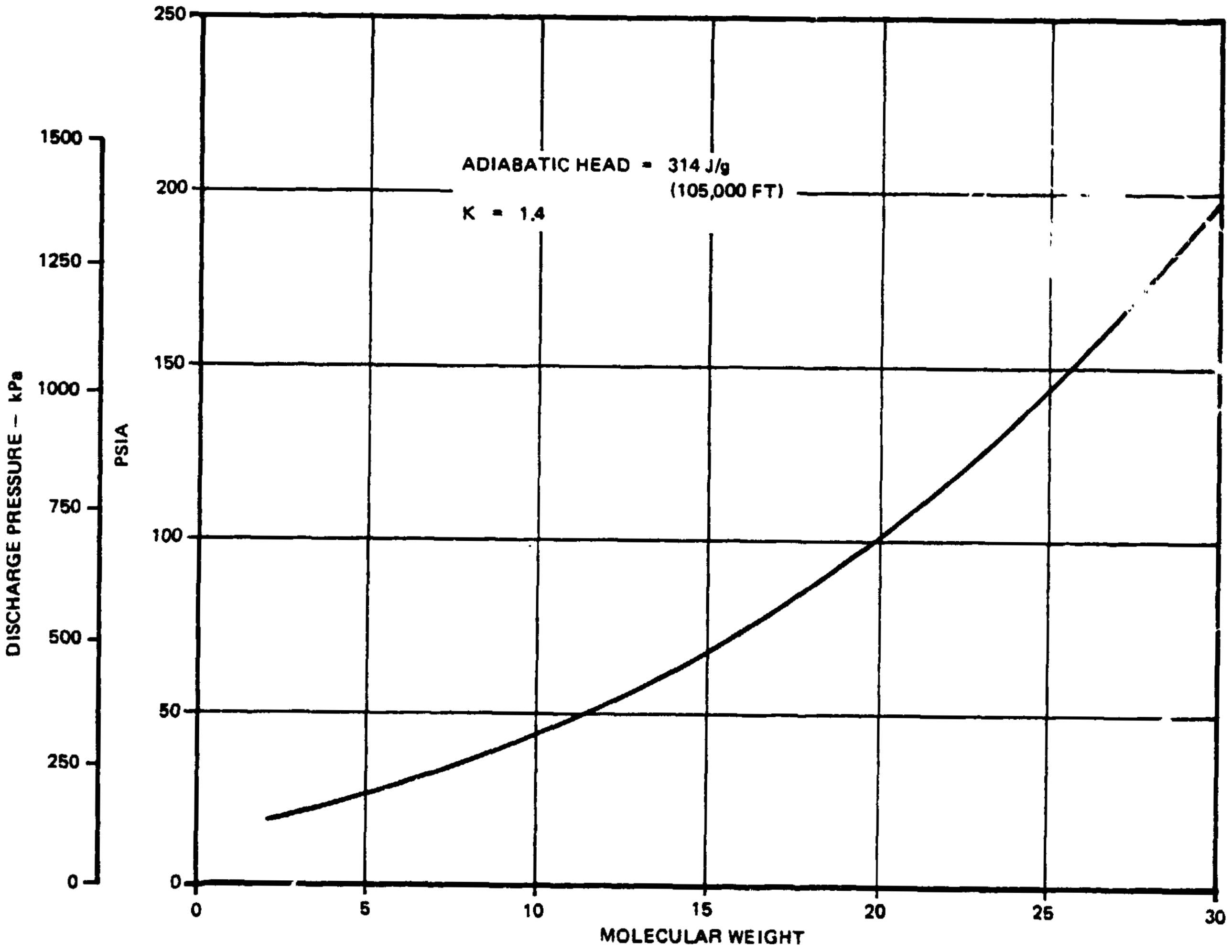

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FIGURE 1

MOLECULAR WEIGHT OF HYDROGEN PROPANE MIXTURES

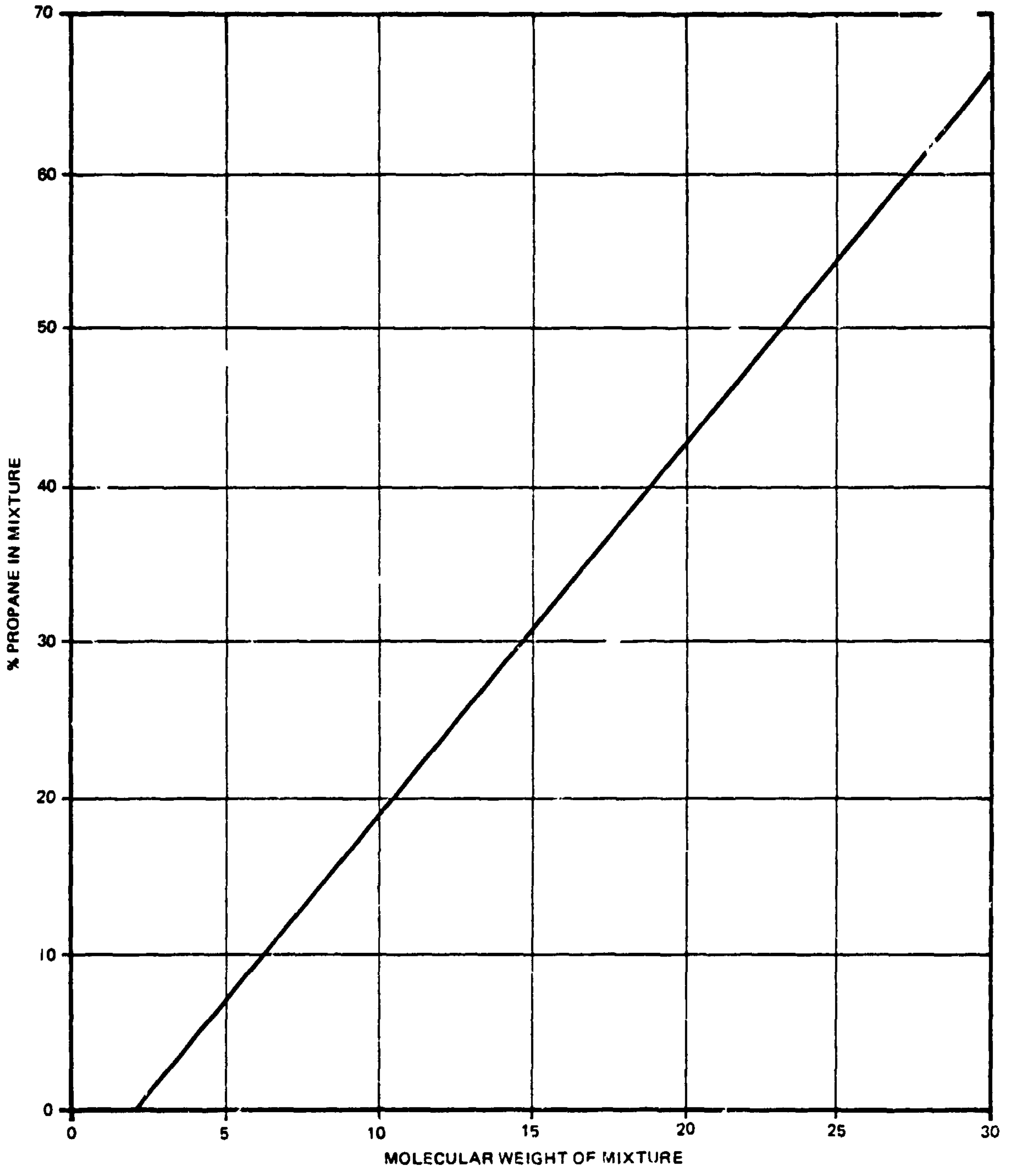

FIGURE 2

IMPELLER REQUIREMENTS
FOR
CENTRIFUGAL COMPRESSION OF
HYDROGEN – PROPANE MIXTURES

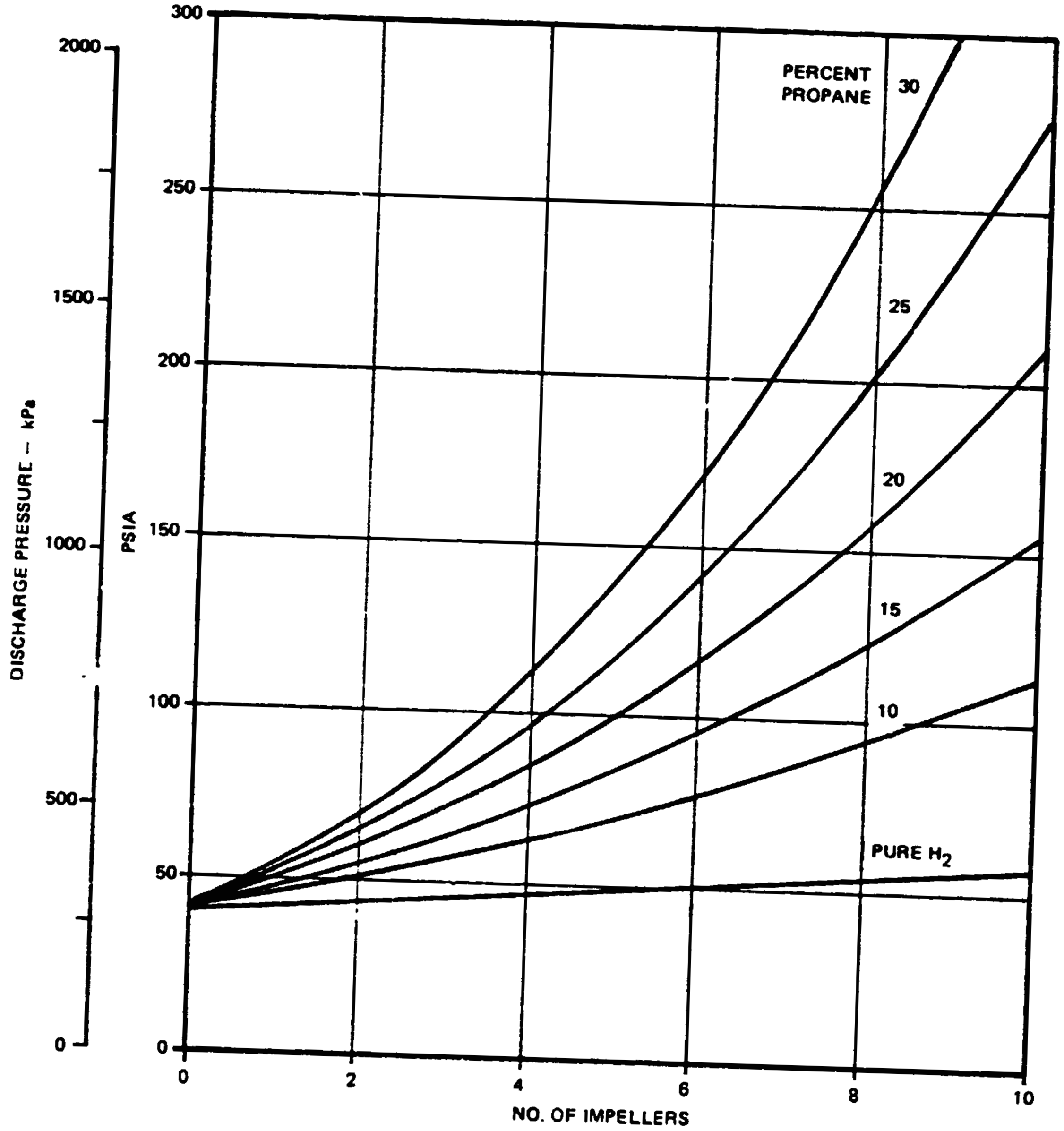

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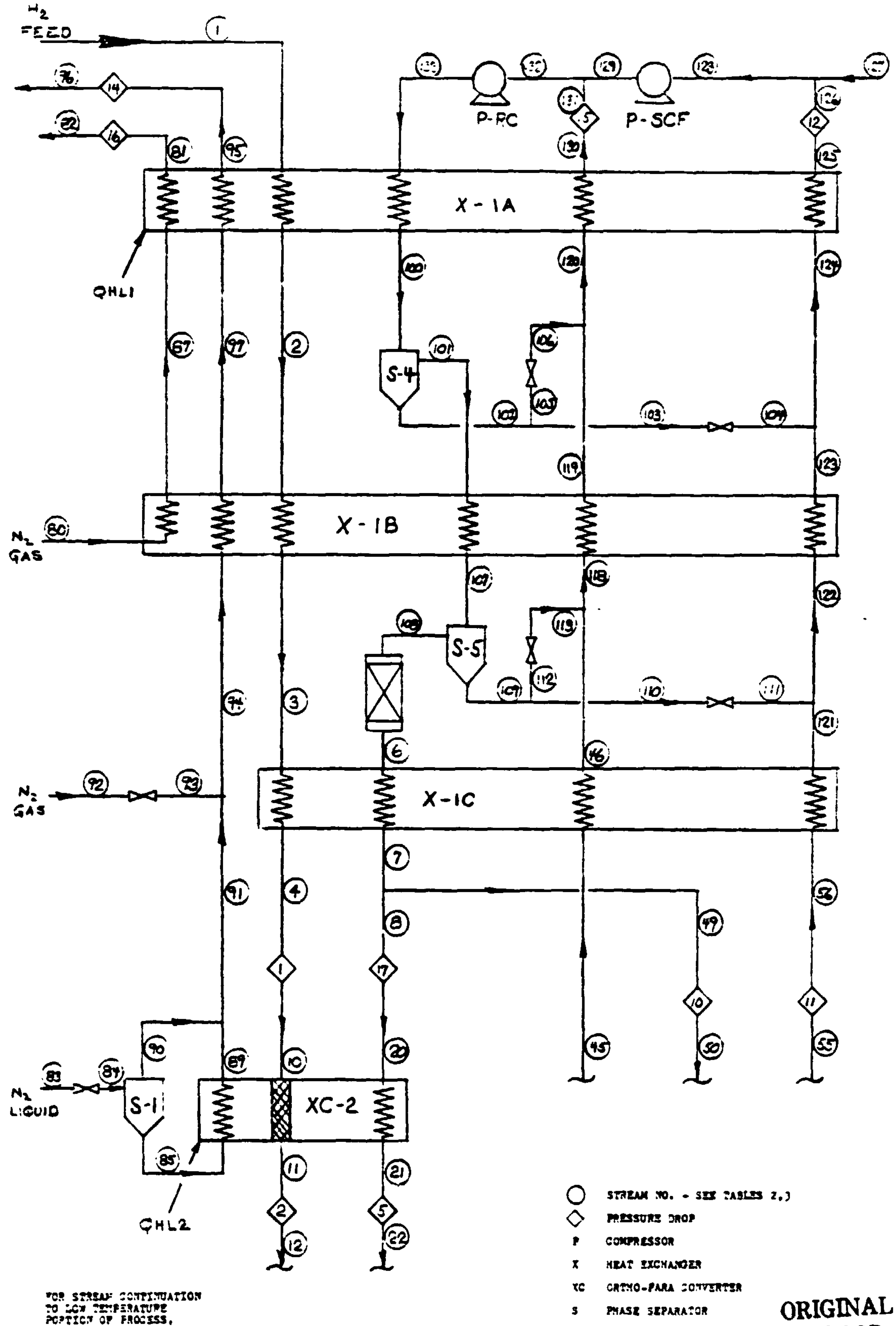
FIGURE 3



○ STREAM NO. - SEE TABLES 2,3
◇ PRESSURE DROP
P COMPRESSOR
X HEAT EXCHANGER
XC ORTHO-PARA CONVERTER
S PHASE SEPARATOR

FOR STREAM CONTINUATION TO LOW TEMPERATURE PORTION OF PROCESS, SEE FIGURE 5

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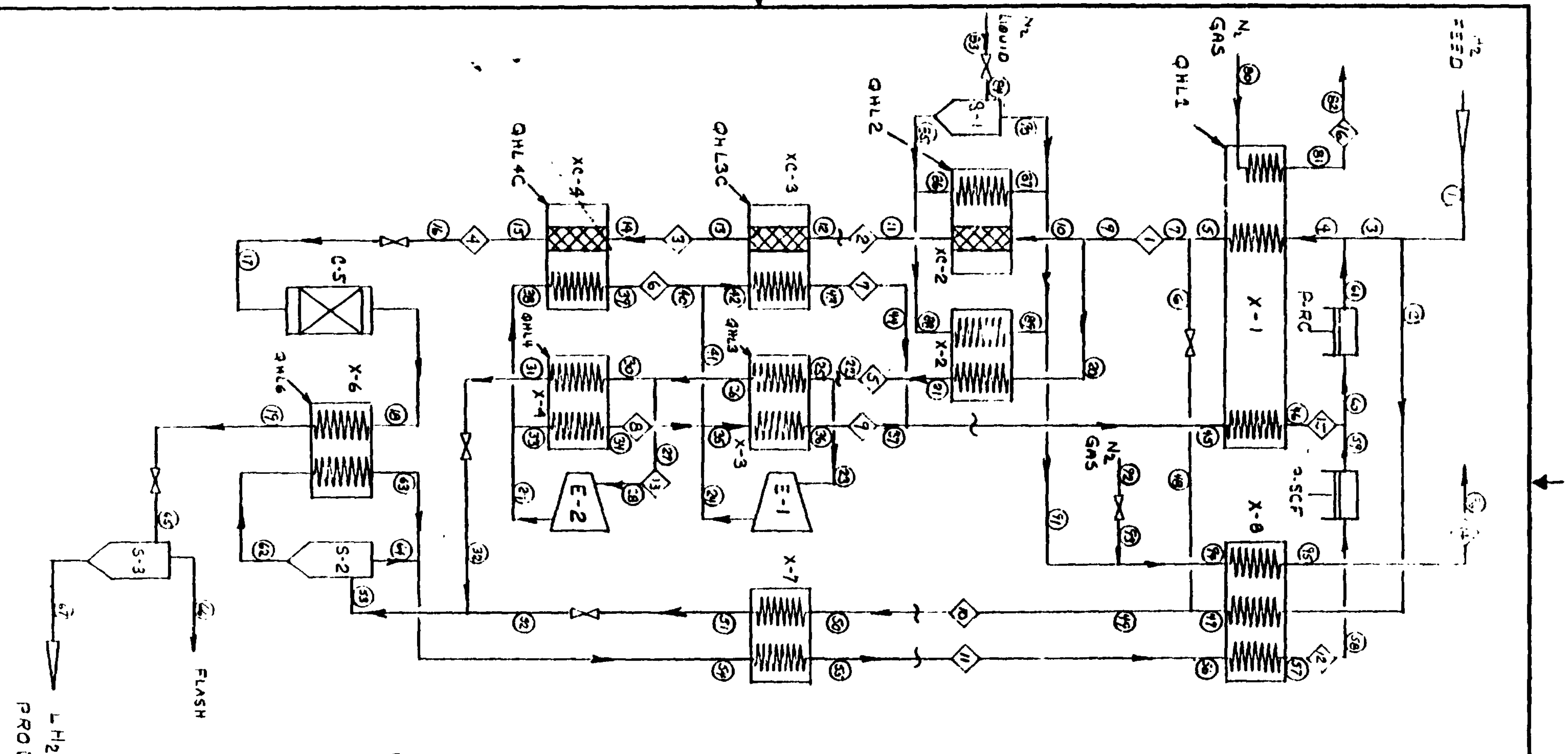
HYDROGEN LIQUEFIER PROCESS MODEL - PROCESS CRC FIGURE 4



Ⓝ STREAM No. SEE TABLES 6,7

◇ PRESSURE DROP

P COMPRESSOR

E EXPANDER

X HEAT EXCHANGER

S PHASE SEPARATOR

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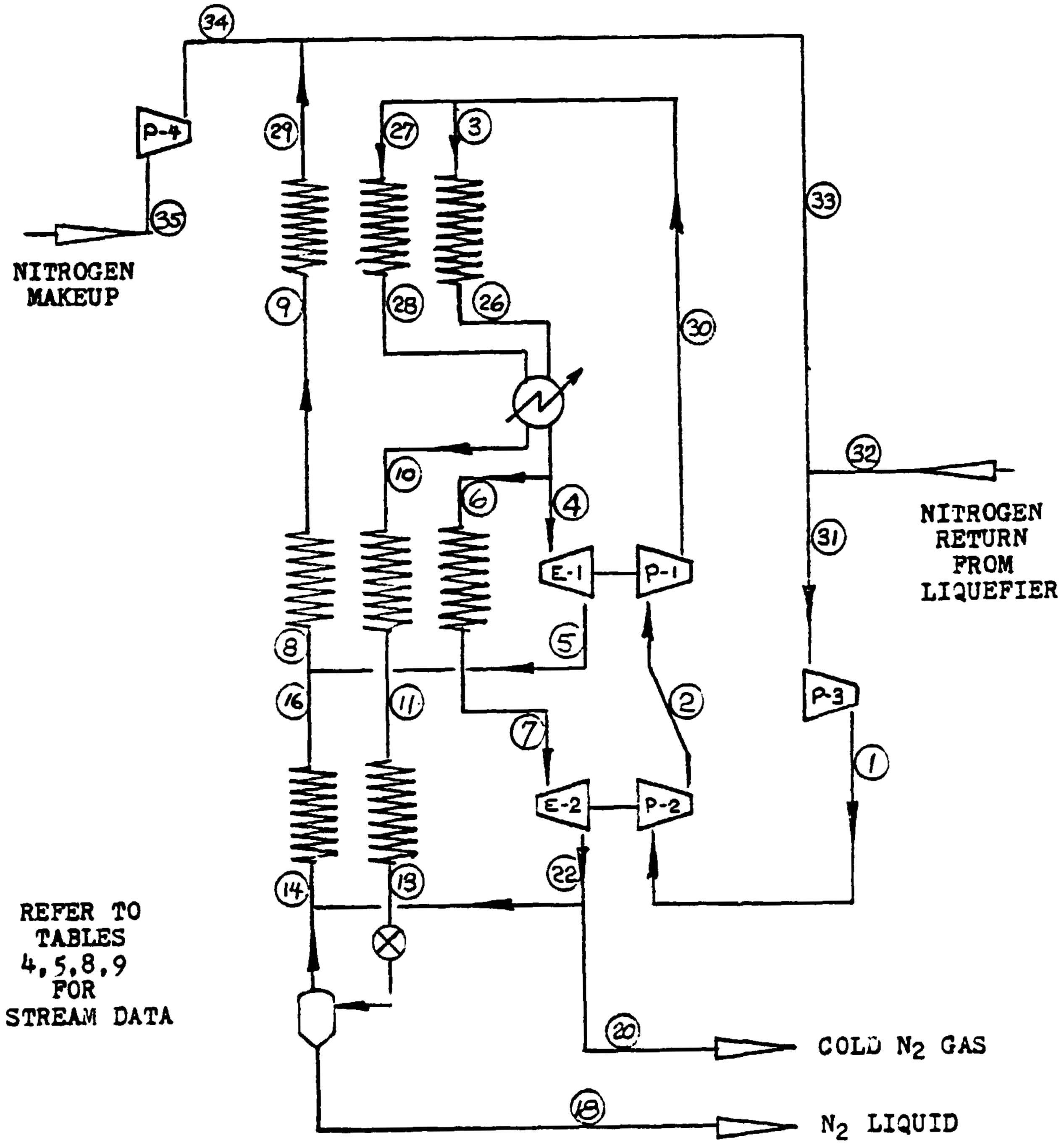
ENGINEERING DEPARTMENT • TONAWANDA, NEW YORK

H_2 LIQUEFIER PROCESS MODEL

PROCESS RRC

FIGURE 5



NITROGEN REFRIGERATOR PROCESS MODEL

FIGURE 6

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BARREL DESIGN REQUIREMENTS H_2 COMPRESSORS

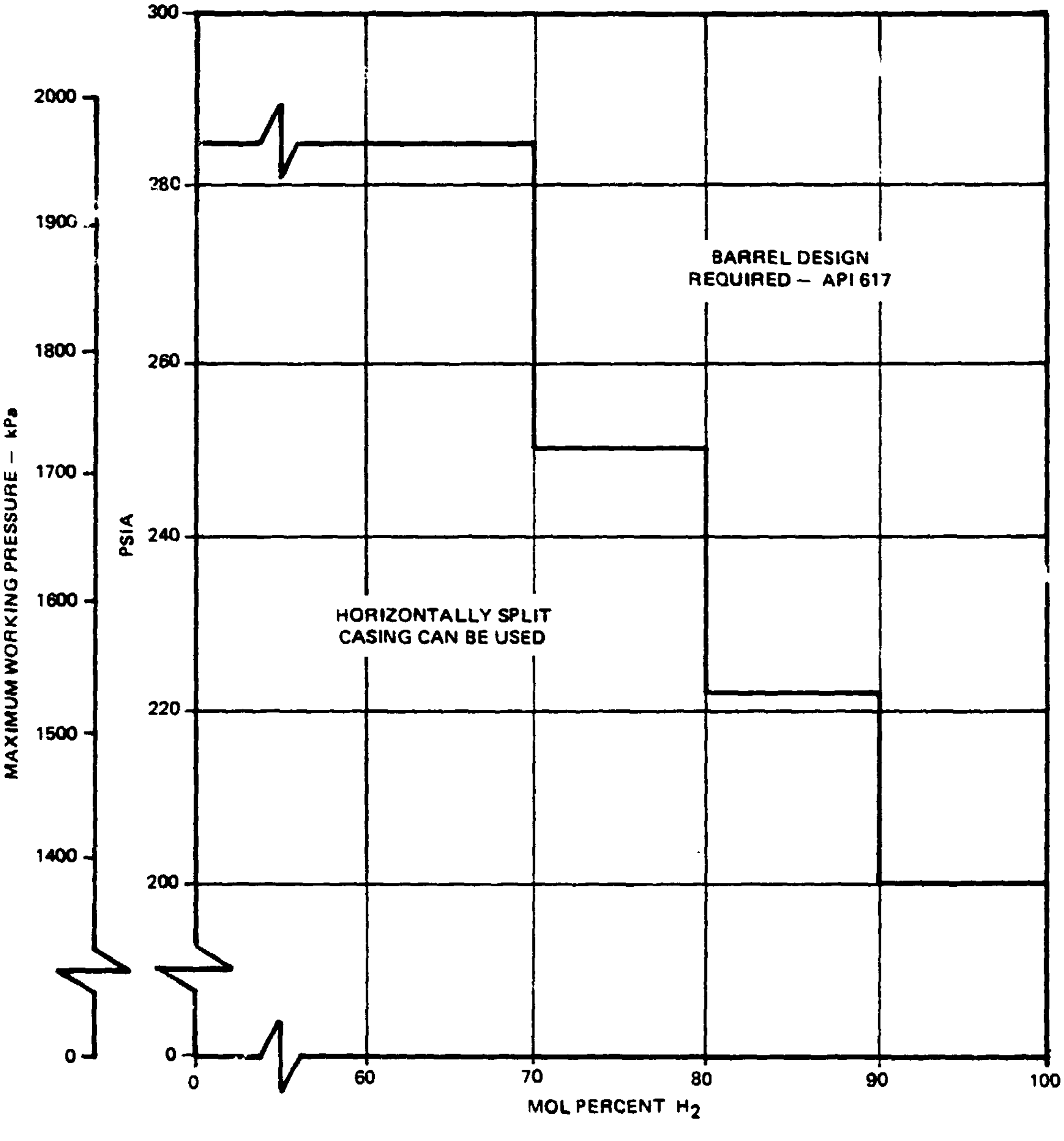
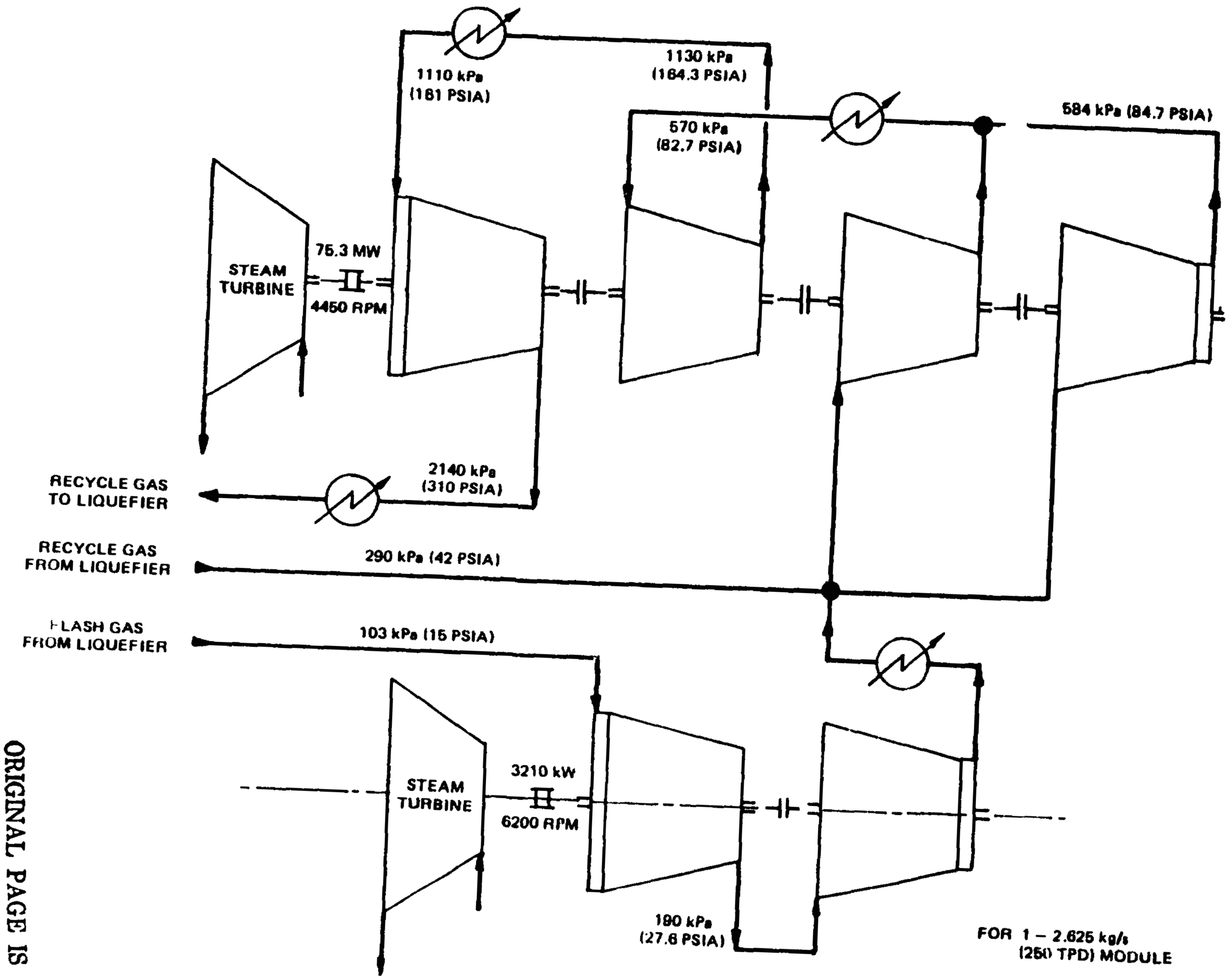


FIGURE 7



RECYCLE COMPRESSOR CONFIGURATION PROCESS CRC

FIGURE 8

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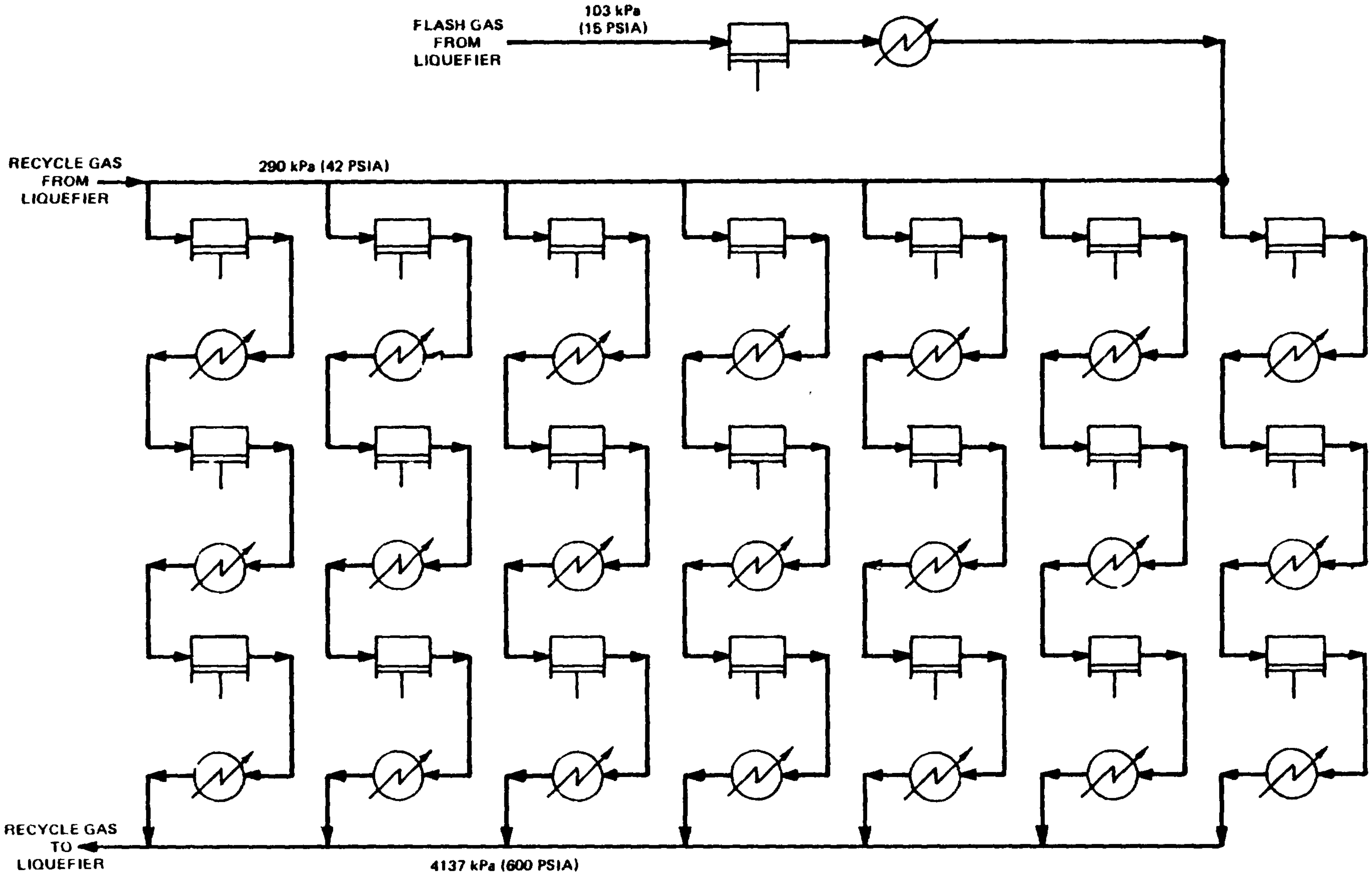
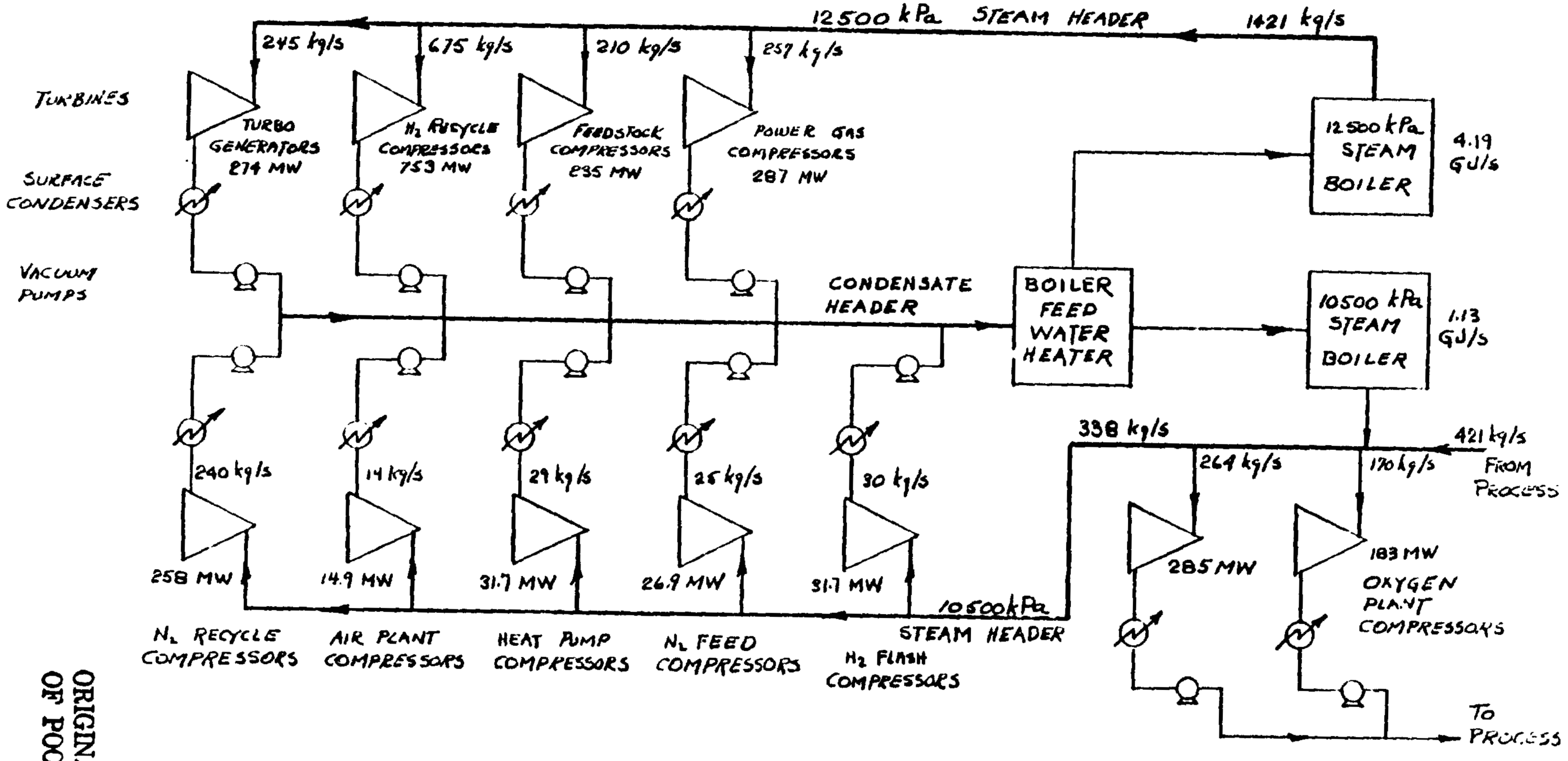


FOR 1 – 2.625 kg/s MODULE
(250 TPD)

RECIPROCATING COMPRESSOR CONFIGURATION PROCESS RRC

FIGURE 9



STEAM POWER SYSTEM - PROCESS CRC FIGURE 10

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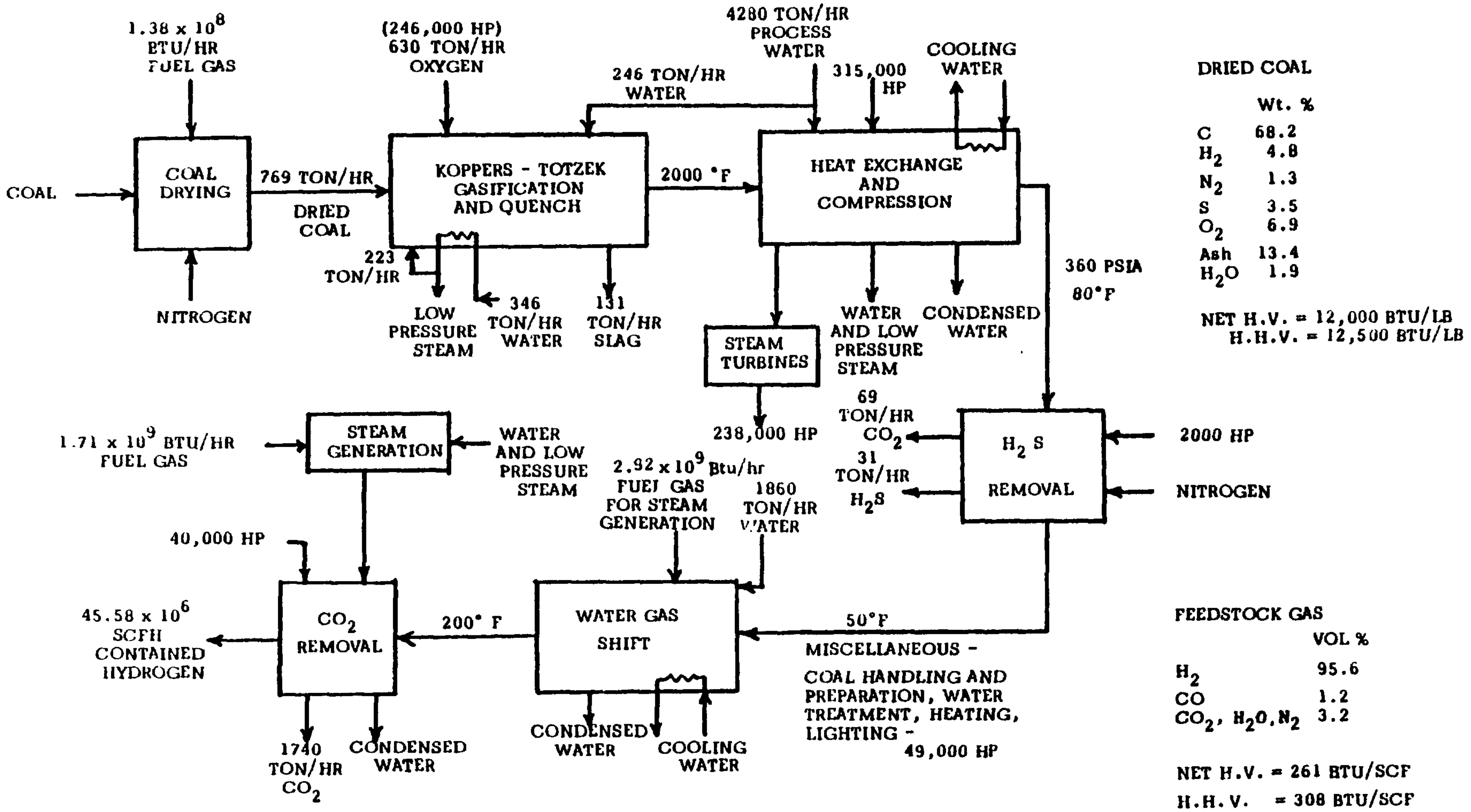


DRIED COAL

	Wt. %
C	68.2
H_2	4.8
N_2	1.3
S	3.5
O_2	6.9
Ash	13.4
H_2O	1.9

NET H.V. = 12,000 BTU/LB
H.H.V. = 12,500 BTU/LB

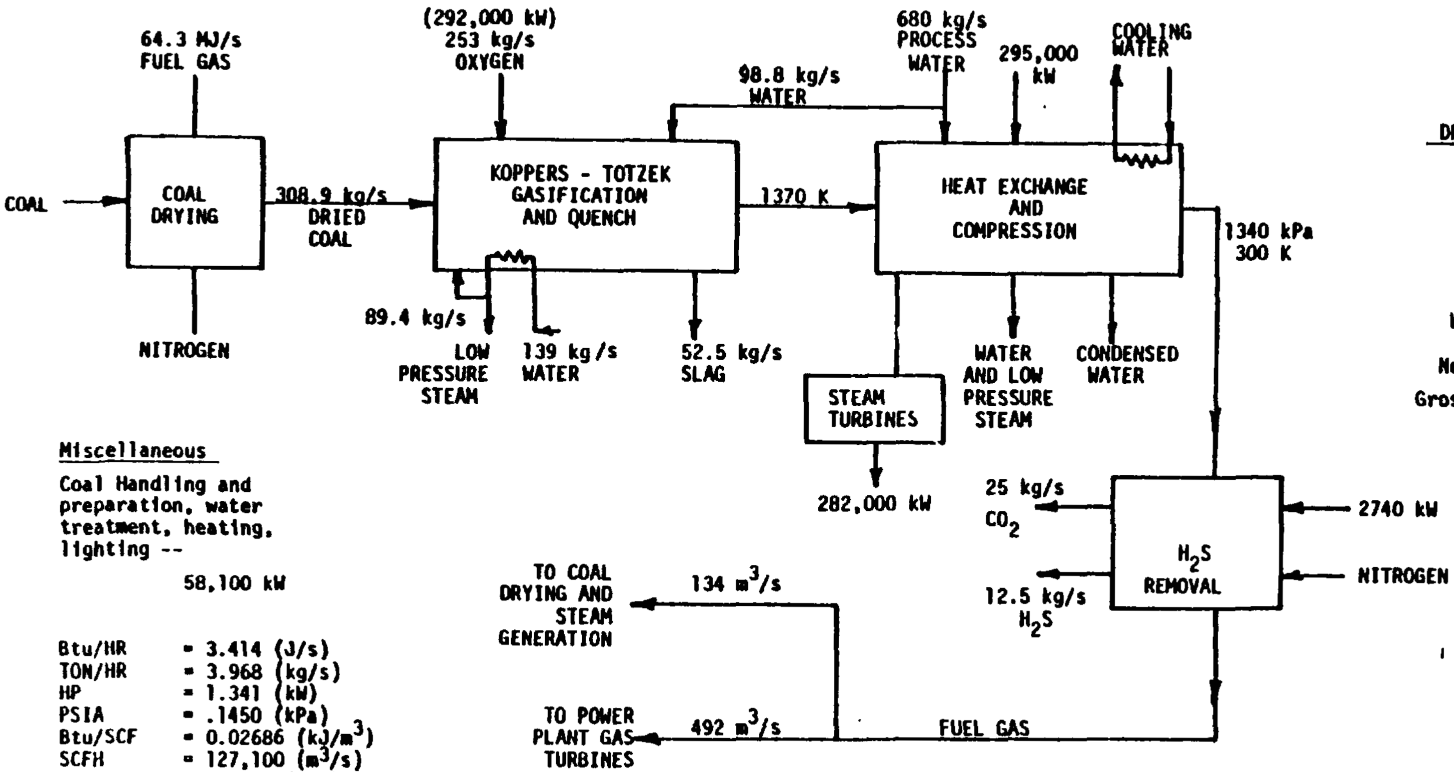
FEEDSTOCK GAS

	VOL %
H_2	95.6
CO	1.2
CO_2, H_2O, N_2	3.2

NET H.V. = 261 BTU/SCF
H.H.V. = 308 BTU/SCF

REFER TO FIGURE 12
FOR CONVERSION FACTORS

FIGURE 11 FEEDSTOCK GASIFICATION



DRIED COAL ANALYSIS

	Wt. %
C	68.2
H	4.8
N	1.3
S	3.5
O	6.9
Ash	13.4
Water	;/0

Net H.V. = 27,900 J/g
Gross H.V. = 29,100 J/g

Miscellaneous

Coal Handling and preparation, water treatment, heating, lighting --

58,100 kW

Btu/HR = 3.414 (J/s)
TON/HR = 3.968 (kg/s)
HP = 1.341 (kW)
PSIA = .1450 (kPa)
Btu/SCF = 0.02686 (kJ/m^3)
SCFH = 127,100 (m^3/s)
Btu/LB = 0.4302 (J/g)

FUEL GAS ANALYSIS

	Vol. %
CO	57.2
H_2	35.3
H_2S	500 ppm
CO_2, N_2	7.5
H_2O	

Net H.V. = 10,200 kJ/m^3
Gross H.V. = 10,900 kJ/m^3

FIGURE 12 POWER PLANT GASIFICATION - PROCESS RRC

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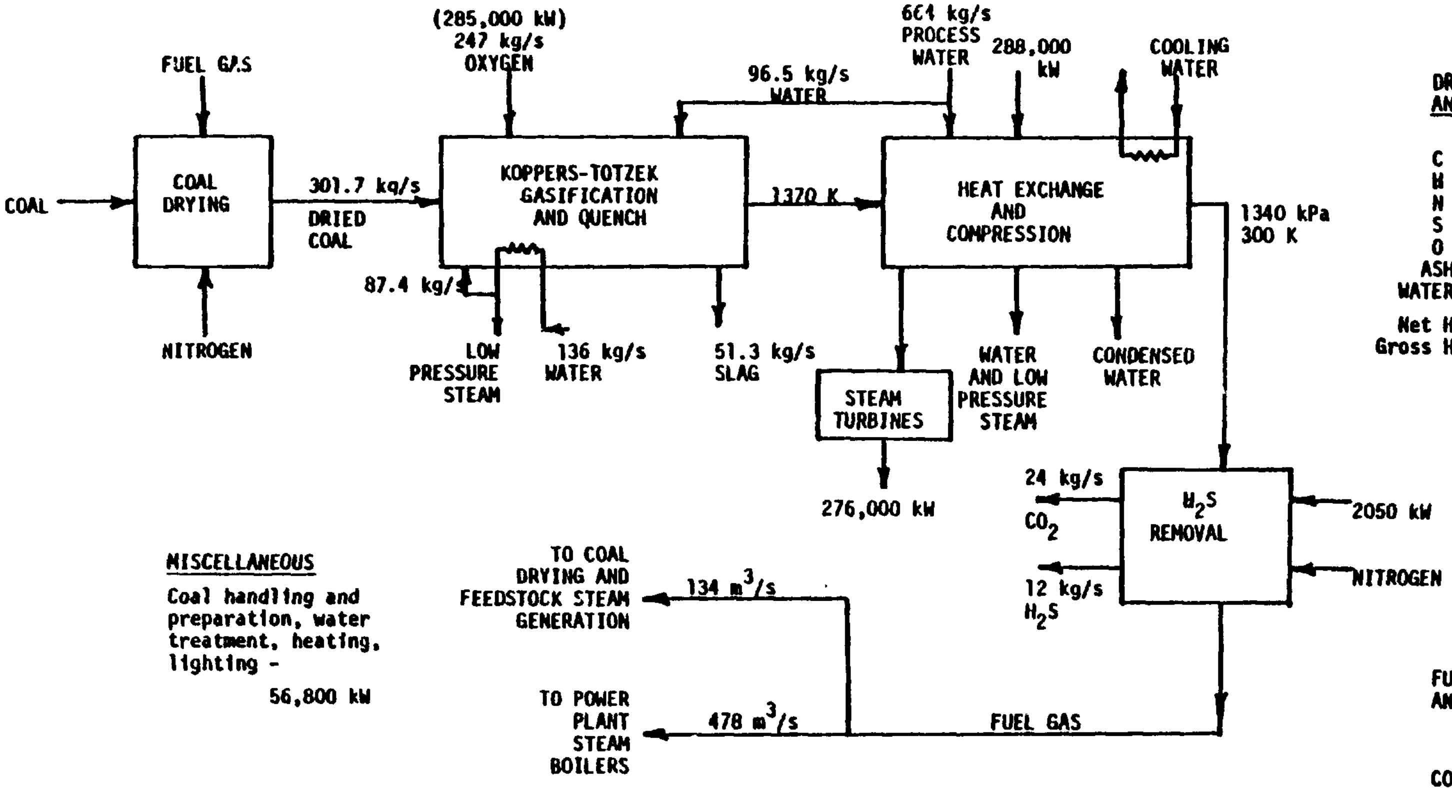


MISCELLANEOUS

Coal handling and preparation, water treatment, heating, lighting -

56,800 kW

DRIED COAL ANALYSIS

	Wt %
C	68.2
H	4.8
N	1.3
S	3.5
O	6.9
ASH	13.4
WATER	1.9

Net H.V. = 27,900 J/g
Gross H.V. = 29,100 J/g

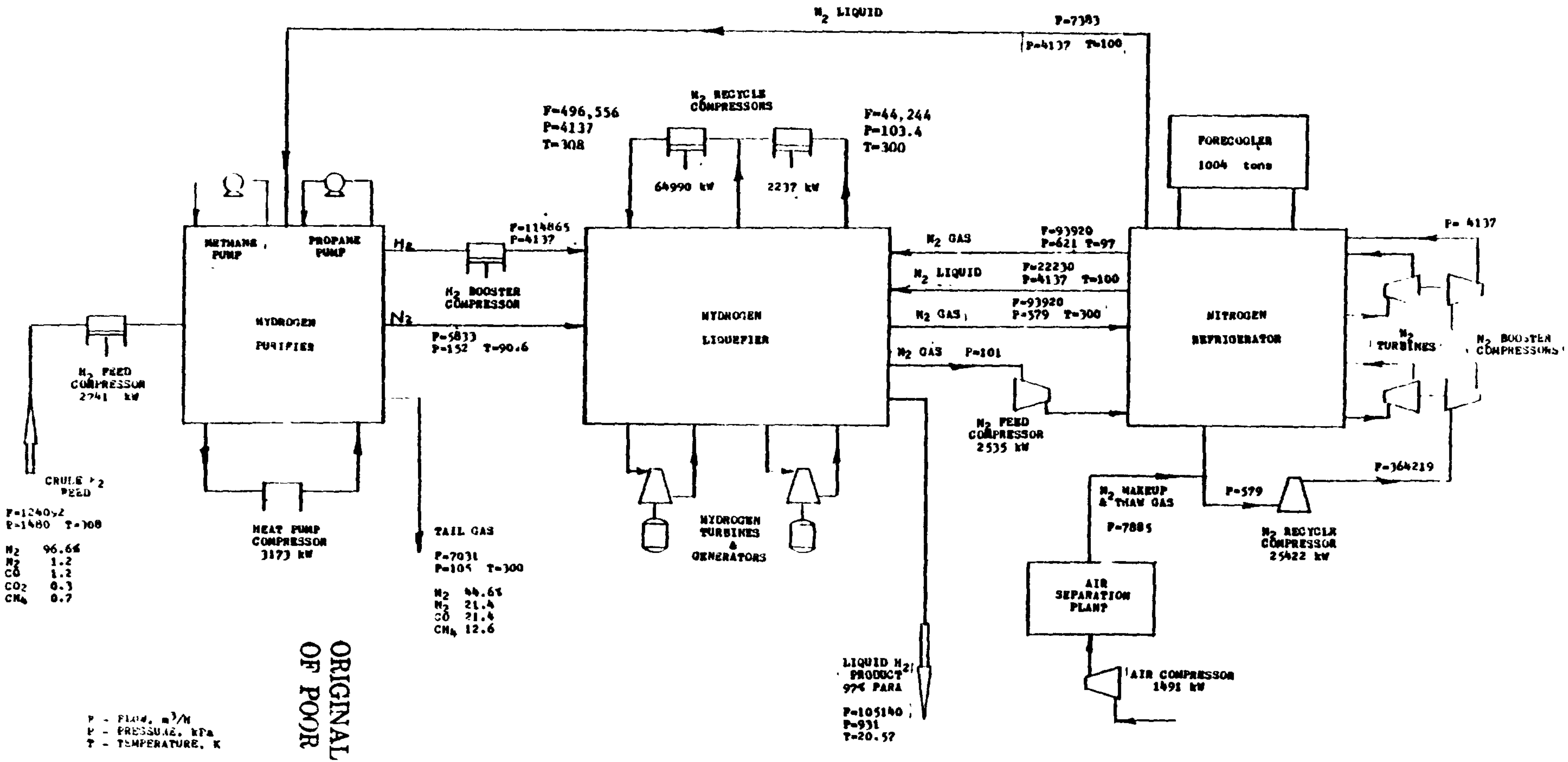
FUEL GAS ANALYSIS

	Vol. %
CO	57.2
H_2	35.3
H_2S	500 ppm
CO_2, N_2 H_2	7.5

Net H.V. = 10,200 kJ/m^3
Gross H.V. = 10,900 kJ/m^3

FIGURE 13 POWER PLANT GASIFICATION - PROCESS CRC

REFER TO FIGURE 12
FOR CONVERSION FACTORS



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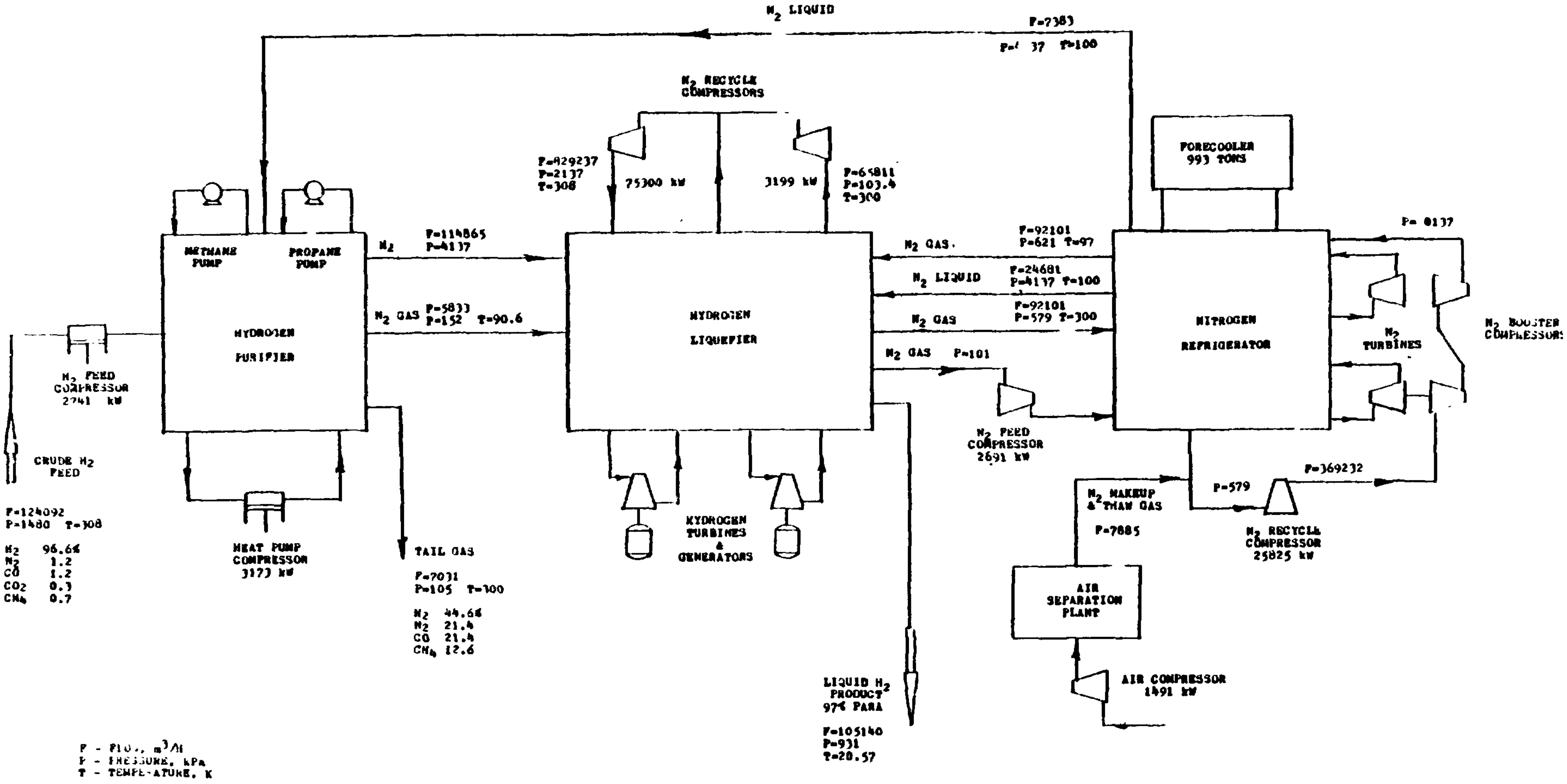
BLOCK FLOW DIAGRAM - HYDROGEN LIQUEFACTION COMPLEX
PROCESS RRC - 2.625 kg/s (250 TPD) MODULE

FIGURE 14



BLOCK FLOW DIAGRAM - HYDROGEN LIQUEFACTION COMPLEX
PROCESS CRC - 2.625 kg/s (250 TPD) MODULE

FIGURE 15

SUMMARY OF TOTAL
PLANT INVESTMENT
RRC & CRC PROCESSES

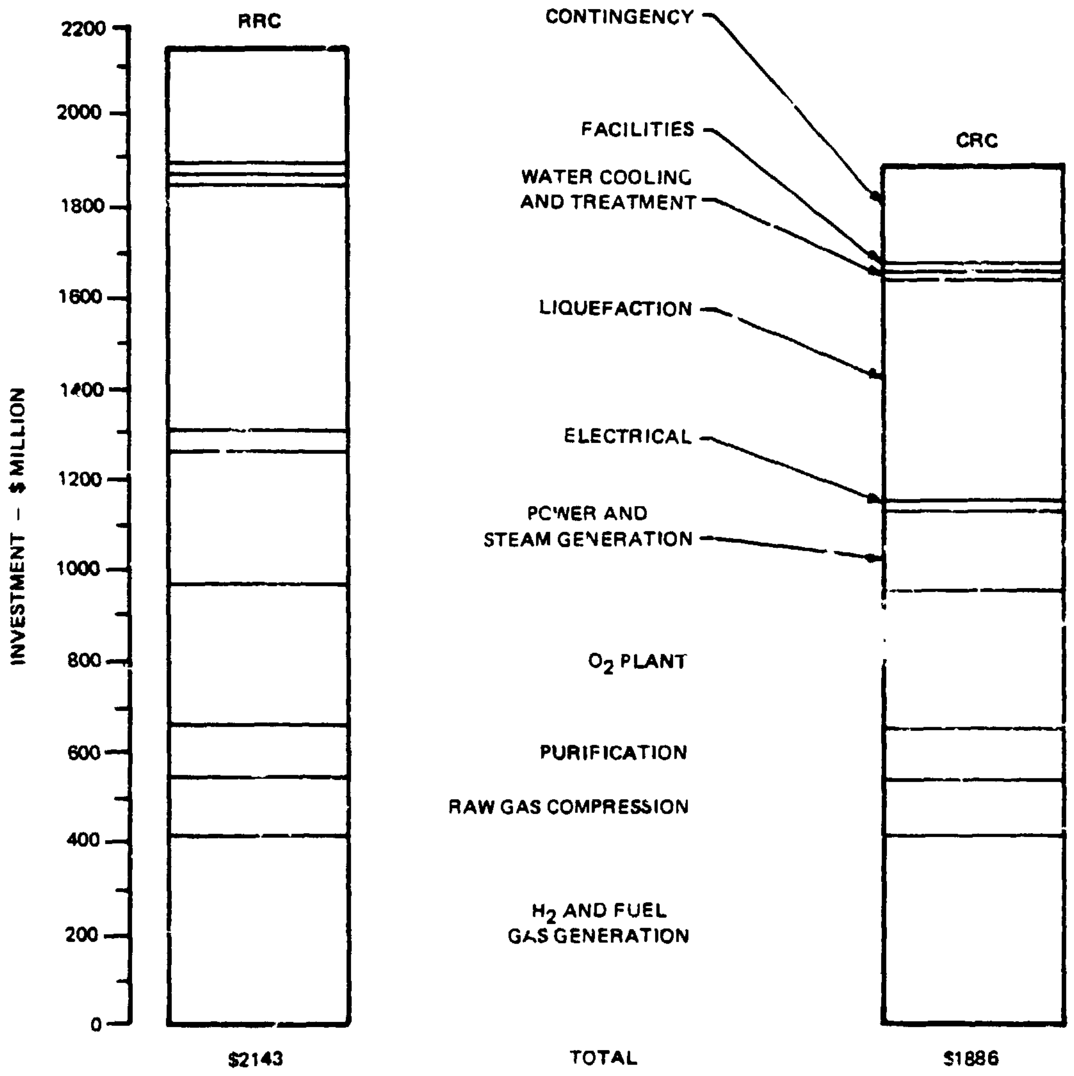


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FIGURE 16

TOTAL UNIT COST OF
LIQUID H_2
RECIPROCATING VS CENTRIFUGAL
RECYCLE COMPRESSORS

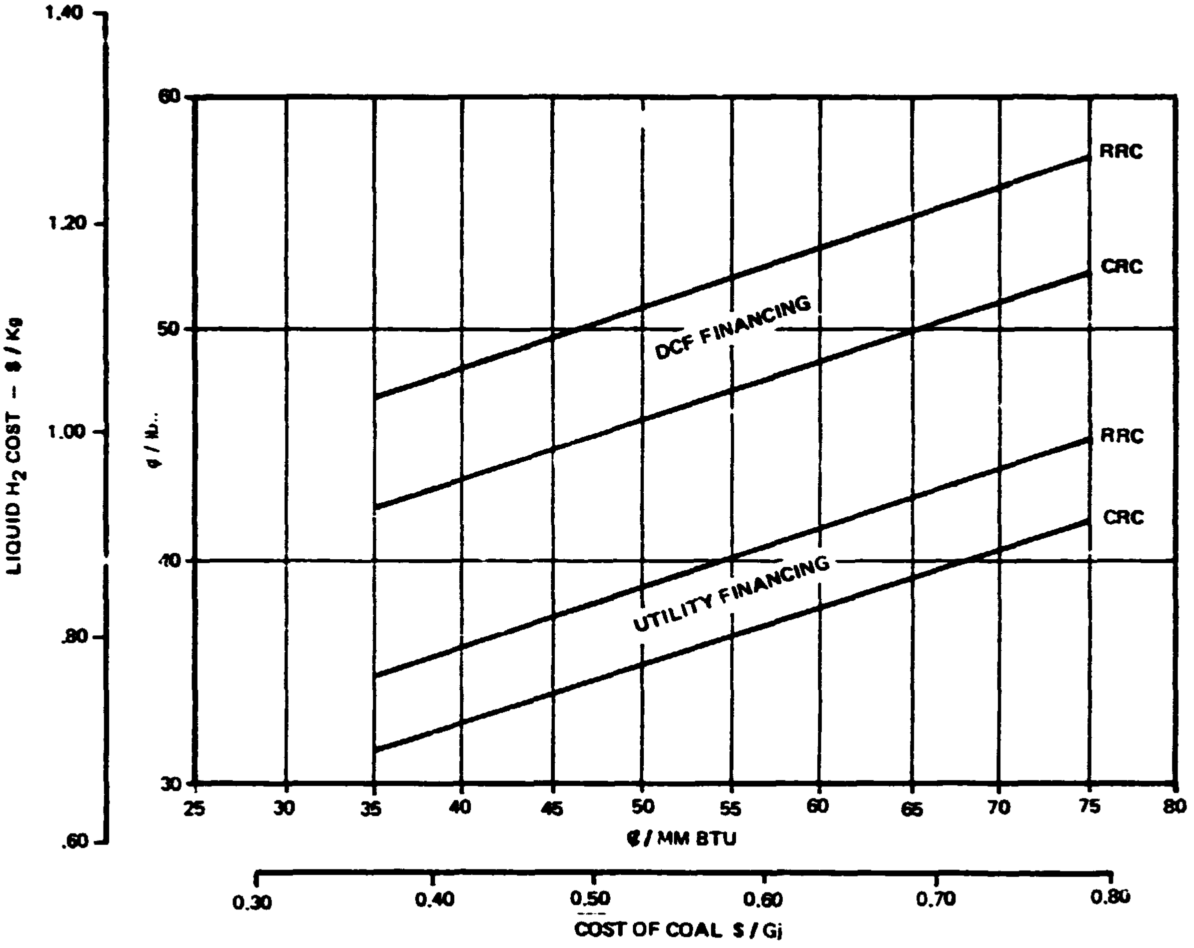

FIGURE 17

DISTRIBUTION OF PROCESS ENERGY

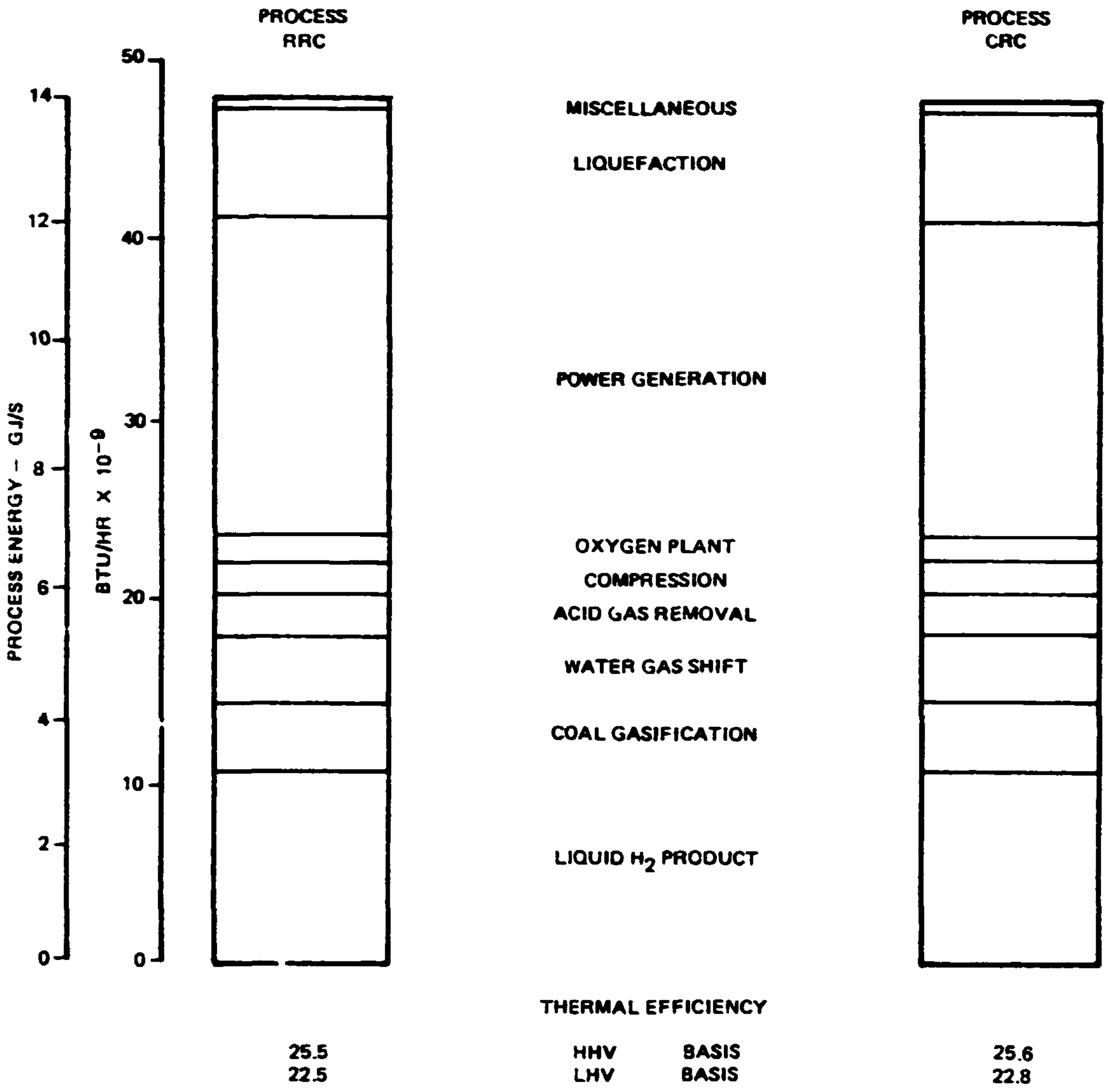

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FIGURE 18

CHEMICAL EQUILIBRIUM CONSTANT FOR REACTION HD + H_2O = HDO + H_2

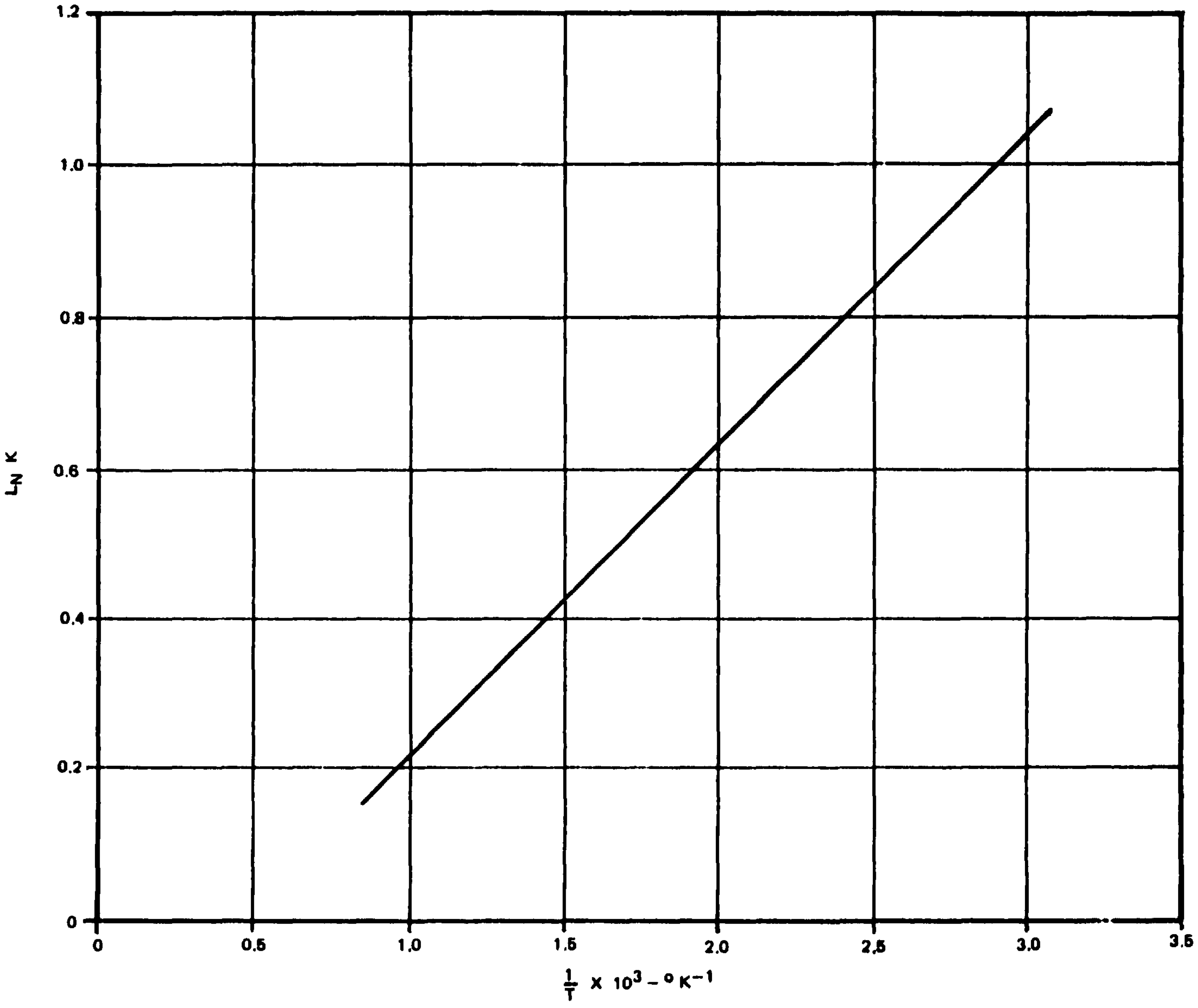
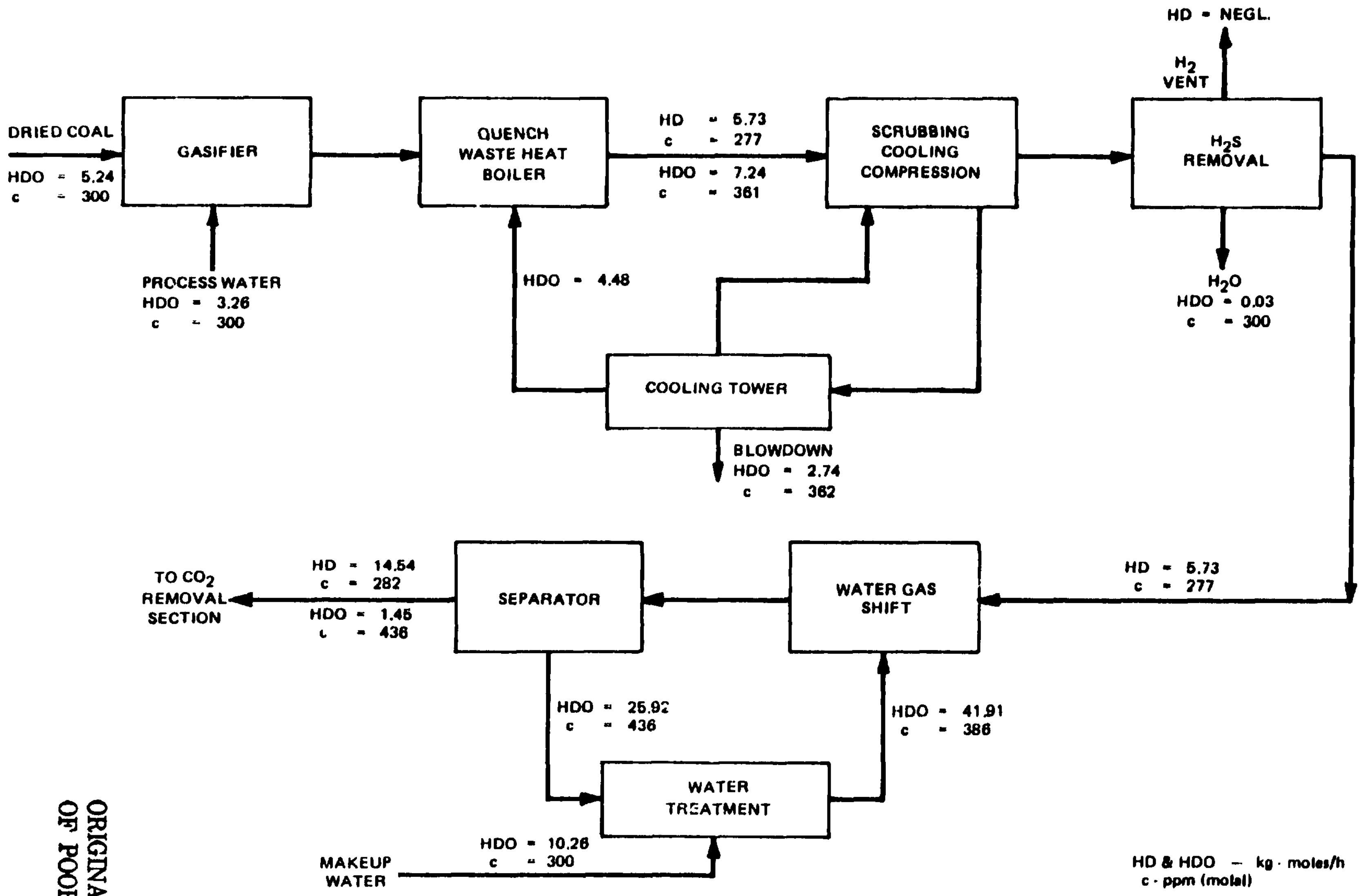


FIGURE 19



BLOCK DIAGRAM – DEUTERIUM BALANCE FEEDSTOCK GASIFIER

FIGURE 20

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RELATIVE VOLATILITY
FOR
eH_2 – HD BINARY SYSTEM

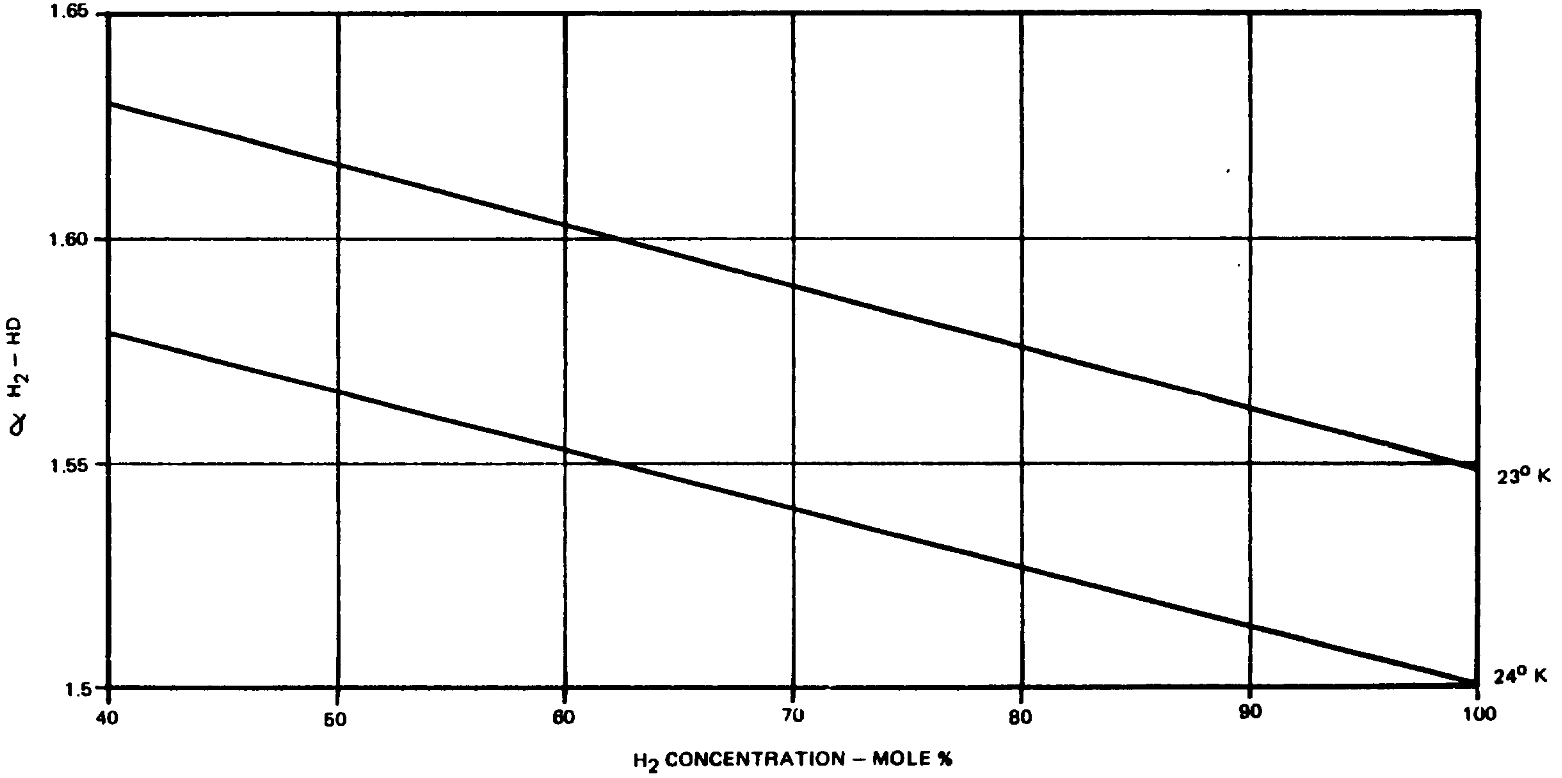
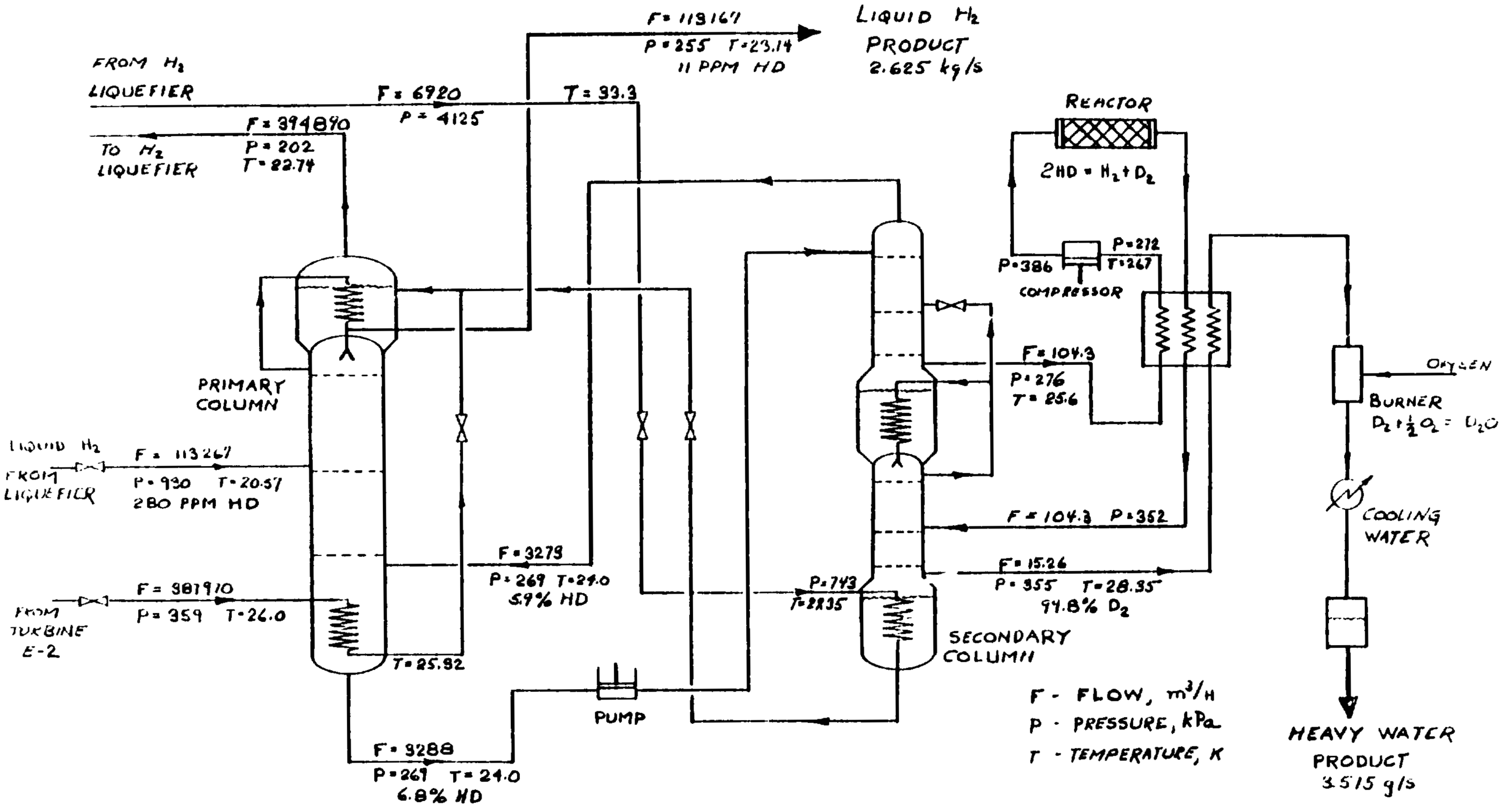


FIGURE 21



FLOW DIAGRAM - DEUTERIUM RECOVERY

FIGURE 22

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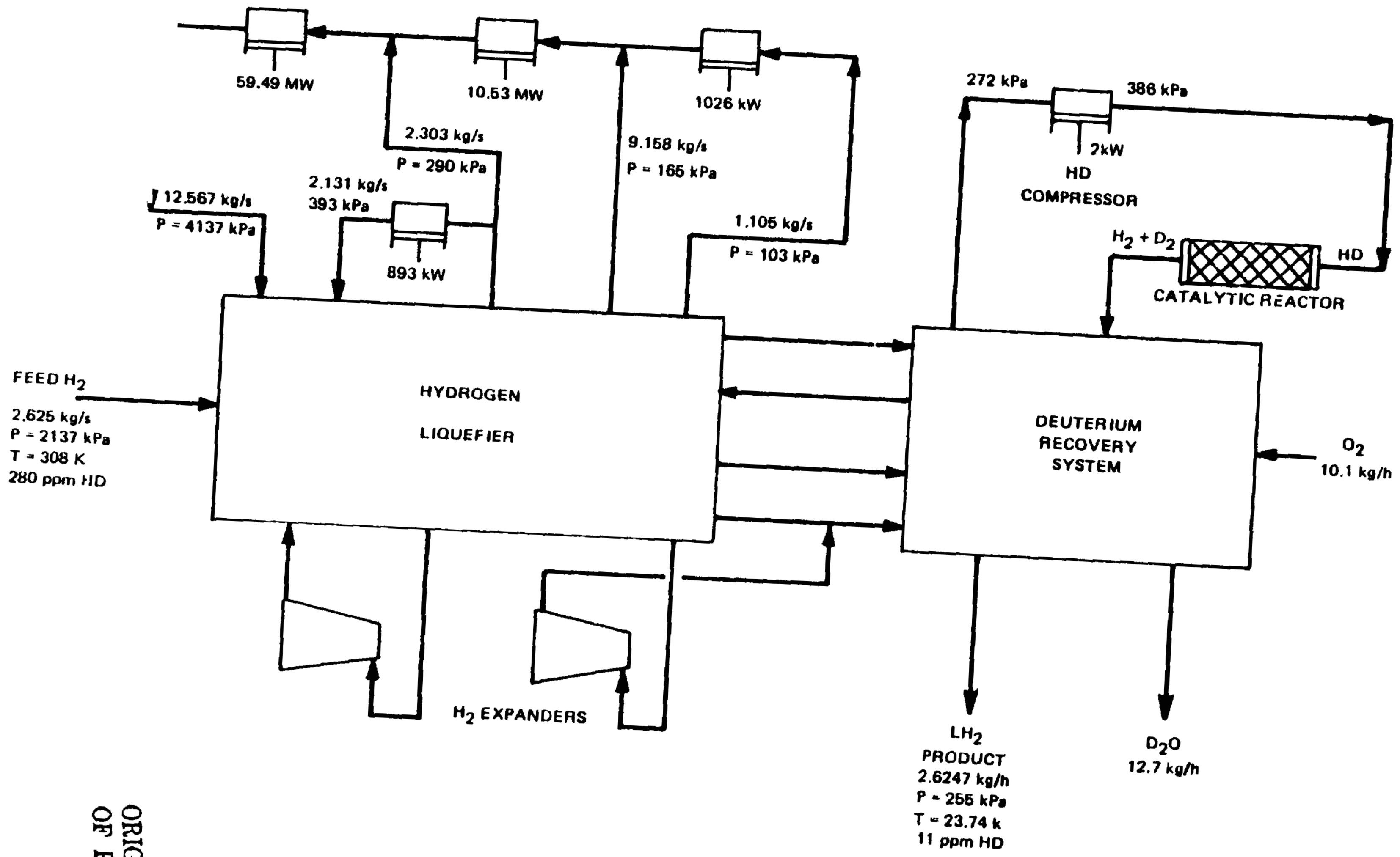

BLOCK DIAGRAM FOR DEUTERIUM RECOVERY

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FIGURE 23

UNIT COST OF HEAVY WATER FROM H_2 LIQUEFACTION PLANT VIA COAL GASIFICATION

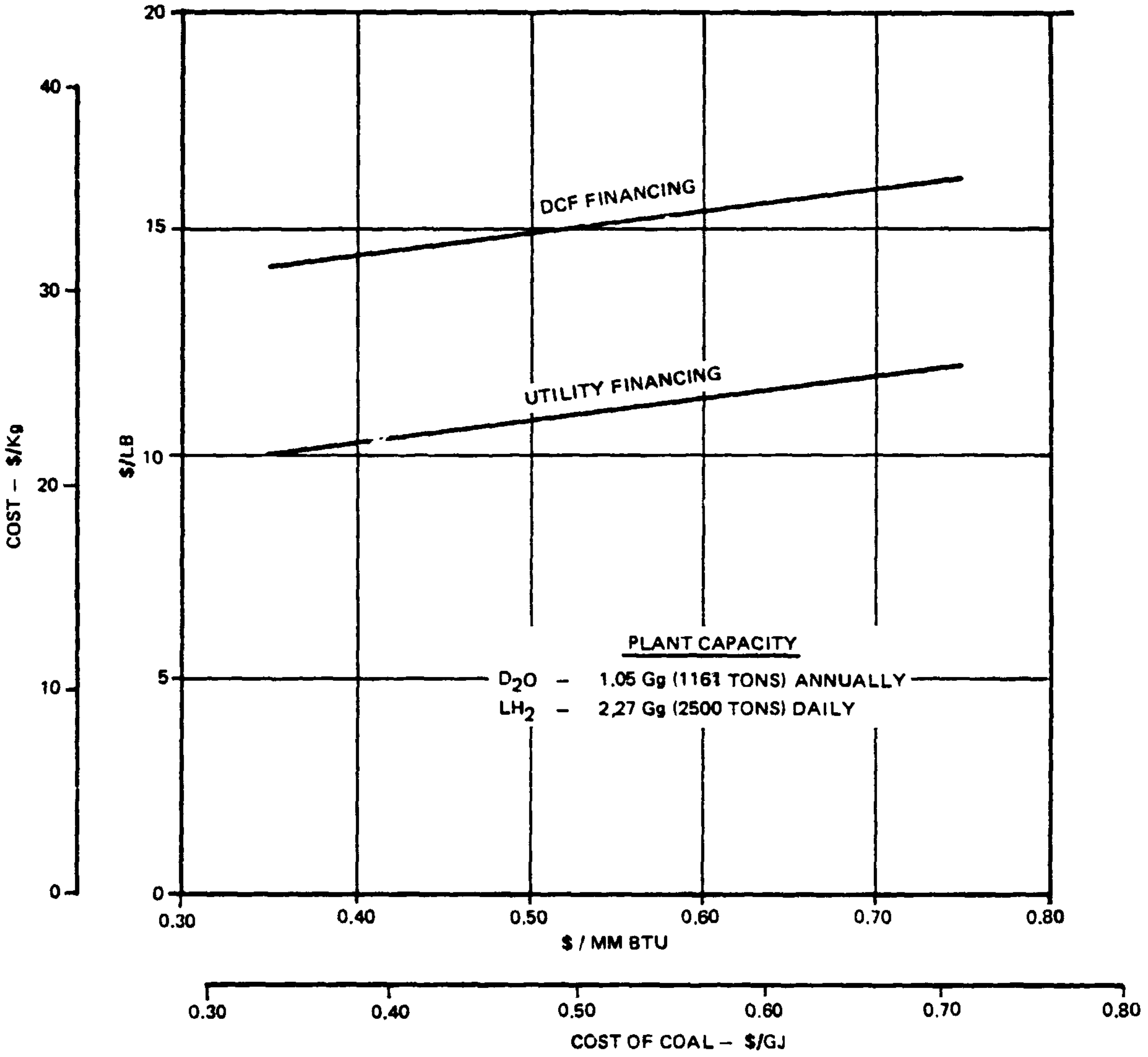

FIGURE 24

NET REVENUE FROM SALE OF HEAVY WATER

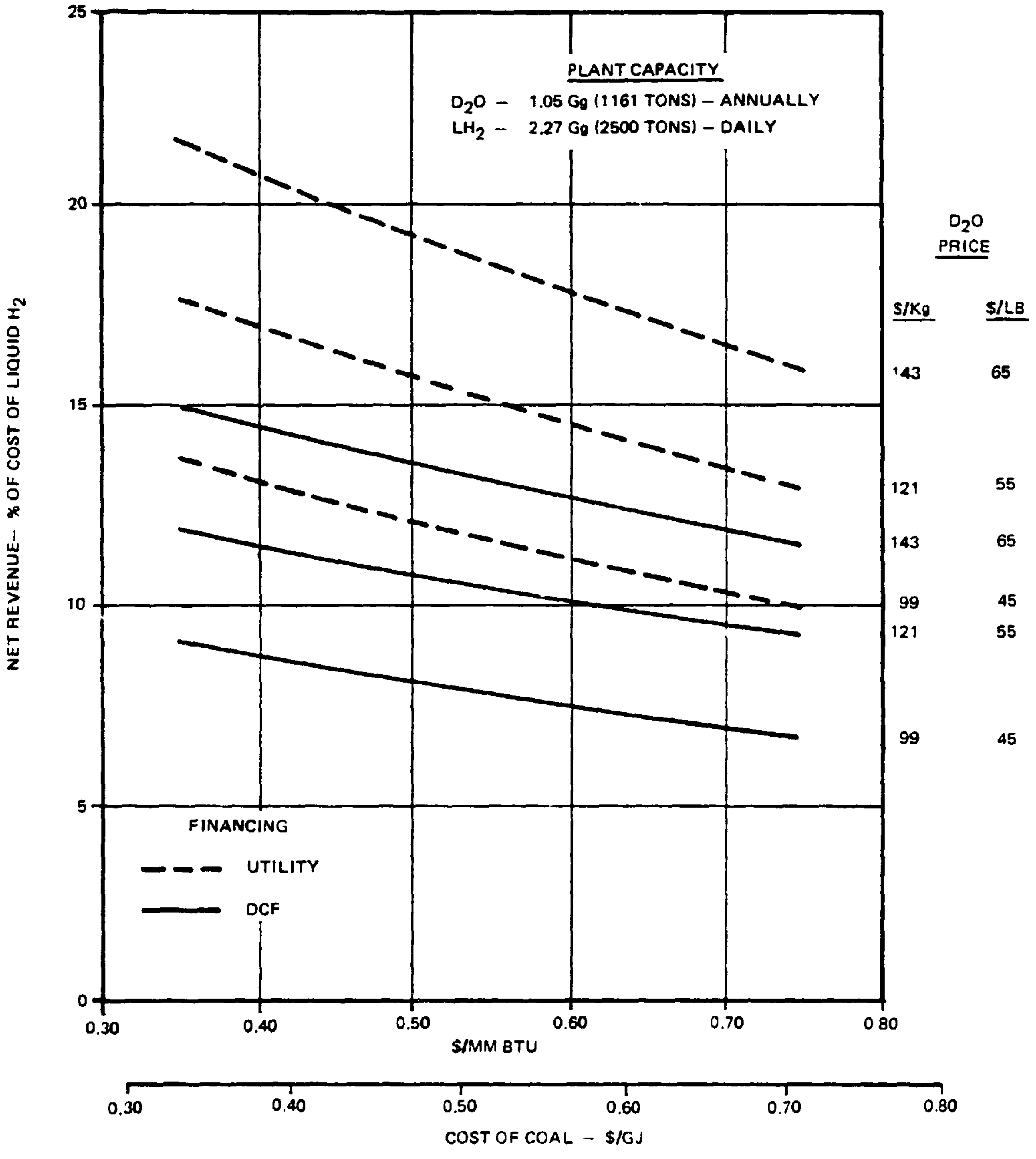

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FIGURE 25