NASA CONTRACTOR REPORT 158991

.SA-CR-158991) STUDY OF HYDROGEN RECOVERY N79-18057
STEMS FOR GAS VENTED WHILE REFUELING
.IQUID-HYDROGEN FUELED AIRCRAFT (Union
Carbide Corp.) 70 p HC A04/MF A01 CSCL 21P Unclas
G3/28 16316

STUDY OF HYDROGEN RECOVERY SYSTEMS FOR GAS VENTED WHILE REFUELING LIQUID-HYDROGEN FUELED AIRCRAFT.

C. R. BAKER

CONTRACT NAS 1-14698, MOD. 1
FEBRUARY, 1979

National Aeronautics and
Space Administration

Langley Research Center
Hampton, Virginia 23665
AC 804 827-3966

TABLE OF CONTENTS

LIST OF TABLES

LIST OF TABLES (Continued)

TABLE

LIST OF FIGURES

FORWARD

This study is a continuation of previous investigations into the use of liquid hydrogen as a jet aircraft fuel. It is directed toward methods for recovery of cold hydrogen vapor which is vented during aircraft refueling operations and from losses in the fuel storage and distribution system. A preceding study of fueling operations at the San Francisco International Airport revealed that these venting losses were quite sizeable and that provision should be made for recovery of the vent gas.

The current study examines several techniques for vent gas recovery, selects the optimum method, and determines the economics of the selected recovery method.

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, under the direction of Charles R. Baker, Consultant. 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.

Gas volumes are presented in cubic feet (or multiples thereof) at NTP conditions for U.S. Customary units and in cubic meters at STP conditions for SI units. The conversion factor is 38.044 cubic feet per cubic meter.

To be consistent with the preceding study, "LH_2 Airport Requirements Study", NASA CR-2700, all economic values are given in terms of mid-1975 dollars and are based on liquefaction technology for the year 2000 AD.

1.0 INTRODUCTION

This study is an outgrowth of previous work[1] in which the necessary facilities at a representative major U. S. air terminal to service liquid hydrogen-fueled long range commercial aircraft in the 1990 decade were defined. The San Francisco International Airport (SFO) was the selected air terminal and a projected schedule of flight traffic in the year 2000 AD for liquid hydrogen-fueled aircraft was established. This consisted of up to 70 flights per day to 9 different domestic and 4 different foreign destinations which would have the servicing facilities for return flights. Based on this forecast, a liquid hydrogen production facility having a capacity of 10.5 kg/s (1000 TPD) was recommended for SFO. This would provide enough capacity to meet the 6.681 kg/s (636.3 tons/day) average production rate throughout the year or the 8.886 kg/s (846.2 tons/day) peak month rate, plus some overcapacity to meet future expansion needs.

In the liquid hydrogen distribution and fueling system which was devised for the San Francisco Airport, there would be a substantial quantity of liquid hydrogen lost due to such factors as heat leak into the liquid hydrogen distribution lines, boiloff during cooldown of on-board aircraft fueling lines, vapor displacement while filling aircraft fuel tanks, pump work losses, and vapor displacement in the liquid hydrogen storage tanks. These losses were shown to amount to 15.7% of the block fuel requirements, which represents a substantial increase in the amount of liquid hydrogen fuel which had to be produced. If the hydrogen vapor which is evolved during fueling operations could be recovered and recycled, there would result a reduction in the amount of hydrogen feedstock delivered to the hydrogen liquefaction complex. The hydrogen, moreover, is evolved as a cold vapor, very near its saturation temperature, and possesses a definite refrigeration value. Recovery of the refrigeration would result in a reduction in the net energy required for hydrogen liquefaction.

In the SFO study, the hydrogen distribution system was provided with a hydrogen vent return line. Preliminary estimates were also made of the value of the recovered hydrogen vent gas. However, a detailed study of the optimum method for recovering the vent gas and its refrigeration value was considered to be outside the scope of that study and was not performed. Because of the significant impact of vent gas recovery on total fueling requirements and cost, the present study was authorized to determine the most suitable method for recovering cold hydrogen vent gas and recycling it to the liquefaction complex.

2.0 OBJECTIVES

The objectives of this study are to:

1. Investigate methods of capturing and reliquefying the cold hydrogen vapor produced during the fueling of aircraft designed to utilize liquid hydrogen fuel.
2. Provide an assessment of the most practical, most economic and most energy efficient of the hydrogen recovery methods which were investigated.

3.0 SUMMARY OF ACCOMPLISHMENTS

As a logical continuation of the study concerning LH_2 airport requirements (1), the quantity and variability of cold hydrogen vapor produced from the fueling of liquid hydrogen fueled aircraft were determined based on the projected flight schedules at the San Francisco International Airport in the year 2000 AD. There was found to be two types of hydrogen vent gas release, categorized by the nature of the release rate. One type had a highly variable release rate caused by the intermittent nature of the aircraft fueling schedule. Release rates varied quite frequently in response to the 70-per-day fueling schedule, varying from no flow to peak flow on the average of eight times daily. The other type of release resulted from heat leak into the liquid hydrogen distribution and fueling lines and from vapor displacement while filling the large liquid hydrogen storage tanks; this occurred at a steady rate. The total hydrogen vent gas which had to be recovered varied from a minimum rate of 4.085 m^3/s (559,500 CFH) to a maximum of 15.575 m^3/s (2,133,500 CFH) and averaged 8.429 m^3/s (1,154,500 CFH).

To achieve a steady flow of hydrogen vent gas return to the liquefier, an accumulator capacity of 95,200 m^3 (3,622,000 cu ft) was found to be required. A 3 1/2-fold reduction in accumulator size could be achieved, however, by returning the vent gas to the hydrogen liquefier at non-steady rates which were adjusted, infrequently, to minimize gas in storage. Such changes were within the acceptance capability of the hydrogen liquefier and, although not steady, the return rates represented a high degree of smoothing of the original vent gas rates.

The hydrogen vent gas has economic value, not only the material value as feedstock substitute, but also refrigeration value resulting from its low temperature. Some of the processing operations which were investigated resulted in a loss of part of the refrigeration value. Consequently, for purposes of economic comparisons in subsequent studies, the refrigeration value of the hydrogen as a function of temperature was determined, as well as the material value. Hydrogen has a material value of \$0.3627 per kg (16.45¢/lb) and a refrigeration value which varies from \$0.2130 per kg (9.66¢/lb) at 25°K to \$0.0712 per kg (3.23¢/lb) at 100°K to

\$0.0139 per kg (0.63¢/lb) at 300 K.

Performance of the hydrogen vent gas pipeline collector was determined. Over the range of vent gas flows, pressure drop through 3353 m (11,000 ft) of 10-inch pipe varies from 2.8 kPad (0.4 psid) at minimum flow to 35.0 kPad (5.1 psid) at maximum flow and amounts to 9.65 kPad (1.4 psid) at average flow. Heat leak into the pipeline will raise the temperature over a range of 0.90 K to 3.45 K with an increase of 1.75 K at average flow.

The following vent gas recovery schemes were investigated and compared on an econimic as well as on an energy conservation basis.

1. Accumulation of hydrogen in a constant pressure cold accumulator followed by refrigeration recovery and compression.
2. Accumulation of hydrogen by cold compression into a constant volume cold accumulator followed by refrigeration recovery and further compression.
3. Accumulation of hydrogen by the preceding scheme (No. 2) using the hydrogen vent gas pipeline as the constant volume accumulator.
4. Accumulation of hydrogen by refrigeration recovery and compression into a constant volume warm accumulator.
5. Accumulation of hydrogen by refrigeration recovery and a constant pressure warm accumulator followed by compression.

The use of cold accumulators was found to be unattractive, both economically and in terms of energy conservation, because of the loss of a significant fraction of the refrigeration content of the cold gas. This loss resulted from both heat leak into the accumulator and from temperature rise during compression. The latter factor was shown to make cold compression unattractive even with the greatly reduced power requirements associated with compression at cryogenic suction temperatures.

Of the three cold accumulator processes, the constant pressure accumulator scheme proved to be the most attractive. Both investment and operating costs were less for this than for either of the other two processes; energy consumption was also lower. Application of this process, however, would be based on the presumption that the cryogenic gasholder, which is a necessary item of process equipment, can be developed, engineered and fabricated to perform as assumed at the cost which was estimated for it. To our knowledge, such equipment does not exist nor has ever been built.

Such considerations, however, have no great importance in the selection of an optimum recovery process because both of the process arrangements which use warm accumulation of vent gas were found to be preferred over those using cold accumulation. Of the two warm accumulator processes, the one which utilizes the constant volume accumulator has a lower investment than the one using a constant pressure accumulator, providing that the latter operates with a rate of hydrogen withdrawal from the accumulator to the liquefier which remains constant. If the withdrawal rate is permitted to vary, and this need be only infrequently, then the gas holder becomes smaller and the investment becomes less. In such case, both of the warm accumulator processes become equivalent with respect to cost and energy consumption. For practicability of operation, the constant pressure gasholder process is to be preferred because of a simpler compressor arrangement and operating schedule. The compressors for the constant volume accumulator process must operate to accommodate the highly variable vent gas release rate and consequently must function on a complex, highly variable schedule. The competing process requires a change in operation only three times daily. For these reasons, the process of choice in this study is Process No. 5, featuring warm compression of accumulated gas, withdrawn variably from a warm constant pressure gas holder.

Examination and study of the placement of heat exchangers for refrigeration recovery indicated that such exchangers should be an integral portion of the heat exchange system of the liquefier where the effect of variation in the vent gas flow rates could be absorbed by the much greater thermal ballast provided by the liquefier streams. Other non-integrated arrangements merely exchanged flow variations in the feed stream for temperature variations in the smoothed stream.

Finally, the hydrogen vent gas recovery system of choice was integrated into a hydrogen liquefier of the capacity required for SFO. The recovered vent gas which is returned to the liquefier recycle stream emerges from the liquefier as additional liquid hydrogen product, permitting a reduction in hydrogen feedstock rate. The recovered refrigeration permits a reduction in the liquefier recycle rate and reduces the liquefier power requirement. The variation in vent gas flow rates produces only minor perturbations in the operating characteristics of the liquefier and these are easily accommodated.

Total plant investment for a liquefier with an integrated vent gas system and having a capacity of 907 MTD (1000 TPD) is $235,300,000, a reduction of $3,700,000 compared with a liquefier without recovery. Reduction in total capital requirement is $4,710,000. This liquefier, operating at a 577.2 MTD (636.3 TPD) average year-round daily capacity, with recovery, has a power requirement of 241.535 MWe compared with 248.70 MWe for operation without recovery. Annual operating costs have been reduced by $7,893,000 to $125,668,600. Total unit cost for liquid hydrogen with recovery amounts to \$0.9008 per kg (\$0.4086/lb) compared with \$0.9429 per kg (\$0.4277/lb), a reduction of \$0.0421 per kg (1.91¢/lb). At the total annual production rate of 210690 Mg (232,250 Tons), the total annual savings amount to \$8,872,000.

Economics for this analysis are based on a DCF analysis, with hydrogen feedstock valued at \$0.3627 per kg (\$0.1645/lb), electricity at \$0.02 per kWh, with capital investment adjusted to mid-1975 dollars and using 2000 AD technology. These assumptions are consistent with those used in the SFO airport facilities study(1).

4.0 HYDROGEN VENT GAS LOSSES

The initial task was to establish not only the quantity of hydrogen vented but also the variation in hydrogen venting rates with time. If the hydrogen gas were vented at a steady rate, its recovery would be greatly simplified. The only requirement would be to re-introduce the cold hydrogen gas stream into the hydrogen liquefier at the approximate temperature level as part of the low pressure hydrogen recycle stream. However, the hydrogen will not be evolved at a steady rate, but rather in a sporadic manner as a result of fueling operations. High venting rates would occur with simultaneous fueling of two or more aircraft and, as will be shown, the projected flight schedule at San Francisco specifies as many as four planes refueling at the same time. In addition, the hydrogen venting rate will increase with the amount of fuel loaded aboard the aircraft. The flight destination and its distance from the point of departure govern the block fuel requirements and the amount of fuel which must be loaded. A great diversity of destination cities would further increase the variability in the venting rate. At other times of the day there will be no refueling activity at all and the venting rate will then drop to zero.

The necessity for establishing the variability of the venting rate with time lies in the nature of the processing equipment. Such equipment, compressors in particular, tend to have steady state operating characteristics. Some variations in capacity can be accommodated but large excursions from design capacities cannot usually be handled without special consideration. Techniques which can be used include multiple compressors which can be turned on or off as required, recycle of part of the compressor effluent back to suction to maintain steady throughput (an inefficient procedure), or the installation of a gas accumulator to smooth the flow of gas to the compressor.

The variability of the venting rate will therefore have an effect upon the nature and size of the process recovery equipment which is required as well as influencing the complexity of the recovery scheme. A venting schedule must therefore be established to form a basis of comparison for the various recovery schemes which are to be evaluated.

4.1 Determination of Venting Losses

The schedule of venting losses which forms the basis of this study is based on refueling operations during an average day in the peak month and was derived from the projected flight schedule at SFO in the year 2000 AD as given in Table VII of Ref. 1 and from the refueling losses as given in Tables 1 and 2, Appendix A, of the same reference. Adoption of a venting schedule on this basis provides a continuity of information and a consistency which relates back to the original study[1].

Not all of the hydrogen fuel which is lost is recoverable. For example, in-flight boiloff due to heat leak into the fuel tanks is not recoverable, nor is ground-time boiloff due to heat leak recoverable unless the aircraft is connected to the vent recovery system. For purposes of this study, it was assumed that only those venting losses which occurred during refueling would be recovered. These include:

- Vapor displacement while filling fuel tanks
- Heat leak boiloff during refueling operations
- Piping system losses
- Pump work
- Saving by warmup of liquid withdrawn from the storage tank which, on the average, is not saturated but somewhat subcooled.

These losses were applied to each flight for all destination cities. The block fuel requirement for each city as given in Table V of Ref. 1 formed the basis of the variable losses. Displacement, pump work, and saving from liquid warmup were taken to be proportional to the block fuel requirement as determined from the sample mission in Table 1 of Appendix A[1]. Heat leak and piping system losses are assumed to remain constant for all refuelings. Not all of the potentially recoverable hydrogen vent gas has been included. For example, there would be some boiloff from on-board tankage from fueled aircraft parked for extended periods of time in maintenance facilities.

A tabulation of the vent losses from each of the preceding sources and for all scheduled flights, both domestic and international is given in Table 1. These data were then used to construct a table of daily fuel consumption and the corresponding daily venting losses, Table 2. The total daily venting rate for 59 domestic flights is 24,642 kg (54,327 lb) and for the 11 international flights is

9,113 kg (20,090 lb) for an overall total of 33,755 kg (74,417 lb). This corresponds to an average volumetric venting rate of 4.344 m^3/s (595,000 CFH).

The average vent rate amounts to 418 kg (921 lb) for the 59 daily domestic flights with deviations of +22% for the 10 flights to Honolulu and -37% for the 7 flights to Los Angeles. For the 11 daily international flights, the average vent rate is 828 kg (1,826 lb) with deviations of +5% for the flight to Rome and -6% for the 3 flights to London. For purposes of simplification in constructing a schedule of venting rates throughout the day, the assumption was made that the venting rates would remain at a constant average value of 418 kg (921 lb) for all domestic flights and 828 kg (1,826 lb) for all foreign flights. This provides a venting schedule which does not exactly represent the release rates from the fueling schedule adopted for SFO. However, it is probably sufficiently realistic in view of the likelihood of changes in flight schedules from time to time.

To establish an hour by hour schedule for hydrogen vent gas release, the airline destinations and departure times given in Table VII of Ref. 1 were adopted along with the preceding values of constant average vent rate per flight. This schedule is shown as Figure 1 and also in Tables 3A and 3B with tabulated values of vent gas release rates. There is also a column of values (see also Figure 2) showing the volume of vent gas in storage based on a constant withdrawal rate equal to the daily average vent rate of 4.344 m^3/s (595,000 CFH). The maximum volume of gas to be stored determines the size of the gas accumulator required to store it. This occurs at period No. 31 where the gas storage volume is 95.20×10^3 m^3 (3,622 MCF). An accumulator having this volume would be completely full at 2215 hours and completely empty at 0600 hours.

An accumulator large enough to contain 95.20 x 10^3 m^3 (3622 MCF) of hydrogen vent gas would be extremely large. The size can be reduced if, instead of withdrawing hydrogen for return to the hydrogen liquefaction system at a constant rate, it is withdrawn at a number of different rates which are, nevertheless, constant over periods of time less than the length of one day. The withdrawal rate would be high during periods of high venting rate and low during periods of low venting rate. This procedure is applied in Tables 4A and 4B, where three different withdrawal rates are used. From 2215 hours to 0420 hours, a low withdrawal rate of 1.10 m^3/s (150 MCFH) is used, from 0420 hours to 0720 hours, an intermediate withdrawal rate of 2.97 m^3/s (407 MCFH) is used and for the remaining 14.92 hours of the day, the withdrawal rate is increased to a maximum of 5.95 m^3/s (814 MCFH). By making use of such a procedure, the volume of the vent gas accumulator can be reduced by approximately 73%. The disadvantage of the procedure is that the recovered hydrogen vent gas is now returned to the hydrogen liquefaction complex at a variable rate rather than at a steady rate and the ability of the hydrogen liquefier to accommodate the variation must be considered. There is also the need to use multiple compressors to accommodate the variable flow with the result that some of the installed compressor capacity must stand idle part of the time. These considerations will be explored later in this report.

4.2 Steady Rate Vent Losses

In addition to the variable hydrogen gas venting rates resulting from individual plane refuelings, there will also be recoverable hydrogen vent gas originating from the following sources.

- Heat leak into the liquid hydrogen airport supply and distribution piping system.
- Vapor displacement while filling liquid hydrogen storage tanks

These steady venting rates are quantified in Table 5. Of the total of 0.3674 kg/s (34.99 TPD), about three-fourths is accounted for as storage tank vapor displacement. Of the total piping system losses, about 61% had been previously accounted for in the variable refueling losses for the individual flights and only the difference is included here. The average steady venting rate of 4.085 m^3/s (559,500 CFH) is nearly as great as the 4.344 m^3/s (595,000 CFH) average of the variable venting rate. The total recoverable hydrogen vent gas rate for return to liquefaction is the sum of the variable and steady flow rates, making the total stream vent rate also variable with a minimum of 4.085 m^3/s (559,500 CFH), a maximum of 15.557 m^3/s (2,133,500 CFH), and an average of 8.429 m^3/s (1,154,500 CFH). Capability must be provided for handling the maximum hydrogen vent rate as well as this variability in vent rate.

The average vent gas recovered, on a weight basis, is 0.758 kg/s (72.21 TPD). Total hydrogen losses have been shown(1) to amount to 15.7% of the block fuel requirements. With block fuel requirements of 7.680 kg/s (731.4 TPD) on an average peak-month day, the hydrogen losses, if unrecovered, would amount to 1.205 kg/s (114.8 TPD). Therefore, 63% of the total hydrogen loss is available for recovery. The remaining 37% represents unrecoverable boiloff loss which is incurred during flight and while on the ground.

4.3 Value of Hydrogen Gas

Comparative studies which were made between competing processes for vent gas recovery required a material value for the hydrogen gas (feedstock value) and an energy value for the refrigeration potential of the cold gas. These values were developed based on the breakdown, shown in Table 6, for the cost of liquid hydrogen produced at the SFO facility as obtained from Section 5.1.2.3 of Ref. 1.

The refrigeration value of the cold vent gas was determined from the cost of liquefaction by proportioning it to the ratio of available energy function for the cold hydrogen gas to the available energy function of liquid hydrogen. The availability function[2] is defined as

$$A_1 - A_2 = (H_1 - H_2) - T_0 (S_1 - S_2) \qquad (1)$$

The symbols H and S represent thermodynamic properties of enthalpy and entropy, respectively, while the subscripts 1 and 2 refer to the initial and final states. The symbol T_0 is the heat sink temperature at which heat is rejected to the surroundings. The change in available energy between states 1 and 2 is the minimum theoretical work required to produce the change. The assumed proportionality implies a constant efficiency for the hydrogen liquefier while producing either liquid hydrogen or cold hydrogen gas which is a satisfactory assumption for the comparative studies of this project.

The value of the cold hydrogen gas, determined in the preceding manner as a function of temperature, is presented in Figure 3. The energy value curve represents the energy required to produce the refrigeration in the cold gas. The energy required to produce cold low pressure hydrogen vent gas at 25°K amounts to 4.89 kWh/kg (2.22 kWh/lb) and drops rapidly as the temperature increases.

The other two curves represent economic value. The total value curve is the sum of the refrigeration value curve and the constant material value of the hydrogen feedstock. At 25°K, the refrigeration value amounts to 37% of the total value of the vent gas while at 300°K it accounts for only 4%. The slight positive refrigeration value of the gas at 300°K results from its pressure energy at a slightly superatmospheric pressure of 20 psia.

5.0 HYDROGEN VENT GAS PIPELINE COLLECTOR

5.1 Description

The vent gas pipeline has been defined in the previous airport study(1). It consists of a 25.4 cm (10 in.) vacuum jacketed pipe running the perimeter of the airport, from one fueling gate to the next, plus a return line running from the last fueling gate to the hydrogen liquefaction site. The perimeter loop is estimated to be 1829 m (6000 ft) in length and the return line to the liquefier is estimated to be 1524 m (5000 ft) in length.

5.2 Hydrogen Vent Gas Condition

It has been established in the airport study(1) that the operating pressure of the LH_2 supply line will be 241.3 kPa (35 psia). This will allow a 48.3 kPad (7 psid) loss through the hydrant fueling valve and the fueler vehicle, to insure a 193.1 kPa (28 psia) aircraft interface pressure when fueling at the design rate of 11,354 l/m (3000 gpm) to the design aircraft tank pressure of 144.8 kPa (21 psia). Thus, the vent gas will be available at the fuel tank at a pressure of 21 psia and at its saturation temperature of 21.45°K. The composition will be the same as that of the liquid hydrogen product from the liquefier, 97% para.

5.3 Vent Line Pressure Drop

The calculated pressure drop through 4023 m (13,200 ft) (equivalent length for pressure drop) of 26.62 cm (10.482 in.) i.d. vent line is presented in Figure 4 as a function of vent gas flow. The pressure drop at the maximum expected flow rate of 15.557 m^3/s (2,133,500 CFH) is 35.0 kPad (5.1 psid) so that the gas will always be delivered to the liquefier site at a pressure of at least 109.6 kPa (15.9 psia). At the average flow rate, the pressure drop will be only 9.7 kPad (1.4 psid).

5.4 Vent Line Heat Leak

The calculated heat leak into the 4724 m (15,500 ft) (equivalent length for heat leak) of vent line is presented in Figure 5 as a function of vent flow. The heat leak at the minimum flow rate of 4.085 m^3/s (559,500 CFH) is sufficient to raise the temperature of the vent gas to 24.9 K at the end of the pipeline. On the average, the gas will be available at the liquefier site at 23.2 K.

6.0 COMPARATIVE STUDY OF VENT RECOVERY PROCESSES

6.1 General Process Considerations

This portion of the study was devoted to the determination of the optimum process and/or procedure for recovery of the hydrogen vent gas together with its refrigeration content. As cited previously, the single factor which greatly complicates the procedure and causes most of the difficulty in the recovery of the vent gas is the highly variable nature of the vent gas supply. A steady-state supply would pose only minimal problems because it could be returned directly to the liquefier for straightforward processing. A variable rate supply has to be smoothed, however, in order to convert it to steady state and, in order to do this, an accumulator of some sort must be used to store the gas in a peak-leveling operation. This may have to be combined with compressors, used either singly or in multiple.

An important process factor which was included in the study was the temperature level for accumulating and storing the vent gas. The gas may be stored at the cryogenic level or it may first be warmed to ambient temperature, and then stored. In the first instance, the accumulator will be much smaller to store the same quantity of vent gas because of the greater gas density. However, because of increased heat leak, much of the refrigeration content of the cold gas will be lost.

Another process factor which was considered was whether the vent gas accumulation and storage should be at constant pressure or at constant volume. Both types of accumulator for ambient temperature application are well-known and are in widespread commercial usage. The major problem with either accumulator would be the compression of a highly variable supply of warm hydrogen vent gas to the constant volume accumulator.

In the case of cryogenic storage, a constant volume accumulator would present no particular technical problem; it would consist of a well-insulated pressure vessel. However, it would require cold compression of the vent gas which would destroy a portion of the refrigeration content, although the work of compression would be reduced. Special consideration would also have to be given to materials of construction for the cryogenic compressor.

A constant pressure accumulator (cryogenic gasholder) is the other alternative for storage of gas at cryogenic levels. Such a device would be quite difficult to build and operate in a cryogenic environment. There would be problems associated with moving parts, with effective gas sealing and with efficient insulation. We have never encountered such a device and were unable to discover any information concerning one. It is almost certain that such a piece of equipment has never been designed and built. Nevertheless, for the purposes of this study, a hypothetical cryogenic gasholder was considered as one means for accumulating hydrogen vent gas.

In summary, the hydrogen vent gas recovery process variations, which were comparatively studied, are listed as follows:

1. Constant pressure accumulator at cryogenic temperature level.
2. Constant volume accumulator at cryogenic temperature level.
3. Constant volume accumulator at cryogenic temperature level using the vent gas pipeline as the accumulator.
4. Constant volume accumulator at ambient temperature level.
5. Constant pressure accumulator at ambient temperature level.

6.2 Economic Basis

The economics for this study are based on the discounted cash flow (DCF) method of accounting. The DCF method accounts for the time value of money and converts all expenditures and revenues occurring at different periods of time to a common basis which is the "present value". It is through present value comparisons that equitable economic judgements can be made on combined investment and operating costs.

The following relationship for present value, which was derived in the airport study[1], is as follows:

$$PV = 4.1887\,(AOC) + 0.95956\,(I) + 0.52\,(S) + 0.9666\,(W) \qquad (2)$$

where
PV = Total present value
AOC = Annual operating cost
I = Investment
S = Startup costs
W = Working capital

This relationship is based on the following assumptions.

- Discounted cash flow financing
- 30 year project life
- 16 year sum of the years' digits depreciation of investment
- 100% equity capital
- 12% discounted rate of return
- 48% Federal income tax
- Mid-1975 dollars, no escalation
- Investment and working capital treated as capital costs in year 0
- Startup costs are treated as an expense in year 0
- Return on investment during construction based on 1.875 years

The equation for present value was rearranged to the following expression for the required annual income to cover costs of operation plus return on investment.

$$AI = AOC + 0.22908\ (I) + 0.12414\ (S) + 0.23067\ (W) \qquad (3)$$

This relationship tells us that the trade-off between operating costs and capital investment is $1.00 of investment for 22.908 cents per year of operating cost. Alternatively, it can be stated that an expenditure for capital investment and expenditures for 4.365 years of operating cost are equivalent.

The economic analyses for competing vent gas recovery systems were based on this relationship. Only operating costs and capital investment were considered; startup costs and working capital were ignored. However, when the optimum recovery process is finally selected and compared with the standard liquefaction process which does not provide for vent gas recovery, all cost factors are considered.

6.3 Cold Compression vs. Warm Compression

A process which uses constant volume accumulation of vent gas requires that the gas first be compressed and the option exists to compress the gas either cold or warm. With the aid of heat exchangers, this option can be extended to either cold or warm accumulators. There are advantages to cold compression. Because the work of compression is directly proportional to the absolute temperature of the gas at the suction to the compressor. compression at very low cryogenic temperature levels greatly reduces the work of compression. Furthermore, the density of hydrogen vapor at temperatures close to saturation is sufficiently high to permit the use of centrifugal compressors. On the other hand, cold compression raises the temperature of the gas and destroys a portion of its refrigeration content. This represents an energy loss. In addition, a cold compressor is a higher investment piece of equipment than is a warm compressor.

The comparison between warm and cold compression is based on the process configurations sketched in Figure 13. For cold compression, vent gas is delivered to the suction of a centrifugal compressor at a pressure of 131 kPa (19 psia) and a temperature of 25°K. The flow rate is assumed to be the daily average value. The gas is compressed to some discharge pressure and warmed to 300 K in a heat exchanger which has a pressure drop of 28 kPad (4 psid) when the average pressure in the exchanger is 117 kPa (17 psia). This pressure drop varies as a function of the average gas density in the heat exchanger and is accounted for in this analysis. The vent gas is delivered at a final pressure, P-2.

For warm compression, the vent gas is available at the same process conditions as for the cold compression and is warmed to 300°K in a heat exchanger having a pressure drop of 28 kPad (4 psid). It is then compressed to final pressure, P-2, in a reciprocating compressor.

In either case, it is assumed that the refrigeration content of the cold gas is recovered in the heat exchanger and effectively used, as, for example, in the hydrogen liquefier. However, as the discharge pressure of the cold compressor increases, so does the discharge temperature and less refrigeration remains for recovery. The value of this lost refrigeration between the suction and discharge temperatures was determined with the aid of the "refrigeration value" curve presented on Figure 3.

The total power consumption for each of the two alternatives is shown in Figure 14. The power consumption for cold compression exceeds that for warm compression by 85-100% over most of the pressure range. A breakdown of individual power consumption presented in Table 7 shows that over 90% of the total power for cold compression is the equivalent power for the refrigeration content of the cold gas. The power of compression of the cold gas is generally of the order of 10% of the power of compression of the warm gas.

Figure 15 presents the results of the economic comparison between the two alternatives. Power is the major cost item and greatly outweighs the much smaller investment required for the cold compressors. The comparison makes clearly evident that warm compression is the preferred alternative on both an energy utilization basis and a cost basis.

6.4 Cold Accumulator Processes

6.4.1 Constant Pressure Cold Accumulator

The constant pressure cold accumulator process is shown schematically in Figure 6. Cold vent gas is collected from the 19 fueling gates at 25°K and 138 kPa (20 psia) and delivered to a cold accumulator, or cryogenic gasholder, which is at constant atmospheric pressure. Cold gas is withdrawn from the gasholder at a constant average daily rate, warmed to 300°K in a heat exchanger and compressed in a warm reciprocating compressor to the 4137 kPa (600 psia) recycle pressure of the hydrogen liquefier and combined with the liquefier recycle stream.

In this process, energy is consumed via two mechanisms:

1. Energy of compression of the warm vent gas
2. Energy equivalent of the refrigeration loss due to heat leak into the cold accumulator

The second item could be of considerable magnitude, especially if the accumulator were very large and present a large amount of surface area for heat transfer.

One of the first problems encountered in this process was to determine the nature of the cryogenic gasholder. Several types of warm gasholders have been described[3] but we were unaware of any such equipment suitable for use in a cryogenic environment. A search of the literature also failed to disclose any information concerning cryogenic gasholders or even their existence. We are almost certain that such a device has never been designed or built. Nevertheless, we were committed to an evaluation of the process which uses constant pressure cryogenic accumulation of vent gas and so a hypothetical gas holder was assumed and a set of characteristics was assigned to it. Had the process appeared to be favorable after evaluation, the development of the gasholder would have been a recommended course of action.

It was decided to pattern the hypothetical gasholder after those used in ambient temperature service. These are broadly categorized into two types: a water-sealed type in which the gas is trapped beneath a cylindrical bell within a shell, and a dry type which is of shell and piston configuration with the gas trapped beneath the piston. In both types, the bell, or the piston, rises and falls within the shell to accommodate the increasing or decreasing volume of gas. The difficulty of adapting current designs for cryogenic service lies in providing a suitable gas seal which retains its integrity at low temperatures. Liquid seals, which are often used with warm holders, must be rejected as unsuitable for use at temperatures in the region of 25°K. Even seals used with dryholders are normally fabricated from rubberized fabric, usually lubricated, and these would lose their flexibility, and therefore their sealing characteristics, when used at low temperatures.

The Wiggins type of gas holder uses a fabric impregnated and coated with synthetic rubber to seal the annular space between the piston and shell. The seal is distended upward in a loop when gas is confined beneath the piston. This seemed to be the easiest to adapt to cryogenic service. A conceptual sketch of the cryogenic gasholder is shown in Figure 7. The basic configuration is the inverted bell within a loose fitting shell. The seal between the floating roof and the shell is not intended to be gas tight but is primarily for the purpose of preventing convective circulation of cold hydrogen into the region where it would contact the flexible enclosure, or Wiggins-type seal, which must be kept warm to retain its flexibility. The flexible seal is attached to a warm spot on the floating roof and to the top edge of the shell, and is guided, as appropriate, with such devices as rollers and guide channels The entire gasholder would require an insulation layer between the tank and the foundation and an electrically heated foundation to prevent heaving. This is a type of construction used with cryogenic liquid storage tanks.

The derivation of a generalized relationship to determine the number and size of the gasholders will be developed first. Based on the maximum volume of gas to be stored, shown in Table 3 to equal 95,200 m^3 (3622 MCF), the actual volume of gas in storage is

$$\text{Volume} = 3{,}622{,}000 \times \frac{38.117}{386.79} = 357{,}000 \text{ cu ft} = 10{,}109 \text{ m}^3 \qquad (4)$$

where 38.117 and 386.79 are the molal volumes of hydrogen at 1 atm, 30°K and at NTP conditions, respectively. This is then equated to the gasholder volume.

$$357{,}000 = \frac{\pi}{4} D^2 H N \qquad (5)$$

where D = gasholder diameter, ft.

H = gasholder height, ft.

N = number of gasholders

Standardizing on a height-to-diameter ratio,

$$H = 1.5 D, \qquad (6)$$

substituting (6) into (5), and solving for the gasholder diameter, gives

$$D = 98.377 (\pi N)^{-1/3} \qquad (7)$$

The surface area of the gasholders may be expressed as follows:

$$A = N\left[\pi D\psi H + 2\frac{\pi}{4} D^2\right] \tag{8}$$

In this expression, ψ is the fraction extension of the floating roof which also represents the fraction of total capacity of the gasholder. For the design which was adopted, the maximum value of ψ is 1.0 and the minimum value is 0.5. The expression also recognizes that the surface of the bottom of the tank is effective for heat transfer. Substituting Equations (6) and (7) into (8), and simplifying, the following relation is obtained.

$$A = 9678\ (\pi N)^{1/3}\ (1.5\psi + 0.5) \tag{9}$$

The performance of the gasholder depends on the quality of insulation provided. The performance of an uninsulated gasholder was first investigated to determine whether <u>any</u> insulation would be required. For an uninsulated gasholder, the heat leak into the holder is governed by the following basic relation for heat flux as a function of temperature difference.

$$\frac{q}{A} = U\ (T_w - T_c) \tag{10}$$

where
- q = Heat transfer rate into holder
- A = Surface area of holder
- T_w = Ambient temperature
- T_c = Temperature of gas inside holder
- U = Heat transfer coefficient

The overall heat transfer coefficient, U, is determined by the natural convection of hydrogen inside the holder and air outside the holder. The relationship for the individual film heat transfer coefficient under natural convection forces is given by McAdams [4] with symbols as defined in the reference.

$$h = 0.13\ k_f \left[\frac{\rho_f^{\,2}\ g\beta_f\ \Delta_t}{u_f^{\,2}} \left(P_f \right) \right]^{1/3} \tag{11}$$

Substituting the appropriate values for the physical properties of hydrogen and air at their corresponding film temperature yields the following values for the heat transfer coefficients.

	SI $J/sec, m^2, °K$	U.S. Customary $Btu/hr, ft^2, °F$
h , hydrogen	29.5	5.2
h, air	11.3	2.0
U, overall	8.5	1.5

Assuming that 4 gasholders, each 12.9 m (42.3 ft) diam. by 19.3 m (63.5 ft) high, will be required, the total surface area for heat transfer is 2613 m^2 (28125 ft^2) with the tank half full. The total heat transfer into the gasholder from the surroundings under a temperature difference of 270°K will amount to 6004 kJ/sec (20,500 ,000 Btu/hr). A heat balance based on steady flow conditions of 4.344 m^3/s (595,000 CFH) vent gas through the holder shows that this heat leak is sufficient to warm the effluent gas to a temperature of 245°K. These calculations consider only convective heat transfer and ignore radiation. Superimposing the additional heat transfer resulting from radiation would only make an uninsulated gasholder look even more unattractive.

Having established that insulation of the cold gasholder is necessary, it becomes necessary to establish the performance of an insulated gasholder. The heat transfer rate through the insulated wall is given by the conduction equation

$$q = k\, A \frac{dT}{dL} \qquad (12)$$

where q = heat transfer rate

k = thermal conductivity

A = heat transfer surface

$\frac{dT}{dL}$ = temperature gradient across insulated wall.

Substituting Equation (9) into (12) and allowing a 30.48 cm (1 ft) thickness of insulation with a thermal conductivity of 0.0346 watt/m, K (0.02 Btu/hr,ft,F) gives the following:

$$q = B\,(300 - T_2)(\pi N)^{1/3}\,(1.5\psi + 0.5) \qquad (13)$$

This relationship is based on a 300 K ambient temperature. The symbol B is a constant with the following values.

	SI	U.S. Customary
Units of q	j/s	Btu/hr
Value of B	0.10203	348.4

The thermal performance of the gasholder is based on a model in which cold vent hydrogen gas is flowing into the gasholder at a variable rate, M_f, and at a constant temperature assumed to be 30 K. Over a short interval of time, heat leak into the gasholder raises the temperature of the gas from T_i to T_f. Hydrogen is withdrawn from the gasholder at a constant rate, M_w, and at the average temperature, T_{av}, of the gas in the gasholder during the short time interval, $\Delta\Theta$.

Heat leak into the holder is given by Equation (14).

$$q = B\,(300 - T_{av})\,(\pi N)^{1/3}\,(1.5\,\psi + 0.5)\,(\Delta\Theta) \qquad (14)$$

Some of this heat is imparted to the entering feed gas and warms it to T_f.

$$q = M_f\,C_p\,(T_f - 30)\,(\Delta\Theta) \qquad (15)$$

Another portion of the heat warms the withdrawn gas from T_i to T_{av}.

$$q = M_w\,C_p\,(T_{av} - T_i)\,(\Delta\Theta) \qquad (16)$$

The remainder of the heat warms the residual tank content from T_i to T_f.

$$q = M_T\,C_p\,(T_f - T_i)\,(\Delta\Theta) \qquad (17)$$

The residual tank content, M_T, is defined as the difference between the contents of the tank at the beginning of the interval and the gas withdrawn from the tank

$$M_T = (M_i - M_w)\,(\Delta\Theta) \qquad (18)$$

Equations (14) through (18) were solved over a 24-hour period for the temperature history of the gas within the tank. The results are presented in Figure 8. The temperature remains within the limits of 36-50°K during most of the active portion of the day while refueling operations are being carried out. However, during the interval between midnight and 4:00 AM when no planes are refueling and there is no vent gas flow into the tank, the temperature rises to as high as 74 K. At 9:00 AM during peak refueling, the temperature drops to its minimum level of 36 K.

The economics for recovery of vent gas using the constant pressure cold accumulator process is given in Table 8. The analysis is based on the use of 4 gasholders, one for each of the 4 liquefier modules, and each having a capacity of 23,800 m^3 (905.5 MCF) of vent gas. The holder dimensions are 12.89 m (42.3 ft) diameter by 19.35 m (63.5 ft) high when full. The cost of the cryogenic accumulator was estimated from the cost of a warm gasholder and known relationships between the cost of cryogenic vs. warm tankage. The total annual cost of $1,827,000 for recovering the vent gas by this process is equivalent to a unit cost of $0.1483/kg (6.73¢/lb). This cost, when deducted from the value of the cold vent gas of $0.5756/kg (26.11¢/lb), leaves a net value of $0.4274/kg (19.38¢/lb).

Table 8 also tabulates the power cost for the process so that the several competing processes may be compared for energy consumption.

6.4.2 Constant Volume Cold Accumulator

The process for recovery of hydrogen vent gas using a constant volume cold accumulator is shown in Figure 9. Vent gas is collected at each fueling gate into the pipeline and transported to the hydrogen liquefier site. Here it is compressed, by means of a centrifugal compressor, into a constant volume accumulator, or tank, at some intermediate pressure. It is then withdrawn from the tank at a constant average rate, warmed to ambient temperature in a heat exchanger and compressed the rest of the way to the hydrogen liquefier recycle pressure by means of a reciprocating compressor.

There is nothing unique about the constant volume cold accumulator. It is simply a well-insulated cryogenic pressure vessel. However, the operating pressure for the accumulator is an additional variable which must be determined by an optimization procedure. Increasing the pressure reduces the volume of tankage required as well as the size and energy consumption of the warm compressor. However, the size and energy consumption of the cold compressor increases and, simultaneously, increasing amounts of the refrigeration value of the cold gas are lost.

Figure 10 plots the results of the optimization as total annual cost as a function of compressor discharge pressure. Minimum cost occurs at a storage pressure of 1655 kPa (240 psia) although the curve is relatively flat over the range 689-2068 kPa (100-300 psia). The discontinuities in the curve result from the use of a discrete number of accumulators of fixed size of 3.66 m (12 ft) diameter by 30.48 m (100 ft) long. These are judged to be maximum practical dimensions for such tankage.

The economic analysis for the process at the optimum storage pressure is presented in Table 9. The total annual cost is about 37% more than for the cold constant pressure accumulator. Although increased investment accounts for a small portion of the total increase, the increased power requirement accounts for most of it. Loss of refrigeration value of the gas from the cold compression is the major power item, representing over 90% of the total. Total energy consumption is more than 50% greater than the energy consumption for the cold constant pressure accumulator process which is consistent with previous findings in the comparison between warm and cold compression. In the present case, only 27% of the compression is at low temperature.

Heat leak into the cold accumulator was investigated and found to be negligible, causing a temperature rise of only 0.4 K.

6.4.3 Vent Gas Pipeline as Cold Accumulator

This is a variation of the preceding process in which the vent gas collecting pipeline is sized for use as a constant volume accumulator. Figure 11 shows the process. With this configuration, each fueling gate requires its own compressor, discharging into an enlarged pipeline which terminates at the hydrogen liquefier. The remainder of the process is similar to the preceding process shown in Figure 9.

As with the preceding process, an optimization of the storage pressure in the pipeline is necessary. The specified diameter of 25.4 cm (10 in) of the pipe does not provide sufficient volume to accumulate the excess gas flow at high venting rates, even with its length of slightly in excess of 3.22 km (2 miles). For example, a storage pressure of 172 kPag (25 psig) would require a pipe diameter of approximately 1.32 m (52 in.) to provide the necessary storage volume.

With increasing pressure, the pipe diameter and cost will decrease, but the proportion of the compression which occurs at the low temperature level increases and causes an increase in operating cost. The results of the optimization are presented in Figure 12 and show a decrease in total annual cost with increasing storage pressure. The optimum occurs in the 1793-2068 kPa (260-300 psia) range and over this range differences are too small to be significant. An economic analysis based on a 1793 kPa (260 psia) pressure is shown in Table 10. This process is found to be inferior to both of the other two cold accumulator processes, both in cost and in energy consumption. The greatest difference lies in the incremental investment of over $6,000,000 for the 80.0 cm (31.5 in.) diameter vacuum jacketed pipeline. Other cost items are comparable to costs encountered for the cold constant volume accumulator of the tankage variety.

In this case, heat leak into the pipeline was considered and found to be a small but, nevertneless, significant amount. It is much greater for the pipeline accumulator than for the tank-type accumulator primarily because the surface area to volume ratio is much greater for the pipeline.

6.5 Warm Accumulator Processes

Previous comparisons between warm and cold compression lead to the expectation that warm accumulator processes would be more economical than cold accumulator processes. The high value possessed by the refrigeration content of the cold gas at the very low temperatures involved override all other cost factors when the bulk of this refrigeration content is lost by compressing the cold gas. The advantage gained by not compressing the cold gas was shown by the constant pressure cold accumulator process, which, so far, has proven to be,not only the most economical process, but also the lowest in energy consumption. It might be expected that further gains would accrue if the refrigeration loss caused by heat leak into the accumulator could also be eliminated. Detailed process studies of both constant pressure and constant volume warm accumulator processes were conducted to either verify or reject these expectations.

6.5.1 Constant Volume Warm Accumulator

The constant volume warm accumulator process is given on Figure 16. Hydrogen vent gas is collected into the vent gas pipeline collector, transmitted to the hydrogen liquefier site, and warmed to ambient temperature. It is then compressed to the hydrogen liquefier recycle pressure, stored in a warm accumulator, which is nothing more than a pressurized tank, and then fed to the hydrogen liquefier. In this process, there is no optimization of the storage pressure. All compression is in the warm state and in order to minimize tankage volume, maximum storage pressure is used.

Special attention was directed to the warm compressor configuration. The main requirement was to provide capability for handling the highly variable vent gas flow rate using a minimum number of compressors. It was assumed that each compressor was capable of being turned back 10% by either speed control or the use of automatically adjusted clearance pockets. The result was a bank of four compressors each of a different size.

Compressor No.	FLOW SI m^3/s	FLOW U.S. Customary (MCFH)
1	0.767	(105)
2	1.533	(210)
3	3.067	(420)
4	6.133	(840)

The schedule for operating these compressors is shown in Table 11. For example, for the first 20 minutes of the day, No. 3 compressor would operate at 94% of capacity. All compressors would then be down until 4:00 AM at which time compressors No. 1 and No. 3 would be turned on and operate for 20 minutes at full capacity. For the next 20 minutes, No's 2 and 4 would be turned on at full capacity and No's 1 and 3 turned off. This would be followed by another 20 minute period when only compressors No. 1 and No. 3 would operate and then from 5:00 AM to 6:00 AM, all compressors would be down again. Operation in this variable manner continues throughout the remainder of the day.

Bringing compressors in and out of service in accordance with such a schedule would have to be accomplished automatically. A technique for doing this was not devised but a control scheme to bring compressors on line using sensors to detect pressure build up at the suction nozzle should be workable. Compressors to be taken off line could simply be unloaded automatically in a suitable manner and kept turning over during the period that they are off line, to be maintained in readiness for return to operation when needed.

The results of the economic analysis for the constant volume warm accumulator process are presented in Table 12. This process is economically superior to any of the cold accumulator processes examined, both in capital investment and in operating cost. Energy consumption is also much less. The major disadvantage with this process, as revealed by the high capital investment requirement for compressors, is that the compressors are required to assume part of the function of smoothing the vent gas flow. As a result, much of the compression capacity is standing idle for a large percentage of the time. This represents an inefficient utilization of equipment.

6.5.2 Constant Pressure Warm Accumulator

This process, depicted on Figure 17, uses a conventional gasholder for the accumulator. Cold hydrogen vent gas is delivered, via pipeline, to the liquefier site, warmed to ambient temperature and fed to the gasholder. The gasholder expands or contracts, as required to smooth the flow variations in the vent stream. A steady flow of hydrogen is withdrawn from the holder, compressed to liquefier recycle pressure in warm reciprocating compressors, and delivered to the liquefier.

Results of the economic analysis for this process are presented in Table 13. This process is also more attractive economically than any of the cold accumulator processes but less attractive than the warm constant volume accumulator process. Although the operating costs are identical in each case, the high cost of the gasholder causes the investment for the constant pressure accumulator process to be nearly double the investment for the constant volume accumulator process.

The economics for this process can be improved, however, if the compressors are allowed to share part of the function of smoothing the hydrogen vent gas flow. As described in Section 4.1 and shown in Table 4, a variable withdrawal rate from the gasholder permits a decrease in gasholder size. The withdrawal schedule shown in Table 4 permits a 73% reduction in gasholder volume, from 95.20×10^3 m^3 (3622 MCF) to 26.13×10^3 m^3 (994 MCF). As with the constant volume accumulator process, the compressors are required to operate on a varying flow schedule. However, in contrast, only three compressors rather than four are used, and with only three daily compressor changes, the compressors are turned on and off much less frequently. It is possible to accommodate the operating schedule on a time basis which is simpler to effect than one based on detecting an accumulation of gas at the compressor suction.

The economics for this process, as shown in Table 14, are also favorable and comparable, in both investment and operating cost, to the best previous case, the constant volume warm accumulator process. It is, moreover, a process which uses conventional equipment and one which presents no particular technical nor operating difficulties or problems. Therefore, it has been selected as the best of the competing processes for recovery of the hydrogen vent gas.

7.0 HEAT EXCHANGER PLACEMENT AND PERFORMANCE

All vent gas recovery processes assumed that the vent gas was, at some point in the process, warmed to ambient temperature in a heat exchanger, with recovery of refrigeration from the vent gas, before delivering the vent gas to the hydrogen liquefier. The proper placement for this heat exchanger will now be examined in more detail. Only the selected vent gas recovery process was considered in the study. Figure 18 shows the inclusion of the heat exchanger in the process. The cold vent gas, delivered from the pipeline, is warmed to ambient temperature in countercurrent heat exchange with the compressed gas. The cold compressed gas is throttled to 112 kPa (16.2 psia), the pressure of the flash gas in the hydrogen liquefier process. A phase separator makes the appropriate separation of the stream into liquid and vapor streams which are, in turn, delivered to the hydrogen liquefier at the appropriate process location.

Performance of the liquefier under varying flow conditions was examined. The low-pressure hydrogen vent gas flow is highly variable while the flow rate for the compressed return hydrogen is more nearly constant. For purposes of this study, it was assumed to be constant at the average daily flow rate of 8.43 m^3/s (1154.5 MCFH). (At this flow rate, the steady rate vent losses (Section 4.2) as well as the variable rate vent losses (Section 4.1) are included.)

Heat exchanger performance is highly sensitive to the flow ratio between the two streams in heat exchange. A flow for the warming stream which greatly exceeds that for the cooling stream will provide an excess of refrigeration and will cool and liquefy a large fraction of the compressed stream. A warming stream flow which is deficient, however, may not cool the compressed gas stream sufficiently to liquefy it.

Performance of the heat exchanger, under three different conditions of flow unbalance and over a range of pressures for the compressed gas stream, is shown in Tables 15A through C. These tables show that the cold compressed gas would be returned to the liquefier under conditions equally as variable as the flow of the cold vent gas. One extreme would exist during periods of maximum vent flow, when 81% of the compressed gas would be liquefied while the other extreme would occur during periods of minimum vent flow, when the compressed gas would be cooled to only 168 K. Obviously, such a heat exchanger arrangement serves no useful purpose inasmuch as it merely exchanges flow variability for thermal variability, and it was concluded that the preferred technique would be the introduction of the cold vent gas directly into the hydrogen liquefier at the appropriate point in the process.

8.0 INTEGRATION INTO HYDROGEN LIQUEFIER

8.1 Process Arrangement

Flow variations in the vent gas recovery scheme are smoothed by addition of an accumulator. This, in itself, is inadequate because the recovery of the refrigeration content of the vent gas imposes a varying thermal load on the refrigeration recovery portion of the process. The inadequacy of a simple heat exchanger arrangement such as the one depicted in Figure 18 was shown in Section 7. This study led to the conclusion that the varying thermal load could best be absorbed directly by the recycle stream of the liquefier process. As an indication of this capability, the flow of the vent gas stream can be compared with the recycle flow. Such a comparison shows that the $\pm$ 5.75 m^3/s (787 MCFH) variation in vent gas flow is only about 5% of the 119.15 m^3/s (16,319 MCFH) recycle flow and indicates that the resulting variation in refrigeration supply to the liquefier might be readily assimilated by the combination of the recycle flow stream and the thermal ballast of the liquefier equipment.

To pursue the investigation further, the process model of the hydrogen liquefier which was prepared in a previous study[2] for automatic computation was revised to include the addition of the hydrogen vent gas stream. The revised process model is presented as Figure 20 which may be compared with the original process model, Figure 19. Cold hydrogen vent gas is introduced as process stream No. 133 and is warmed to ambient temperature in a separate pass in each of three exchangers in series, X-4, X-3 and X-1. The refrigeration content of this stream is imparted to the high pressure recycle gas or portions of it. For example, in heat exchanger X-4, it provides part of the cooling for the subcooling liquid, stream 30-31, while a portion of the turbine exhaust stream 33-34, provides the remainder.

The warmed vent gas stream after leaving heat exchanger X-1, is delivered to a warm gasholder from which a steady stream of hydrogen, No. 147, is withdrawn, compressed in vent gas compressor, P-VNT, and delivered to the suction of the main recycle compressor, P-RC as stream No. 148.

8.2 Process Performance

Tables 18 and 19 contain process stream data for the Figure 20 arrangement, which features recovery of the vent gas. In comparison, Tables 16 and 17 contain the process stream data for the Figure 19 arrangement, which does not feature vent gas recovery. In each case, the plant is sized to produce 192.1 metric tons per day (211.7 TPD) of hydrogen product (as saturated liquid at 1 atm) from each of the four plant modules, yielding a total production rate of 768.2 MTD (846.8 TPD) of liquid hydrogen. This represents operation of the plant to supply the fuel requirements of SFO for an average day during peak month activity. The vent gas recovery case, for which stream data are presented in Tables 18 and 19, is for the daily average vent gas flow rate of 4.344 m^3/s (595.0 MCFH) plus the steady vent gas rate of 4.085 m^3/s (559.5 MCFH), for a total of 8.429 m^3/s (1154.5 MCFH). Thus all available hydrogen vent gas is recovered. Moreover it is recovered as product, and comparison of stream No. 1 flows for the two cases shows that the feed gas flow rate is diminished by the exact amount of the flow rate of the vent gas stream. Comparison of flow rates through the two turbines and of the liquid and cold nitrogen gas flow rates reveals that the refrigeration content of the vent gas stream is being effectively recovered.

The refrigeration value of the para hydrogen is assumed to be non-recoverable. This is a conservative assumption and probably realistic in that the conversion rate from para to ortho is sufficiently slow that, at most, only a small amount of conversion will occur. It is further assumed that complete conversion to normal hydrogen occurs in the recycle compressor and that only normal hydrogen enters the liquefier as the feed/recycle stream.

Stream data for the nitrogen refrigerator, which supplies liquid nitrogen and cold nitrogen gas to the purifier and liquefier, are presented in Tables 20 and 21 for operation without vent gas recovery and in Tables 22 and 23 for operation with vent gas recovery, based on the process model, Figure 21.

In addition to conducting process evaluations for the condition of average vent gas flow rate, the effect of variations in the flow rate was also explored by completing process balances for the conditions of maximum and minimum vent gas flow rate. The effect of these variations on the power requirements for the major items of liquefier process machinery is presented in Table 24. Also shown is the no recovery "standard" process.

Comparisons of power requirement for the three vent gas flow conditions show only modest variations. The recycle compressor, which is the largest item of machinery, varies + 1.5% and - 2.4%, respectively with minimum and maximum vent gas flow. The greatest variation lies with the nitrogen refrigerator, wherein the capacity drops 5.5% with maximum vent flow, and the nitrogen refrigerator can readily handle capacity variations of this magnitude. The hydrogen turbines also undergo only slight changes in power output.

Comparison of power requirements shows, on the average, a 4.0% reduction in power for the power-consuming items of equipment listed when recovering vent gas.

Comparisons were also made among the various cases based on flow rates through the various pieces of equipment. Since, in these process studies, flow and power are proportional, the results, Table 25, are similar to the results of the power comparison.

8.3 Utility Requirements

Utility requirements for hydrogen purification and liquefaction, with and without vent gas recovery, are summarized in Table 26, for a capacity of 577.2 MTD (636.3 TPD), representative of the daily average capacity over a period of one year. The utilities are also based on 2000 AD technology, which has been previously described[(2)]

Total electrical power for hydrogen liquefaction with vent gas recovery, including production and plant auxiliaries and contingency, is 7165 kW (2.88%) less than for liquefaction without recovery. The unit power of liquefaction amounts to 10.04 kWh/kg (4.555 kWh/lb) with vent gas recovery compared with 10.340 kWh/kg (4.690 kWh/lb) without recovery.

8.4 Economics

Investment and other capital requirements for the hydrogen liquefaction complex with vent gas recovery are compared, in Table 27, to the liquefaction complex without recovery. Investment, in mid-1975 dollars, and based on 2000 AD technology, is $3,700,000 (1.5%) less for the system which recovers vent gas.

Annual operating costs for a 577.2 MTD (636.3 TPD) capacity are shown, Table 28, to favor the vent gas recovery system by $7,893,000 (5.91%) annually.

The bulk of the saving (80%) originates from the reduction in feedstock requirements followed by savings resulting from power reduction at 16%.

Table 29 contains a breakdown of unit costs for comparative cases with and without vent gas recovery at a capacity of 577.2 MTD (636.3 TPD). The total unit cost for hydrogen liquefaction with vent gas recovery is $0.9008/kg ($0.4086/lb) compared with $0.9429/kg ($0.4277/lb) without recovery, a reduction of 4.47%. The saving, amounts to $0.0421/kg (1.91¢/lb), which on an annual basis amounts to $8,872,000.

Unit costs, in this analysis, were derived from the discounted cash flow relationship for annual income which was presented in Section 6.2 as Equation 3. Annual income divided by total annual production produces the value for unit cost.

REFERENCES

1. "LH_2 Airport Requirements Study", NASA CR-2700, G. D. Brewer, Editor, Contract NAS 1-14137 by Lockheed-California Co. for NASA, Langley Research Center, Hampton, VA, October 1976.

2. "Survey Study of the Efficiency and Economics of Hydrogen Liquefaction", NASA CR-132631, Contract NAS 1-13395, by Linde Division, Union Carbide Corp. for NASA, Langley Research Center, Hampton, VA, April 8, 1975.

3. Gas Engineers Handbook, American Gas Association, C. George Segeler, Editor in Chief, Section 10, Industrial Press, NY, 1974

4. McAdams, W. A., Heat Transmission, 3rd Ed., p 172, Mc Graw Hill, NY, 1954.

TABLE 1

H_2 VENT LOSSES FOR SCHEDULED FLIGHTS

kg (LB)

AIRPORT	DISPLACE-MENT	HEATLEAK[1]	PIPING	PUMP	SAVED	TOTAL PER FUELING
DOMESTIC						
ORD	204(450)	65(143)	187(412)	49(109)	98(216)	407(898)
HNL	305(673)	"	"	74(163)	120(265)	511(1126)
JFK	265(585)	"	"	64(142)	111(245)	470(1037)
DFW	174(383)	"	"	42(93)	91(201)	376(830)
ATL	228(503)	"	"	55(122)	103(227)	432(953)
IAD	252(556)	"	"	61(135)	108(239)	457(1007)
MIA	276(608)	"	"	67(147)	113(250)	481(1060)
MCI	171(377)	"	"	41(91)	91(200)	373(823)
LAX	62(137)	"	"	15(33)	67(147)	262(578)
INTERNATIONAL						
TYO	650(1433)	65(143)	187(412)	157(347)	196(432)	863(1903)
LHR	564(1243)	"	"	137(301)	177(390)	775(1709)
CDG	587(1295)	"	"	142(314)	182(402)	799(1762)
FCO	659(1453)	"	"	160(352)	198(436)	873(1924)

(1) Heat leak boiloff during refuel - 30 min. refueling

TABLE 2

SUMMARY OF DAILY VENT LOSSES DURING FUELING

AVERAGE DAY DURING PEAK MONTH

AIRPORT	BLOCK FUEL PER FLIGHT kg	(lb)	FLIGHTS PER DAY	BLOCK FUEL PER DAY Mg	(klb)	VENT PER FUELING kg	(lb)	DAILY VENT Mg/day	(lb/day)
DOMESTIC:									
ORD	6985	(15400)	14	97.79	(215.6)	407	(898)	5.703	(12572)
HNL	10433	(23000)	10	104.33	(230.0)	511	(1126)	5.107	(11260)
JFK	9072	(20000)	9	81.65	(180.0)	470	(1037)	4.233	(9333)
DFW	5942	(13100)	9	53.48	(117.9)	376	(830)	3.388	(7470)
ATL	7802	(17200)	3	23.41	(51.6)	432	(953)	1.297	(2859)
IAD	8618	(19000)	3	25.85	(57.0)	457	(1007)	1.370	(3021)
MIA	9435	(20800)	2	18.87	(41.6)	481	(1060)	0.962	(2120)
MCI	5851	(12900)	2	11.70	(25.8)	373	(823)	0.747	(1646)
LAX	2132	(4700)	7	14.92	(32.9)	262	(578)	1.835	(4046)
TOTAL			59	432.00	(952.4)			24.642	(54327)
INTERNATIONAL:									
TYO	22225	(49000)	5	111.13	(245.0)	863	(1903)	4.316	(9515)
LHR	19280	(42500)	3	57.83	(127.5)	775	(1709)	2.326	(5127)
CDG	20095	(44300)	2	40.19	(88.6)	799	(1762)	1.598	(3524)
FCO	22545	(49700)	1	22.54	(49.7)	873	(1924)	0.873	(1924)
TOTAL			11	231.69	(510.8)			9.113	(20090)

AVERAGE LOSS: For 59 Daily domestic flights = 24.642/59 = 418 kg (921 lb)
For 11 Daily international flights = 9.113/11 = 828 kg (1826 lb)

AVERAGE OVERALL VENT RATE = (24.642 + 9.113)/24 = 1.406 Mg/hr (3100.7 lb/hr)
= 4.344 m^3/s (595,000 cfh)

TABLE 3A
FUELING SCHEDULE AND VENT GAS RELEASE
SAN FRANCISCO INTERNATIONAL AIRPORT
REF. NASA CR-2700
SI UNITS

NO	HOUR START	HOUR FINISH	TIME MIN.	NO PLANES REFUELING	TYPE OF FLIGHT	VENT GAS RELEASED KG.	VENT GAS OFFRATE M3/S	VENT GAS STORED M3 X 10^3
1	0000	0020	20	1	DOM	310.0	2.87	77.26
2	0020	0400	220	0	DOM	0.0	0.0	19.92
3	0400	0420	20	1	INT	414.2	3.84	19.31
4	0420	0440	20	2	INT	828.4	7.68	23.31
5	0440	0500	20	1	INT	414.2	3.84	22.70
6	0500	0600	60	0	DOM	0.0	0.0	7.06
7	0600	0720	80	1	DOM	1240.1	2.87	0.0
8	0720	0740	20	3	DOM	930.1	8.62	5.13
9	0740	0820	40	4	DOM	2480.2	11.49	22.28
10	0820	0920	60	3	DOM	2790.2	8.62	37.66
11	0920	1020	60	1	DOM	930.1	2.87	32.36
12	1020	1040	20	0	DOM	0.0	0.0	27.15
13	1040	1100	20	1	DOM	310.0	2.87	25.36
14	1100	1130	30	2	DOM	930.1	5.74	27.90
15	1130	1200	30	3	DOM	1395.1	8.62	35.59
16	1200	1400	120	2	BOTH	4083.6	6.31	49.72
17	1400	1430	30	3	BOTH	1758.5	10.86	61.45
18	1430	1520	50	2	BOTH	1913.5	7.09	69.69
19	1520	1530	10	0	DOM	0.0	0.0	67.08
20	1530	1545	15	2	DOM	465.0	5.74	68.34
21	1545	1620	35	2	BOTH	1448.5	7.67	75.33
22	1620	1645	25	0	DOM	0.0	0.0	68.81
23	1645	1745	60	1	DOM	930.1	2.87	63.51
24	1745	1915	90	2	BOTH	3153.6	6.49	75.11
25	1915	1930	15	0	DOM	0.0	0.0	71.20
26	1930	2000	30	3	BOTH	1758.5	10.86	82.94
27	2000	2015	15	0	DOM	0.0	0.0	79.03
28	2015	2045	30	1	BOTH	828.4	5.12	80.42
29	2045	2120	35	3	BOTH	1991.0	10.54	93.43
30	2120	2200	40	1	BOTH	983.4	4.56	93.94
31	2200	2215	15	2	DOM	465.0	5.74	95.20
32	2215	2320	65	0	DOM	0.0	0.0	78.26
33	2320	2345	25	2	DOM	775.0	5.74	80.36
34	2345	2400	15	1	DOM	232.5	2.87	79.03

AVERAGE FLOW RATE TO H2 LIQUEFIER, M3/S, = 4.34

DOM = DOMESTIC *** INT = INTERNATIONAL

TABLE 3B
FUELING SCHEDULE AND VENT GAS RELEASE
SAN FRANCISCO INTERNATIONAL AIRPORT

REF. NASA CR-2700

U.S. CUSTOMARY UNITS

NO	HOUR START	HOUR FINISH	TIME MIN.	NO PLANES REFUELING	TYPE OF FLIGHT	VENT GAS RELEASED LB.	VENT GAS OFFRATE MCFH	VENT GAS STORED MCF
1	0000	0020	20	1	DOM	683.	393.	2940.
2	0020	0400	220	0	DOM	0.	0.	758.
3	0400	0420	20	1	INT	913.	526.	735.
4	0420	0440	20	2	INT	1825.	1051.	887.
5	0440	0500	20	1	INT	913.	526.	864.
6	0500	0600	60	0	DOM	0.	0.	269.
7	0600	0720	80	1	DOM	2734.	393.	0.
8	0720	0740	20	3	DOM	2050.	1180.	195.
9	0740	0820	40	4	DOM	5468.	1574.	647.
10	0820	0920	60	3	DOM	6151.	1180.	1433.
11	0920	1020	60	1	DOM	2050.	393.	1231.
12	1020	1040	20	0	DOM	0.	0.	1033.
13	1040	1100	20	1	DOM	683.	393.	966.
14	1100	1130	30	2	DOM	2050.	787.	1061.
15	1130	1200	30	3	DOM	3076.	1180.	1354.
16	1200	1400	120	2	BOTH	9003.	864.	1891.
17	1400	1430	30	3	BOTH	3877.	1488.	2338.
18	1430	1520	50	2	BOTH	4219.	971.	2651.
19	1520	1530	10	0	DOM	0.	0.	2552.
20	1530	1545	15	2	DOM	1025.	787.	2600.
21	1545	1620	35	2	BOTH	3193.	1050.	2866.
22	1620	1645	25	0	DOM	0.	0.	2618.
23	1645	1745	60	1	DOM	2050.	393.	2416.
24	1745	1915	90	2	BOTH	6952.	889.	2858.
25	1915	1930	15	0	DOM	0.	0.	2709.
26	1930	2000	30	3	BOTH	3477.	1498.	3155.
27	2000	2015	15	0	DOM	0.	0.	3005.
28	2015	2045	30	1	BOTH	1825.	701.	3059.
29	2045	2120	35	3	BOTH	4389.	1444.	3555.
30	2120	2200	40	1	BOTH	2168.	624.	3574.
31	2200	2215	15	2	DOM	1025.	787.	3622.
32	2215	2320	65	0	DOM	0.	0.	2977.
33	2320	2345	25	2	DOM	1709.	787.	3057.
34	2345	2400	15	1	DOM	913.	393.	3007.

AVERAGE FLOW RATE TO H2 LIQUEFIER, MCFH, = 594.

DOM = DOMESTIC *** INT = INTERNATIONAL

TABLE 4A
FUELING SCHEDULE AND VENT GAS RELEASE
SAN FRANCISCO INTERNATIONAL AIRPORT
VARIABLE GAS WITHDRAWAL RATE FROM ACCUMULATOR
REF. NASA CR-2700
SI UNITS

NO	HOUR START	HOUR FINISH	TIME MIN.	NO PLANES REFUELING	TYPE OF FLIGHT	VENT GAS RELEASED KG.	VENT GAS OFFRATE M3/S	TO H2 LIQUEFIER M3/S	VENT GAS STORED M3 $\times 10^{-3}$
1	0000	0020	20	1	DOM	310.0	2.87	1.10	15.62
2	0020	0400	220	0	DOM	0.0	0.0	1.10	1.16
3	0400	0420	20	1	INT	414.2	3.84	1.10	4.45
4	0420	0440	20	2	INT	828.4	7.68	2.97	10.10
5	0440	0500	20	1	INT	414.2	3.84	2.97	11.14
6	0500	0600	60	0	DOM	0.0	0.0	2.97	0.46
7	0600	0720	80	1	DOM	1240.1	2.87	2.97	0.0
8	0720	0740	20	3	DOM	930.1	8.62	5.95	3.21
9	0740	0820	40	4	DOM	2480.2	11.49	5.95	16.51
10	0820	0920	60	3	DOM	2790.2	8.62	5.95	26.13
11	0920	1020	60	1	DOM	930.1	2.87	5.95	15.86
12	1020	1040	20	0	DOM	0.0	0.0	5.95	7.93
13	1040	1100	20	1	DOM	310.0	2.87	5.95	4.24
14	1100	1130	30	2	DOM	930.1	5.74	5.95	3.87
15	1130	1200	30	3	DOM	1395.1	8.62	5.95	8.68
16	1200	1400	120	2	BOTH	4083.6	6.31	5.95	11.27
17	1400	1430	30	3	BOTH	1758.5	10.86	5.95	20.12
18	1430	1520	50	2	BOTH	1913.5	7.09	5.95	23.56
19	1520	1530	10	0	DOM	0.0	0.0	5.95	19.99
20	1530	1545	15	2	DOM	465.0	5.74	5.95	19.81
21	1545	1620	35	2	BOTH	1448.5	7.67	5.95	23.43
22	1620	1645	25	0	DOM	0.0	0.0	5.95	14.51
23	1645	1745	60	1	DOM	930.1	2.87	5.95	3.44
24	1745	1915	90	2	BOTH	3153.6	6.49	5.95	6.40
25	1915	1930	15	0	DOM	0.0	0.0	5.95	1.05
26	1930	2000	30	3	BOTH	1758.5	10.86	5.95	9.89
27	2000	2015	15	0	DOM	0.0	0.0	5.95	4.54
28	2015	2045	30	1	BOTH	828.4	5.12	5.95	3.05
29	2045	2120	35	3	BOTH	1991.0	10.54	5.95	12.70
30	2120	2200	40	1	BOTH	983.4	4.56	5.95	9.36
31	2200	2215	15	2	DOM	465.0	5.74	5.95	9.18
32	2215	2320	65	0	DOM	0.0	0.0	1.10	4.91
33	2320	2345	25	2	DOM	775.0	5.74	1.10	11.89
34	2345	2400	15	1	DOM	232.5	2.87	1.10	13.49

DOM = DOMESTIC *** INT = INTERNATIONAL

PAGE IS
QUALITY

TABLE 4B
FUELING SCHEDULE AND VENT GAS RELEASE
SAN FRANCISCO INTERNATIONAL AIRPORT
VARIABLE GAS WITHDRAWAL RATE FROM ACCUMULATOR
REF. NASA CR-2700
U.S. CUSTOMARY UNITS

NO	HOUR START	HOUR FINISH	TIME MIN.	NO PLANES REFUELING	TYPE OF FLIGHT	VENT GAS RELEASED LB.	VENT GAS OFFRATE MCFH	TO H2 LIQUEFIER MCFH	VENT GAS STORED MCFH
1	0000	0020	20	1	DOM	683.	393.	150.	594.
2	0020	0400	220	0	DOM	0.	0.	150.	44.
3	0400	0420	20	1	INT	913.	526.	150.	169.
4	0420	0440	20	2	INT	1826.	1051.	407.	384.
5	0440	0500	20	1	INT	913.	526.	407.	424.
6	0500	0600	60	0	DOM	0.	0.	407.	17.
7	0600	0720	80	1	DOM	2734.	393.	407.	0.
8	0720	0740	20	3	DOM	2050.	1180.	814.	122.
9	0740	0820	40	4	DOM	5468.	1574.	814.	628.
10	0820	0920	60	3	DOM	6151.	1180.	814.	994.
11	0920	1020	60	1	DOM	2050.	393.	814.	573.
12	1020	1040	20	0	DOM	0.	0.	814.	302.
13	1040	1100	20	1	DOM	683.	393.	814.	161.
14	1100	1130	30	2	DOM	2050.	787.	814.	147.
15	1130	1200	30	3	DOM	3076.	1180.	814.	330.
16	1200	1400	120	2	BOTH	9603.	864.	814.	429.
17	1400	1430	30	3	BOTH	3877.	1488.	814.	766.
18	1430	1520	50	2	BOTH	4219.	971.	814.	896.
19	1520	1530	10	0	DOM	0.	0.	814.	761.
20	1530	1545	15	2	DOM	1025.	787.	814.	754.
21	1545	1620	35	2	BOTH	3193.	1050.	814.	891.
22	1620	1645	25	0	DOM	0.	0.	814.	552.
23	1645	1745	60	1	DOM	2050.	393.	814.	131.
24	1745	1915	90	2	BOTH	6952.	889.	814.	243.
25	1915	1930	15	0	DOM	0.	0.	814.	40.
26	1930	2000	30	3	BOTH	3877.	1488.	814.	376.
27	2000	2015	15	0	DOM	0.	0.	814.	173.
28	2015	2045	30	1	BOTH	1826.	701.	814.	116.
29	2045	2120	35	3	BOTH	4389.	1444.	814.	483.
30	2120	2200	40	1	BOTH	2168.	624.	814.	356.
31	2200	2215	15	2	DOM	1025.	787.	814.	349.
32	2215	2320	65	0	DOM	0.	0.	150.	187.
33	2320	2345	25	2	DOM	1709.	787.	150.	452.
34	2345	2400	15	1	DOM	513.	393.	150.	513.

DOM = DOMESTIC *** INT = INTERNATIONAL

TABLE 5
STEADY VENTING LOSSES NOT ASSOCIATED WITH REFUELING OPERATIONS

	SI	Customary
Block Fuel	7.680	(731.4)
Refueling Operations Loss @ 12.2%	0.937	(89.2)
SUBTOTAL	8.616	(820.6)
Supply Tank Displacement @ 3.156%	0.2719	(25.90)
Piping System Losses		
Total, 304.7 kg (671.7 lb) @ 70 refuelings/day	0.2469	(23.51)
Less piping system losses during refueling previously accounted for 186.9 kg (412 lb) @ 70 refuelings/day	0.1514	(14.42)
Net Piping System Loss	0.0954	(9.09)
Total Steady Venting Losses	0.3674	(34.99)

All tabulated values in following units

SI - kg/s

Customary - TPD

Venting Rate: 0.3674 kg/s = 4.085 m^3/s (0°C, 1 atm)

34.99 TPD = 559,500 cfh (70°F, 1 atm)

TABLE 6

VALUE OF H_2 VENT GAS

	SI $/kg		Customary ¢/lb
OPERATING COSTS			
Feedstock	0.3627		(16.45)
Electric Power @ $0.02/kwh	0.2066		(9.37)
Labor, Administration, Overhead	0.0108		(0.49)
Chemicals, Supplies, Water, Taxes and Insurance	0.0540		(2.45)
SUBTOTAL	0.6341		(28.76)
CAPITAL INVESTMENT			
Liquefaction and Storage Facility Investment	0.2600		(11.79)
Startup Costs	0.0040		(0.18)
Working Capital	0.0101		(0.46)
SUBTOTAL	0.2741		(12.43)
TOTAL	0.9082		(41.19)
Less Feedstock Cost	0.3627		(16.45)
Cost of Liquefaction	0.5455		(24.74)
Available Energy Ratio, Cold Gas/Liquid		0.3906	
Value of Refrigeration	0.2130		(9.66)
Total Value, H_2 Vent Gas	0.5757		(26.11)

TABLE 7

COMPARISON OF COLD

AND WARM COMPRESSION

OF VENT H_2 GAS

P_2		Cold Compressor Power				Warm Compressor Power
kPa	psia	Td °K	Compression kW	Refrigeration Loss, kW	Total Power kW	Compression kW
138	(20)	25.6	5	75	80	280
276	(40)	37.4	50	1460	1510	850
414	(60)	46.1	90	2160	2250	1100
552	(80)	52.1	110	2560	2670	1350
689	(100)	57.4	130	2870	3000	1590
827	(120)	62.3	150	3190	3340	1710
965	(140)	67.8	170	3430	3600	1820
1103	(160)	71.6	190	3520	3710	1930
1241	(180)	74.8	200	3660	3860	2050
1379	(200)	78.2	210	3800	4010	2160
1517	(220)	81.5	230	3920	4150	2230
1655	(240)	84.8	240	4040	4280	2290
1793	(260)	88.3	250	4170	4420	2360
1930	(280)	92.0	270	4300	4570	2430
2068	(300)	95.2	280	4410	4690	2490

TABLE 8

EVALUATION OF CONSTANT

PRESSURE COLD ACCUMULATOR

INVESTMENT		
Gas holders (4)		$2,500,000
Compressor		600,000
TOTAL INVESTMENT		$3,100,000
Annual Cost of Investment		$ 710,000
OPERATING COST*	kW	$/YEAR
Compressor Power	2708	$ 474,000
Refrigeration Loss	3670	$ 643,000
TOTAL OPERATING	6378	$1,117,000
TOTAL ANNUAL COST		$1,827,000

* Based on 24 hr/day, 365 day/year

Electricity at $0.02/kWh

Annual Investment at 22.9% of total investment

TABLE 9

EVALUATION OF CONSTANT VOLUME COLD ACCUMULATOR AT OPTIMUM STORAGE PRESSURE

CONDITIONS:

Storage Pressure	1655 kPa (240 PSIA)
No. Accumulators	6
Accumulator Diameter	3.66 m (12 ft)
Accumulator Length	30.48 m(100 ft)
Cold Compressor Power	246 kW (330 BHP)
Warm Compressor Power	652 kW (874 BHP)
Accumulator Storage Temperature	84.9°K

INVESTMENT:

Accumulators	$3,180,000
Cold Compressor	87,000
Warm Compressor	200,000
TOTAL INVESTMENT	$3,467,000
TOTAL ANNUAL INVESTMENT COST	$ 794,000

OPERATING COST:

	Power-kW	$/Year
Cold Compressor Power	246	43,100
Warm Compressor Power	652	114,200
Refrigeration Loss	8830	1,547,000
TOTAL OPERATING	9728	1,704,000
TOTAL ANNUAL COST		$2,498,000

TABLE 10

EVALUATION OF H_2 VENT GAS PIPELINE USED AS COLD ACCUMULATOR AT OPTIMUM PIPELINE PRESSURE

CONDITIONS:

Storage Pressure	1795 kPa	(260 PSIA)
Pipeline Diameter	80.0 cm	(31.5 in.)
Pipeline Volume	1682 m^3	(59500 cu ft)
Cold Compressor Power	245 kW	(328 BHP)
Warm Compressor Power	595 kW	(798 BHP)
Compressor Discharge Temperature	84.7°K	
Pipeline Discharge Temperature	87.2°K	

INVESTMENT

Incremental Pipeline Diameter	$6,150,000
Cold Compressors	87,000
Warm Compressor	187,000
TOTAL INVESTMENT	$6,424,000
TOTAL ANNUAL INVESTMENT COST	$1,472,000

OPERATING COST

	Power-KW	$/Year
Cold Compressor	245	42,900
Warm Compressor	595	104,300
Refrigeration Loss		
Due to compression	8815	1,544,000
Due to heat leak	192	33,700
TOTAL OPERATING	9847	$1,725,000
TOTAL ANNUAL COST		$3,197,000

TABLE 11

COMPRESSOR OPERATING SCHEDULE
FOR
COMPRESSING VENT GAS TO WARM CONSTANT VOLUME TANK

Period	Vent Gas Flow m^3/s	Compressors in Operation - No's	% of Capacity
0000-0020	2.87	3	94
00020-0400	0	None	--
0400-0420	3.84	1 & 3	100
0420-0440	7.68	2 & 4	100
0440-0500	3.84	1 & 3	100
0500-0600	0	None	--
0600-0720	2.87	3	94
0720-0740	8.62	3 & 4	94
0740-0820	11.49	1,2,3 & 4	100
0820-0920	8.62	3 & 4	94
0920-1020	2.87	3	94
1020-1040	0	None	--
1040-1100	2.87	3	94
1100-1130	5.74	4	94
1130-1200	8.62	3 & 4	94
1200-1400	6.31	1 & 4	91
1400-1430	10.86	1,2,3 & 4	94
1430-1520	7.09	2 & 4	92
1520-1530	0	None	--
1530-1545	5.74	4	94
1545-1620	7.67	2 & 4	100
1620-1645	0	None	--
1645-1745	2.87	3	94
1745-1915	6.49	1 & 4	94
1915-1930	0	None	--
1930-2000	10.86	1,2,3 & 4	94
2000-2015	0	None	--
2015-2045	5.12	1,2 & 3	95
2045-2120	10.54	2,3 & 4	98
2120-2200	4.56	2 & 3	100
2200-2215	5.74	4	94
2215-2320	0	None	--
2320-2345	5.74	4	94
2345-2400	2.87	3	94

TABLE 12

EVALUATION OF CONSTANT VOLUME WARM ACCUMULATOR

CONDITIONS		
Storage Pressure	4137 kPa	(600 PSIA)
No. Accumulators	1	
Accumulator Diameter	3.05 m	(10 ft)
Accumulator Length	15.25 m	(50 ft)
Compressor Power		
No. 1 Compressor	455 kW	(610 BHP)
No. 2 Compressor	911 kW	(1221 BHP)
No. 3 Compressor	1822 kW	(2442 BHP)
No. 4 Compressor	3643 kW	(4884 BHP)
Accumulator Storage Temperature	300°K	

INVESTMENT		
Accumulator		$ 92,000
Compressors		$2,140,000
	TOTAL INVESTMENT	$2,232,000
	TOTAL ANNUAL INVESTMENT COST	$ 511,300
OPERATING		
Compressor Power	2626 kW	$ 460,500
TOTAL ANNUAL COST		$ 972,000

TABLE 13

EVALUATION OF CONSTANT PRESSURE

WARM ACCUMULATOR

CONSTANT WITHDRAWAL RATE

CONDITIONS			
Storage Pressure		101 kPa	(14.7 PSIA)
No. Accumulators		4	
Accumulator Capacity (ea)		26,300 m^3	(1000 MCF)
Accumulator Storage Temperature		300°K	
Compressor Discharge Pressure		4137 kPa	(6000 PSIA)
Compressor Power		2628 kW	(3523 BHP)
INVESTMENT			
Accumulators			\$3,670,000
Compressors			\$ 580,000
	TOTAL INVESTMENT		\$4,250,000
	TOTAL ANNUAL INVESTMENT COST		\$ 973,600
OPERATING			
Compressor Power	2628 kW		\$ 460,500
TOTAL ANNUAL COST			\$1,434,100

TABLE 14

EVALUATION OF CONSTANT PRESSURE WARM ACCUMULATOR VARIABLE WITHDRAWAL RATE

CONDITIONS			
Storage Pressure		101 kPa	(14.7 PSIA)
No. Accumulators		1	
Accumulator Capacity		26,300 m^3	(1000 MCF)
Accumulator Storage Temperature		300°K	
Compressor Power			
No. 1 Compressor		664 kW	(890 BHP)
No. 2 Compressor		1798 kW	(2410 BHP)
No. 3 Compressor		3596 kW	(4820 BHP)
Average		2628 kW	(3523 BHP)
INVESTMENT			
Accumulator			$ 917,000
Compressors			1,370,000
	TOTAL INVESTMENT		$2,287,000
	TOTAL ANNUAL INVESTMENT COST		$ 523,900
OPERATING			
Compressor Power	2628 kW		460,500
TOTAL ANNUAL COST			$ 984,400

TABLE 15A

USE OF VENT GAS

HEAT EXCHANGERS

Condition A - AVERAGE VENT GAS FLOW RATE

Stream 1 Flow = 8.43 m^3/s (1154.5 MCFH)

Stream 3 Flow = 8.43 m^3/s (1154.5 MCFH)

Pressure -Stream 1		Temp. - Stream 2	LH_2 Flow-Stream 7		Vapor Temp. Stream 6
kPa	(PSIA)	°K	m^3/s	(MCFH)	°K
689	(100)	31.4	0	0	24.7
1379	(200)	36.1	0	0	23.1
2068	(300)	39.6	0	0	21.9
2758	(400)	42.1	0	0	20.8
3447	(500)	43.7	0.1245	(17.05)	20.5
4137	(600)	44.9	0.2812	(38.51)	20.5

TABLE 15B

USE OF VENT GAS

HEAT EXCHANGERS

Condition B - MINIMUM VENT GAS FLOW RATE

Stream 1 Flow = 8.43 m^3/s (1154.5 MCFH)

Stream 3 Flow = 4.09 m^3/s (559.5 MCFH)

Pressure-Stream 1 kPa	(PSIA)	Temp. Stream 2 °K	Vapor Temp. Stream 6 °K
689	(100)	166.1	166.0
1379	(200)	166.5	166.3
2068	(300)	166.8	166.7
2758	(400)	167.2	167.0
3447	(500)	167.6	167.3
4137	(600)	168.0	167.6

NOTE: There was no liquefaction of the compressed Hydrogen Stream; i.e. Flow, Stream 7 = 0.0

TABLE 15C

USE OF VENT GAS HEAT EXCHANGERS

Condition C - MAXIMUM VENT GAS FLOW RATE

Stream 1 Flow = 8.43 m^3/s (1154.5 MCFH)

Stream 3 Flow = 15.58 m^3/s (2133.5 MCFH)

Pressure-Stream 1 kPa	(PSIA)	Temp. Stream 2 °K	LH_2 Flow - Stream 7 m^3/s	(MCFH)
689	(100)	28.9	1.539	(210.72)
1379	(200)	33.2	3.562	(487.90)
2068	(300)	32.6	4.933	(675.65)
2758	(400)	29.8	5.930	(812.11)
3447	(500)	26.5	6.671	(913.60)
4137	(600)	25.1	6.853	(938.59)

TABLE 16

H2 LIQUEFIER - NO VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR NITROGEN STREAMS
SI UNITS

STREAM NUMBER	080	081	082	083	084	085
FLOW, CMH(STP)	81942.8	81942.8	81942.8	19052.8	19052.8	15011.4
FLOW, GM.MOLE/SEC	1015.54	1015.54	1015.54	236.127	236.127	186.041
PRESSURE, KILOPASCAL	620.528	620.528	594.328	4136.86	153.064	153.064
TEMPERATURE, DEG K	97.0000	300.000	299.949	99.9000	81.1034	81.1034
ENTHALPY, JOULES/GM.MOLE	6693.07	12908.8	12908.8	2191.67	2191.67	1037.40
ENTROPY, JOULES/GM.MOLE-DEG C	73.4695	108.626	108.985	25.5956	28.6789	14.4487
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	0.78789	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)	7032.78	7032.78	7978.64	7978.64	4041.33	19052.8
FLOW, GM.MOLE/SEC	87.1595	87.1595	98.8818	98.8818	50.0854	236.127
PRESSURE, KILOPASCAL	153.064	153.064	153.064	153.064	153.064	153.064
TEMPERATURE, DEG K	81.1034	81.1034	81.1034	81.1034	81.1034	81.1034
ENTHALPY, JOULES/GM.MOLE	1037.40	6479.17	1037.40	6479.17	6479.17	6479.17
ENTROPY, JOULES/GM.MOLE-DEG C	14.4486	81.5364	14.4486	81.5364	81.5364	81.5364
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)	5057.23	5057.23	24110.0	24110.0	24110.0
FLOW, GM.MOLE/SEC	62.6759	62.6759	298.802	298.802	298.802
PRESSURE, KILOPASCAL	153.064	153.064	153.064	153.064	103.421
TEMPERATURE, DEG K	90.6000	90.6000	83.0827	300.000	299.902
ENTHALPY, JOULES/GM.MOLE	6772.16	6772.16	6540.63	12935.7	12935.7
ENTROPY, JOULES/GM.MOLE-DEG C	84.9531	84.9531	82.2849	128.339	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 16
H2 LIQUEFIER - NO VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	001	002	003	004	005	006
FLOW, CMH(STP)	91166.9	63090.1	28076.8	457028.	457028.	39124.6
FLOW, GM.MOLE/SEC	1129.86	781.895	347.964	5664.09	5664.09	484.884
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)	496152.	91166.9	496152.	91166.9	91166.9	91166.9
FLOW, GM.MOLE/SEC	6148.96	1129.86	6148.96	1129.86	1129.86	1129.86
PRESSURE, KILOPASCAL	4136.86	4136.86	4128.58	4128.58	4128.58	4051.36
TEMPERATURE, DEG K	85.5000	308.000	85.4899	85.4899	81.3034	61.2016
ENTHALPY, JOULES/GM.MOLE	2697.04	0.0	2697.04	2697.04	2308.36	2308.35
ENTROPY, JOULES/GM.MOLE-DEG C	77.6278	0.0	77.6428	77.6428	73.8113	73.9502
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	0.54554	0.54554
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.45446	0.45446

STREAM NUMBER	013	014	015	016	017	018
FLOW, CMH(STP)	91166.9	91166.9	91166.9	91166.9	91166.9	91166.9
FLOW, GM.MOLE/SEC	1129.86	1129.86	1129.86	1129.86	1129.86	1129.86
PRESSURE, KILOPASCAL	4051.36	4005.16	4005.16	3958.97	930.792	930.792
TEMPERATURE, DEG K	40.9798	40.9193	28.0000	28.0252	28.6783	29.0946
ENTHALPY, JOULES/GM.MOLE	457.964	457.962	-220.085	-220.086	-220.089	-220.089
ENTROPY, JOULES/GM.MOLE-DEG C	41.9483	42.0137	23.3868	23.4366	26.9020	26.9038
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	0.20371	0.20371	0.04132	0.04132	0.04132	0.03032
PARA -HYDROGEN	0.79629	0.79629	0.95868	0.95868	0.95868	0.96968

TABLE 16
H2 LIQUEFIER - NO VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	019	020	021	022	023	024
FLOW, CMH(STP)	91166.9	404985.	404985.	404985.	149500.	149500.
FLOW, GM.MOLE/SEC	1129.86	5019.11	5019.11	5019.11	1852.81	1852.81
PRESSURE, KILOPASCAL	930.792	4128.58	4128.58	4114.79	4114.79	335.775
TEMPERATURE, DEG K	20.5509	85.4899	81.3034	81.2849	81.2849	38.9802
ENTHALPY, JOULES/GM.MOLE	-451.402	2697.04	2589.83	2589.83	2589.83	1820.78
ENTROPY, JOULES/GM.MOLE-DEG C	17.6649	77.6428	76.3775	76.4021	76.4025	82.0680
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	0.03032	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.96968	0.25000	0.25000	0.25000	0.25000	0.25000
STREAM NUMBER	025	026	027	028	029	030
FLOW, CMH(STP)	255485.	255485.	240933.	240933.	240933.	14552.3
FLOW, GM.MOLE/SEC	3166.31	3166.31	2985.95	2985.95	2985.95	180.352
PRESSURE, KILOPASCAL	4114.79	4114.79	4114.79	4114.79	348.185	4114.79
TEMPERATURE, DEG K	81.2849	58.7669	58.7669	58.7668	26.0000	58.7669
ENTHALPY, JOULES/GM.MOLE	2589.83	1957.73	1957.73	1957.73	1492.12	1957.73
ENTROPY, JOULES/GM.MOLE-DEG C	76.4021	67.2211	67.2211	67.2210	71.9989	67.2211
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)	14552.3	14552.3	50756.7	50756.7	190772.	190772.
FLOW, GM.MOLE/SEC	180.352	180.352	629.043	629.043	2364.29	2364.29
PRESSURE, KILOPASCAL	4114.79	111.006	348.185	348.185	335.774	335.775
TEMPERATURE, DEG K	29.5000	20.5009	26.0000	39.0678	38.9798	77.7849
ENTHALPY, JOULES/GM.MOLE	822.009	822.009	1492.12	1820.77	1820.77	2670.89
ENTROPY, JOULES/GM.MOLE-DEG C	40.7975	47.9247	71.9989	81.9024	82.0678	97.8766
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	0.69667	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 16
H2 LIQUEFIER - NO VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	037	038	039	040	041	042
FLOW, CMH(STP)	190772.	190176.	190176.	190176.	-9485.39	199661.
FLOW, GM.MOLE/SEC	2364.29	2356.91	2356.91	2356.91	-117.555	2474.47
PRESSURE, KILOPASCAL	317.159	348.185	348.185	335.775	335.775	335.775
TEMPERATURE, DEG K	77.7429	26.0000	39.0678	38.9798	38.9798	38.9798
ENTHALPY, JOULES/GM.MOLE	2670.89	1492.12	1820.77	1820.77	1820.77	1820.77
ENTROPY, JOULES/GM.MOLE-DEG C	98.3128	71.9989	81.9024	82.0677	82.0677	82.0677
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)	199661.	199661.	390433.	390433.	63090.1	39124.6
FLOW, GM.MOLE/SEC	2474.47	2474.47	4838.76	4838.76	781.895	484.884
PRESSURE, KILOPASCAL	335.775	317.159	317.159	317.159	4136.86	4136.86
TEMPERATURE, DEG K	77.7016	77.6596	77.7003	300.000	85.5000	85.5000
ENTHALPY, JOULES/GM.MOLE	2669.10	2669.10	2669.98	8505.89	2697.04	2697.04
ENTROPY, JOULES/GM.MOLE-DEG C	97.8514	98.2874	98.2998	132.618	77.6278	77.6279
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, CMH(STP)	23965.4	23965.4	23965.4	23965.4	38517.8	38517.8
FLOW, GM.MOLE/SEC	297.011	297.011	297.011	297.011	477.363	477.363
PRESSURE, KILOPASCAL	4136.86	4127.20	4127.20	111.006	111.006	111.006
TEMPERATURE, DEG K	85.5000	85.4882	33.2811	20.5009	20.5009	20.5009
ENTHALPY, JOULES/GM.MOLE	2697.04	2697.04	928.872	928.871	888.497	1443.96
ENTROPY, JOULES/GM.MOLE-DEG C	77.6278	77.6453	44.1913	53.1488	51.1750	78.3294
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.C. LIQ	0.57697	0.62219	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 16
H2 LIQUEFIER - NO VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	055	056	057	058	059	060
FLOW, CMH(STP)	38517.8	38517.8	38517.8	38517.8	38517.8	428951.
FLOW, GM.MOLE/SEC	477.363	477.363	477.363	477.363	477.363	5316.12
PRESSURE, KILOPASCAL	111.006	107.558	107.558	103.421	289.580	289.580
TEMPERATURE, DEG K	71.3882	71.3790	300.000	300.002	308.000	300.730
ENTHALPY, JOULES/GM.MOLE	2544.10	2544.10	8503.40	8503.39	8732.93	8526.27
ENTROPY, JOULES/GM.MOLE-DEG C	105.385	105.580	141.674	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)	428951.	23965.4	23965.4	14552.3	91166.9	2194.86
FLOW, GM.MOLE/SEC	5316.12	297.011	297.011	180.352	1129.86	27.2015
PRESSURE, KILOPASCAL	4136.86	111.006	111.006	111.006	101.353	101.353
TEMPERATURE, DEG K	308.000	20.5009	20.5009	20.5009	20.2337	20.2337
ENTHALPY, JOULES/GM.MOLE	8779.20	551.206	1443.96	1443.96	-451.402	422.916
ENTROPY, JOULES/GM.MOLE-DEG C	112.301	34.6863	78.3294	78.3294	18.8048	62.0292
LIQUID FRACTION (MOLE L/F)	S.H. VAP	SAT. LIQ	SAT. VAP	SAT. VAP	0.97592	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.03032	0.03032
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.96968	0.96968

STREAM NUMBER	067
FLOW, CMH(STP)	88972.0
FLOW, GM.MOLE/SEC	1102.66
PRESSURE, KILOPASCAL	101.353
TEMPERATURE, DEG K	20.2337
ENTHALPY, JOULES/GM.MOLE	-472.970
ENTROPY, JOULES/GM.MOLE-DEG C	17.7385
LIQUID FRACTION (MOLE L/F)	SAT. LIQ
COMPOSITION, MOLE FRACTION:	
ORTHO-HYDROGEN	0.03032
PARA -HYDROGEN	0.96968

TABLE 17
H2 LIQUEFIER - NO VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR NITROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	080	081	082	083	084	085
FLOW, CFH(NTP)	3117474.	3117474.	3117474.	724853.	724853.	571102.
FLOW, LB.MOLE/HR	8059.86	8059.86	8059.86	1874.02	1874.02	1476.52
PRESSURE, PSIA	90.0000	90.0000	86.2000	600.000	22.2000	22.2000
TEMPERATURE, DEG K	97.0000	300.000	299.949	99.9000	81.1034	81.1034
ENTHALPY, BTU/LB.MOLE	2879.48	5553.63	5553.63	942.897	942.896	446.308
ENTROPY, BTU/LB.MOLE-DEG K	31.6080	46.7331	46.8873	11.0117	12.3382	6.21608
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	0.78789	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, CFH(NTP)	267559.	267559.	303543.	303543.	153750.	724853.
FLOW, LB.MOLE/HR	691.742	691.742	784.776	784.776	397.504	1874.02
PRESSURE, PSIA	22.2000	22.2000	22.2000	22.2000	22.2000	22.2000
TEMPERATURE, DEG K	81.1034	81.1034	81.1034	81.1034	81.1034	81.1034
ENTHALPY, BTU/LB.MOLE	446.310	2787.46	446.310	2787.46	2787.46	2787.46
ENTROPY, BTU/LB.MOLE-DEG K	6.21606	35.0785	6.21606	35.0785	35.0785	35.0785
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, CFH(NTP)	192400.	192400.	917253.	917253.	917253.
FLOW, LB.MOLE/HR	497.428	497.428	2371.45	2371.45	2371.45
PRESSURE, PSIA	22.2000	22.2000	22.2000	22.2000	15.0000
TEMPERATURE, DEG K	90.6000	90.6000	83.0827	300.000	299.902
ENTHALPY, BTU/LB.MOLE	2913.51	2913.51	2813.90	5565.20	5565.20
ENTROPY, BTU/LB.MOLE-DEG K	36.5484	36.5484	35.4005	51.7720	53.1731
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 17
H2 LIQUEFIER - NO VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	001	002	003	004	005	006
FLOW, CFH(NTP)	3468400.	2400232.	1068168.	17387392.	17387392.	1488478.
FLOW, LB.MOLE/HR	8967.14	6205.52	2761.62	44953.1	44953.1	3848.29
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)	18875856.	3468400.	18875856.	3468400.	3468400.	3468400.
FLOW, LB.MOLE/HR	48801.3	8967.14	48801.3	8967.14	8967.14	8967.14
PRESSURE, PSIA	600.000	600.000	598.800	598.800	598.800	587.600
TEMPERATURE, DEG K	85.5000	308.000	85.4899	85.4899	81.3034	81.2016
ENTHALPY, BTU/LB.MOLE	1160.32	0.0	1160.32	1160.32	993.099	993.097
ENTROPY, BTU/LB.MOLE-DEG K	33.3969	0.0	33.4034	33.4034	31.7550	31.8147
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	0.54554	0.54554
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.45446	0.45446

STREAM NUMBER	013	014	015	016	017	018
FLOW, CFH(NTP)	3468400.	3468400.	3468400.	3468400.	3468400.	3468400.
FLOW, LB.MOLE/HR	8967.14	8967.14	8967.14	8967.14	8967.14	8967.14
PRESSURE, PSIA	587.600	580.899	580.899	574.199	135.000	135.000
TEMPERATURE, DEG K	40.9798	40.9193	28.0000	28.0252	28.6783	29.0946
ENTHALPY, BTU/LB.MOLE	197.025	197.024	-94.6846	-94.6852	-94.6862	-94.6862
ENTROPY, BTU/LB.MOLE-DEG K	18.0469	18.0751	10.0615	10.0829	11.5737	11.5745
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	0.20371	0.20371	0.04132	0.04132	0.04132	0.03032
PARA -HYDROGEN	0.79629	0.79629	0.95868	0.95868	0.95868	0.96968

TABLE 17
H2 LIQUEFIER - NO VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	019	020	021	022	023	024
FLOW, CFH(NTP)	3468400.	15407471.	15407471.	15407471.	5687671.	5687671.
FLOW, LB.MOLE/HR	8967.14	39834.2	39834.2	39834.2	14704.8	14704.8
PRESSURE, PSIA	135.000	598.800	598.800	596.800	596.800	48.7000
TEMPERATURE, DEG K	20.5509	85.4899	81.3034	81.2849	81.2849	38.9802
ENTHALPY, BTU/LB.MOLE	-194.202	1160.32	1114.19	1114.19	1114.19	783.333
ENTROPY, BTU/LB.MOLE-DEG K	7.59975	33.4034	32.8590	32.8696	32.8698	35.3072
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	0.03032	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.96968	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	025	026	027	028	029	030
FLOW, CFH(NTP)	9719800.	9719800.	9166164.	9166164.	9166164.	553637.
FLOW, LB.MOLE/HR	25129.4	25129.4	23698.0	23698.0	23698.0	1431.36
PRESSURE, PSIA	596.800	596.800	596.800	596.800	50.5000	596.800
TEMPERATURE, DEG K	81.2849	58.7669	58.7669	58.7668	26.0000	58.7669
ENTHALPY, BTU/LB.MOLE	1114.19	842.253	842.253	842.251	641.938	842.253
ENTROPY, BTU/LB.MOLE-DEG K	32.8696	28.9198	28.9198	28.9197	30.9753	28.9198
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, CFH(NTP)	553637.	553637.	1931013.	1931013.	7257817.	7257817.
FLOW, LB.MOLE/HR	1431.36	1431.36	4992.41	4992.41	18764.2	18764.2
PRESSURE, PSIA	596.800	16.1000	50.5000	50.5000	48.7000	48.7000
TEMPERATURE, DEG K	29.5000	20.5009	26.0000	39.0678	38.9798	77.7849
ENTHALPY, BTU/LB.MOLE	353.644	353.644	641.938	783.327	783.329	1149.07
ENTROPY, BTU/LB.MOLE-DEG K	17.5519	20.6181	30.9753	35.2359	35.3071	42.1084
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	0.69667	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 17
H2 LIQUEFIER - NO VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	037	038	039	040	041	042
FLOW, CFH(NTP)	7257817.	7235151.	7235151.	7235151.	-360867.	7596018.
FLOW, LB.MOLE/HR	18764.2	18705.6	18705.6	18705.6	-932.979	19638.6
PRESSURE, PSIA	46.0000	50.5000	50.5000	48.7000	48.7000	1.7000
TEMPERATURE, DEG K	77.7429	26.0000	39.0678	38.9798	38.9798	.9798
ENTHALPY, BTU/LB.MOLE	1149.07	641.938	783.327	783.329	783.329	783.329
ENTROPY, BTU/LB.MOLE-DEG K	42.2960	30.9753	35.2359	35.3071	35.3071	35.3071
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)	7596018.	7596018.	14853835.	14853835.	2400232.	1488478.
FLOW, LB.MOLE/HR	19638.6	19638.6	38402.9	38402.9	6205.52	3848.29
PRESSURE, PSIA	48.7000	46.0000	46.0000	46.0000	600.000	600.000
TEMPERATURE, DEG K	77.7016	77.6596	77.7003	300.000	85.5000	85.5000
ENTHALPY, BTU/LB.MOLE	1148.30	1148.30	1148.67	3659.39	1160.32	1160.32
ENTROPY, BTU/LB.MOLE-DEG K	42.0975	42.2851	42.2904	57.0546	33.3969	33.3970
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, CFH(NTP)	911753.	911753.	911753.	911753.	1465390.	1465390.
FLOW, LB.MOLE/HR	2357.23	2357.23	2357.23	2357.23	3788.60	3788.60
PRESSURE, PSIA	600.000	598.600	598.600	16.1000	16.1000	16.1000
TEMPERATURE, DEG K	85.5000	85.4882	33.2811	20.5009	20.5009	20.5009
ENTHALPY, BTU/LB.MOLE	1160.32	1160.32	399.618	399.618	382.248	621.220
ENTROPY, BTU/LB.MOLE-DEG K	33.3969	33.4045	19.0119	22.8656	22.0164	33.6988
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.C. LIQ	0.57697	0.62219	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 17
H2 LIQUEFIER - NO VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	055	056	057	058	059	060
FLOW, CFH(NTP)	1465390.	1465390.	1465390.	1465390.	1465390.	16319225.
FLOW, LB.MOLE/HR	3788.60	3788.60	3788.60	3788.60	3788.60	42191.5
PRESSURE, PSIA	16.1000	15.6000	15.6000	15.0000	42.0000	42.0000
TEMPERATURE, DEG K	71.3862	71.3790	300.000	300.002	308.000	306.730
ENTHALPY, BTU/LB.MOLE	1094.52	1094.52	3658.32	3658.32	3757.07	3668.16
ENTROPY, BTU/LB.MOLE-DEG K	45.3386	45.4226	60.9507	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)	16319225.	911754.	911753.	553636.	3468400.	83502.3
FLOW, LB.MOLE/HR	42191.5	2357.23	2357.23	1431.36	8967.14	215.885
PRESSURE, PSIA	600.000	16.1000	16.1000	16.1000	14.7000	14.7000
TEMPERATURE, DEG K	308.000	20.5009	20.5009	20.5009	20.2337	20.2337
ENTHALPY, BTU/LB.MOLE	3776.98	237.139	621.220	621.220	-194.201	181.946
ENTROPY, BTU/LB.MOLE-DEG K	48.3138	14.9227	33.6988	33.6988	8.09017	26.6861
LIQUID FRACTION (MOLE L/F)	S.H. VAP	SAT. LIQ	SAT. VAP	SAT. VAP	0.97592	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.03032	0.03032
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.96968	0.96968

STREAM NUMBER	067
FLOW, CFH(NTP)	3384897.
FLOW, LB.MOLE/HR	8751.25
PRESSURE, PSIA	14.7000
TEMPERATURE, DEG K	20.2337
ENTHALPY, BTU/LB.MOLE	-203.481
ENTROPY, BTU/LB.MOLE-DEG K	7.63142
LIQUID FRACTION (MOLE L/F)	SAT. LIQ
COMPOSITION, MOLE FRACTION:	
ORTHO-HYDROGEN	0.03032
PARA -HYDROGEN	0.96968

TABLE 18
H2 LIQUEFIER - WITH VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR NITROGEN STREAMS
SI UNITS

STREAM NUMBER	080	081	082	083	084	085
FLOW, CMH(STP)	73505.1	73505.1	73505.1	18728.9	18728.9	14756.3
FLOW, GM.MOLE/SEC	910.972	910.972	910.972	232.113	232.113	182.879
PRESSURE, KILOPASCAL	620.528	620.528	594.328	4136.86	153.064	153.064
TEMPERATURE, DEG K	97.0000	300.000	299.949	99.9000	81.1034	81.1034
ENTHALPY, JOULES/GM.MOLE	6693.07	12908.8	12908.8	2191.67	2191.67	1037.40
ENTROPY, JOULES/GM.MOLE-DEG C	73.4695	108.626	108.985	25.5956	28.6788	14.4487
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	0.78789	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)	7032.76	7032.76	7723.50	7723.50	3972.63	18728.9
FLOW, GM.MOLE/SEC	87.1592	87.1592	95.7197	95.7197	49.2341	232.113
PRESSURE, KILOPASCAL	153.064	153.064	153.064	153.064	153.064	153.064
TEMPERATURE, DEG K	81.1034	81.1034	81.1034	81.1034	81.1034	81.1034
ENTHALPY, JOULES/GM.MOLE	1037.40	6479.17	1037.40	6479.17	6479.17	6479.17
ENTROPY, JOULES/GM.MOLE-DEG C	14.4486	81.5364	14.4486	81.5364	81.5364	81.5364
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)	5057.23	5057.23	23786.1	23786.1	23786.1
FLOW, GM.MOLE/SEC	62.6759	62.6759	294.789	294.789	294.789
PRESSURE, KILOPASCAL	153.064	153.064	153.064	153.064	103.421
TEMPERATURE, DEG K	90.6000	90.6000	83.1097	300.000	299.902
ENTHALPY, JOULES/GM.MOLE	6772.16	6772.16	6541.46	12935.7	12935.7
ENTROPY, JOULES/GM.MOLE-DEG C	84.9531	84.9531	82.2951	120.339	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

ORIGINAL PAGE IS OF POOR QUALITY

TABLE 18
H2 LIQUEFIER - WITH VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	001	002	003	004	005	006
FLOW, CMH(STP)	83580.3	62746.1	20834.3	444419.	444419.	38780.7
FLOW, GM.MOLE/SEC	1035.84	777.632	258.206	5507.82	5507.82	480.622
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	77.6279
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)	483200.	83580.3	483200.	91166.5	91166.5	91166.5
FLOW, GM.MOLE/SEC	5988.45	1035.84	5988.45	1129.85	1129.86	1129.86
PRESSURE, KILOPASCAL	4136.86	4136.86	4128.58	4128.58	4128.58	4051.36
TEMPERATURE, DEG K	85.5000	308.000	85.4899	85.4899	81.3034	81.2016
ENTHALPY, JOULES/GM.MOLE	2697.04	0.0	2697.04	2697.04	2308.36	2308.35
ENTROPY, JOULES/GM.MOLE-DEG C	77.6278	0.0	77.6428	77.6428	73.8113	73.9502
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	0.54554	0.54554
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.45446	0.45446

STREAM NUMBER	013	014	015	016	017	018
FLOW, CMH(STP)	91166.5	91166.5	91166.5	91166.5	91166.5	91166.5
FLOW, GM.MOLE/SEC	1129.86	1129.86	1129.86	1129.86	1129.86	1129.86
PRESSURE, KILOPASCAL	4051.36	4005.16	4005.16	3958.97	930.792	930.792
TEMPERATURE, DEG K	40.9798	40.9193	28.0000	28.0252	28.6783	29.0946
ENTHALPY, JOULES/GM.MOLE	457.964	457.962	-220.085	-220.086	-220.089	-220.089
ENTROPY, JOULES/GM.MOLE-DEG C	41.9483	42.0137	23.3868	23.4366	26.9020	26.9038
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	0.20371	0.20371	0.04132	0.04132	0.04132	0.03032
PARA -HYDROGEN	0.79629	0.79629	0.95868	0.95868	0.95868	0.96968

TABLE 18
H2 LIQUEFIER - WITH VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	019	020	021	022	023	024
FLOW, CMH(STP)	91166.5	392033.	392033.	392033.	143794.	143794.
FLOW, GM.MOLE/SEC	1129.86	4858.59	4858.59	4858.59	1782.09	1782.09
PRESSURE, KILOPASCAL	930.792	4128.58	4128.58	4114.79	4114.79	335.775
TEMPERATURE, DEG K	20.5509	85.4899	81.3034	81.2849	81.2849	38.9802
ENTHALPY, JOULES/GM.MOLE	-451.402	2697.04	2589.83	2589.83	2589.83	1820.78
ENTROPY, JOULES/GM.MOLE-DEG C	17.6649	77.6428	76.3775	76.4021	76.4025	82.0680
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	0.03032	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.96968	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	025	026	027	028	029	030
FLOW, CMH(STP)	248239.	248239.	233687.	233687.	233687.	14552.3
FLOW, GM.MOLE/SEC	3076.51	3076.51	2896.15	2896.15	2896.15	180.351
PRESSURE, KILOPASCAL	4114.79	4114.79	4114.79	4114.79	348.185	4114.79
TEMPERATURE, DEG K	81.2849	58.7669	58.7669	58.7668	26.0000	58.7669
ENTHALPY, JOULES/GM.MOLE	2589.83	1957.73	1957.73	1957.73	1492.12	1957.73
ENTROPY, JOULES/GM.MOLE-DEG C	76.4021	67.2211	67.2211	67.2210	71.9989	67.2211
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)	14552.3	14552.3	43511.6	43511.6	177820.	177820.
FLOW, GM.MOLE/SEC	180.351	180.351	539.252	539.252	2203.78	2203.78
PRESSURE, KILOPASCAL	4114.79	111.006	348.185	348.185	335.774	335.775
TEMPERATURE, DEG K	29.5000	20.5009	26.0000	39.0678	38.9798	77.7849
ENTHALPY, JOULES/GM.MOLE	822.009	822.009	1492.12	1820.77	1820.77	2670.89
ENTROPY, JOULES/GM.MOLE-DEG C	40.7975	47.9247	71.9989	81.9024	82.0678	97.8766
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	0.69667	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 18
H2 LIQUEFIER - WITH VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	037	038	039	040	041	042
FLOW, CMH(STP)	177820.	190175.	190175.	190175.	-9485.36	199661.
FLOW, GM.MOLE/SEC	2203.78	2356.90	2356.90	2356.90	-117.555	2474.46
PRESSURE, KILOPASCAL	317.159	348.185	348.185	335.775	335.775	335.775
TEMPERATURE, DEG K	77.7429	26.0000	39.0678	38.9798	38.9798	38.9798
ENTHALPY, JOULES/GM.MOLE	2670.89	1492.12	1820.77	1820.77	1820.77	1820.77
ENTROPY, JOULES/GM.MOLE-DEG C	98.3128	71.9989	81.9024	82.0677	82.0677	82.0677
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)	199661.	199661.	377481.	377481.	62746.1	38780.7
FLOW, GM.MOLE/SEC	2474.46	2474.46	4678.24	4678.24	777.632	480.622
PRESSURE, KILOPASCAL	335.775	317.159	317.159	317.159	4136.86	4136.86
TEMPERATURE, DEG K	77.7016	77.6596	77.6989	300.000	85.5000	85.5000
ENTHALPY, JOULES/GM.MOLE	2669.10	2669.10	2669.95	8505.89	2697.04	2697.04
ENTROPY, JOULES/GM.MOLE-DEG C	97.8514	98.2874	98.2994	132.618	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	0.79000	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)	23965.4	23965.4	23965.4	23965.4	38517.6	38517.6
FLOW, GM.MOLE/SEC	297.010	297.010	297.010	297.010	477.361	477.361
PRESSURE, KILOPASCAL	4136.86	4127.20	4127.20	111.006	111.006	111.006
TEMPERATURE, DEG K	85.5000	85.4882	33.2811	20.5009	20.5009	20.5009
ENTHALPY, JOULES/GM.MOLE	2697.04	2697.04	928.872	928.871	888.497	1443.96
ENTROPY, JOULES/GM.MOLE-DEG C	77.6278	77.6453	44.1913	53.1488	51.1751	78.3294
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.C. LIQ	0.57697	0.62219	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 18
H2 LIQUEFIER - WITH VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	055	056	057	058	059	060
FLOW, CMH(STP)	38517.6	38517.6	38517.6	38517.6	38517.6	423585.
FLOW, GM.MOLE/SEC	477.361	477.361	477.361	477.361	477.361	5249.63
PRESSURE, KILOPASCAL	111.006	107.558	107.558	103.421	289.580	289.580
TEMPERATURE, DEG K	71.3882	71.3790	300.000	300.002	308.000	300.883
ENTHALPY, JOULES/GM.MOLE	2544.10	2544.10	8503.40	8503.39	8732.93	8530.60
ENTROPY, JOULES/GM.MOLE-DEG C	105.385	105.580	141.674	142.020	133.767	133.196
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)	423585.	23965.3	23965.4	14552.3	91166.5	2194.85
FLOW, GM.MOLE/SEC	5249.63	297.010	297.010	180.351	1129.86	27.2014
PRESSURE, KILOPASCAL	4136.86	111.006	111.006	111.006	101.353	101.353
TEMPERATURE, DEG K	308.000	20.5009	20.5009	20.5009	20.2337	20.2337
ENTHALPY, JOULES/GM.MOLE	8779.20	551.206	1443.96	1443.96	-451.402	422.916
ENTROPY, JOULES/GM.MOLE-DEG C	112.301	34.6863	78.3294	78.3294	18.8048	62.0292
LIQUID FRACTION (MOLE L/F)	S.H. VAP	SAT. LIQ	SAT. VAP	SAT. VAP	0.97592	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.03032	0.03032
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.96968	0.96968

STREAM NUMBER	067	133	134	135	136	146
FLOW, CMH(STP)	88971.7	7586.50	7586.50	7586.50	7586.50	7586.50
FLOW, GM.MOLE/SEC	1102.65	94.0220	94.0220	94.0220	94.0220	94.0220
PRESSURE, KILOPASCAL	101.353	124.106	124.106	120.658	120.658	120.658
TEMPERATURE, DEG K	20.2337	25.0000	39.0678	39.0435	77.7849	300.000
ENTHALPY, JOULES/GM.MOLE	-472.970	522.196	836.046	836.046	1683.60	8450.26
ENTROPY, JOULES/GM.MOLE-DEG C	17.7385	65.3264	75.1800	75.3710	90.4026	130.313
LIQUID FRACTION (MOLE L/F)	SAT. LIQ	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.03032	0.03029	0.03029	0.03029	0.03029	0.03029
PARA -HYDROGEN	0.96968	0.96971	0.96971	0.96971	0.96971	0.96971

TABLE 18
H2 LIQUEFIER - WITH VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
SI UNITS

STREAM NUMBER	147	148
FLOW, CMH(STP)	7586.50	7586.50
FLOW, GM.MOLE/SEC	94.0220	94.0220
PRESSURE, KILOPASCAL	101.353	289.580
TEMPERATURE, DEG K	308.000	308.000
ENTHALPY, JOULES/GM.MOLE	8730.67	8732.93
ENTROPY, JOULES/GM.MOLE-DEG C	142.834	133.767
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:		
ORTHO-HYDROGEN	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000

TABLE 19
H2 LIQUEFIER - WITH VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR NITROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	080	081	082	083	084	085
FLOW, CFH(NTP)	2796468.	2796468.	2796468.	712532.	712532.	561395.
FLOW, LB.MOLE/HR	7229.94	7229.94	7229.94	1842.17	1842.17	1451.42
PRESSURE, PSIA	90.0000	90.0000	86.2000	600.000	22.2000	22.2000
TEMPERATURE, DEG K	97.0000	300.000	299.949	99.9000	81.1034	81.1034
ENTHALPY, BTU/LB.MOLE	2879.48	5553.63	5553.63	942.897	942.896	446.308
ENTROPY, BTU/LB.MOLE-DEG K	31.6080	46.7331	46.8873	11.0117	12.3382	6.21608
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.H. VAP	S.C. LIQ	0.78789	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, CFH(NTP)	267558.	267558.	293837.	293837.	151137.	712532.
FLOW, LB.MOLE/HR	691.740	691.740	759.680	759.680	390.747	1842.17
PRESSURE, PSIA	22.2000	22.2000	22.2000	22.2000	22.2000	22.2000
TEMPERATURE, DEG K	81.1034	81.1034	81.1034	81.1034	81.1034	81.1034
ENTHALPY, BTU/LB.MOLE	446.310	2787.46	446.310	2787.46	2787.46	2787.46
ENTROPY, BTU/LB.MOLE-DEG K	6.21606	35.0785	6.21606	35.0785	35.0785	35.0785
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, CFH(NTP)	192400.	192400.	904932.	904932.	904932.
FLOW, LB.MOLE/HR	497.428	497.428	2339.60	2339.60	2339.60
PRESSURE, PSIA	22.2000	22.2000	22.2000	22.2000	15.0000
TEMPERATURE, DEG K	90.6000	90.6000	83.1097	300.000	299.902
ENTHALPY, BTU/LB.MOLE	2913.51	2913.51	2814.26	5565.20	5565.20
ENTROPY, BTU/LB.MOLE-DEG K	36.5484	36.5484	35.4049	51.7720	53.1731
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 19
H2 LIQUEFIER - WITH VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	001	002	003	004	005	006
FLOW, CFH(NTP)	3179775.	2387145.	792630.	16907712.	16907712.	1475394.
FLOW, LB.MOLE/HR	8220.94	6171.68	2049.25	43712.9	43712.9	3814.46
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	33.3970
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)	18383104.	3179775.	18383104.	3468386.	3468387.	3468387.
FLOW, LB.MOLE/HR	47527.4	8220.94	47527.4	8967.11	8967.11	8967.11
PRESSURE, PSIA	600.000	600.000	598.800	598.800	598.800	587.600
TEMPERATURE, DEG K	85.5000	308.000	85.4899	85.4899	81.3034	81.2016
ENTHALPY, BTU/LB.MOLE	1160.32	0.0	1160.32	1160.32	993.099	993.097
ENTROPY, BTU/LB.MOLE-DEG K	33.3969	0.0	33.4034	33.4034	31.7550	31.8147
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	0.54554	0.54554
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.45446	0.45446

STREAM NUMBER	013	014	015	016	017	018
FLOW, CFH(NTP)	3468387.	3468387.	3468387.	3468387.	3468387.	3468387.
FLOW, LB.MOLE/HR	8967.11	8967.11	8967.11	8967.11	8967.11	8967.11
PRESSURE, PSIA	587.600	580.899	580.899	574.199	135.000	135.000
TEMPERATURE, DEG K	40.9798	40.9193	28.0000	28.0252	28.6783	29.0946
ENTHALPY, BTU/LB.MOLE	197.025	197.024	-94.6846	-94.6852	-94.6862	-94.6862
ENTROPY, BTU/LB.MOLE-DEG K	18.0469	18.0751	10.0615	10.0829	11.5737	11.5745
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	0.20371	0.20371	0.04132	0.04132	0.04132	0.03032
PARA -HYDROGEN	0.79629	0.79629	0.95868	0.95868	0.95868	0.96968

TABLE 19
H2 LIQUEFIER - WITH VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	019	020	021	022	023	024
FLOW, CFH(NTP)	3468397.	14914718.	14914718.	14914718.	5470582.	5470582.
FLOW, LB.MOLE/HR	8967.11	38560.3	38560.3	38560.3	14143.6	14143.6
PRESSURE, PSIA	135.000	598.800	598.800	596.800	596.800	48.7000
TEMPERATURE, DEG K	20.5509	85.4899	81.3034	81.2849	81.2849	38.9802
ENTHALPY, BTU/LB.MOLE	-194.202	1160.32	1114.19	1114.19	1114.19	783.333
ENTROPY, BTU/LB.MOLE-DEG K	7.59975	33.4034	32.8590	32.8696	32.8698	35.3072
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	0.03032	0.75000	0.75000	0.75000	0.75000	0.75000
PARA -HYDROGEN	0.96968	0.25000	0.25000	0.25000	0.25000	0.25000

STREAM NUMBER	025	026	027	028	029	030
FLOW, CFH(NTP)	9444136.	9444136.	8890501.	8890501.	8890501.	553635.
FLOW, LB.MOLE/HR	24416.7	24416.7	22985.4	22985.4	22985.4	1431.36
PRESSURE, PSIA	596.800	596.800	596.800	596.800	50.5000	596.800
TEMPERATURE, DEG K	81.2849	58.7669	58.7669	58.7668	26.0000	58.7669
ENTHALPY, BTU/LB.MOLE	1114.19	842.253	842.253	[illegible]2.251	641.938	842.253
ENTROPY, BTU/LB.MOLE-DEG K	32.8696	28.9198	28.9198	28.9197	30.9753	28.9198
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, CFH(NTP)	553635.	553635.	1655377.	1655377.	6765093.	6765093.
FLOW, LB.MOLE/HR	1431.36	1431.36	4279.78	4279.78	17490.4	17490.4
PRESSURE, PSIA	596.800	16.1000	50.5000	50.5000	48.7000	48.7000
TEMPERATURE, DEG K	29.5000	20.5009	26.0000	39.0678	38.9798	77.7849
ENTHALPY, BTU/LB.MOLE	353.644	353.644	641.938	783.327	783.329	1149.07
ENTROPY, BTU/LB.MOLE-DEG K	17.5519	20.6181	30.9753	35.2359	35.3071	42.1084
LIQUID FRACTION (MOLE L/F)	S.C. LIQ	0.69667	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 19
H2 LIQUEFIER - WITH VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	037	038	039	040	041	042
FLOW, CFH(NTP)	6765093.	7235124.	7235124.	7235124.	-360866.	7595990.
FLOW, LB.MOLE/HR	17490.4	18705.6	18705.6	18705.6	-932.977	19638.5
PRESSURE, PSIA	46.0000	50.5000	50.5000	48.7000	48.7000	48.7000
TEMPERATURE, DEG K	77.7429	26.0000	39.0678	38.9798	38.9798	38.9798
ENTHALPY, BTU/LB.MOLE	1149.07	641.938	783.327	783.329	783.329	783.329
ENTROPY, BTU/LB.MOLE-DEG K	42.2960	30.9753	35.2359	35.3071	35.3071	35.3071
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)	7595990.	7595990.	14361083.	14361083.	2387145.	1475394.
FLOW, LB.MOLE/HR	19638.5	19638.5	37128.9	37128.9	6171.68	3814.46
PRESSURE, PSIA	48.7000	46.0000	46.0000	46.0000	600.000	600.000
TEMPERATURE, DEG K	77.7016	77.6596	77.6989	300.000	85.5000	85.5000
ENTHALPY, BTU/LB.MOLE	1148.30	1148.30	1148.66	3659.39	1160.32	1160.32
ENTROPY, BTU/LB.MOLE-DEG K	42.0975	42.2851	42.2902	57.0546	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, CFH(NTP)	911751.	911751.	911751.	911751.	1465385.	1465385.
FLOW, LB.MOLE/HR	2357.23	2357.23	2357.23	2357.23	3788.58	3788.58
PRESSURE, PSIA	600.000	598.600	598.600	16.1000	16.1000	16.1000
TEMPERATURE, DEG K	85.5000	85.4882	33.2811	20.5009	20.5009	20.5009
ENTHALPY, BTU/LB.MOLE	1160.32	1160.32	399.618	399.618	382.248	621.220
ENTROPY, BTU/LB.MOLE-DEG K	33.3969	33.4045	19.0119	22.8656	22.0165	33.6988
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP	S.C. LIQ	0.57697	0.62219	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 19
H2 LIQUEFIER - WITH VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	055	056	057	058	059	060
FLOW, CFH(NTP)	1465385.	1465385.	1465385.	1465385.	1465385.	16115093.
FLOW, LB.MOLE/HR	3788.58	3788.58	3788.58	3788.58	3788.58	41663.7
PRESSURE, PSIA	16.1000	15.6000	15.6000	15.0000	42.0000	42.0000
TEMPERATURE, DEG K	71.3882	71.3790	300.000	300.002	308.000	300.883
ENTHALPY, BTU/LB.MOLE	1094.52	1094.52	3658.32	3658.32	3757.07	3670.02
ENTROPY, BTU/LB.MOLE-DEG K	45.3386	45.4226	60.9507	61.0995	57.5489	57.3033
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)	16115093.	911750.	911751.	553635.	3468387.	83501.9
FLOW, LB.MOLE/HR	41663.7	2357.22	2357.23	1431.36	8967.11	215.885
PRESSURE, PSIA	600.000	16.1000	16.1000	16.1000	14.7000	14.7000
TEMPERATURE, DEG K	308.000	20.5009	20.5009	20.5009	20.2337	20.2337
ENTHALPY, BTU/LB.MOLE	3776.98	237.139	621.220	621.220	-194.201	181.94[illegible]
ENTROPY, BTU/LB.MOLE-DEG K	48.3138	14.9227	33.6988	33.6988	8.09017	26.686[illegible]
LIQUID FRACTION (MOLE L/F)	S.H. VAP	SAT. LIQ	SAT. VAP	SAT. VAP	0.97592	SAT. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.75000	0.75000	0.75000	0.75000	0.03032	0.03032
PARA -HYDROGEN	0.25000	0.25000	0.25000	0.25000	0.96968	0.96968

STREAM NUMBER	067	133	134	135	136	146
FLOW, CFH(NTP)	3384885.	288625.	288625.	288625.	288625.	288625.
FLOW, LB.MOLE/HR	8751.22	746.206	746.206	746.206	746.206	746.206
PRESSURE, PSIA	14.7000	18.0000	18.0000	17.5000	17.5000	17.5000
TEMPERATURE, DEG K	20.2337	25.0000	39.0678	39.0435	77.7849	300.000
ENTHALPY, BTU/LB.MOLE	-203.481	224.658	359.683	359.683	724.316	3635.46
ENTROPY, BTU/LB.MOLE-DEG K	7.63142	28.1046	32.3438	32.4260	38.8929	56.0631
LIQUID FRACTION (MOLE L/F)	SAT. LIQ	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:						
ORTHO-HYDROGEN	0.03032	0.03029	0.03029	0.03029	0.03029	0.03029
PARA -HYDROGEN	0.96968	0.96971	0.96971	0.96971	0.96971	0.96971

TABLE 19
H2 LIQUEFIER - WITH VENT GAS RECOVERY
192.1 MTD (211.7 TPD) LIQUID HYDROGEN PRODUCT
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR HYDROGEN STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	147	148
FLOW, CFH(NTP)	288625.	288625.
FLOW, LB.MOLE/HR	746.206	746.206
PRESSURE, PSIA	14.700	42.0000
TEMPERATURE, DEG K	308.000	308.000
ENTHALPY, BTU/LB.MOLE	3756.10	3757.07
ENTROPY, BTU/LB.MOLE-DEG K	61.4497	57.5489
LIQUID FRACTION (MOLE L/F)	S.H. VAP	S.H. VAP
COMPOSITION, MOLE FRACTION:		
ORTHO-HYDROGEN	0.75000	0.75000
PARA -HYDROGEN	0.25000	0.25000

TABLE 20
N2 REFRIGERATOR - NO VENT GAS RECOVERY
NASA CONTRACT NAS1-14698,MOD.1

STREAM DATA FOR N2 REFRIGERATOR STREAMS
SI UNITS

STREAM NUMBER	1	2	3	4	5	6
FLOW, CMH(STP)	275549.	275549.	256496.	110228.	110228.	146269.
FLOW, GM.MOLE/SEC	3414.96	3414.96	3178.84	1366.09	1366.09	1812.75
PRESSURE, KILOPASCAL	2785.77	3308.39	4136.86	4136.86	615.012	4136.86
TEMPERATURE, DEG K	308.000	308.000	308.000	235.000	147.990	235.000
ENTHALPY, JOULES/GM.MOLE	13029.5	13002.8	12961.1	10650.9	8376.42	10650.9
ENTROPY, JOULES/GM.MOLE-DEG C	96.5690	95.0568	93.0685	84.4956	87.5696	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	7	8	9	10	11	12
FLOW, CMH(STP)	146269.	174553.	174553.	19052.8	19052.8	19052.8
FLOW, GM.MOLE/SEC	1812.75	2163.29	2163.29	236.127	236.127	236.127
PRESSURE, KILOPASCAL	4136.86	615.012	615.012	4136.86	4136.86	4136.86
TEMPERATURE, DEG K	162.431	148.615	232.000	235.000	162.431	140.000
ENTHALPY, JOULES/GM.MOLE	8001.06	8395.78	10905.4	10650.9	8001.06	6763.73
ENTROPY, JOULES/GM.MOLE-DEG C	70.8914	87.7001	101.126	84.4956	70.8914	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	13	14	15	16	17	18
FLOW, CMH(STP)	19052.8	64325.7	64325.7	64325.7	19052.8	19052.8
FLOW, GM.MOLE/SEC	236.127	797.209	797.209	797.209	236.127	236.127
PRESSURE, KILOPASCAL	4136.86	620.528	620.528	615.012	4136.86	4136.86
TEMPERATURE, DEG K	99.0000	97.0000	138.000	149.687	100.650	100.650
ENTHALPY, JOULES/GM.MOLE	2137.69	6693.07	8063.26	8428.96	2236.92	2236.92
ENTROPY, JOULES/GM.MOLE-DEG C	25.0528	73.4695	85.3076	87.9227	26.0468	26.0468
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 20
N2 REFRIGERATOR - NO VENT GAS RECOVERY
NASA CONTRACT NAS1-14698,MOD.1

STREAM DATA FOR N2 REFRIGERATOR STREAMS
SI UNITS

STREAM NUMBER	19	20	21	22	23	25
FLOW, CMH(STP)	0.0	81942.8	81942.8	146269.	0.0	256496.
FLOW, GM.MOLE/SEC	0.0	1015.54	1015.54	1812.75	0.0	3178.84
PRESSURE, KILOPASCAL	4136.86	620.528	620.528	620.528	620.528	4136.86
TEMPERATURE, DEG K	100.650	97.0000	97.0000	97.0000	63.1363	235.000
ENTHALPY, JOULES/GM.MOLE	0.0	6693.07	6693.07	6693.07	-0.02724	10650.9
ENTROPY, JOULES/GM.MOLE-DEG C	0.0	73.4695	73.4695	73.4695	-0.25022	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	26	27	28	29	30	31
FLOW, CMH(STP)	256496.	19052.8	19052.8	174553.	275549.	275549.
FLOW, GM.MOLE/SEC	3178.84	236.127	236.127	2163.29	3414.96	3414.96
PRESSURE, KILOPASCAL	4136.86	4136.86	4136.86	607.428	4136.86	594.328
TEMPERATURE, DEG K	265.565	308.000	265.565	303.000	308.000	302.436
ENTHALPY, JOULES/GM.MOLE	11636.2	12961.1	11636.2	12997.7	12961.1	12981.8
ENTROPY, JOULES/GM.MOLE-DEG C	88.4387	93.0685	88.4387	109.098	93.0685	109.227
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	32	33	34	35
FLOW, CMH(STP)	81942.8	193606.	19052.8	19052.8
FLOW, GM.MOLE/SEC	1015.54	2399.42	236.127	236.127
PRESSURE, KILOPASCAL	594.328	607.428	607.428	101.353
TEMPERATURE, DEG K	300.000	303.492	308.000	300.000
ENTHALPY, JOULES/GM.MOLE	12910.3	13012.1	13144.4	12938.7
ENTROPY, JOULES/GM.MOLE-DEG C	108.989	109.146	109.579	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 21
N2 REFRIGERATOR - NO VENT GAS RECOVERY
NASA CONTRACT NAS1-14698,MOD.1

STREAM DATA FOR N2 REFRIGERATOR STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	1	2	3	4	5	6
FLOW, CFH(NTP)	10483124.	10483124.	9758271.	4193556.	4193556.	5564715.
FLOW, LB.MOLE/HR	27102.9	27102.9	25228.9	10841.9	10841.9	14386.9
PRESSURE, PSIA	404.041	479.841	600.000	600.000	89.2000	600.000
TEMPERATURE, DEG K	308.000	308.000	308.000	235.000	147.990	235.000
ENTHALPY, BTU/LB.MOLE	5605.55	5594.04	5576.10	4582.21	3603.69	4582.21
ENTROPY, BTU/LB.MOLE-DEG K	41.5458	40.8952	40.0398	36.3516	37.6741	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 NUMBER	7	8	9	10	11	12
FLOW, CFH(NTP)	5564715.	6640797.	6640797.	724853.	724853.	724853.
FLOW, LB.MOLE/HR	14386.9	17169.0	17169.0	1874.02	1874.02	1874.02
PRESSURE, PSIA	600.000	89.2000	89.2000	600.000	600.000	600.000
TEMPERATURE, DEG K	162.431	148.615	232.000	235.000	162.431	140.000
ENTHALPY, BTU/LB.MOLE	3442.20	3612.02	4691.72	4582.21	3442.20	2909.88
ENTROPY, BTU/LB.MOLE-DEG K	30.4988	37.7302	43.5063	36.3516	30.4988	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 NUMBER	13	14	15	16	17	18
FLOW, CFH(NTP)	724853.	2447241.	2447241.	2447241.	724853.	724853.
FLOW, LB.MOLE/HR	1874.02	6327.05	6327.05	6327.05	1874.02	1874.02
PRESSURE, PSIA	600.000	90.0000	90.0000	89.2000	600.000	600.000
TEMPERATURE, DEG K	99.0000	97.0000	138.000	149.687	100.650	100.650
ENTHALPY, BTU/LB.MOLE	919.675	2879.48	3468.97	3626.30	962.365	962.365
ENTROPY, BTU/LB.MOLE-DEG K	10.7782	31.6080	36.7009	37.8260	11.2058	11.2058
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 21
N2 REFRIGERATOR - NO VENT GAS RECOVERY
NASA CONTRACT NAS1-14698,MOD.1

STREAM DATA FOR N2 REFRIGERATOR STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	19	20	21	22	23	25
FLOW, CFH(NTP)	0.0	3117474.	3117474.	5564715.	0.0	9758271.
FLOW, LB.MOLE/HR	0.0	8059.86	8059.86	14386.9	0.0	25228.9
PRESSURE, PSIA	600.000	90.0000	90.0000	90.0000	90.0000	600.000
TEMPERATURE, DEG K	100.650	97.0000	97.0000	97.0000	63.1363	235.000
ENTHALPY, BTU/LB.MOLE	0.0	2879.48	2879.48	2879.48	-0.01172	4582.20
ENTROPY, BTU/LB.MOLE-DEG K	0.0	31.6080	31.6080	31.6080	-0.10765	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 NUMBER	26	27	28	29	30	31
FLOW, CFH(NTP)	9758271.	724853.	724853.	6640797.	10483124.	10483124.
FLOW, LB.MOLE/HR	25228.9	1874.02	1874.02	17169.0	27102.9	27102.9
PRESSURE, PSIA	600.000	600.000	600.000	88.1000	600.000	86.2000
TEMPERATURE, DEG K	265.565	308.000	265.565	303.000	308.000	302.436
ENTHALPY, BTU/LB.MOLE	5006.10	5576.10	5006.10	5591.83	5576.10	5585.02
ENTROPY, BTU/LB.MOLE-DEG K	38.0480	40.0398	38.0480	46.9360	40.0398	46.9915
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	32	33	34	35
FLOW, CFH(NTP)	3117474.	7365650.	724853.	724853.
FLOW, LB.MOLE/HR	8059.86	19043.0	1874.02	1874.02
PRESSURE, PSIA	86.2000	88.1000	88.1000	14.7000
TEMPERATURE, DEG K	300.000	303.492	308.000	300.000
ENTHALPY, BTU/LB.MOLE	5554.27	5598.04	5654.97	5566.43
ENTROPY, BTU/LB.MOLE-DEG K	46.8893	46.9566	47.1427	53.249[illegible]
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 22
N2 REFRIGERATOR - WITH VENT GAS RECOVERY
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR N2 REFRIGERATOR STREAMS
SI UNITS

STREAM NUMBER	1	2	3	4	5	6
FLOW, CMH(STP)	256439.	256439.	237710.	100973.	100973.	136737.
FLOW, GM.MOLE/SEC	3178.13	3178.13	2946.02	1251.39	1251.39	1694.63
PRESSURE, KILOPASCAL	2793.19	3319.63	4136.86	4136.86	615.012	4136.86
TEMPERATURE, DEG K	308.000	308.000	308.000	235.000	147.990	235.000
ENTHALPY, JOULES/GM.MOLE	13029.2	13002.2	12961.1	10650.9	8376.42	10650.9
ENTROPY, JOULES/GM.MOLE-DEG C	96.5458	95.0267	93.0685	84.4956	87.5696	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	7	8	9	10	11	12
FLOW, CMH(STP)	136737.	164205.	164205.	18728.9	18728.9	18728.9
FLOW, GM.MOLE/SEC	1694.63	2035.05	2035.05	232.113	232.113	232.113
PRESSURE, KILOPASCAL	4136.86	615.012	615.012	4136.86	4136.86	4136.86
TEMPERATURE, DEG K	162.431	148.643	232.000	235.000	162.431	140.000
ENTHALPY, JOULES/GM.MOLE	8001.06	8396.65	10905.4	10650.9	8001.06	6763.73
ENTROPY, JOULES/GM.MOLE-DEG C	70.8914	87.7062	101.126	84.4956	70.8914	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	13	14	15	16	17	18
FLOW, CMH(STP)	18728.9	63232.3	63232.3	63232.3	18728.9	18728.9
FLOW, GM.MOLE/SEC	232.113	783.657	783.657	783.657	232.113	232.113
PRESSURE, KILOPASCAL	4136.86	620.528	620.528	615.012	4136.86	4136.86
TEMPERATURE, DEG K	99.0000	97.0000	138.000	149.687	100.679	100.679
ENTHALPY, JOULES/GM.MOLE	2137.69	6693.07	8063.26	8428.96	2238.64	2238.64
ENTROPY, JOULES/GM.MOLE-DEG C	25.0528	73.4695	85.3076	87.9227	26.0640	26.0640
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 22
N2 REFRIGERATOR - WITH VENT GAS RECOVERY
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR N2 REFRIGERATOR STREAMS
SI UNITS

STREAM NUMBER	19	20	21	22	23	25
FLOW, CMH(STP)	0.0	73505.1	73505.1	136737.	0.0	237710.
FLOW, GM.MOLE/SEC	0.0	910.972	910.972	1694.63	0.0	2946.02
PRESSURE, KILOPASCAL	4136.86	620.528	620.528	620.528	620.528	4136.86
TEMPERATURE, DEG K	100.679	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	26	27	28	29	30	31
FLOW, CMH(STP)	237710.	18728.9	18728.9	164205.	256439.	256439.
FLOW, GM.MOLE/SEC	2946.02	232.113	232.113	2035.05	3178.13	3178.13
PRESSURE, KILOPASCAL	4136.86	4136.86	4136.86	607.428	4136.86	594.328
TEMPERATURE, DEG K	265.113	308.000	265.113	303.000	308.000	302.487
ENTHALPY, JOULES/GM.MOLE	11621.9	12961.1	11621.9	12997.7	12961.1	12983.3
ENTROPY, JOULES/GM.MOLE-DEG C	88.3846	93.0685	88.3846	109.098	93.0685	109.232
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	32	33	34	35
FLOW, CMH(STP)	73505.1	182934.	18728.9	18728.9
FLOW, GM.MOLE/SEC	910.972	2267.16	232.113	232.113
PRESSURE, KILOPASCAL	594.328	607.428	607.428	101.353
TEMPERATURE, DEG K	300.000	303.511	308.000	300.000
ENTHALPY, JOULES/GM.MOLE	12910.3	13012.7	13144.4	12928.7
ENTROPY, JOULES/GM.MOLE-DEG C	108.989	109.148	109.579	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 23
N2 REFRIGERATOR - WITH VENT GAS RECOVERY
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR N2 REFRIGERATOR STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	1	2	3	4	5	6
FLOW, CFH(NTP)	9756110.	9756110.	9043578.	3841468.	3841468.	5202110.
FLOW, LB.MOLE/HR	25223.3	25223.3	23381.1	9931.67	9931.67	13449.4
PRESSURE, PSIA	405.117	481.472	600.000	600.000	89.2000	600.000
TEMPERATURE, DEG K	308.000	308.000	308.000	235.000	147.990	235.000
ENTHALPY, BTU/LB.MOLE	5605.39	5593.80	5576.10	4582.21	3603.69	4582.21
ENTROPY, BTU/LB.MOLE-DEG K	41.5358	40.8823	40.0398	36.3516	37.6741	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 NUMBER	7	8	9	10	11	12
FLOW, CFH(NTP)	5202110.	6247110.	6247110.	712532.	712532.	712532.
FLOW, LB.MOLE/HR	13449.4	16151.2	16151.2	1842.17	1842.17	1842.17
PRESSURE, PSIA	600.000	89.2000	89.2000	600.000	600.000	600.000
TEMPERATURE, DEG K	162.431	148.643	232.000	235.000	162.431	140.000
ENTHALPY, BTU/LB.MOLE	3442.21	3612.39	4691.72	4582.21	3442.21	2909.68
ENTROPY, BTU/LB.MOLE-DEG K	30.4988	37.7328	43.5063	36.3516	30.4988	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 NUMBER	13	14	15	16	17	18
FLOW, CFH(NTP)	712532.	2405642.	2405642.	2405642.	712532.	712532.
FLOW, LB.MOLE/HR	1842.17	6219.50	6219.50	6219.50	1842.17	1842.17
PRESSURE, PSIA	600.000	90.0000	90.0000	89.2000	600.000	600.000
TEMPERATURE, DEG K	99.0000	97.0000	138.000	149.687	100.679	100.679
ENTHALPY, BTU/LB.MOLE	919.675	2879.48	3468.97	3626.30	963.104	963.104
ENTROPY, BTU/LB.MOLE-DEG K	10.[illegible]782	31.6080	36.7009	37.8260	11.2132	11.2132
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 23
N2 REFRIGERATOR - WITH VENT GAS RECOVERY
NASA CONTRACT NAS1-14698, MOD.1

STREAM DATA FOR N2 REFRIGERATOR STREAMS
US CUSTOMARY UNITS

STREAM NUMBER	19	20	21	22	23	25
FLOW, CFH(NTP)	0.0	2796468.	2796468.	5202110.	0.0	9043578.
FLOW, LB.MOLE/HR	0.0	7229.94	7229.94	13449.4	0.0	23381.1
PRESSURE, PSIA	600.000	90.0000	90.0000	90.0000	90.0000	600.000
TEMPERATURE, DEG K	100.679	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 NUMBER	26	27	28	29	30	31
FLOW, CFH(NTP)	9043578.	712532.	712532.	6247110.	9756110.	9756110.
FLOW, LB.MOLE/HR	23381.1	1842.17	1842.17	16151.2	25223.3	25223.3
PRESSURE, PSIA	600.000	600.000	600.000	88.1000	600.000	86.2000
TEMPERATURE, DEG K	265.113	308.000	265.113	303.000	308.000	302.487
ENTHALPY, BTU/LB.MOLE	4999.94	5576.10	4999.94	5591.83	5576.10	5585.67
ENTROPY, BTU/LB.MOLE-DEG K	38.0247	40.0398	38.0247	46.9360	40.0398	46.9937
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

C-2

STREAM NUMBER	32	33	34	35
FLOW, CFH(NTP)	2796468.	6959642.	712532.	712532.
FLOW, LB.MOLE/HR	7229.94	17993.3	1842.17	1842.17
PRESSURE, PSIA	86.2000	88.1000	88.1000	14.7000
TEMPERATURE, DEG K	300.000	303.511	308.000	300.000
ENTHALPY, BTU/LB.MOLE	5554.27	5598.29	5654.97	5566.48
ENTROPY, BTU/LB.MOLE-DEG K	46.8893	46.9574	47.1427	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

ORIGINAL PAGE IS
OF POOR QUALITY

TABLE 24

POWER REQUIREMENTS FOR MAJOR ITEMS OF MACHINERY

For 192.1 MTD (211.7 TPD) Peak Month Output
From one - 226.8 MTD (250 TPD) Liquefaction Module

PLANT	No Vent Gas Recovery "Standard"	With Vent Gas Recovery Vent Gas Flow Avg.	Max.	Min
FLOWS: m^3/s (MCFH)				
LH_2 Product	24.715 (3384.9)	24.715 (3384.9)	24.715 (3384.9)	24.715 (3384.9)
Vent Gas	0 0	2.11 (288.63)	3.89 (533.40)	1.02 (139.88)
LN_2 Added	5.29 (724.85)	5.20 (712.53)	5.13 (702.87)	5.25 (718.41)
Cold N_2 Gas	22.76 (3,117.47)	20.42 (2,796.47)	18.61 (2,548.45)	21.52 (2,947.17)
POWER: kW				
Compressors				
H_2 Recycle	50,731	50,097	48,898	50,826
H_2 Flash	1,878	1,878	1,878	1,878
Vent Gas	0	377	377	377
H_2 Feed/Booster	18,725	17,177	17,177	17,177
N_2 Heat Pump	11,005	10,089	10,089	10,089
N_2 Refrigeration	22,537	21,047	19,888	21,752
SUBTOTAL	104,876	100,665	98,307	102,099
Turbine Return				
E-1	-1,425	-1,371	-1,328	-1,397
E-2	-1,390	-1,348	-1,316	-1,368
NET POWER	102,061	97,946	95,663	99,334

TABLE 25

FLOW VARIATION FOR H_2

LIQUEFIER MACHINERY

WITH VARIATION IN VENT GAS FLOW

For 192.1 MTD (211.7 TPD) Peak Month Output

PLANT	No Vent Gas Recovery "Standard"	With Vent Gas Recovery Vent Gas Flow Avg.	Max.	Min.
H_2 LIQUEFIER:				
Stream Flows, m^3/s (MCFH)				
LH_2 Product	24.715 (3384.9)	24.715 (3384.9)	24.715 (3384.9)	24.715 (3384.9)
H_2 Recycle Compressor	119.15 (16,319.2)	117.68 (16,116.9)	114.86 (15,731.1)	119.39 (16,351.4)
H_2 Flash Compressor	10.70 (1,465.4)	10.69 (1,463.8)	10.69 (1,463.8)	10.60 (1,463.8)
H_2 Turbine, E-1	41.53 (5,687.7)	40.05 (5,484.6)	38.80 (5,314.6)	40.80 (5,587.9)
H_2 Turbine, E-2	70.54 (9,661.6)	65.00 (8,902.1)	63.42 (8,686.3)	65.96 (9,033.3)
N_2 LIQUEFIER:				
Stream Flows, m^3/s (MCFH)				
Liquid N_2	5.29 (724.85)	5.20 (712.53)	5.13 (702.87)	5.25 (718.41)
Cold N_2 Gas	22.76 (3,117.47)	20.42 (2,796.47)	18.61 (2,548.45)	21.52 (2,947.17)

TABLE 26

UTILITY SUMMARY: HYDROGEN LIQUEFACTION/STORAGE

COMPLEX

For 577.2 MTD (636.3 TPD) Average Output

ELECTIRCAL POWER - kW	VENT GAS RECOVERY	
	WITHOUT	WITH
Production		
Hydrogen Compressors	132,350	131,700
Nitrogen Recycle Compressors	54,660	50,980
Forecooler	6,900	6,300
Air Compressor, Nitrogen Plant	3,320	3,320
Purifier Heat Pump Compressor	6,970	6,380
Hydrogen Feed/Booster Compressor	11,740	10,740
Nitrogen Feed Compressor	5,570	5,470
Hydrogen Drier	1,730	1,730
Pumps	430	430
		217,050
Subtotal	223,670	217,050
Hydrogen Turbine Return	- 6,640	- 6,410
Net Subtotal	217,030	210,640
Production Auxiliaries		
Cooling Tower and Water Supply	10,325	9,950
Plant Air Compressor and Drier	655	655
Purge Blower and Thaw Heater	3,190	3,190
Miscellaneous	4,250	4,250
Subtotal	18,420	18,045
Process Contingency (5%)	11,800	11,400
Subtotal	247,250	240,085
Plant Auxiliaries		
Road and Exterior Lighting	300	300
Building Lighting, Heating, Air Conditioning	750	750
Cranes	250	250
Fueling Pumps	150	150
Subtotal	1,450	1,450
TOTAL, ELECTRICAL POWER	248,700	241,535

TABLE 26 (Continued)

UTILITY SUMMARY: HYDROGEN LIQUEFACTION/STORAGE COMPLEX

		VENT GAS RECOVERY	
WATER		WITHOUT	WITH
Cooling water makeup	- m^3/s	0.196	0.186
	(gpm)	(3100)	(2950)
Potable Water	- m^3/s	0.001095	0.001095
	(gal/day)	(25,000)	(25,000)
CHEMICALS			
Sulfuric acid for water treatment	kg/s	0.105	0.105
	(lb/hr)	(835)	(835)
Dessicants and Adsorbents	kg/yr	47,000	47,000
	(lb/yr)	(105,000)	(105,000)
ANNUAL THAW			
N_2 for Plant Purge and Thaw	km^3	710	710
	(MMCF)	(27.0)	(27.0)
Heating Fuel	Gd	696	696
	(MMBtu)	(660)	(660)

TABLE 27

CAPITAL INVESTMENT
LIQUEFACTION COMPLEX

PLANT CAPACITY: 907 MTD (1000 TPD)
MID-1975 DOLLARS
YEAR 2000 TECHNOLOGY

	$ THOUSANDS	
	VENT GAS RECOVERY	
	WITHOUT	WITH
TOTAL PLANT INVESTMENT	239,000	235,300
INTEREST DURING CONSTRUCTION(1)	53,800	52,900
STARTUP COSTS	6,570	6,500
WORKING CAPITAL(2)	9,250	9,210
TOTAL CAPITAL REQUIREMENT	$308,620	$303,910

(1) At 12% interest rate on total plant investment for 1.875 years.

(2) Sum of (1) materials and supplies at 0.9% of total plant investment plus (2) net receivables on product hydrogen at 1/24 of annual production at 80.82¢/kg (36.66 ¢/lb)/

TABLE 28

ANNUAL OPERATING COST

FOR 577.2 MTD (636.3 TPD) Average Output

	$ THOUSANDS	
	VENT GAS RECOVERY	
	WITHOUT	WITH
RAW MATERIALS		
Feedstock (GH_2 @ $0.1645/lb)	76,415	69,899
CHEMICALS		
Sulfuric Acid	243	243
Dessicants and Adsorbents	93	93
UTILITIES		
Electricity (at $0.02/kWh)	43,572	42,317
Cooling Water Makeup	662	639
Potable Water	4.5	4.5
LABOR		
Operating Labor	1,092	1,092
Supervision	250.6	250.6
ADMINISTRATION AND OVERHEAD	919.9	919.9
SUPPLIES		
Operating	327.6	327.6
Maintenance	3,585	3,530
TAXES AND INSURANCE	6,453	6,353
TOTAL ANNUAL OPERATING COST	133,561.6	125,688.6

TABLE 29

UNIT COST OF LIQUID HYDROGEN
WITH AND WITHOUT VENT GAS RECOVERY

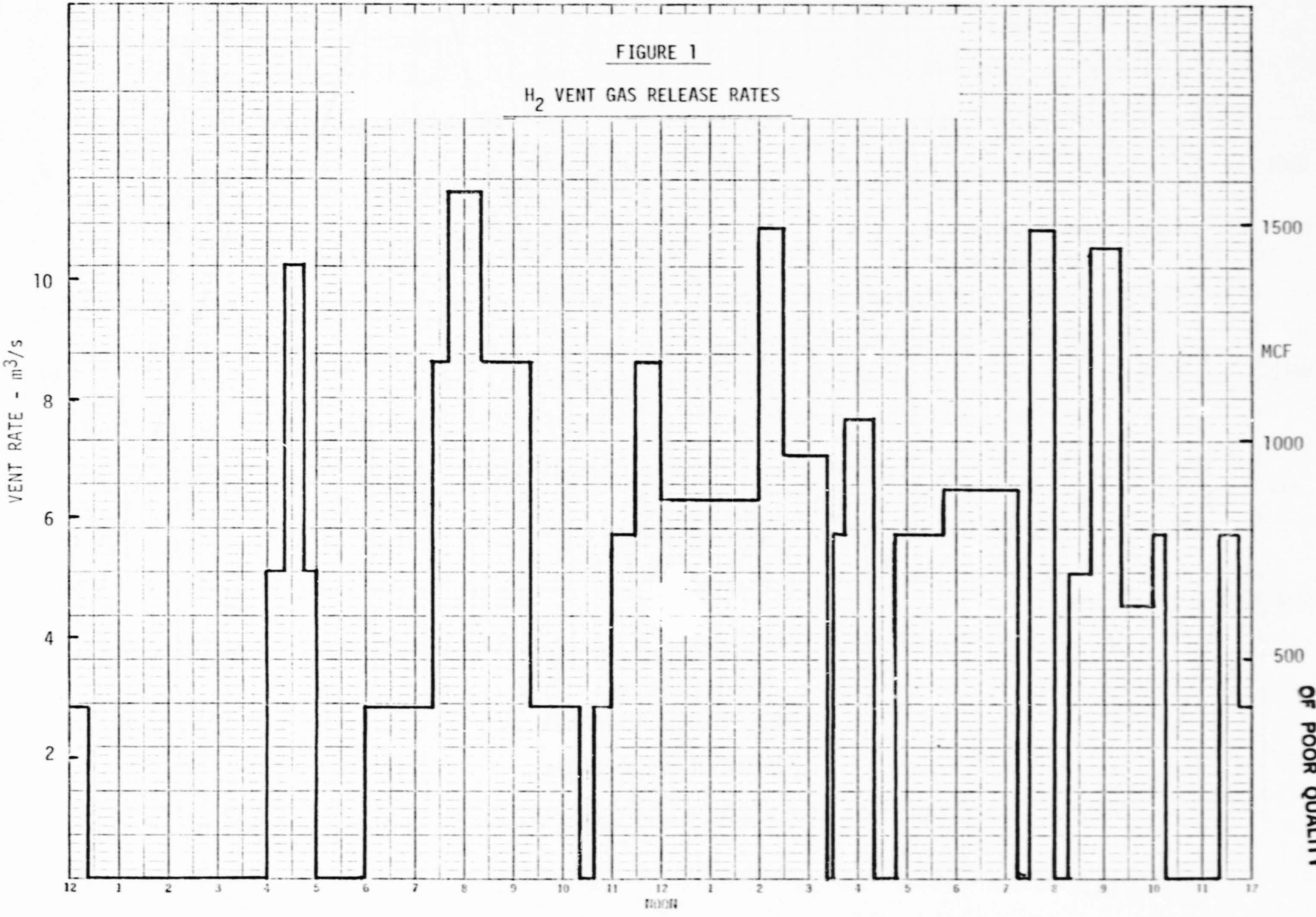
AVERAGE PRODUCTION = 577.2 MTD (636.3 TPD)

	VENT GAS RECOVERY			
	WITHOUT		WITH	
	$/kg	($/lb)	$/kg	($/lb)
OPERATING COST				
Feedstock	0.3627	(0.1645)	0.3318	(0.1505)
Electric Power	0.2068	(0.0938)	0.2008	(0.0911)
Labor, Administration and Overhead	0.0108	(0.0049)	0.0108	(0.0049)
Chemicals, Supplies, Water, Taxes and Insurance	0.0540	(0.0245)	0.0531	(0.0241)
SUBTOTAL	0.6343	(0.2877)	0.5966	(0.2706)
CAPITAL INVESTMENT:				
Liquefaction and Storage Facility Investment	0.2599	(0.1179)	0.2527	(0.1160)
Distribution System Investment	0.0346	(0.0157)	0.0346	(0.0157)
Startup Costs	0.0040	(0.0018)	0.0037	(0.0017)
Working Capital	0.0101	(0.0046)	0.0101	(0.0046)
SUBTOTAL	0.3086	(0.1400)	0.3042	(0.1380)
TOTAL	0.9429	(0.4277)	0.9008	(0.4086)

REDUCTION IN COST	0.0421	(0.0191)

FIGURE 1

H_2 VENT GAS RELEASE RATES



ORIGINAL PAGE IS OF POOR QUALITY

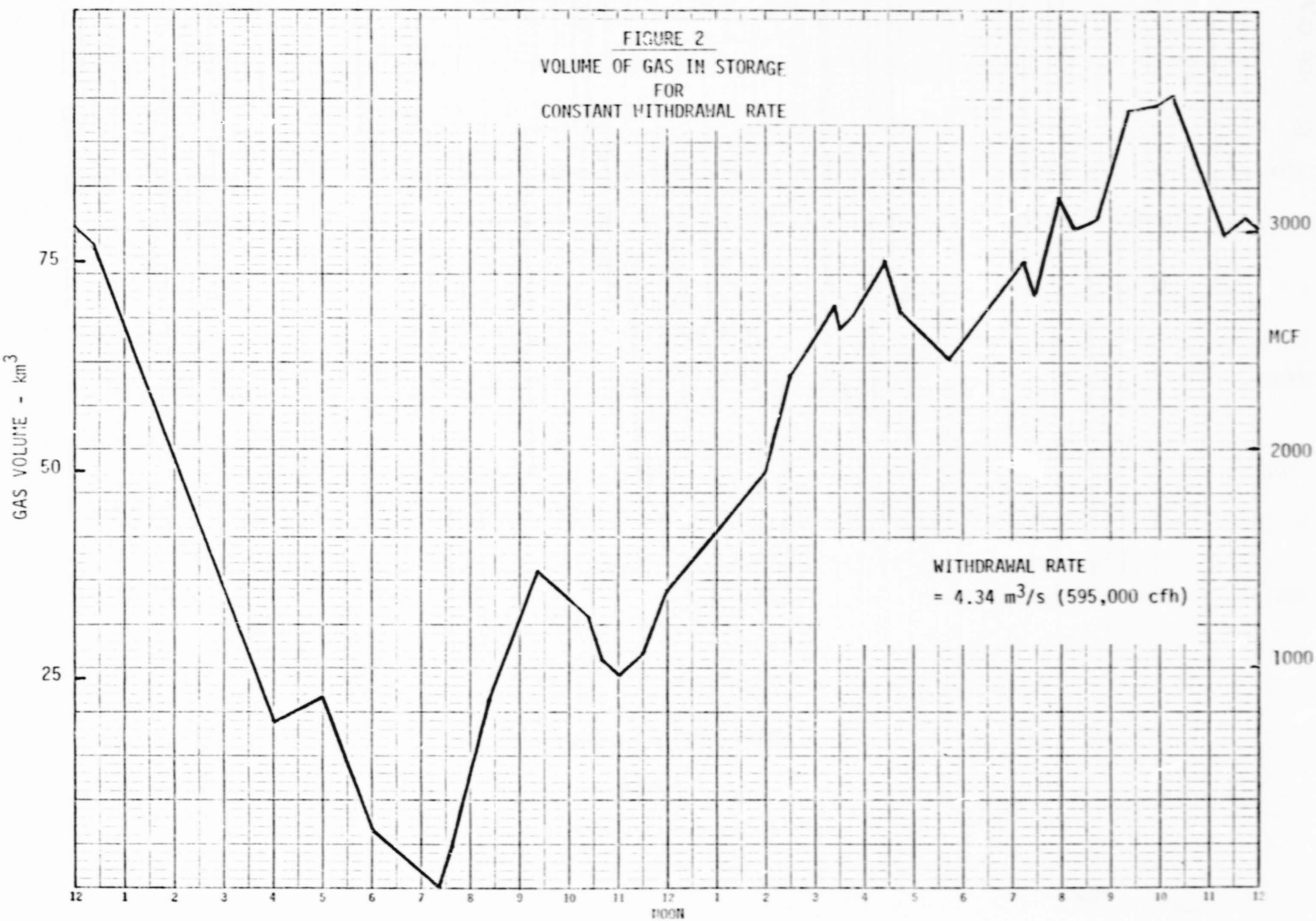


FIGURE 2
VOLUME OF GAS IN STORAGE
FOR
CONSTANT WITHDRAWAL RATE
GAS VOLUME - km^3
75
50
25
3000
MCF
2000
1000
WITHDRAWAL RATE
= 4.34 m^3/s (595,000 cfh)
12 1 2 3 4 5 6 7 8 9 10 11 12 1 2 3 4 5 6 7 8 9 10 11 12
NOON

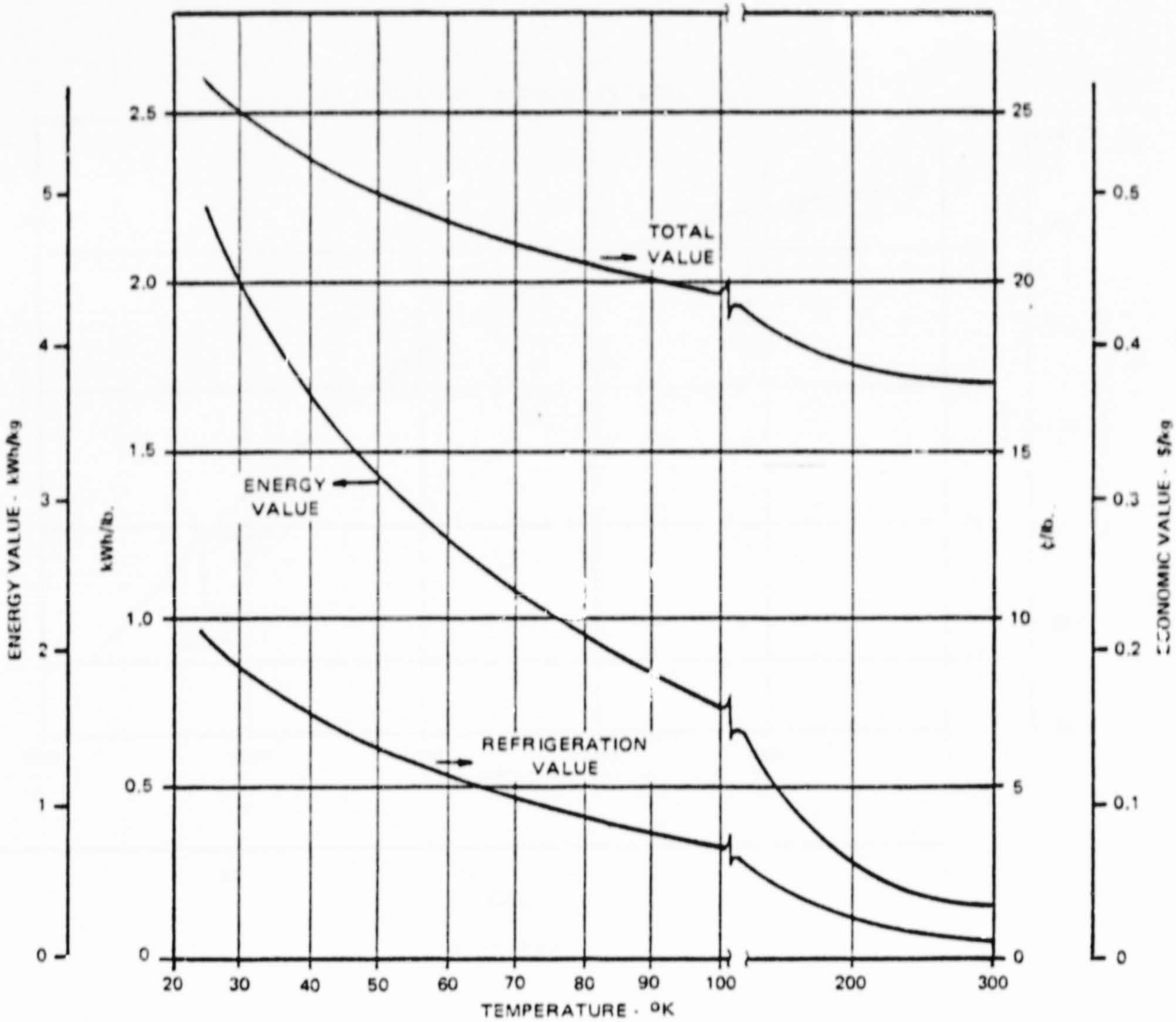


FIGURE 3
VALUE OF COLD H_2 VENT GAS

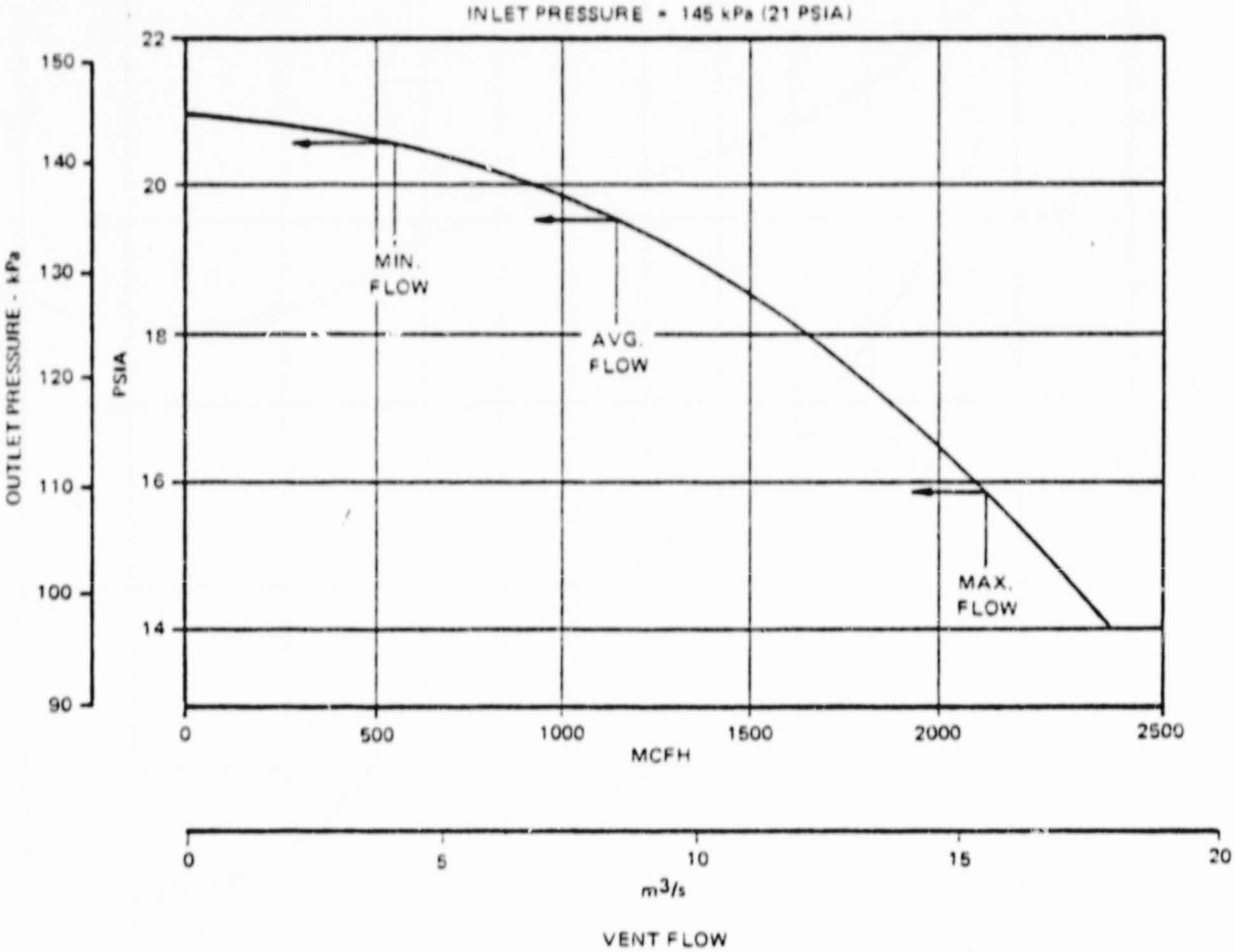


FIGURE 4
PRESSURE DROP THROUGH
H_2 VENT LINE

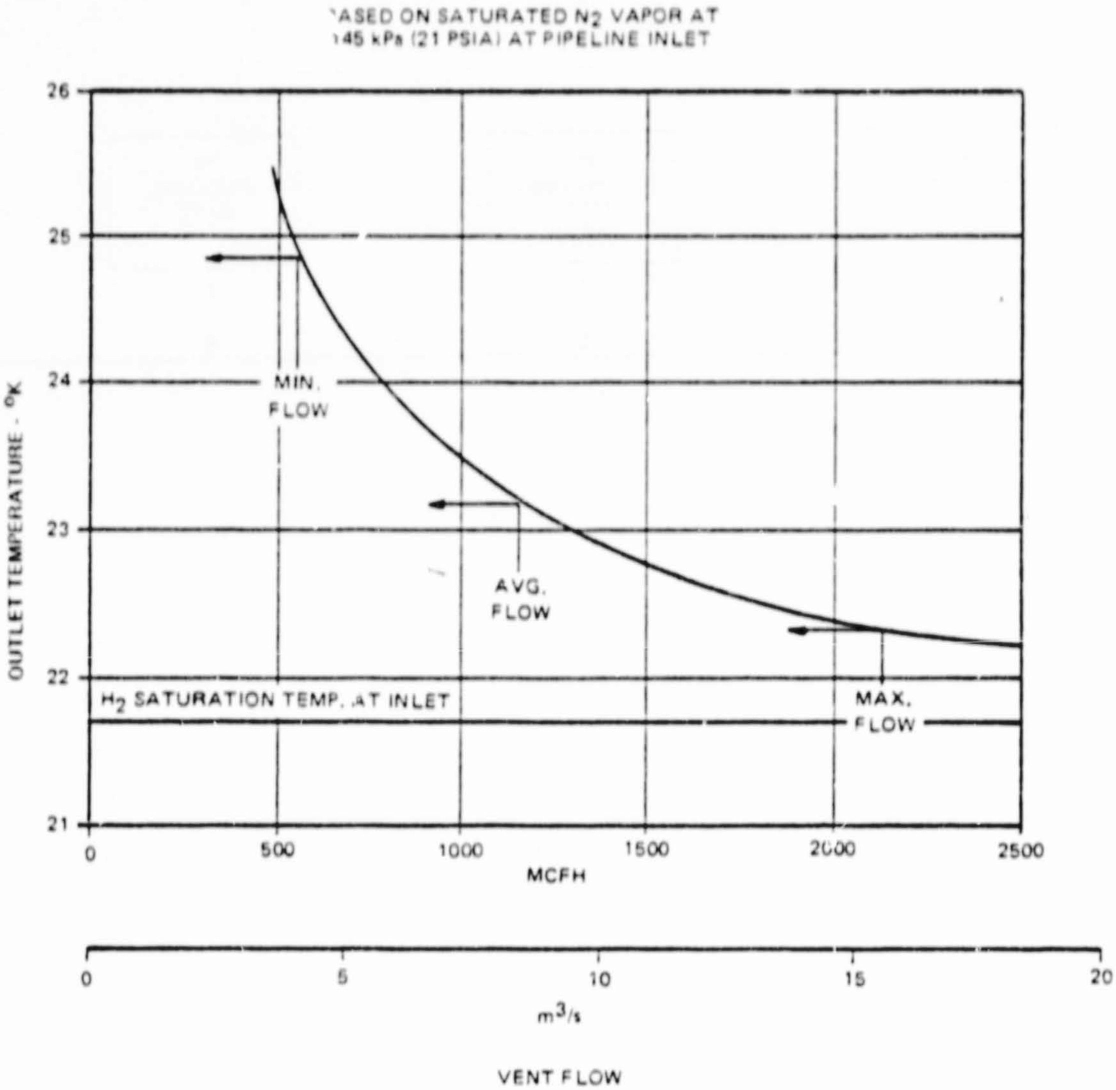


FIGURE 5
TEMPERATURE OF H_2 VENT GAS
AT PIPELINE OUTLET RESULTING FROM
HEAT LEAK TO PIPELINE

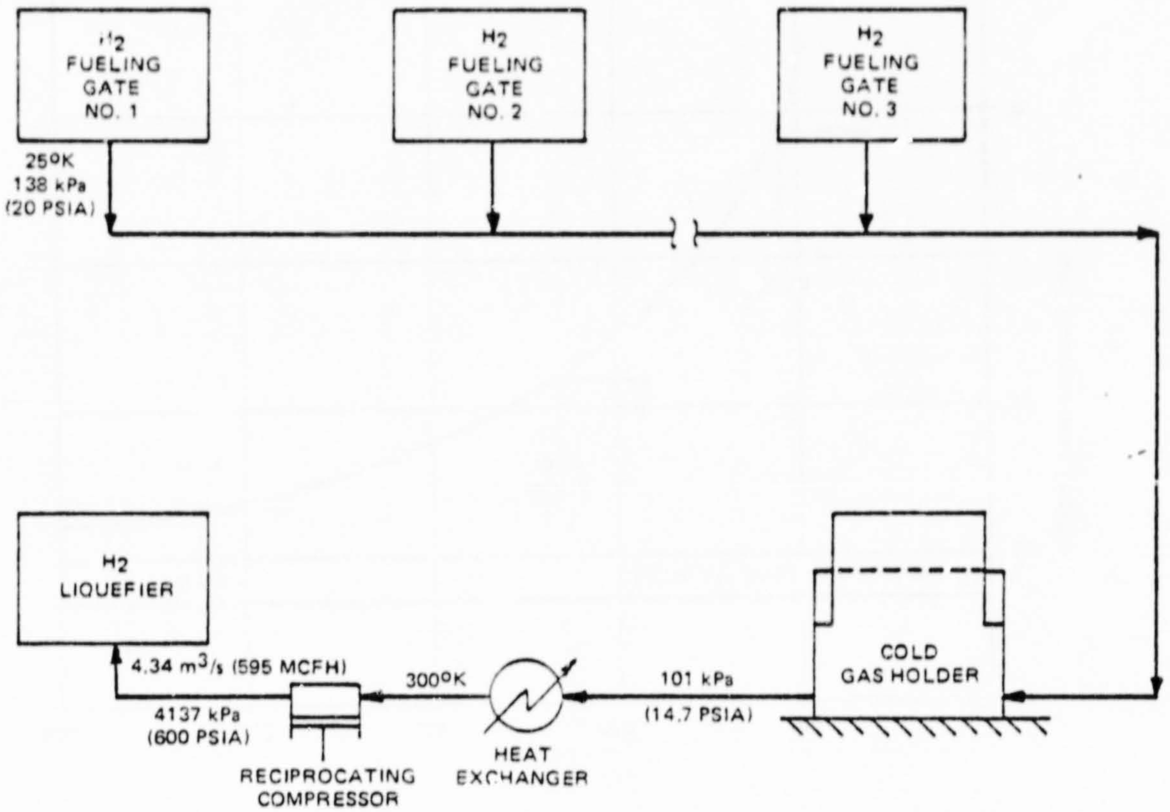


FIGURE 6
FLOW DIAGRAM
H_2 VENT GAS RECOVERY
CONSTANT PRESSURE COLD ACCUMULATOR

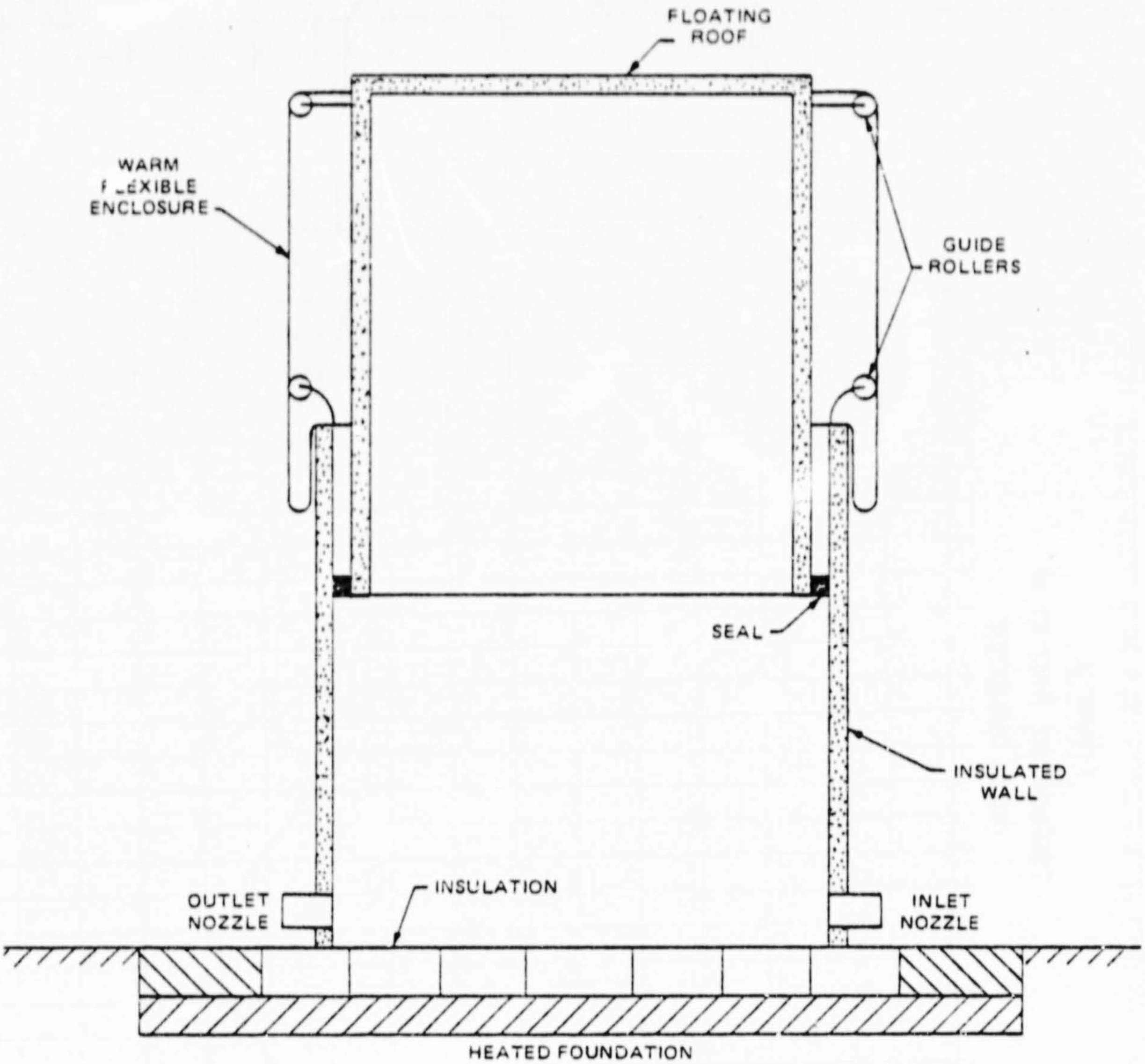
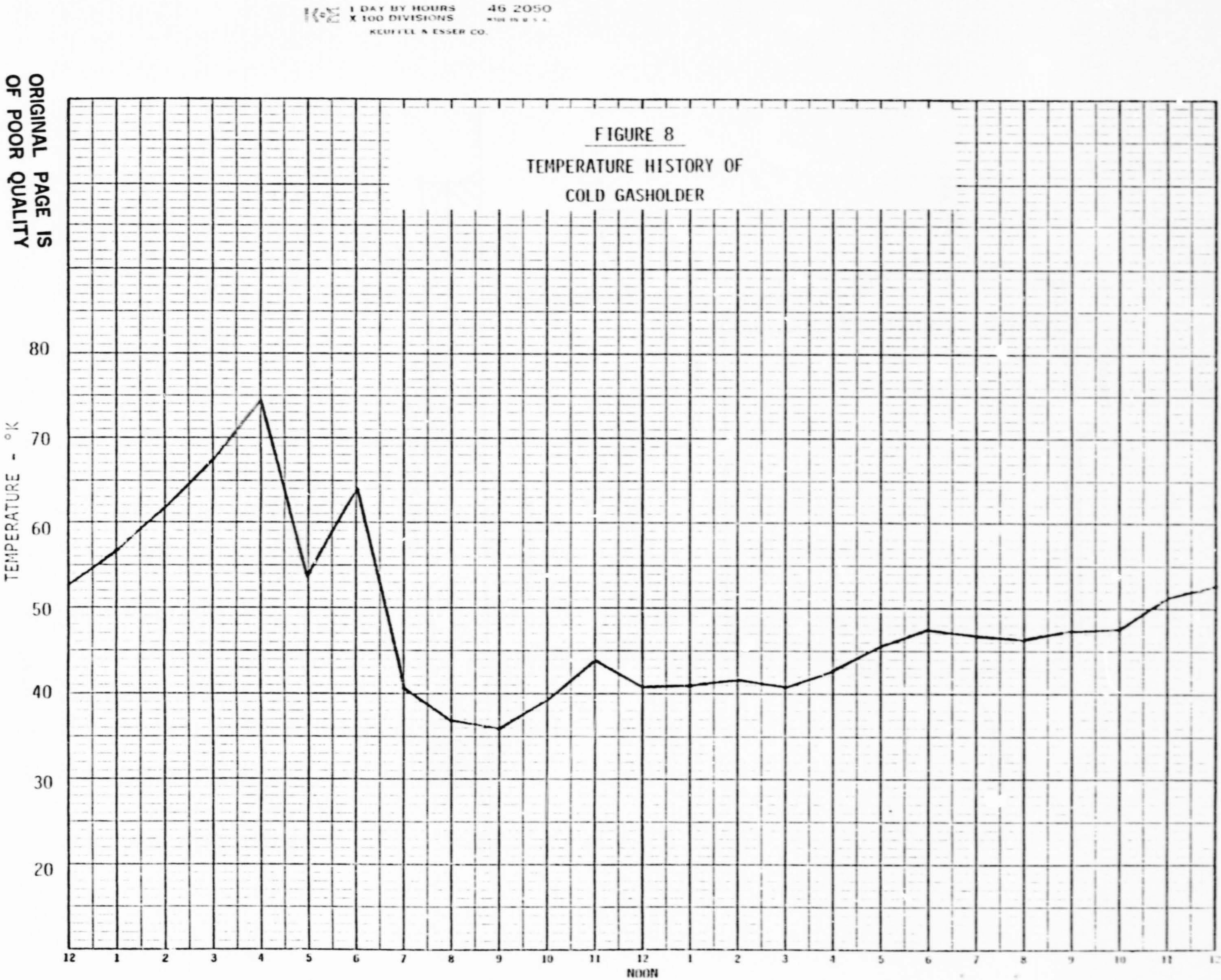


FIGURE 7
CONCEPTUAL SKETCH
CRYOGENIC GASHOLDER


K&E 1 DAY BY HOURS X 100 DIVISIONS 46 2050
KEUFFEL & ESSER CO.
FIGURE 8
TEMPERATURE HISTORY OF
COLD GASHOLDER
TEMPERATURE - °K
80
70
60
50
40
30
20
12
1
2
3
4
5
6
7
8
9
10
11
12
NOON
1
2
3
4
5
6
7
8
9
10
11
12

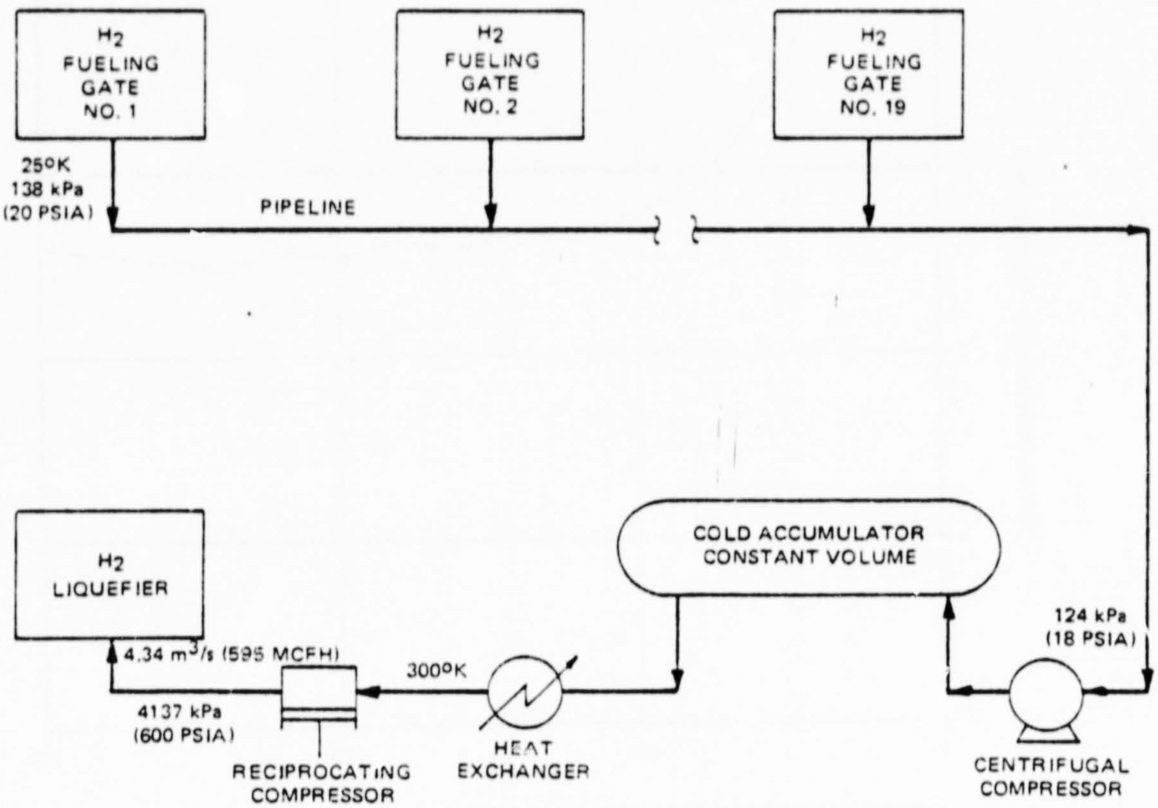


FIGURE 9
FLOW DIAGRAM
H_2 VENT GAS RECOVERY
USING CONSTANT VOLUME COLD ACCUMULATOR

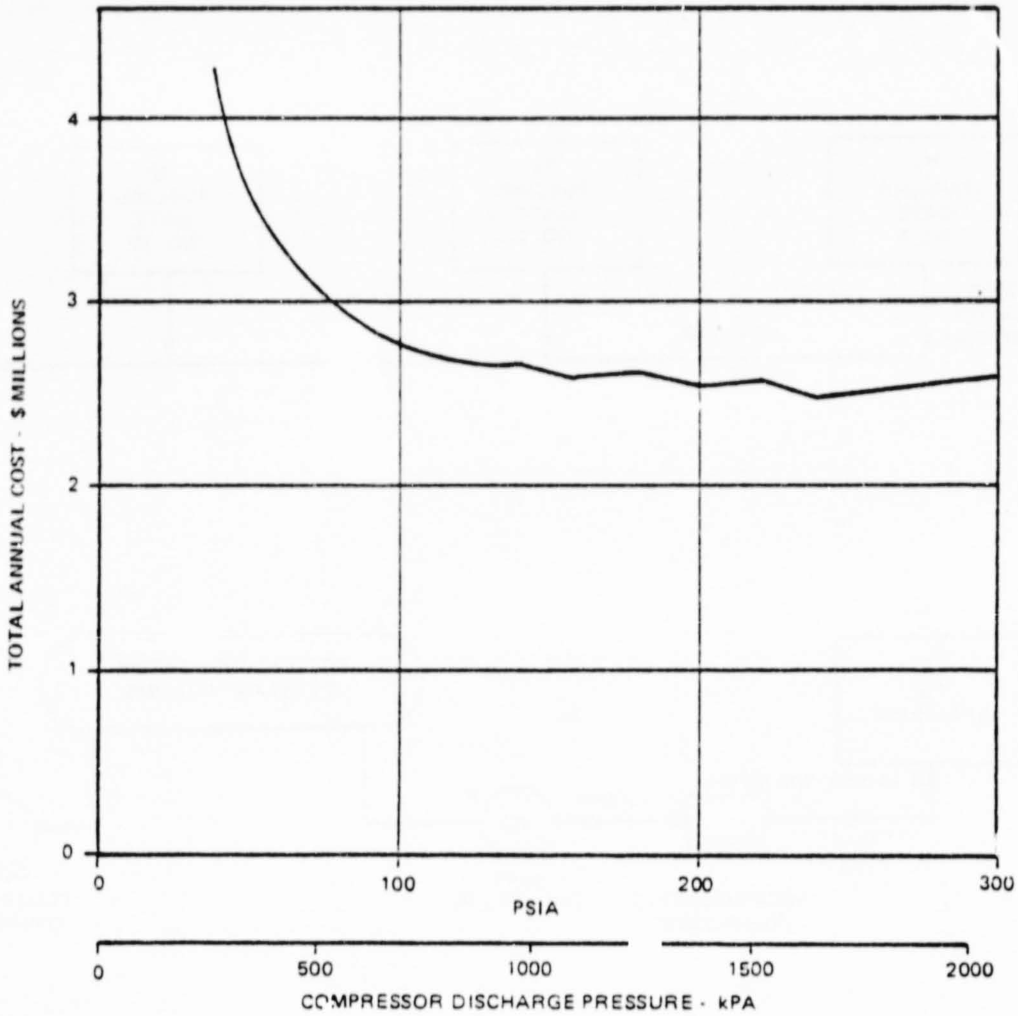


FIGURE 10
OPTIMIZATION OF COMPRESSOR
DISCHARGE PRESSURE
CONSTANT VOLUME COLD ACCUMULATOR

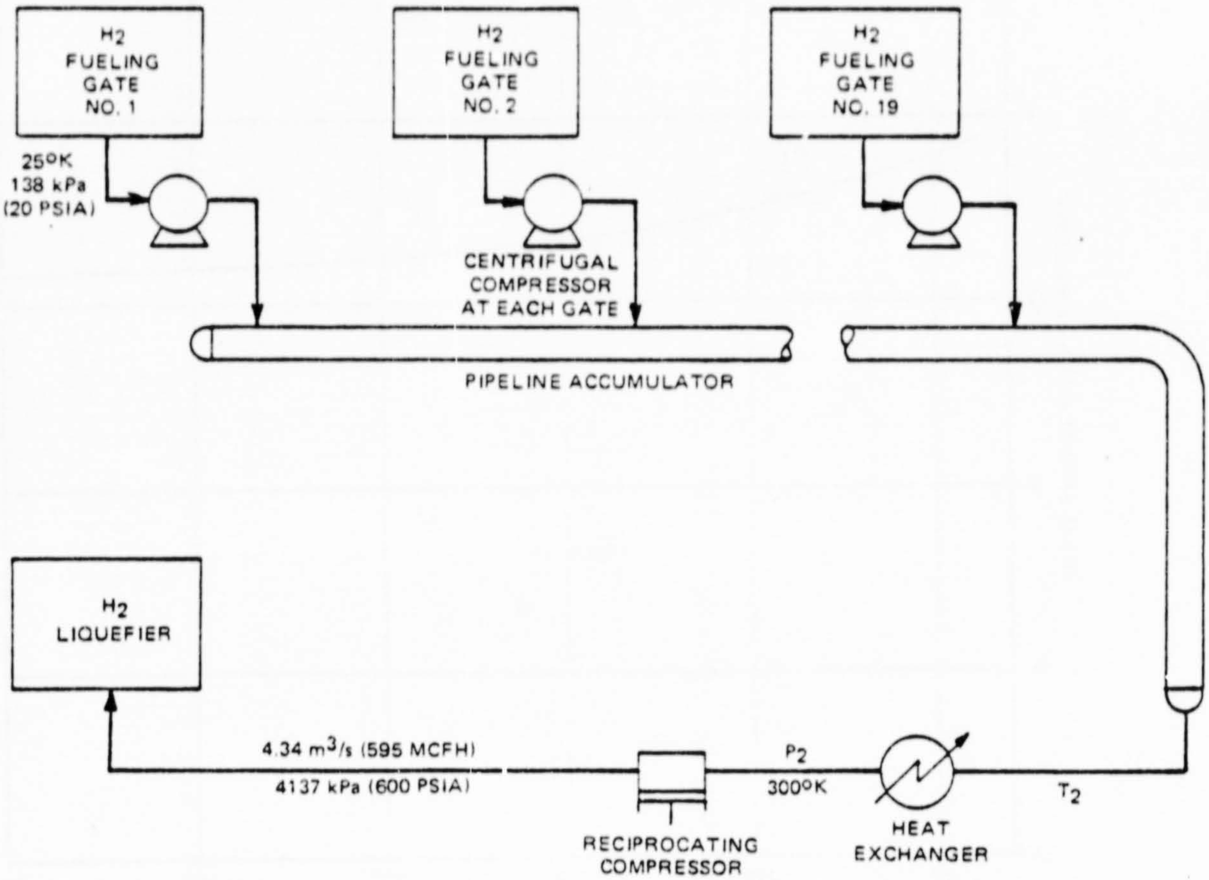


FIGURE 11
FLOW DIAGRAM
H_2 VENT PIPELINE
AS COLD ACCUMULATOR

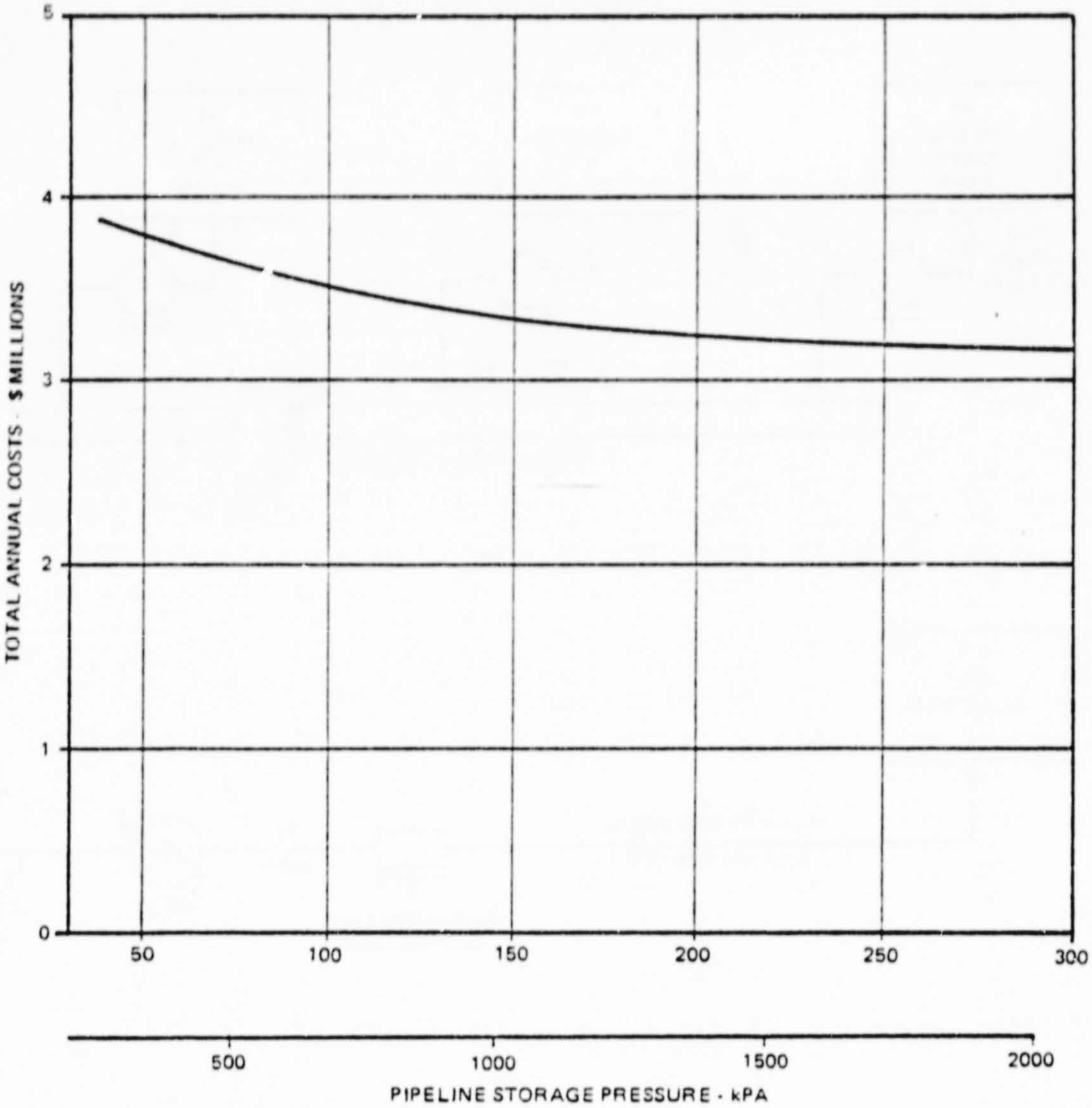
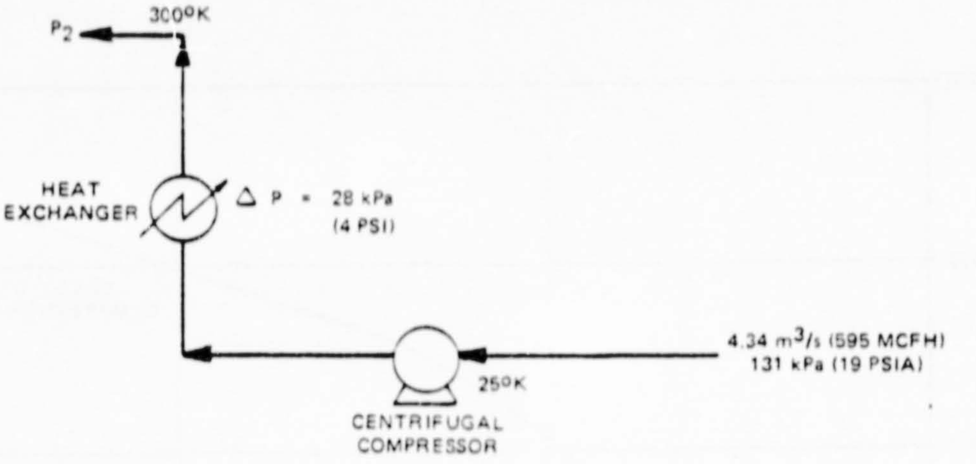
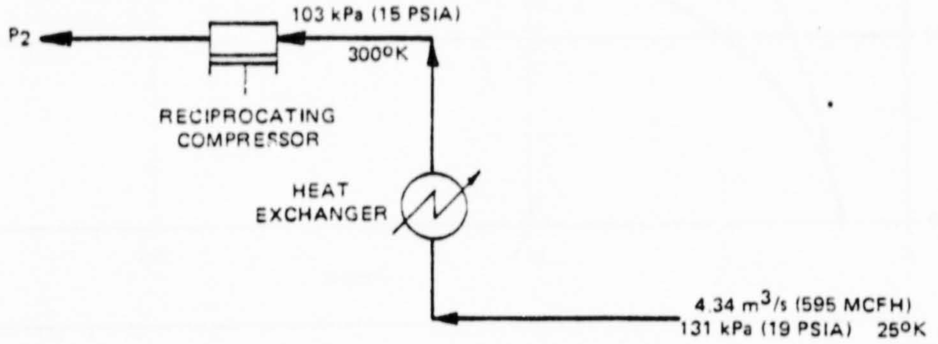


FIGURE 12
OPTIMIZATION OF PIPELINE
STORAGE PRESSURE
H_2 VENT PIPELINE USED AS ACCUMULATOR



COLD COMPRESSION



WARM COMPRESSION

FIGURE 13
WARM VS COLD COMPRESSION
OF H_2 VENT GAS

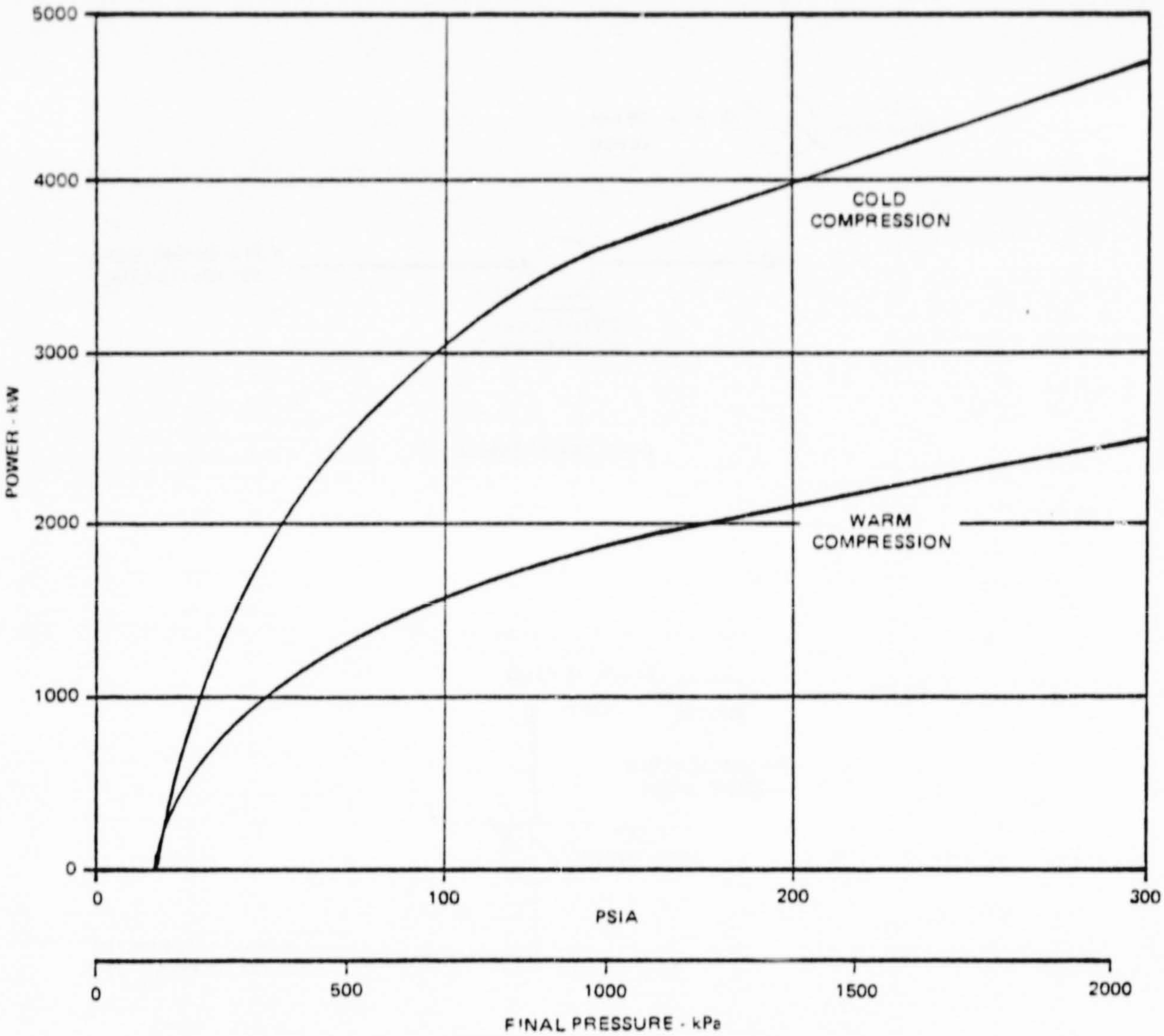


FIGURE 14
POWER CONSUMPTION
FOR WARM AND COLD COMPRESSION
OF H_2 VENT GAS

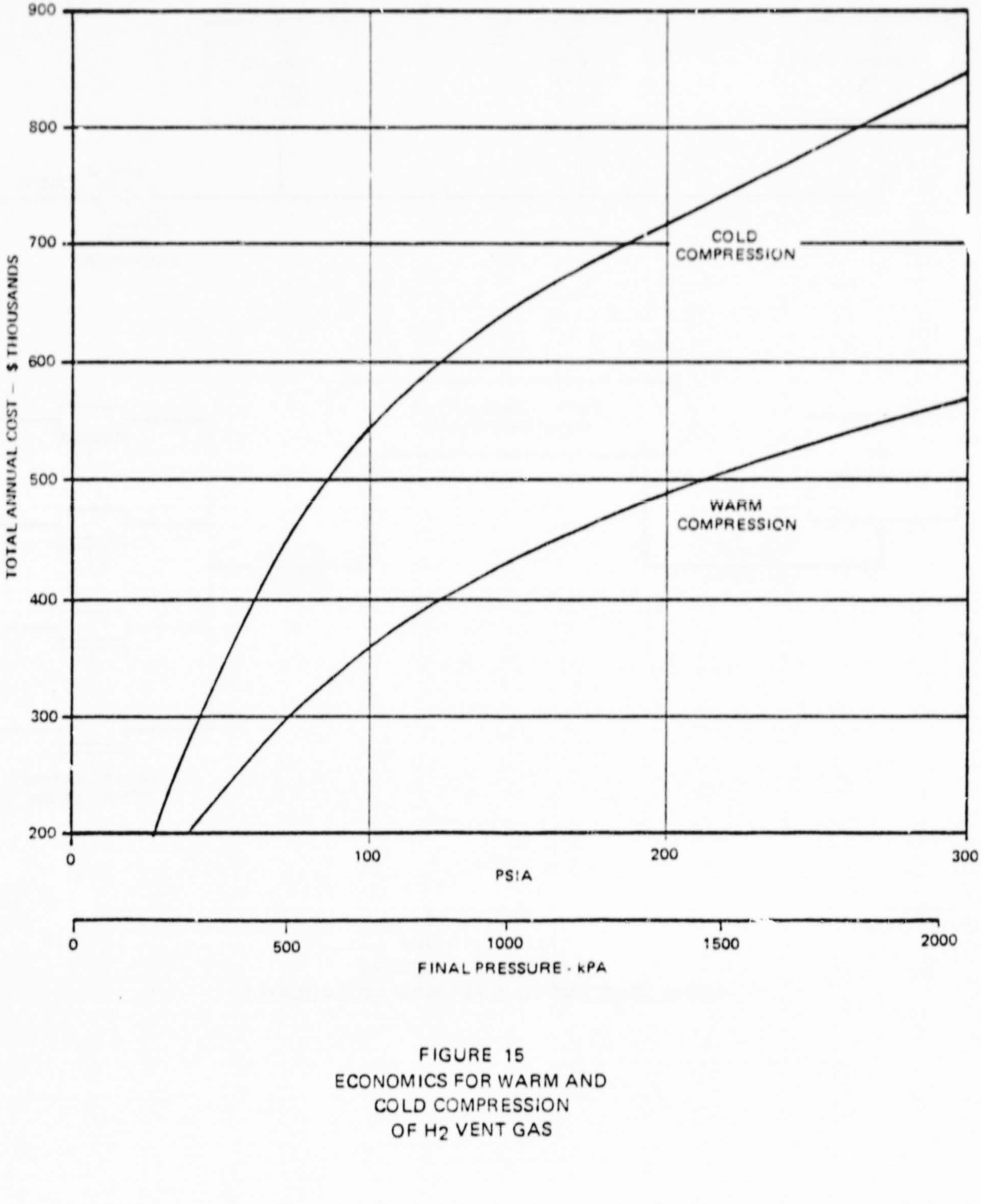


FIGURE 15
ECONOMICS FOR WARM AND
COLD COMPRESSION
OF H_2 VENT GAS

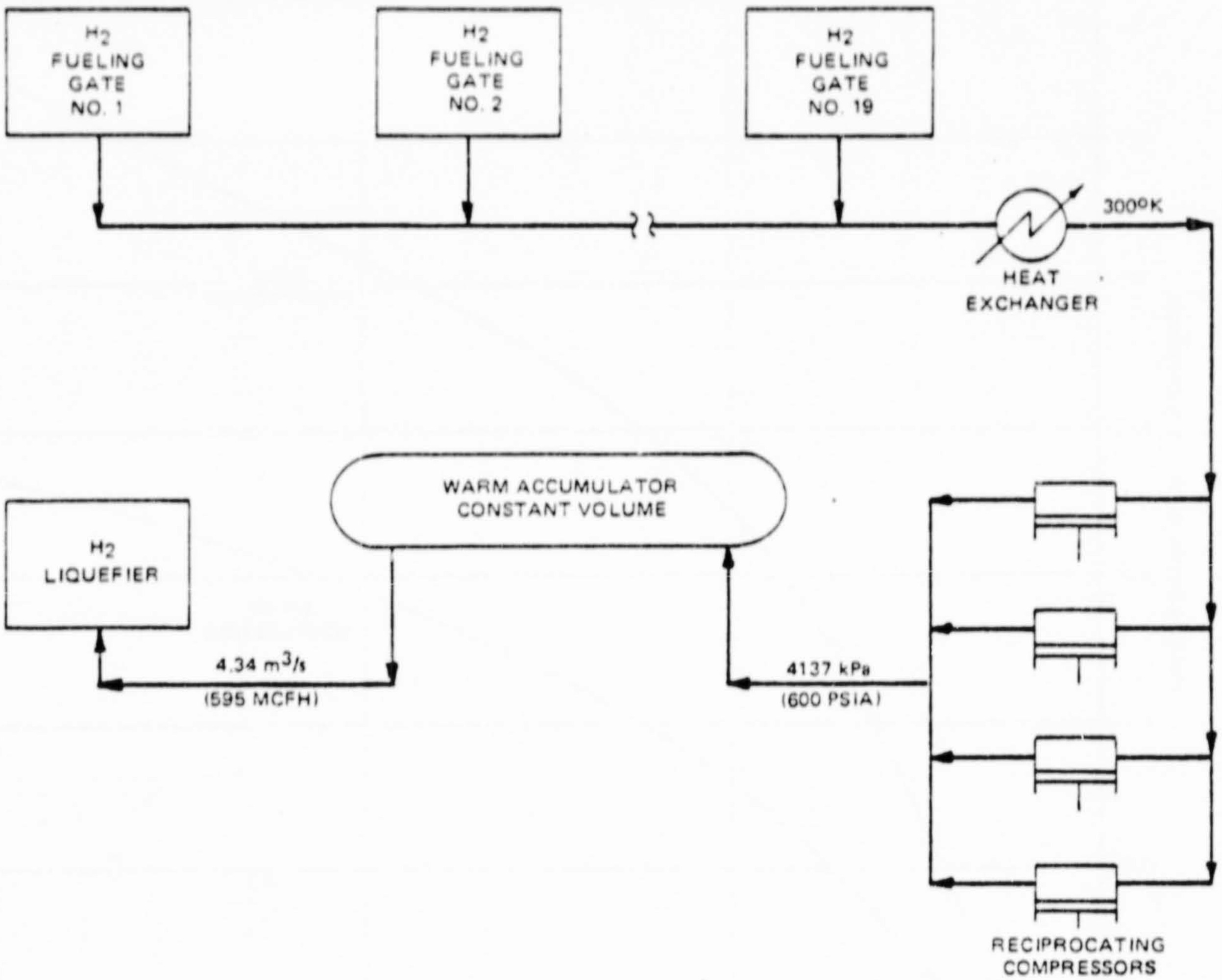


FIGURE 16
FLOW DIAGRAM
H_2 VENT GAS RECOVERY
USING CONSTANT VOLUME WARM ACCUMULATOR

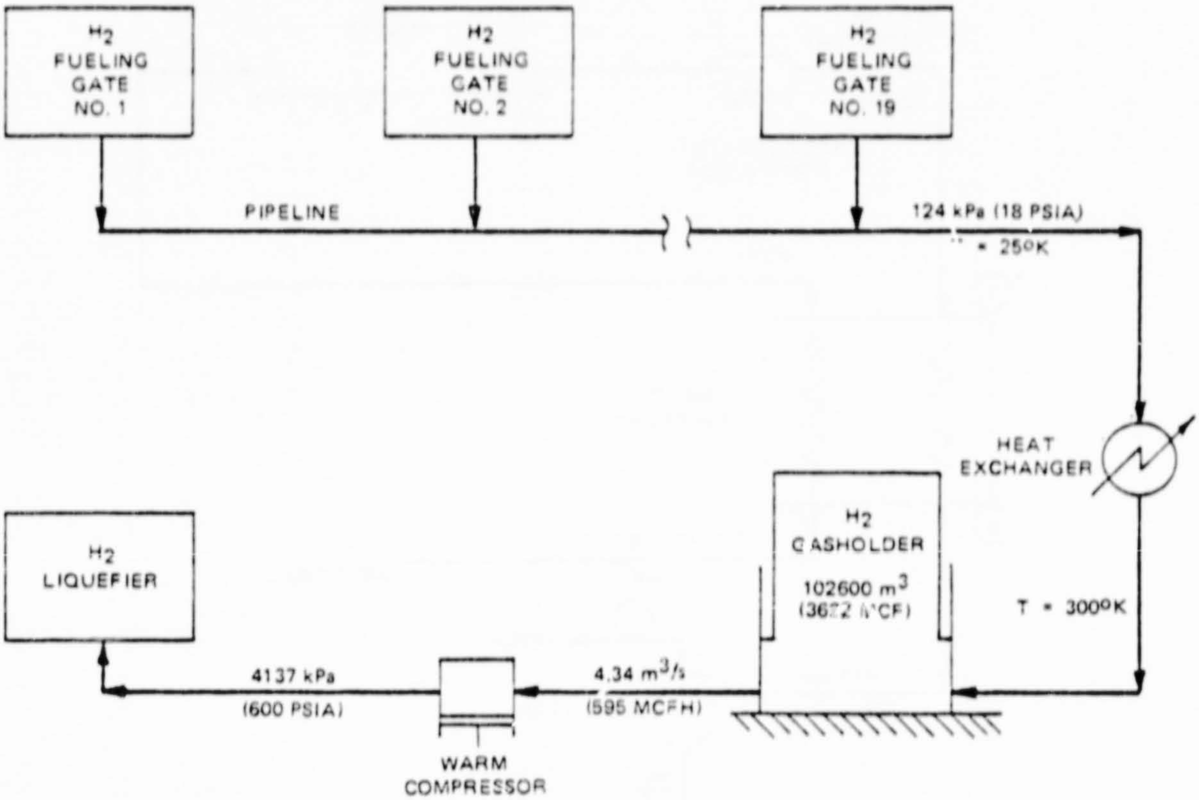


FIGURE 17
FLOW DIAGRAM
H_2 VENT GAS RECOVERY
USING CONSTANT PRESSURE WARM ACCUMULATOR

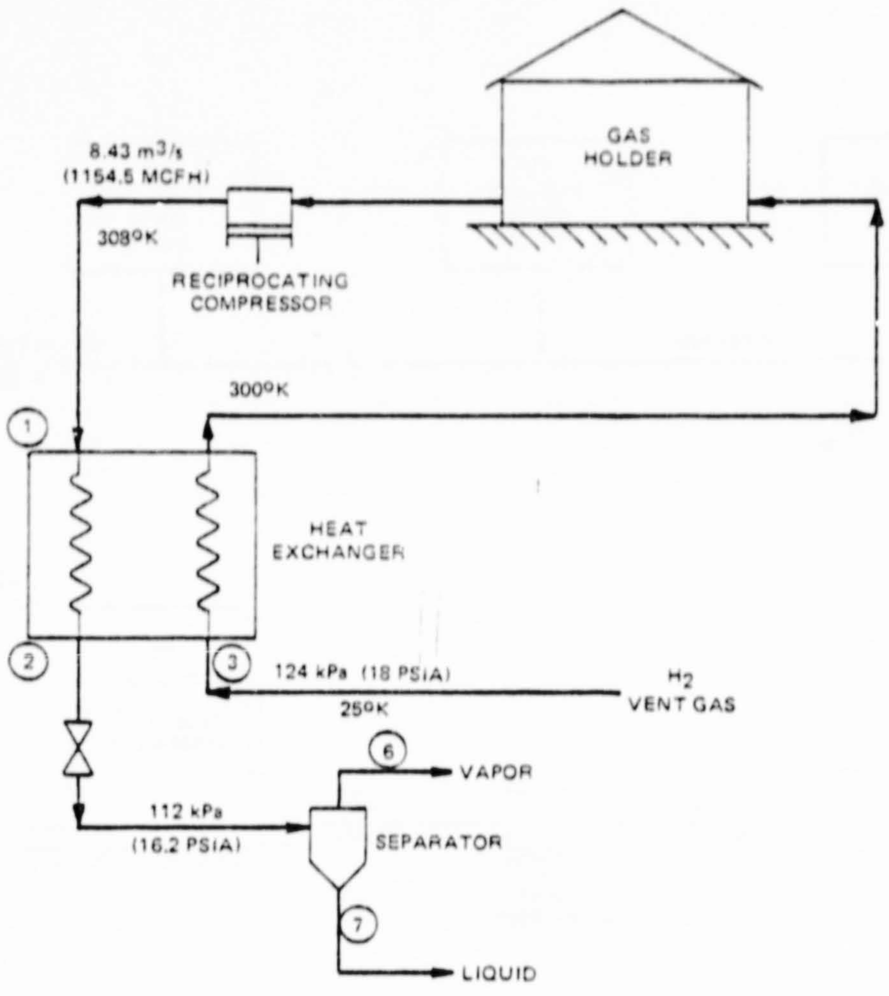
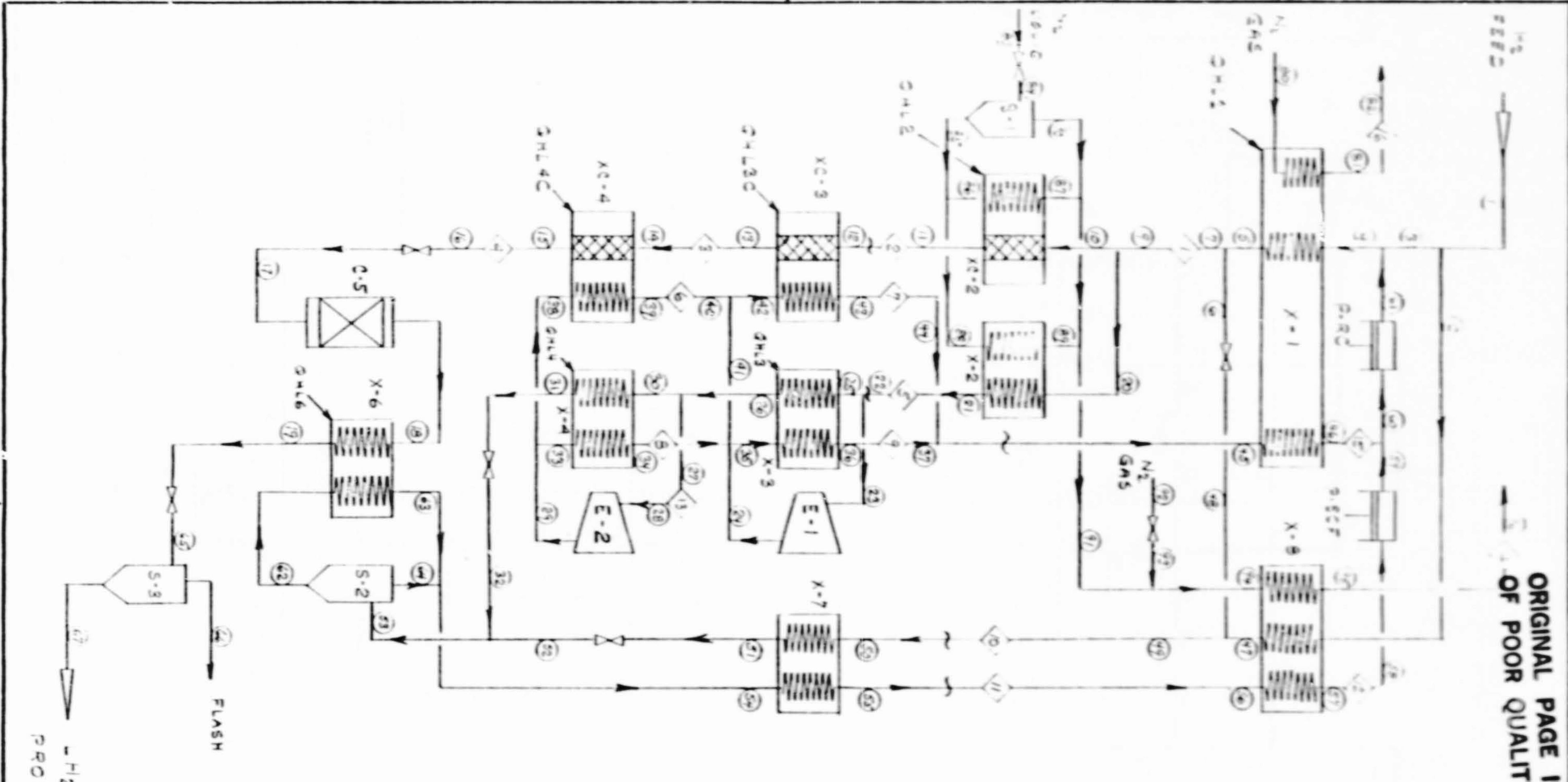


FIGURE 18
FLOW DIAGRAM
EVALUATION OF HEAT EXCHANGERS



(N) STREAM No. SEE TABLES 6,7

◇ PRESSURE DROP

P COMPRESSOR

E EXPANDER

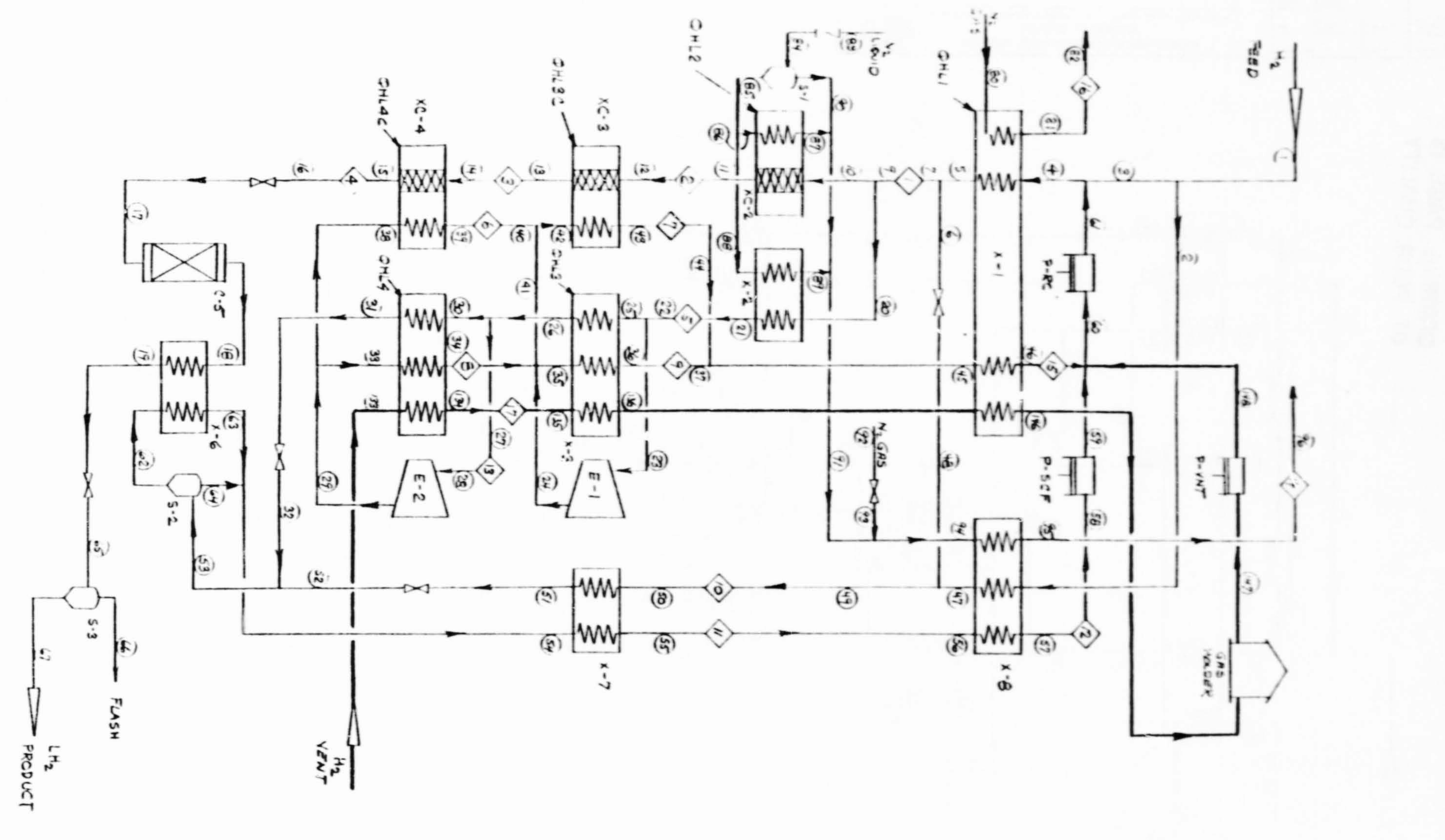
X HEAT EXCHANGER

S PHASE SEPARATOR

UNION CARBIDE CORPORATION
LINDE DIVISION
ENGINEERING DEPARTMENT • TONAWANDA, NEW YORK

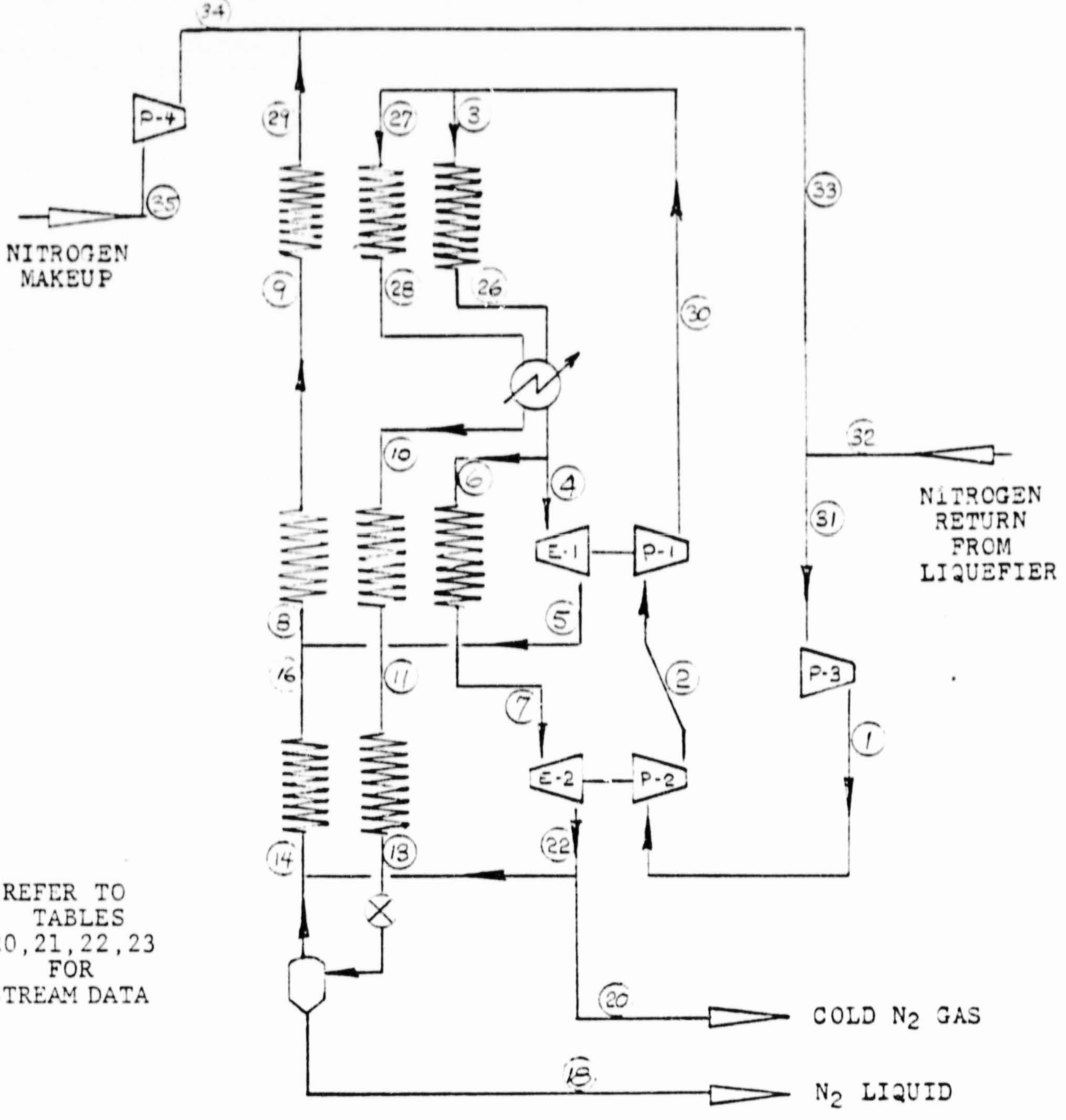
H_2 LIQUEFIER PROCESS MODEL WITHOUT H_2 VENT GAS RECOVERY

FIGURE 19



H2 LIQUEFIER PROCESS MODEL WITH H2 VENT GAS RECOVERY

FIGURE 2C



REFER TO
TABLES
20, 21, 22, 23
FOR
STREAM DATA

NITROGEN REFRIGERATOR PROCESS MODEL

FIGURE 21