1968

Natural gas pressure let-down utilization

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NATURAL GAS PRESSURE LET-DOWN UTILIZATION

A MASTERS DEGREE THESIS BY -

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OCTOBER 9, 1968

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ABSTRACT

In order to determine the feasibility for recovering energy at natural gas pressure let-down installations, a simple turbo-expander-generator refrigeration system is studied. Work and refrigeration capacity calculations are made for a pure methane stream of 20 MMscfd at pressures of 750 psia, 550 psia, and 350 psia which after preheating to +140°F or precooling to -120°F is expanded to 70 psia. After cost estimates for the system are made, a value/ton of refrigeration for each of the 123 cases studied is calculated and compared with the value/ton for a conventional refrigeration system operating at the same temperature level. Using this comparison it is found that let-down energy recovery appears advantageous at delivery points offering pipeline pressures from 200 psia to 800 psia, flow rates above 10 MMscfd and after precooling to expander inlet temperatures below -60°F but not so low as to produce an excess amount of liquid in the expander exhaust gas.

INTRODUCTION

In the transmission of natural gas by pipeline there is a considerable waste of potential energy at pressure let-down installations. These pressure let-downs usually occur at city-gate reduction stations and large industrial and utility delivery points. Through the use of turbo-expanders to generate electric power from this "free" expansion and utilization of the cold expander exhaust gas for refrigeration, moderate success has been achieved in the operation of LNG peak-shaving plants and ethane-propane recovery plants.[1]

It is the purpose of this study to examine the pressure let-down phenomenon with these four objectives in mind.

1) Conduct a literature search in an attempt to uncover any previous uses of the pressure let-down and define the current state-of-the-art as far as utilization is concerned. The collection of natural gas availability data, where availability is defined as the pressure, flow rate, and composition at which gas is supplied to major utility and industrial customers, should also be undertaken. This data would prove useful in estimating the potential for energy recovery and the range of let-down pressures available.

2) Since the project is also educational in purpose, the theory underlying expander operation as well as the calculations involved should

be studied in an attempt to develop a firm understanding of the isentropic expansion as applied to natural gas mixtures. This also includes a study of the APCI computer program for expander calculations.

3) Define the performance criteria for a simple turbo-expander refrigeration system. This takes into account not only the range of feasible operating conditions, but also the effect of these conditions on the system performance when they are varied.

4) Describe several of the best uses of the pressure let-down phenomenon and make suggestions for later use by APCI.

Using these four objectives as a guideline, a generalized study of pressure let-down utilization is conducted with the purpose of finding some economic incentive for using expanders to recover the wasted energy. A reduction in the scope of the study was eventually made by eliminating the fourth objective.

PROCEDURE AND RESULTS

The first step toward the objectives of this study is a literature search. The search is directed mainly towards studying previous uses of the pressure let-down and collecting natural gas availability data. Most of the literature dealing with pressure let-down utilization is found in the petroleum and cryogenic engineering trade journals. These sources revealed that the two most common types of pressure let-down utilization are ethane-propane recovery plants and LNG peak-shaving installations. In these installations the expansion energy is usually recovered through the use of compressor or generator loaded turbo-expanders while the cold expander exhaust gas is used to precool the inlet gas or for other refrigeration purposes. The efficiency of these plants, evidenced by their low power costs, appears to be their major economic advantage over conventional low temperature cycles.[2]

The question of what are "typical" pressures, flow rates, and compositions found at pipeline gas let-down installations is perhaps the greatest motivation behind the undertaking of a gas availability study. In the four expander-LNG installations presently operating or under construction in the United States, flow rates range from 50 MMscfd to 12 MMscfd while pressures range from 450 psig to as low as 180 psig.[3] Most of the gas availability information obtained is the result of direct correspondence with over 100 of the major gas transmission companies in the United States and Canada. A majority of the data is in the form of schematic pipeline maps and tabulations of gas pressures, flow rates and compositions. A study of data reveals what sort of operating conditions are available at pressure letdown points. Without resorting to statistical analysis it is apparent that pipeline delivery pressures range from as high as 850 psig to as low as 50 psig, with pressures between 350 psig and 100 psig comprising the majority. Flow rates range from 400 MMscfd down to thousands of standard cubic feet per day, with flow rates of 50 MMscfd to 5 MMscfd the most common. Compositions vary from one point to the next and are therefore difficult to specify. Natural gas components are usually methane, ethane, propane, butane, carbon dioxide, nitrogen, and a wide variety of heavier hydrocarbons. The pipeline operating conditions also fluctuate greatly due to seasonal and daily demand changes. As a result, most of the data obtained is either annual average or peak-design day data.[4]

The next step in the study is the selection and analysis of a simple turbo-expander refrigeration system. A turbo-expander rather than a Joule-Thompson expansion is chosen because it is desirable to include expansion work output in the economic analysis of the system. The simplest expander system is considered because of the general intentions of the study and the understanding that refinements to the system may be made at a later date. Figure 1 is a simplified flow diagram (for the cases in which the inlet gas is precooled) showing the carbon dioxide removal unit, dehydrator, precooling core, turbo-expander - generator system, and the refrigeration core. In the cases where preheated or ambient temperature gas is considered the precooling core is replaced by a furnace or is non-existent. The carbon dioxide removal unit is also removed for several cases of preheating.

The selection of a set of realistic expander operating conditions is aided by a study of the natural gas availability data. Three expander pressure ratios of 750 psia/70 psia, 550 psia/70 psia, and 350 psia/70 psia are considered. The inlet pressures all lie within the range indicated by the gas availability data. The outlet pressure of 70 psia is used because it is more or less a "typical" municipal distribution pressure. With the three pressure ratios set the gas inlet temperature is varied in 20°F internals from +140°F down to a temperature producing less than 10% liquid in the exhaust gas or about -100°F. The flow rate for each case was originally set at 1 MMscfd. This, however, has been increased to 20 MMscfd in an effort to use a more realistic figure and bring the calculations in line with the equipment cost data available. Calculations made at the 20 MMscfd rate are then extended to rates of 50 and 10 MMscfd to determine the effect of flow rate on the system's economics. In lieu of natural gas mixtures only pure methane is considered because of the general nature of the study and the complications involved with calculations considering natural gas mixtures.

With the operating conditions set the next step in the study is the calculation of the expander system output in terms of horsepower and the tons of refrigeration capacity along with the temperature level of that refrigeration. With inlet conditions and flow rate specified and assuming a 75% expander efficiency the gross work output and expander exhaust conditions may be determined for each case at the three pressure ratios and flow rates considered. By assuming system inlet and outlet conditions and the pressure drops through the auxiliary pieces of equipment the refrigeration capacity and temperatures at other points in the system are determined. Tables I - A, B, C list the refrigeration capacity and gross horsepower output for the three pressure ratios and flow rates while the temperatures and pressures at various points in the three systems are given in tables II - A, B, C.

With the completion of the system energy calculations it is now possible to consider the economics of the system. By sizing and pricing the major pieces of equipment as well as determining the annual direct costs and capital charges, the annual value per ton of refrigeration may be calculated for each of the 123 cases studied. These annual values may then be compared with the annual value per ton for a conventional refrigeration system operating at the same tonnage capacity and temperature level as the expander system.

The first step toward this objective is the sizing and pricing of the major pieces of process equipment. The number of major pieces varies according to the case being studied but generally consists of a carbon dioxide removal unit, dehydrator, precooling or preheating equipment, expander-gearbox-generator system, and a refrigeration core. Because of limitations in the cost data available the sizing and pricing is done for a flow rate of 20 MMscfd and later scaled up and down for 50 and 10 MMscfd.

The first peices of equipment to be considered are the precooling cores and preheating furnaces. For the cases requiring precooling (expander inlet temperature less than 80°F) Stewart-Warner brazed aluminum cores with a basic core size of 18" x 29" x 125" and a UA value of 125,000 BTU/HR^{o}F are used.[5] Core costs per cubic inch are available from APCI files.[6] With this information, the number of cores necessary and the total core cost for each case may be determined. Preheating furnaces (for expander inlet temperatures above 80^{o}F) are priced according to the heat duty. Details of these calculations are given in the sample calculations of Appendix II and the results are tabulated in Tables III - A,B,C.

For the removal of carbon dioxide from the process stream to prevent expander frosting a mono-ethanol amine type unit is used. This type of unit is chosen because of the ready availability of cost data and the general nature of the study does not warrant lengthy cost studies on individual pieces of equipment. Cost information for an MEA unit operating at 600 psia and handling 20 MMscfd is available from APCI files and is upgraded and down-graded to provide data for the other pressure ratios and flow rates considered.[7] The utility costs and investment costs are given in Table IV, along with the estimated utility and investment costs for a porous bed desiccant dehydrator.

In sizing and pricing the expander-generator system for each case, there are two requirements which must be fulfilled before a particular expander arrangement may be assigned. First, the gas enthalpy change per expander stage should not exceed 50 BTU/lb, or for a flow rate of 20 MMscfd the actual gross horsepower output must not exceed 678 HP. Secondly, the actual volumetric flow rate must meet specified criteria for a given expander wheel size.[8] Tables V - A,B,C, give the operating temperatures, actual volumetric flow rates and gross horsepower for all the 20 MMscfd cases studied. From the data it is apparent that all except some of the low temperature cases require a two stage expansion to comply with the maximum horsepower requirement. By splitting the single expansions into two equal pressure ratio expansions the horsepower requirement is fulfilled, and the expander wheel size may then be determined with the use of APCI expander specifications.

The most common type of expander-generator system used consists of a 6" expander in series with a 9" expander, both of which are connected by means of a gearbox to a generator. In several cases, however, a single 9" expander is connected via a gearbox to a generator. In all the systems considered prices include a lubrication system and explosion proof controls. These prices are given in Table VI. In assigning an expander-generator system to a particular case the horsepower and volumetric flow rate requirements must be complied with as well as the requirement that the actual design horsepower output must not exceed 80% of the rated gear horsepower for the expander-generator system. Tables VII, - A,B,C, give the design and gear horsepower as well as the unit cost for each of the 20 MMscfd cases studied.

With the sizing and pricing of the expanders completed all of the major pieces of equipment have been considered. The pricing of the refrigeration cores is omitted because of an inability to specify the conditions of all the streams passing through it. The stream being refrigerated (warm stream) has been considered as part of another process system. This creates a problem in specifying the temperature level of the refrigeration. The true temperature level is the exit temperature of the warm stream. Since this stream is unspecified, making it impossible to determine the temperatures, it is necessary to assign the exit temperature of the cold natural gas stream as the temperature level. The result is that the temperature levels specified are too high.

With all the process equipment sized and priced the next step is calculating the utility costs and capital charges for the purpose of determining the annual value per ton of refrigeration. First, the total installed cost (I) is computed using a Lange factor of 5.0 times the total cost of all the major pieces of process equipment. The large Lange factor (usually between 2.0 and 5.0) is used because of the cryogenic nature of the equipment. For the purpose of comparison with a conventional refrigeration plant it appears advantageous to use the larger value.

The total annual direct costs are taken as the sum of all the utility costs such as natural gas, water, power, chemicals, and maintenance costs. Since the study is comparative in nature, labor costs are assumed equal for both the expander and conventional refrigeration systems and therefore are neglected. The details of these cost calculations are given in Appendix II along with the capital cost calculations, which include depreciation, tax and insurance costs, and allowances for a 6% net profit after taxes. The electric power generated, assuming a 90% generator efficiency, is used as a credit against the annual direct and capital charges.

In order to obtain similar investment and cost information for the flow rates of 50 MMscfd and 10 MMscfd the results for the 20 MMscfd cases must be scaled up and down. The direct costs are multiplied by the factor 50/20 for the 50 MMscfd cases and by 10/20 for the 10 MMscfd cases. The total installed cost and other directly related costs are multiplied by the cost factors $f_{50} = (50/20)^{.6}$ and $f_{10} = (10/20)^{.6}$. Tables VIII - A-I gives an investment and cost summary for each flow rate and pressure ratio studied.

The next step in the calculations is the determination of the expander system's annual value per ton of refrigeration. Using cost data for conventional industrial vapor compression refrigeration systems, an investment summary including the total installed (I) along with utility and capital costs is developed.[9] This summary, similar to that given for the expander system in Table VIII, is presented in Tables IX-A-I. From the net annual values of the refrigeration produced, given in both sets of tables, the annual value per ton is calculated. The "value per ton" data might also be referred to as "price per ton" data. For the purposes of this study, however, it will be referred to as value per ton, meaning the worth of each ton of refrigeration to the producer with a 6% net profit considered.

With value/ton data available for both the expander and conventional system, a comparison is made in Tables X and Figures 2, 3, and 4. Conclusions drawn from a study of these tables and graphs are now discussed.

DISCUSSION OF THE RESULTS

From a careful study of the data obtained from the calculations, a substantial amount of performance information on natural gas expansion refrigeration systems may be developed. Tables I - I along with Figures 2 - 4 present this information and reveal several interesting trends as well as indicating useful performance criteria.

Tables I - A,B,C provide expander horsepower output and operating temperatures along with the refrigeration capacity and the related temperature level. Several performance trends are evident upon comparing the data for the three pressure ratios and flow rates studied. Aside from the intuitively obvious trends such as the approximately linear decrease in horsepower output with expander inlet temperature and the increase in horsepower and refrigeration capacity with the flow rate and pressure ratio there are some less obvious trends. For example, at a given pressure ratio and flow rate there is a rise in refrigeration capacity to a maximum at about -40°F and then a steady decline. This is probably due to greater heat duties in the precooling core thus reducing the amount of refrigeration available in the expander exhaust. Another trend is evidenced by the fact that as the pressure ratio increases the lowest attainable refrigeration temperature level without liquid in theexpander exhaust rises from about -140°F for the 350/70 ratio to about -100°F for the 750/70 ratio. This temperature remains stable, however, as the flow rate is increased for a given pressure ratio.

Tables II, III, and IV do not present much information of a general nature, but give specific information on system operating conditions as well as providing a comparison of the relative magnitude of equipment costs and indicating the sensitivity of these costs to changes in the system parameters.

Tables V - A,B,C are also not very general in nature, but do present expander inlet and outlet temperatures and actual flow rates along with the gross horsepower output for each case studied. The intermediate conditions for the cases requiring a two stage expansion are not given here because they are not vital to the study.

Tables VI and VII - A,B,C present expander cost information as well as the type of unit used. It is obvious from the data presented that the expander-generator system is one of the major cost items in the plant.

Tables VIII - A-I provide a general investment summary for all the cases in the expander refrigeration system at the three pressure ratios and flow rates studied. They include the total investment (I) as well as utility and capital costs. The credit value of the power generated is also considered so that a net annual cost is obtained. The net annual cost is more accurately described as the net annual value because of the fact that a 6% net profit after taxes has been included among the capital charges. Several trends are also observable here. The installed cost (I) increases, as expected, with increases in the flow rate and/or the pressure ratio. It tends to decrease, however, as the expander inlet temperature is lowered.

The lower total installed cost for the first two cased studied for each flow rate and pressure ratio are due to the higher expander inlet temperatures which permit the elimination of the feed treatment equipment.

Tables IV - A-I provide the same information as Tables VIII-A-I for a conventional industrial vapor compression refrigeration system. From cost data available for these conventional systems the total installed cost and power requirements for systems identical in tonnage capacity and temperature level to the expander systems studied are obtained. Here, the total installed cost (I) tends to increase directly with both the flow rate and pressure ratio considered. The magnitude of I is quite sensitive to changes in the temperature level. As the temperature level is lowered I increases exponentially for a given tonnage capacity. The rise is quite marked below levels of -90°F. The capital costs and utility costs are direct functions of I and therefore behave in a similar manner.

Tables VIII and IX provide the information that is used as the basis for comparing the relative merits of the expander and conventional refrigeration systems. Tables X-A-I and Figures 2-4 make this comparison, by comparing the total installed cost and the value per ton of refrigeration for the expander system with the same figures for a conventional system, an area of economic advantage for one of the systems may be found. With this area determined it is then possible to develop expander performance criteria which will define the conditions of economic feasibility.

Most of the data in Tables X-A-I is presented graphically in Figures 2-4. A study of these graphs is perhaps the best way to observe the economic advantages of the expander system.

It is apparent that the valve/ton curve for the conventional system is approximately identical for each flow rate studied and each pressure ratio as well. This indicates that the only factor affecting the valve/ton for the conventional system is the temperature level. For each pressure ratio studied the expander system value/ton curves at the individual flow rates are presented. At a constant pressure ratio the variation in value/ton with changes in flow rate is quite marked. The value/ton at 50 MMscfd is about half that for the 10 MMscfd flow rate. The variations, with the flow rate held constant while the pressure ratio changes, are more moderate.

The point of intersection of the conventional system and expander system curves marks the feasibility boundary between the two systems. This intersection occurs between the -110^{o}F and -90^{o}F temperature levels and tends to be in the lower end of that range for the 10 MMscfd flow rate and close to -90^{o}F for a 50 MMscfd flow rate. At temperature levels below the intersection temperature the expander system has a far lower value/ton than the conventional system. Since the value/ton figure is really more of a price/ton the expander system would be more competitive at the lower temperature levels. The opposite is true of temperature levels above the intersection. The conventional system holds the competitive edge.

CONCLUSIONS AND PERFORMANCE CRITERIA

From the value/ton plots (Figure 2,3,4) it is apparent that the expander system has an economic advantage over the conventional vapor compression refrigeration system at refrigeration temperature levels in or below the temperature range of -80°F to -110°F. The location of the cut-off temperature within the -80° to -110°F range is affected by the flow rate and to a lesser extent the pressure ratio of the expansion. At a flow rate of 50 MMscfd the cut off temperature is closer to -80°F while it is closer to -110°F for the 10 MMscfd flow rate. The 750 psia/70 psia pressure ratio value/ton data also reveals a slight advantage over the two lower pressure ratios or more specifically the expander inlet pressure. Thus it appears that the best possibilities for expander use occur at flow rates greater than 10 MMscfd, expander inlet pressures above 350 psia, and expander inlet temperatures less than -60°F.

The three system parameters that are varied, namely the flow rate, expander inlet temperature, and the pressure ratio have definite limits as far as expander refrigeration system feasibility is concerned. By defining these limits a set of general performance criteria for natural gas letdown refrigeration systems may be developed.

Expander inlet temperature - The lower limit for this parameter is set by the formation of liquid in the 70 psia expander exhaust gas.

Liquid begins to form in the exhaust at -220°F or expander inlet temperatures below -60°F for the 750 psia cases, -80°F for the 550 psia cases and -110°F for the 350 psia cases. The refrigeration temperature levels corresponding to these inlet temperatures are -110°F, -120°F and -145°F respectively.

The upper limit on the expander inlet temperature appears to be controlled by the cut off temperature level in the value/ton comparison or simply the temperature level below which the expander system has an economic advantage. These temperatures range from -80°F for the 50 MMscfd flow rate to -110°F for the 10 MMscfd flow rate. The corresponding expander inlet temperatures then range from -40°F to -80°F and the lower limits range from -60°F to -110°F. The two inlet temperature ranges overlap for a flow rate of 50 MMscfd and the 750/70 pressure ratio. The largest operating range for the inlet temperature exists for a pressure ratio of 350/70 and a 10 MMscfd flow rate.

Flow Rate - The lower limit on this parameter appears to be around 10 MMscfd. The low refrigeration capacity and horsepower output for lesser flow rates result in a large value/ton for the refrigeration thus making the competitive nature of such a system doubtful.

No upper limit on the flow rate exists since the advantage of the expander system appears to strengthen as the flow rate is increased. This is illustrated by the fact that the value/ton at a given temperature level and 50 MMscfd flow rate is about half that at 10 MMscfd.

Pressure Ratio - The pressure ratio or more specifically the expander inlet pressure poses some problems in attempting to define its limits. The lower limit on inlet pressure appears to be about 200 psia since lower pressures make expansion to a competitive temperature level a physical impossibility. The upper level seems to be about 800 psia and is governed by two factors. The first is the availability of natural gas at pipeline pressures above 800 psia. Gas deliveries above this pressure are extremely rare. The second is the problem of overlapping the temperature ranges of the expander inlet temperature upper and lower limits. The inlet gas must be precooled so that the desired temperature level may be attained after the expansion. At high pressures, however, the inlet temperature at which liquid forms in the exhaust is about -60^{o}F. This raises the maximum temperature level attainable above the cut off temperature and thus affects the economic advantage of the expander system. Sacrificing refrigeration tonnage to lower the temperature level below the cuttoff temperature is the required remedy.

Using these criteria as a guide line it can be stated that natural gas let-down energy recovery appears advantageous at delivery points offering pipeline pressures from 200 psia to 800 psia, flow rates above 10 MMscfd and after precooling the gas to temperatures that will not produce an unmanagable amount of liquid in the turbo-expander exhaust gas.

APPENDIX I

FIGURE 1

NATURAL GAS PRESSURE LET-DOWN TURBO-EXPANDER REFRIGERATION SYSTEM

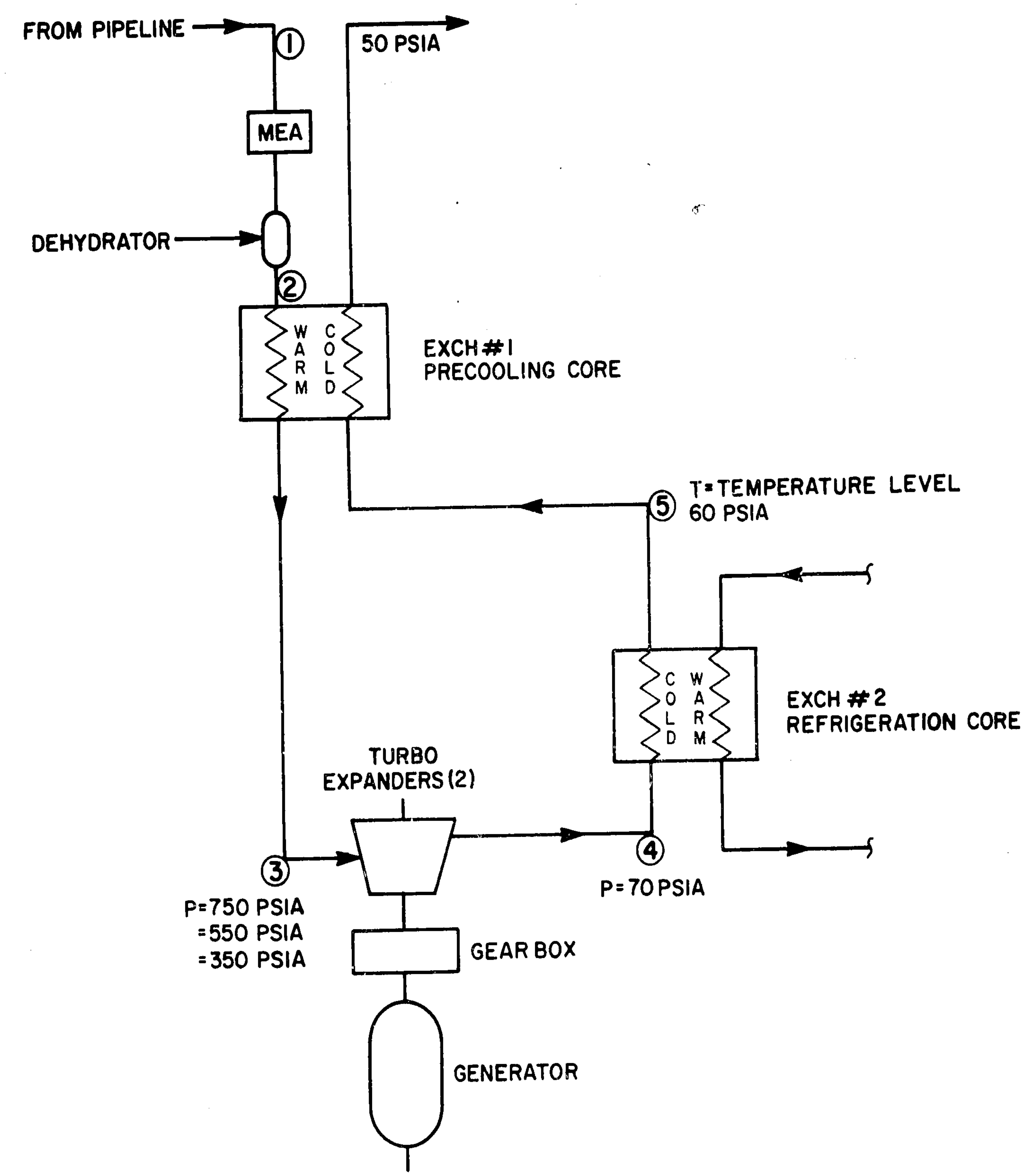


Figure 2

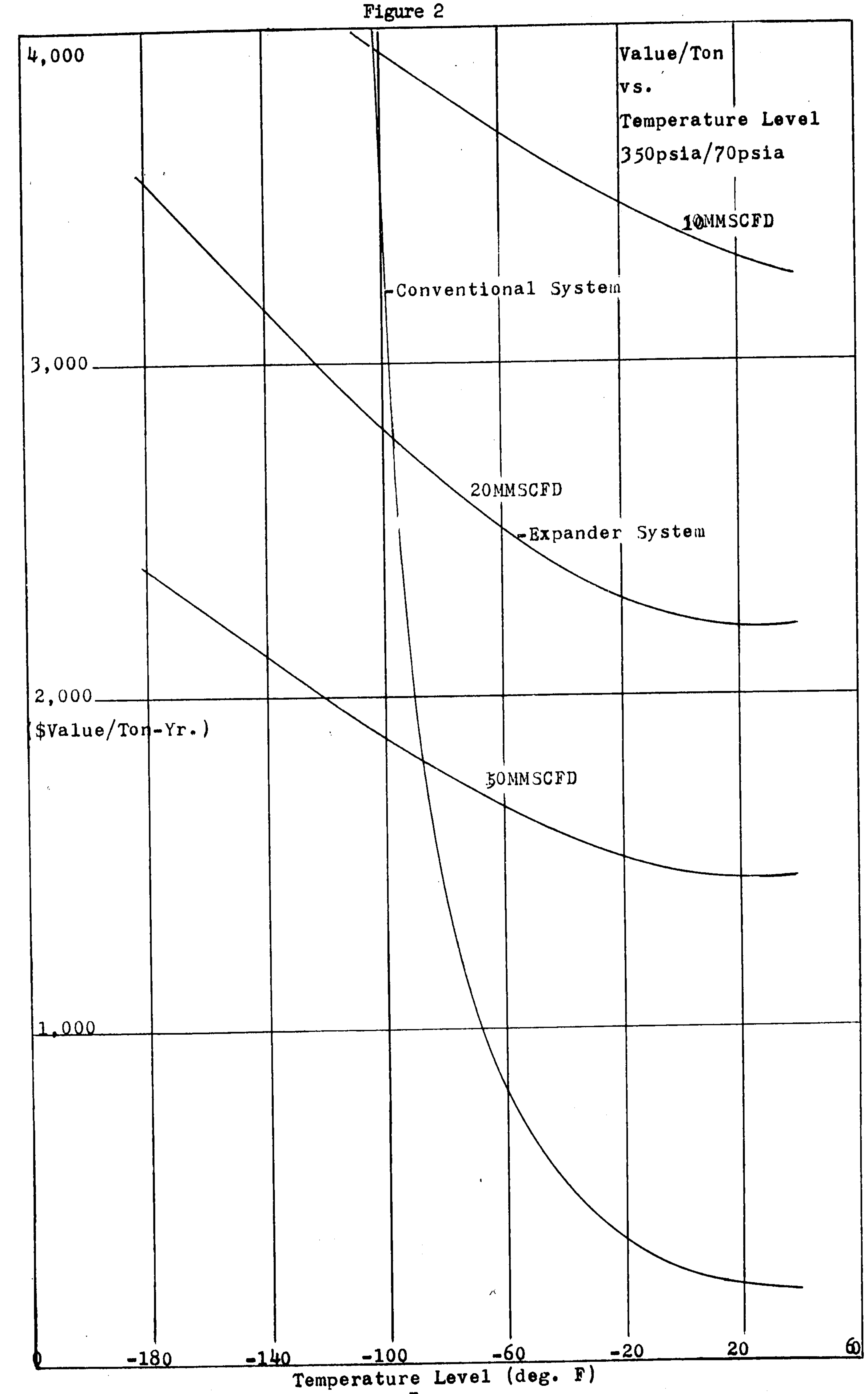
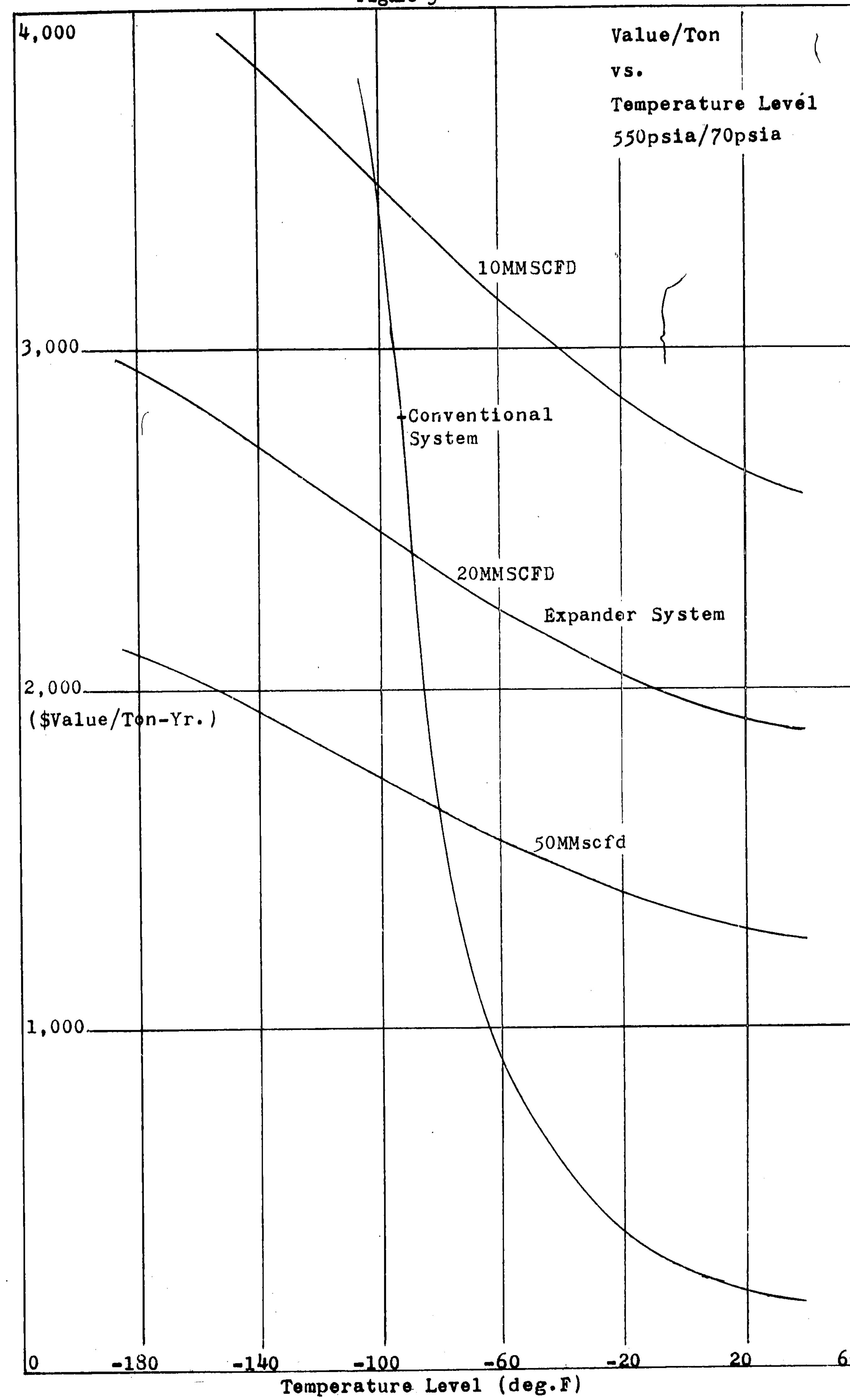

Value/Ton
vs.
Temperature Level
350psia/70psia
10MMSCFD
-Conventional System
20MMSCFD
-Expander System
50MMSCFD
4,000
3,000
2,000
($Value/Ton-Yr.)
1,000
0
-180
-140
-100
-60
-20
20
60
Temperature Level (deg. F)

Figure 3



Value/Ton
vs.
Temperature Level
550psia/70psia
4,000
3,000
2,000
($Value/Ton-Yr.)
1,000
0
10MMSCFD
Conventional System
20MMSCFD
Expander System
50MMscfd
-180
-140
-100
-60
-20
20
60
Temperature Level (deg.F)

Figure 4

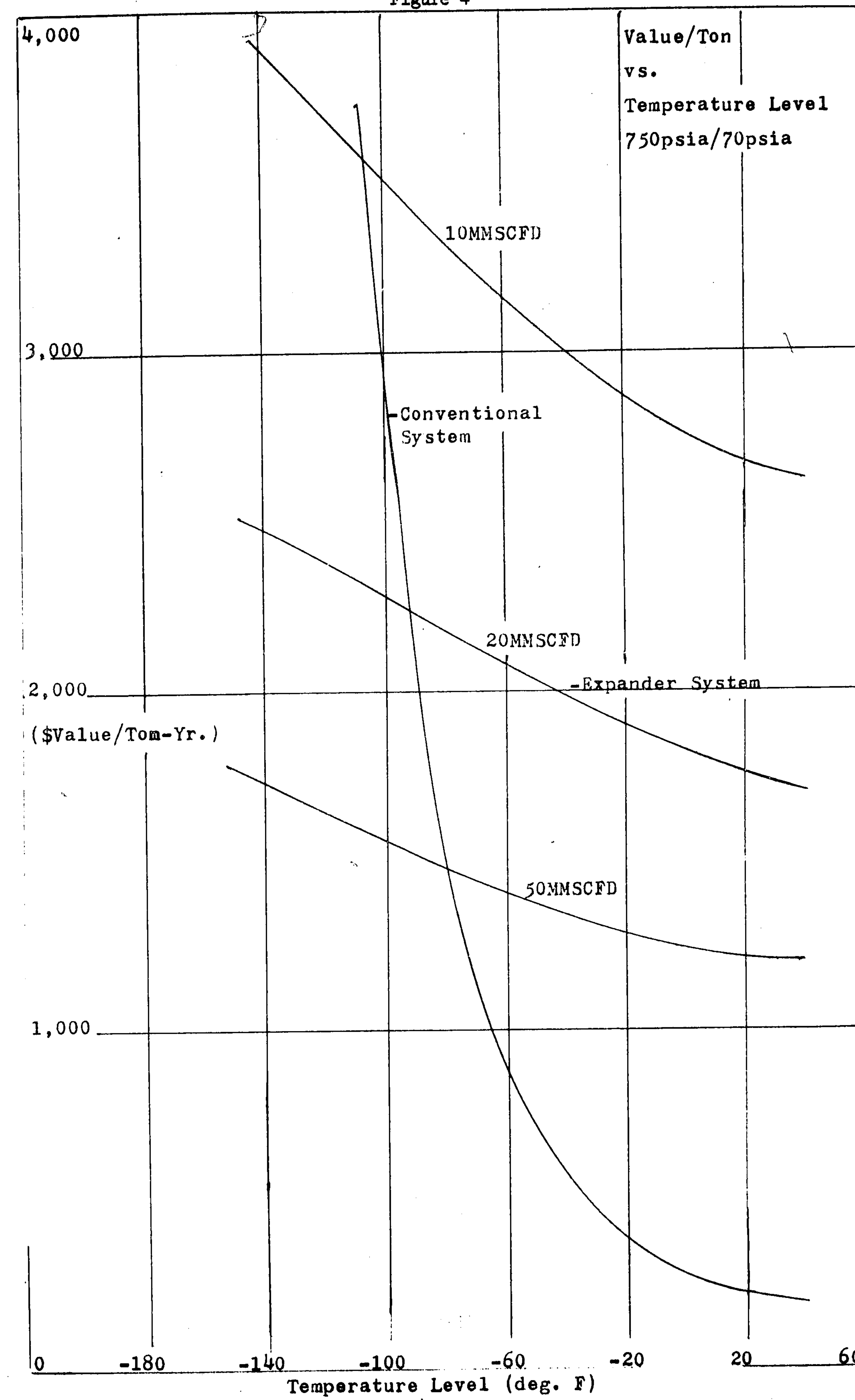

Value/Ton
vs.
Temperature Level
750psia/70psia
4,000
3,000
2,000
1,000
($Value/Tom-Yr.)
10MMSCFD
-Conventional
System
20MMSCFD
-Expander System
50MMSCFD
0
-180
-140
-100
-60
-20
20
60
Temperature Level (deg. F)

TABLE I-A

HORSEPOWER OUTPUT AND REFRIGERATION CAPACITY

Pressure Ratio = 350 PSIA/70 PSIA

Case No.	Temp. Expander Inlet	Temp. Expander Outlet	Temp. Level of Refrigeration	@ 10MMscfd HP	@ 10MMscfd Tons	@ 20MMscfd HP	@ 20MMscfd Tons	@ 50MMscfd HP	@ 50MMscfd Tons
1	140°F	- 8	40	494	36	988	72	2470	179
2	120	-24	40	475	48	950	96	2370	242
3	100	-41	40	455	61	910	122	2280	304
4	80	-57	40	435	73	870	146	2180	366
5	60	-74	40	415	86	830	172	2070	428
6	40	-90	31	397	91	795	182	1985	455
7	20	-107	8	377	87	755	174	1880	435
8	0	-123	-12	357	83	715	166	1780	415
9	-20	-139	-35	338	79	675	158	1690	395
10	-40	-156	-57	317	74	634	148	1585	370
11	-60	-172	-80	295	70	590	140	1475	350
12	-80	-189	-104	275	65	550	130	1375	325
13	-100	-209	-130	267	62	515	124	1286	310
14	-120	-219(L)	-156	229	56	457	112	1140	280
15	-140	-219(L)	-183	218	53	435	106	1085	265

TABLE I-B

HORSEPOWER OUTPUT AND REFRIGERATION CAPACITY

Pressure Ratio = 550 PSIA/70 PSIA

Case No.	Temp. Expander Inlet	Temp. Expander Outlet	Temp. Level of Refrig.	@ 10MMscfd		@ 20MMscfd		@ 50MMscfd	
				HP	Tons	HP	Tons	HP	Tons
1	140	35	40	615	44	1230	87	3075	218
2	120	13	40	600	58	1200	115	3000	288
3	100	-7	40	570	70	1140	140	2850	350
4	80	-97	40	540	79	1080	158	2700	395
5	60	-109	40	515	93	1030	185	2575	462
6	40	-125	27	490	116	980	231	2450	578
7	20	-143	5	453	112	906	224	2265	560
8	0	-159	-18	430	108	860	216	2150	537
9	-20	-173	-43	408	100	815	200	2040	500
10	-40	-190	-68	370	94	740	187	1850	467
11	-60	-209	-95	342	88	683	175	1710	437
12	-80	-219(L)	-121	305	83	610	165	1525	412
13	-100	-219(L)	-152	278	76	556	152	1390	380
14	-120	-219(L)	-187	258	74	515	143	1290	358

TABLE I-C

HORSEPOWER OUTPUT AND REFRIGERATION CAPACITY

Pressure Ratio = 750 PSIA/70 PSIA

	Temp. Expander Inlet	Temp. Expander Outlet	Temp. Level of Refrig.	@ 10MMscfd HP	@ 10MMscfd Tons	@ 20MMscfd HP	@ 20MMscfd Tons	@ 50MMscfd HP	@ 50MMscfd Tons
1	140	-69	40	645	82	1289	164	3220	410
2	120	-84	40	616	93	1232	186	3080	465
3	100	-100	40	587	105	1175	210	2940	525
4	80	-115	40	558	117	1117	234	2790	585
5	60	-131	40	528	138	1059	272	2650	690
6	40	-147	29	500	131	1000	262	2500	655
7	20	-162	3	468	125	938	250	2340	625
8	0	-178	-21	437	118	874	236	2180	590
9	-20	-196	-50	407	112	815	224	2150	560
10	-40	-212	-78	371	104	742	308	1850	520
11	-60	-219(L)	-109	343	98	686	196	1710	490
12	-80	-219(L)	-144	307	91	615	182	1580	455

TABLE II-A

SYSTEM TEMPERATURES (°F)

Pressure Ratio = 350 PSIA/70PSIA

Case No.	System Inlet 1*	Precool or Preheat Inlet 2	Expander Inlet 3	Expander Outlet 4	Refrigeration Core Outlet 5	System Outlet 6
1	80	80	140	-8	40	70
2	80	80	120	-24	40	70
3	80	80	100	-41	40	70
4	80	80	80	-57	40	70
5	80	80	60	-74	40	70
6	80	80	40	-90	31	70
7	80	80	20	-107	8	70
8	80	80	0	-123	-12	70
9	80	80	-20	-139	-35	70
10	80	80	-40	-156	-57	70
11	80	80	-60	-172	-80	70
12	80	80	-80	-189	-104	70
13	80	80	-100	-202	-130	70
14	80	80	-120	-219(L)	-156	70
15	80	80	-140	-219(L)	-183	70

*See diagrams for system points

TABLE II-B

SYSTEM TEMPERATURES (^oF)

Pressure Ratio - 550 PSIA/70 PSIA

Case No.	System Inlet 1	Precool or Preheat Inlet 2	Expander Inlet 3	Expander Outlet 4	Refrigeration Core Outlet 5	System Outlet 6
1	80	80	140	-52	40	70
2	80	80	120	-71	40	70
3	80	80	100	-87	40	70
4	80	80	80	-98	40	70
5	80	80	60	-115	40	70
6	80	80	40	-133	27	70
7	80	80	20	-144	5	70
8	80	80	0	-160	-18	70
9	80	80	-20	-177	-43	70
10	80	80	-40	-192	-68	70
11	80	80	-60	-205	-95	70
12	80	80	-80	-219(L)	-121	70
13	80	80	-100	-219(L)	-152	70
14	80	80	-120	-219(L)	-187	70

TABLE II-C

SYSTEM TEMPERATURES (^{o}F)

Pressure Ratio = 750 PSIA/70PSIA

Case No.	System Inlet 1	Precool or Preheat Inlet 2	Expander Inlet 3	Expander Outlet 4	Refrigeration Core Outlet 5	System Outlet 6
1	80	80	140	-69	40	70
2	80	80	120	-84	40	70
3	80	80	100	-100	40	70
4	80	80	80	-115	40	70
5	80	80	60	-131	40	70
6	80	80	40	-147	29	70
7	80	80	20	-162	3	70
8	80	80	0	-178	-21	70
9	80	80	-20	-196	-50	70
10	80	80	-40	-212	-78	70
11	80	80	-60	-219(L)	-109	70
12	80	80	-80	-219(L)	-144	70

TABLE III-A

PREHEATING FURNACE AND PRECOOLING CORE COSTS @ 20MMscfd

Furnace Cost = $2000/MMBTU/Hr
Unit Core Cost = $9,900/Core

Pressure Ratio = 350 PSIA/70 PSIA

Case No.	Number of Basic Cores	$ Cost / IN^3	Total Core or Furnace Cost
1	Furnace	-	$ 2,600
2	Furnace	-	1,800
3	Furnace	-	1,000
4	-	-	-
5	0.255	.300	4,680
6	0.570	.175	6,500
7	0.885	.160	9,220
8	1.09	.155	11,030
9	1.22	.150	11,930
10	1.38	.145	13,100
11	1.55	.140	14,270
12	1.57	.137	14,050
13	1.61	.135	14,200
14	1.61	.135	14,300
15	1.62	.135	14,600

TABLE III-B

PREHEATING FURNACE AND PRECOOLING CORE COSTS @ 20MMscfd

Pressure Ratio = 550 PSIA/70 PSIA

Case No.	Number of Basic Cores	$ Cost / IN^3	Total Core or Furnace Cost
1	Furnace	-	$ 2,500
2	Furnace	-	1,700
3	Furnace	-	900
4	-	-	-
5	0.248	.31	3,300
6	0.557	.18	6,500
7	0.769	.17	8,500
8	0.934	.158	9,600
9	0.98	.156	10,000
10	1.15	.150	11,300
11	1.20	.147	11,500
12	1.27	.144	11,900
13	1.27	.144	11,900
14	1.24	.146	11,900

TABLE III-C

PREHEATING CURNACE AND PRECOOLING CORE COSTS @ 20MMscfd

Pressure Ratio = 750 PSIA/70 PSIA

Case No.	Number of Basic Cores	$ Cost / IN^3	Total Core or Furnace Cost
1	Furnace	-	$ 2,800
2	Furnace	-	2,000
3	Furnace	-	1,000
4	-	-	-
5	0.264	.300	5,200
6	0.584	.182	6,900
7	0.730	.171	8,100
8	0.904	.162	9,600
9	0.970	.158	10,000
10	1.015	.156	10,300
11	1.053	.155	10,600
12	1.062	.154	10,700

TABLE IV

MEA-CO_2 REMOVAL UNIT AND DEHYDRATOR UNIT

Annual Utility Costs and Unit Cost @ 20MMscfd

Pressure Ratio = 350/70

MEA Unit

Utilities: $ 28,550/Yr
Unit Cost: $151,500

Dehydrator Unit

Utilities: $ 5,000/Yr
Unit Cost: $50,000

Pressure Ratio = 550/70

MEA Unit

Utilities: $ 31,810/Yr
Unit Cost: $168,300

Dehydrator Unit

Utilities: $ 5,000/Yr
Unit Cost: $50,000

Pressure Ratio = 750/70

MEA Unit

Utilities: $ 35,000/Yr
Unit Cost: $185,000

Dehydrator Unit

Utilities: $ 5,000/Yr
Unit Cost: $50,000

TABLE V-A

EXPANDER OPERATING DATA @ 20MMscfd

Pressure Ratio = 350 PSIA/70 PSIA

Case No.	Expander Inlet Temp.	Expander Outlet Temp.	ACFM Inlet	ACFM Outlet	Gross Horsepower
1	140	-8	638	2490	988
2	120	-24	617	2400	950
3	100	-41	593	2300	910
4	80	-57	570	2130	870
5	60	-74	542	2050	830
6	40	-90	518	1985	795
7	20	-107	496	1860	755
8	0	-123	467	1800	715
9	-20	-139	442	1690	675
10	-40	-156	417	1600	634
11	-60	-172	390	1480	590
12	-80	-189	358	1370	550
13	-100	-209	329	1250	515
14	-120	-219(L)	298	1140	457
15	-140	-219(L)	267	1020	435

TABLE V-B

EXPANDER OPERATING DATA @ 20MMscfd

Pressure Ratio = 550 PSIA/70 PSIA

Case No.	Expander Inlet Temp.	Expander Outlet Temp.	ACFM Inlet	ACFM Outlet	Gross Horsepower
1	+140	-52	452	2520	1230
2	120	-71	437	2310	1200
3	100	-87	422	2210	1140
4	80	-98	407	2135	1080
5	60	-115	392	2045	1030
6	40	-133	377	1992	980
7	20	-144	362	1868	906
8	0	-160	347	1780	860
9	-20	-177	332	1684	815
10	-40	-192	317	1595	740
11	-60	-205	302	1488	683
12	-80	-219(L)	286	1387	610
13	-100	-219(L)	271	1230	556
14	-120	-219(L)	256	1029	515

TABLE V-C

EXPANDER OPERATING DATA @ 20MMscfd

Pressure Ratio = 750 PSIA/70 PSIA

Case No.	Expander Inlet Temp.	Expander Outlet Temp.	ACFM Inlet	ACFM Outlet	Gross Horsepower
1	140	-69	292	2130	1289
2	120	-84	279	1985	1232
3	100	-100	267	1920	1175
4	80	-115	254	1800	1117
5	60	-131	241	1740	1059
6	40	-147	227	1640	1000
7	20	-162	213	1550	938
8	0	-178	199	1440	874
9	-20	-196	186	1330	815
10	-40	-212	174	1240	742
11	-60	-219(L)	161	1120	686
12	-80	-219(L)	147	1030	615

TABLE VI

EXPANDER HORSEPOWER AND UNIT COST

Expander - Gearbox - Generator System

6"   9"   Expanders in Series = 2-6"x9"

Unit Number	Gear Horsepower	Unit Cost	Maximum Allowable Design Horsepower
1	750	$110,000	600
2	1000	120,000	800
3	1500	135,000	1200
4	2000	150,000	1600

9" Expander = 1 - 9"

Unit Number	Gear Horsepower	Unit Cost	Maximum Allowable Design Horsepower
1	1000	$ 90,000	800
2	750	82,500	600

TABLE VII-A

EXPANDER UNITS AND COST @ 20MMscfd

Pressure Ratio = 350 PSIA/70 PSIA

Case No.	Actual Design Horsepower	Maximum Gear Horsepower of Unit	Type of Unit	Expander Cost
1	988	1500	2-6" x 9"	$135,000
2	950	1500	2-6" x 9"	135,000
3	910	1500	2-6" x 9"	135,000
4	870	1500	2-6" x 9"	135,000
5	830	1500	2-6" x 9"	135,000
6	795	1000	2-6" x 9"	120,000
7	755	1000	2-6" x 9"	120,000
8	717	1000	2-6" x 9"	120,000
9	675	1000	1-9"	90,000
10	634	1000	1-9"	90,000
11	590	750	1-9"	82,500
12	550	750	1-9"	82,500
13	515	750	1-9"	82,500
14	457	750	1-9"	82,500
15	415	750	1-9"	82,500

TABLE VII-B

EXPANDER UNITS AND COST @ 20MMscfd

Pressure Ratio = 550 PSIA/70 PSIA

Case No.	Actual Design Horsepower	Maximum Gear Horsepower of Unit	Type of Unit	Expander Cost
1	1230	2000	2-6" x 9"	$150,000
2	1200	1500	2-6" x 9"	135,000
3	1140	1500	2-6" x 9"	135,000
4	1080	1500	2-6" x 9"	135,000
5	1030	1500	2-6" x 9"	135,000
6	980	1500	2-6" x 9"	135,000
7	906	1500	2-6" x 9"	135,000
8	860	1500	2-6" x 9"	135,000
9	815	1500	2-6" x 9"	135,000
10	740	1000	2-6" x 9"	120,000
11	683	1000	2-6" x 9"	120,000
12	610	1000	2-6" x 9"	120,000
13	556	750	2-6" x 9"	110,000
14	515	750	2-6" x 9"	110,000

TABLE VII-C

EXPANDER UNITS AND COST @ 20MMscfd

Pressure Ratio = 750 PSIA/70 PSIA

Case No.	Actual Design Horsepower	Maximum Gear Horsepower of Unit	Type of Unit	Expander Cost
1	1289	2000	2-6" x 9"	$150,000
2	1232	2000	2-6" x 9"	150,000
3	1175	1500	2-6" x 9"	135,000
4	1117	1500	2-6" x 9"	135,000
5	1059	1500	2-6" x 9"	135,000
6	1000	1500	2-6" x 9"	135,000
7	938	1500	2-6" x 9"	135,000
8	874	1500	2-6" x 9"	135,000
9	815	1500	2-6" x 9"	135,000
10	742	1000	2-6" x 9"	120,000
11	686	1000	2-6" x 9"	120,000
12	615	1000	2-6" x 9"	120,000

TABLE VIII-A

INVESTMENT SUMMARY-EXPANDER SYSTEM @ 10MMscfd

Case No.	I	Direct Cost	Capital Charges	Gross Annual Costs	Power Value	Net Operating Costs
1	$ 483,000	$16,800/Yr	$102,300/Yr	$119,100/Yr	$23,300/Yr	$ 95,800/Yr
2	487,000	16,300	103,000	119,300	22,400	96,900
3	1,191,000	61,900	253,000	314,900	22,900	292,000
4	1,190,000	61,600	252,000	313,600	20,500	293,100
5	1,208,000	62,700	256,000	318,700	19,600	299,100
6	1,160,000	61,100	246,000	307,100	18,800	298,300
7	1,170,000	61,400	248,000	309,400	17,800	291,600
8	1,177,000	61,500	249,000	310,500	16,900	293,600
9	1,071,000	58,300	227,000	285,300	15,800	269,500
10	1,078,000	58,400	228,000	286,400	14,900	271,500
11	1,051,000	57,300	223,000	280,300	13,900	266,400
12	1,050,000	57,300	222,000	279,300	13,000	266,300
13	1,051,000	57,300	223,000	280,300	12,100	268,200
14	1,053,000	57,700	224,000	281,700	10,800	270,900
15	1,054,000	57,700	224,000	281,700	10,300	270,900

TABLE VIII-B

INVESTMENT SUMMARY-EXPANDER SYSTEM @ 20MMscfd

Pressure Ratio = 350/70

Case No.	I	Direct Costs	Capital Charges	Gross Annual Costs	Value of Power	Net Operating Costs
1	$ 683,000	$23,800/Yr	$144,700/Yr	$168,500/Yr	$46,600/Yr	$121,900/Yr
2	687,000	22,900	145,800	168,700	44,700	124,000
3	1,685,000	87,600	357,000	444,600	43,800	400,800
4	1,681,000	87,200	356,000	443,200	41,000	402,200
5	1,707,000	88,300	362,000	450,300	39,100	411,200
6	1,640,000	86,300	348,000	434,300	37,500	396,800
7	1,653,000	86,700	351,000	437,700	35,600	402,100
8	1,662,000	86,900	352,000	438,900	33,700	405,200
9	1,517,000	82,300	321,000	403,300	31,600	371,700
10	1,522,000	82,500	323,000	405,500	29,800	375,700
11	1,487,000	81,400	315,000	396,400	27,800	368,600
12	1,485,000	81,300	314,000	395,300	25,900	369,400
13	1,487,000	81,400	315,000	396,400	24,200	372,200
14	1,490,000	81,600	316,000	397,600	21,500	376,100
15	1,491,000	81,600	316,000	397,600	20,500	377,100

TABLE VIII-C

INVESTMENT SUMMARY-EXPANDER SYSTEM @ 50MMscfd

Pressure Ratio = 350/70

Case No.	I	Direct Costs	Capital Charges	Gross Annual Costs	Value of Power	Net Operating Costs
1	$1,180,000	$ 31,200/Yr	$251,000/Yr	$282,200/Yr	$11 ,300/Yr	$165,900/Yr
2	1,190,000	29,700	252,000	281,700	112,000	169,700
3	2,910,000	151,500	617,000	768,500	109,500	659,000
4	2,920,000	151,900	615,000	766,900	102,500	664,400
5	2,950,000	152,500	626,000	778,500	97,800	680,700
6	2,840,000	149,100	602,000	751,100	93,900	657,200
7	2,860,000	150,300	607,000	757,300	89,000	668,300
8	2,880,000	150,100	609,000	759,100	84;300	674,800
9	2,620,000	142,600	555,000	697,600	79,000	618,600
10	2,640,000	143,500	559,000	702,500	74,500	628,000
11	2,570,000	140,400	545,000	685,400	69,500	615,900
12	2,570,000	140,400	545,000	683,400	64,700	618,700
13	2,570,000	140,400	545,000	685,400	60,500	624,900
14	2,580,000	141,100	546,000	687,100	53,700	633,400
15	2,580,000	141,100	546,000	687,100	51,200	635,900

TABLE VIII-D

INVESTMENT SUMMARY-EXPANDER SYSTEM @ 10MMscfd

Pressure Ratio = 550/70

Case No.	I	Direct Costs	Capital Charges	Gross Costs	Value of Power	Net Operating Costs
1	$ 707,000	$25,400/Yr	$150,000/Yr	$175,400/Yr	$29,100/Yr	$146,300/Yr
2	653,000	23,100	138,000	161,100	28,300	132,800
3	1,250,000	56,500	265,000	321,500	26,400	295,100
4	1,250,000	55,900	265,000	320,900	25,300	295,600
5	1,261,000	56,300	268,000	324,300	24,300	300,000
6	1,270,000	56,600	269,000	325,600	23,000	302,600
7	1,280,000	56,800	271,000	327,800	23,000	304,800
8	1,282,000	57,000	272,000	329,000	20,200	308,800
9	1,284,000	57,000	272,000	329,000	19,100	309,900
10	1,235,000	56,400	262,000	318,400	17,500	300,900
11	1,240,000	55,600	263,000	318,600	16,000	302,600
12	1,240,000	55,100	263,000	318,100	14,400	303,700
13	1,206,000	55,400	256,000	311,400	13,100	298,300
14	1,206,000	56,400	256,000	312,400	12,000	300,400

TABLE VIII-E

INVESTMENT SUMMARY-EXPANDER SYSTEM @ 20MMscfd

Pressure Ratio = 550/70

Case No.	I	Direct Costs	Capital Charges	Gross Costs	Value of Power	Net Operating Costs
1	$1,000,000	$38,300/Yr	$212,000/Yr	$250,300/Yr	$58,100/Yr	$192,200/Yr
2	925,000	34,900	196,000	230,900	56,500	174,400
3	1,765,000	90,900	374,000	464,900	52,800	412,100
4	1,765,000	89,900	374,000	463,800	50,600	413,200
5	1,784,000	90,300	378,000	468,300	48,600	419,700
6	1,798,000	90,800	381,000	471,800	46,000	425,800
7	1,809,000	91,100	384,000	485,100	42,900	442,200
8	1,815,000	91,300	385,000	486,300	40,300	446,000
9	1,817,000	91,300	385,000	486,300	38,200	448,100
10	1,748,000	89,200	371,000	460,200	35,000	425,200
11	1,755,000	89,400	372,000	461,400	31,900	429,500
12	1,755,000	89,400	372,000	461,400	28,700	432,700
13	1,705,000	89,800	361,000	450,800	26,100	424,700
14	1,705,000	89,800	361,000	450,800	24,000	426,800

TABLE VIII-F

INVESTMENT SUMMARY-EXPANDER SYSTEM @ 50MMscfd

Pressure Ratio = 550/70

Case No.	I	Direct Costs	Capital Charges	Gross Annual Costs	Value of Power	Net Operating Costs
1	$1,730,000	$ 74,600/Yr	$367,000/Yr	$441,600/Yr	$145,200/Yr	$296,400/Yr
2	1,600,000	65,200	339,000	404,200	141,000	263,200
3	3,050,000	189,500	646,000	835,500	132,000	703,500
4	3,050,000	186,500	646,000	832,500	126,500	706,000
5	3,090,000	198,200	655,000	853,200	121,500	731,700
6	3,110,000	188,800	660,000	848,800	115,000	733,800
7	3,120,000	189,600	661,000	850,600	107,000	743,600
8	3,140,000	189,400	665,000	854,400	101,000	753,300
9	3,140,000	189,400	665,000	854,400	95,500	758,900
10	3,020,000	186,000	640,000	826,000	87,500	738,500
11	3,030,000	185,800	642,000	827,800	79,700	747,100
12	3,030,000	186,000	642,000	828,800	71,700	756,100
13	2,950,000	188,000	625,000	813,000	65,200	747,800
14	2,950,000	188,000	625,000	813,000	60,000	753,000

TABLE VIII-G

INVESTMENT SUMMARY-EXPANDER SYSTEM @ 10MMscfd

Pressure Ratio = 750/70

Case No.	I	Direct Costs	Capital Charges	Gross Costs	Gross Value of Power	Net Operating Costs
1	$ 543,000	$18,100/Yr	$115,000/Yr	$133,100/Yr	$30,300/Yr	$102,800
2	538,000	17,500	114,600	132,600	28,500	104,100
3	1,312,000	59,900	278,000	337,900	27,700	310,200
4	1,310,000	58,900	277,000	335,900	26,300	309,600
5	1,327,000	59,700	281,000	340,700	24,900	315,200
6	1,332,000	59,900	283,000	342,900	23,600	319,300
7	1,338,000	60,000	284,000	364,000	22,100	341,900
8	1,342,000	60,200	284,000	364,200	20,600	343,600
9	1,343,000	60,300	285,000	365,300	19,200	346,300
10	1,290,000	58,600	274,000	372,400	17,500	354,900
11	1,290,000	58,800	274,000	372,200	16,200	356,000
12	1,290,000	58,800	274,000	372,200	14,500	357,700

TABLE VIII-H

INVESTMENT SUMMARY-EXPANDER SYSTEM @ 20MMscfd

Pressure Ratio = 750/70

Case No.	I	Direct Costs	Capital Charges	Gross Costs	Gross Value of Power	Net Operating Costs
1	$ 768,000	$26,600/Yr	$162,800/Yr	$189,400/Yr	$60,600/Yr	$128,800/Yr
2	762,500	25,400	161,500	186,900	57,000	219,900
3	1,857,000	97,000	394,000	491,000	55,300	435,700
4	1,850,000	95,000	392,000	487,000	52,500	434,500
5	1,876,000	96,200	398,000	494,200	49,800	444,400
6	1,884,000	96,500	400,000	496,500	47,100	449,400
7	1,890,000	96,600	401,000	497,600	44,200	453,400
8	1,898,000	96,900	402,000	498,900	41,200	457,700
9	1,900,000	97,000	403,000	500,000	38,400	461,600
10	1,826,000	94,700	387,000	481,700	35,000	446,700
11	1,828,000	94,800	387,000	481,800	32,400	449,400
12	1,828,000	94,800	387,000	481,800	29,000	452,800

TABLE VIII-I

INVESTMENT SUMMARY-EXPANDER SYSTEM @ 50MMscfd

Pressure Ratio-= 750/70

Case No.	I	Direct Costs	Capital Charges	Gross Costs	Gross Value of Power	Net Operating Costs
1	$1,329,000	$ 48,800/Yr	$282,000/Yr	$330,800/Yr	$151,500/Yr	$179,300
2	1,318,000	45,900	279,000	324,900	142,500	182,400
3	3,213,000	199,200	681,000	880,200	138,000	742,200
4	3,200,000	195,600	678,000	873,600	131,000	742,600
5	3,245,000	197,700	688,000	885,700	124,500	761,200
6	3,259,000	197,500	692,000	889,500	118,000	771,500
7	3,270,000	198,400	694,000	892,400	110,500	781,900
8	3,283,000	198,200	695,000	893,200	103,000	790,200
9	3,287,000	198,100	696,000	894,100	96,000	798,100
10	3,159,000	194,900	670,000	866,900	87,500	779,400
11	3,162,000	194,800	670,000	866,800	81,000	785,800
12	3,162,000	194,800	670,000	866,800	72,500	790,300

TABLE IX-A

INVESTMENT SUMMARY-CONVENTIONAL SYSTEM @ 10MMscfd

Pressure Ratio = 350 PSIA/70 PSIA

	T. Level	Tons	I	Capital Charges and Maintenance	Power Cost	Total Operating and Capital Charges
1	40	36	$ 15,600	$ 3,770/Yr	$ 3,070	$ 6,840
2	40	48	20,000	4,840	4,100	8,940
3	40	61	24,500	5,930	5,200	11,130
4	40	73	29,000	7,020	6,220	13,240
5	40	85	33,000	7,980	7,250	15,230
6	31	91	39,500	9,550	8,450	18,000
7	-8	87	59,000	14,300	11,100	25,400
8	-12	83	73,000	17,700	13,800	31,500
9	-35	79	110,000	226,600	14,900	41,500
10	-57	74	180,000	43,500	17,600	61,100
11	-80	70	390,000	93,500	21,100	114,600
12	-104	65	~1,400,000	339,000	25,500	364,500
13	-130	62	~2,000,000	483,000	30,800	513,800
14	-156	56	-	-	-	-
15	-183	53	-	-	-	-

TABLE IX-B

INVESTMENT SUMMARY-CONVENTIONAL SYSTEM @ 20MMscfd

Pressure Ratio = 350/70

	T. Level	Tons	I	Capital Charges and Maintenance	Power Cost	Total Operating and Capital Charges
1	40	72	$ 28,500	$ 6,900/Yr	$ 6,150/Yr	$ 13,050/Yr
2	40	96	37,000	8,950	8,200	17,150
3	40	122	45,500	11,000	10,400	21,400
4	40	146	53,000	12,800	12,400	25,200
5	40	171	62,000	15,000	14,600	29,600
6	31	182	72,500	17,500	16,900	34,400
7	-8	174	110,000	26,600	22,200	48,800
8	-12	166	131,000	31,500	23,600	55,100
9	-35	158	210,000	50,800	29,800	81,600
10	-57	148	350,000	84,600	33,700	118,300
11	-80	140	720,000	174,000	42,200	216,200
12	-104	130	~ 1,500,000	363,000	51,300	414,300
13	-130	124	-	-	-	-
14	-156	112	-	-	-	-
15	-183	106	-	-	-	-

TABLE IX-C

INVESTMENT SUMMARY-CONVENTIONAL SYSTEM @ 50MMscfd

Pressure Ratio - 350/70

	T. Level	Tons	I	Capital Charges and Maintenance	Power Cost	Total Operating and Capital Charges
1	40	179	$ 63,000	$ 15,250/Yr	$ 15,280	$ 30,530/Yr
2	40	242	84,000	20,300	20,600	40,900
3	40	304	100,000	24,200	25,900	51,100
4	40	366	120,000	29,000	31,200	60,200
5	40	428	135,000	32,400	36,500	68,900
6	31	455	167,000	40,300	42,200	82,500
7	-8	435	245,000	59,300	55,500	114,800
8	-12	415	300,000	72,500	60,200	132,700
9	-35	395	460,000	111,000	74,500	185,500
10	-57	370	670,000	162,000	88,000	250,000
11	-80	350	1,700,000	411,000	105,600	516,600
12	-104	325	~ 3,000,000	725,000	128,000	853,000
13	-130	310	-	-	-	-
14	-156	280	-	-	-	-
15	-183	265	-	-	-	-

TABLE IX-D

INVESTMENT SUMMARY-CONVENTIONAL SYSTEM @ 10MMscfd

Pressure Ratio = 550/70

	Temp. Level	Tons	I	Capital Charges and Maintenance	Power Cost	Total Operating and Capital Charges
1	40	44	$ 18,200	$ 4,400/Yr	$ 3,850/Yr	$ 8,250/Yr
2	40	58	23,500	5,700	4,950	10,650
3	40	70	27,800	6,700	5,960	12,660
4	40	79	31,000	7,500	6,750	14,250
5	40	93	35,500	8,600	7,930	16,530
6	27	116	51,000	12,300	10,900	23,200
7	5	112	69,000	16,700	13,600	30,300
8	-18	108	100,000	24,200	16,600	40,800
9	-43	100	165,000	39,900	20,600	60,500
10	-68	94	330,000	79,800	24,800	104,600
11	-95	88	900,000	218,000	30,600	248,600
12	-121	83	~ 2,500,000	605,000	38,000	643,000
13	-152	76	-	-	-	-
14	-187	74	-	-	-	-

TABLE IX-E

INVESTMENT SUMMARY-CONVENTIONAL SYSTEM @ 20MMscfd

Pressure Ratio = 550/70

	Temp. Level	Tons	I	Capital Charges and Maintenance	Power Cost	Total Operating and Capital Charges
1	40	87	$ 33,500	$ 8,100/Yr	$ 7,600/Yr	$ 15,700/Yr
2	40	115	43,000	10,400	10,000	20,400
3	40	140	51,000	12,300	12,200	24,500
4	40	158	57,000	13,800	13,800	27,600
5	40	185	65,000	15,800	16,100	31,900
6	27	231	93,000	22,500	22,900	45,400
7	5	224	125,000	30,300	27,200	57,500
8	-18	215	188,000	35,100	33,000	78,500
9	-43	200	308,000	74,700	40,600	115,300
10	-68	187	630,000	152,200	49,800	202,200
11	-95	175	~1,700,000	411,000	57,500	468,500
12	-121	165	-	-	-	-
13	-152	152	-	-	-	-
14	-187	143	-	-	-	-

TABLE IX-F

INVESTMENT SUMMARY-CONVENTIONAL SYSTEM @ 50MMscfd

Pressure Ratio = 550/70

	Temp. Level	Tons	I	Capital Charges and Maintenance	Power Cost	Total Capital Charges and Operating Cost
1	40	218	$ 72,000	$ 17,400/Yr	$ 18,200/Yr	$ 35,600/Yr
2	40	288	95,000	23,000	24,200	47,200
3	40	350	113,000	27,400	29,400	56,800
4	40	395	127,000	30,600	34,500	65,100
5	40	462	145,000	35,000	38,800	73,800
6	27	578	220,000	53,200	54,500	107,700
7	5	560	285,000	69,000	68,000	137,000
8	-18	537	420,000	103,000	81,300	184,300
9	-43	500	700,000	170,000	101,000	271,000
10	-68	467	~1,400,000	339,000	122,000	461,000
11	-95	437	~2,500,000	605,000	200,000	805,000
12	-121	412	-	-	-	-
13	-152	380	-	-	-	-
14	-187	358	-	-	-	-

TABLE IX-G

INVESTMENT SUMMARY-CONVENTIONAL SYSTEM @ 10MMscfd

Pressure Ratio - 750/70

	Temp. Level	Tons	I	Capital Charges and Maintenance	Power Cost	Total Operating and Capital Charges
1	40	82	$ 32,000	$ 7,750/Yr	$ 7,700/Yr	$ 14,750/Yr
2	40	93	36,000	8,700	7,930	16,630
3	40	105	39,000	9,430	8,960	18,390
4	40	117	44,000	10,650	9,980	20,630
5	40	138	50,000	12,100	11,750	23,850
6	29	131	57,000	13,800	12,400	26,200
7	3	125	77,000	18,650	15,500	34,150
8	-21	118	115,000	27,900	18,900	46,800
9	-50	112	220,000	53,200	24,900	88,100
10	-78	104	525,000	127,000	30,800	157,800
11	-109	98	~1,800,000	435,000	39,700	474,700
12	-144	91	-	-	-	-

TABLE IX- H

INVESTMENT SUMMARY-CONVENTIONAL SYSTEM @ 20MMscfd

Pressure Ratio = 750/70

	Temp. Level	Tons	I	Capital Charges and Maintenance	Power Cost	Total Operating and Capital Charges
1	40	164	$ 59,500	$ 14,400/Yr	$ 14,000/Yr	$ 28,400/Yr
2	40	186	65,500	15,800	15,850	31,650
3	40	210	72,000	17,400	17,900	35,300
4	40	234	80,000	19,350	19,950	39,300
5	40	272	91,000	22,000	23,200	45,200
6	29	262	105,000	25,400	24,800	50,200
7	3	250	140,000	33,900	31,000	64,900
8	-21	236	215,000	52,000	37,800	89,800
9	-50	224	410,000	99,300	49,700	149,000
10	-78	208	950,000	230,000	61,500	291,500
11	-109	196	~ 2,700,000	650,000	79,500	729,500
12	-144	182	-	-	-	-

TABLE IX-I

INVESTMENT SUMMARY-CONVENTIONAL SYSTEM @ 50MMscfd

Pressure Ratio = 750/70

	Temp. Level	Tons	I	Capital Charges and Maintenance	Power Cost	Total Operating and Capital Charges
1	40	410	$ 130,000	$ 31,400/Yr	$ 35,000/Yr	$ 66,400/Yr
2	40	465	146,000	35,300	39,600	74,900
3	40	525	162,000	39,200	44,700	83,900
4	40	585	177,000	42,800	49,800	92,600
5	40	690	208,000	50,300	58,800	109,100
6	29	655	228,000	55,100	62,000	117,100
7	3	625	315,000	76,200	77,500	153,700
8	-21	590	480,000	116,000	94,500	210,500
9	-50	560	910,000	220,000	124,400	344,400
10	-78	520	~ 2,500,000	605,000	154,000	759,000
11	-109	490	-	-	-	-
12	-144	455	-	-	-	-

TABLE X-A

COMPARISON OF EXPANDER SYSTEM AND CONVENTIONAL SYSTEM

Total Installed Cost (I) and Value/Ton

Pressure Ratio = 350 PSIA/70 PSIA
Flow Rate = 10MMscfd

Case No.	Temp. Level	Tons	Expander System $I	Expander System $ Value / Ton-Yr.	Conventional System $I	Conventional System $ Value / Ton-Yr.
1	40	36	483,000	2680	15,600	190
2	40	48	487,000	2020	20,000	186
3	40	61	1,191,000	4790	24,500	183
4	40	73	1,190,000	4020	29,000	182
5	40	85	1,208,000	3520	33,000	180
6	31	91	1,160,000	3270	39,500	198
7	8	87	1,170,000	3350	59,000	292
8	-12	83	1,177,000	3530	73,000	379
9	-35	79	1,071,000	3410	110,000	525
10	-57	74	1,078,000	3670	180,000	830
11	-80	70	1,051,000	3810	390,000	1630
12	-104	65	1,050,000	4100	1,400,000	5600
13	-130	62	1,051,000	4320	2,400,000	8650
14	-156	56	1,053,000	4830	-	-
15	-183	53	1,054,000	5110	-	-

TABLE X-B

COMPARISON OF EXPANDER SYSTEM AND CONVENTIONAL SYSTEM

Total Installed Cost (I) and Value/Ton

Pressure Ratio = 350 PSIA/70 PSIA
Flow Rate = 20MMscfd

			Expander System		Conventional System	
Case No.	Temp. Level	Tons	$I	$ Value / Ton-Yr.	$I	$ Value / Ton-Yr
1	40	72	683,000	1690	28,500	181
2	40	96	687,000	1290	37,000	179
3	40	122	1,685,000	3280	45,500	175
4	40	146	1,681,000	2750	53,000	172
5	40	171	1,707,000	2400	62,000	173
6	31	182	1,640,000	2180	72,500	189
7	8	174	1,653,000	2310	110,000	281
8	-12	166	1,662,000	2440	131,000	332
9	-35	158	1,517,000	2350	210,000	517
10	-57	148	1,522,000	2540	350,000	800
11	-80	140	1,487,000	2620	720,000	1542
12	-104	130	1,485,000	2840	1,500,000	3190
13	-130	124	1,487,000	3000	-	-
14	-156	112	1,490,000	3360	-	-
15	-183	106	1,491,000	3560	-	-

TABLE X-C

COMPARISON OF EXPANDER SYSTEM AND CONVENTIONAL SYSTEM

Total Installed Cost (I) and Value/Ton

Pressure Ratio = 350 PSIA/70 PSIA
Flow Rate = 50 MMscfd

Case No.	Temp. Level	Tons	Expander System $I	Expander System $ Value/Ton-Yr	Conventional System $I	Conventional System $ Value/Ton-Yr
1	40	179	1,180,000	925	63,000	171
2	40	242	1,190,000	710	84,000	169
3	40	304	2,910,000	2170	100,000	168
4	40	366	2,920,000	1810	120,000	164
5	40	428	2,950,000	1590	135,000	161
6	31	455	2,840,000	1450	167,000	181
7	-8	435	2,860,000	1560	245,000	264
8	-12	415	2,880,000	1620	300,000	319
9	-35	395	2,620,000	1570	460,000	470
10	-57	370	2,640,000	1700	670,000	675
11	-80	350	2,570,000	1760	1,700,000	1475
12	-104	325	2,570,000	1900	3,000,000	2620
13	-130	310	2,570,000	2010	-	-
14	-156	280	2,580,000	2260	-	-
15	-183	265	2,580,000	2390	-	-

TABLE X-D

COMPARISON OF EXPANDER SYTEM AND CONVENTIONAL SYSTEM

Total Installed Cost (I) and Value/Ton

Pressure Ratio = 550 PSIA/70 PSIA
Flow Rate = 10 MMscfd

Case No.	Temp. Level	Tons	Expander System $I	Expander System $ Value / Ton-Yr	Conventional System $ I	Conventional System $ Value / Ton-Yr
1	40	44	707,000	3330	18,200	188
2	40	58	653,000	2290	23,500	184
3	40	70	1,250,000	4220	27,800	181
4	40	79	1,250,000	3740	31,000	180
5	40	93	1,261,000	3230	35,000	178
6	27	116	1,270,000	2610	51,000	200
7	5	112	1,280,000	2720	69,000	271
8	-18	108	1,282,000	2860	100,000	378
9	-43	100	1,284,000	3010	165,000	695
10	-68	94	1,235,000	3190	330,000	1112
11	-95	88	1,240,000	3440	900,000	2830
12	-121	83	1,240,000	3660	2,500,000	7750
13	-152	76	1,206,000	3930	-	-
14	-187	74	1,206,000	4000	-	-

TABLE X-E

COMPARISON OF EXPANDER SYSTEM AND CONVENTIONAL SYSTEM

Total Installed Cost (I) and Value/Ton

Pressure Ratio = 550 PSIA/70 PSIA
Flow Rate = 20MMscfd

Case No.	Temp. Level	Tons	Expander System $I	Expander System $ Value / Ton-Yr	Conventional System $I	Conventional System $ Value / Ton-Yr
1	40	87	1,000,000	2210	33,500	181
2	40	115	925,000	1520	43,000	177
3	40	140	1,765,000	2940	51,000	175
4	40	158	1,765,000	2610	57,000	175
5	40	185	1,784,000	2270	65,000	172
6	27	231	1,798,000	1840	93,000	196
7	5	224	1,809,000	1970	125,000	256
8	-18	215	1,815,000	2080	188,000	365
9	-43	200	1,817,000	2240	308,000	578
10	-68	187	1,748,000	2280	630,000	1080
11	-95	175	1,755,000	2460	1,700,000	2680
12	-121	165	1,755,000	2620	-	-
13	-152	152	1,705,000	2790	-	-
14	-187	143	1,705,000	2980	-	-

TABLE X-F

COMPARISON OF EXPANDER SYSTEM AND CONVENTIONAL SYSTEM

Total Installed Cost (I) and Value/Ton

Pressure Ratio = 550 PSIA/70 PSIA
Flow Rate = 50MMscfd

			Expander System		Conventional System	
Case No.	Temp. Level	Tons	$I	$ Value Ton-Yr	$ I	$ Value Ton-Yr
1	40	218	1,730,000	1360	72,000	163
2	40	288	1,600,000	915	95,000	164
3	40	350	3,050,000	2010	113,000	162
4	40	395	3,050,000	1790	127,000	165
5	40	462	3,090,000	1580	145,000	160
6	27	578	3,110,000	1270	220,000	186
7	5	560	3,120,000	1330	285,000	245
8	-18	537	3,140,000	1400	420,000	343
9	-43	500	3,140,000	1520	700,000	542
10	-68	467	3,020,000	1580	1,400,000	990
11	-95	437	3,030,000	1710	2,500,000	1840
12	-121	412	3,030,000	1840	-	-
13	-152	380	2,950,000	1970	-	-
14	-187	358	2,950,000	2100	-	-

TABLE X-G

COMPARISON OF EXPANDER SYSTEM AND CONVENTIONAL SYSTEM

Total Installed Cost (I) and Value/Ton

Pressure Ratio = 750 PSIA/70 PSIA
Flow Rate = 10MMscfd

Case No.	Temp. Level	Tons	Expander System $I	Expander System $ Value / Ton-Yr	Conventional System $I	Conventional System $ Value / Ton-Yr
1	40	82	543,000	1255	32,000	180
2	40	93	538,000	1120	36,000	179
3	40	105	1,312,000	2950	39,000	175
4	40	117	1,310,000	2640	44,000	177
5	40	138	1,327,000	2280	50,000	174
6	29	131	1,332,000	2440	57,000	200
7	3	125	1,338,000	2730	77,000	273
8	-21	118	1,342,000	2910	115,000	396
9	-50	112	1,343,000	3090	220,000	787
10	-78	104	1,290,000	3410	525,000	1520
11	-109	98	1,290,000	3630	1,800,000	4850
12	-144	91	1,290,000	3930	-	-

R

TABLE X-H

COMPARISON OF EXPANDER SYSTEM AND CONVENTIONAL SYSTEM

Total Installed Cost (I) and Value/Ton

Pressure Ratio = 750 PSIA/70 PSIA
Flow Rate = 20MMscfd

Case No.	Temp. Level	Tons	Expander System $I	Expander System $ Value Ton-Yr	Conventional System $I	Conventional System $ Value Ton-Yr
1	40	164	768,000	785	59,500	173
2	40	186	762,500	698	65,500	170
3	40	210	1,857,000	2080	72,000	168
4	40	234	1,850,000	1860	80,000	168
5	40	272	1,876,000	1630	91,000	166
6	29	262	1,884,000	1720	105,000	195
7	3	250	1,890,000	1810	140,000	259
8	-21	236	1,898,000	1940	215,000	380
9	-50	224	1,900,000	2060	410,000	665
10	-78	208	1,826,000	2150	950,000	1350
11	-109	196	1,828,000	2290	2,700,000	3720
12	-144	182	1,828,000	2490	-	-

TABLE X-I

COMPARISON OF EXPANDER SYSTEM AND CONVENTIONAL SYSTEM

Total Installed Cost (I) and Value/Ton

Pressure Ratio = 750 PSIA/70 PSIA
Flow Rate = 50MMscfd

			Expander System		Conventional System	
Case No.	Temp. Level	Tons	$I	$ Value / Ton-Yr	$I	$ Value / Ton-Yr
1	40	410	1,329,000	438	130,000	162
2	40	465	1,318,000	392	146,000	161
3	40	525	3,213,000	1415	162,000	160
4	40	585	3,200,000	1270	177,000	158
5	40	690	3,245,000	1105	208,000	158
6	29	655	3,259,000	1180	228,000	179
7	3	625	3,270,000	1250	315,000	246
8	-21	590	3,283,000	1340	480,000	357
9	-50	560	3,287,000	1350	910,000	615
10	-78	520	3,159,000	1500	2,500,000	1460
11	-109	490	3,162,000	1600	-	-
12	-144	455	3,162,000	1740	-	-

APPENDIX II

APPENDIX II

SAMPLE CALCULATION

For this sample calculation, one of the cases studied is chosen for the purpose of detailing the methods and decisions involved in the study.

Case Chosen:

Pressure Ratio = 550 PSIA/70 PSIA

Case Number: #7

Expander Inlet Temperature = 20°F

System Inlet Temperature = 80°F @ 600 PSIA

System Outlet Temperature = 70°F @ 50 PSIA

System Energy Calculations Expander:

Inlet T = 20°F

P = 550 PSIA

H = 5485 BTU/lb. mole

Isentropic Expansion

Outlet T = -184°F

P = 70 PSIA

H = 4075 BTU/lb. mole

Δ H Isent = 1410 BTU/lb. mole

For 75% Expander Efficiency

Δ H Actual = 1058 BTU/lb. mole = WORK

THEREFORE

Actual

Expander T = $-143^{o}F$

Outlet P = 70 PSIA

H = 4427

Refrigeration:

Precooling Exchanger

Warm Stream: T in = $80^{o}F$ P = 600 PSIA

T out = $20^{o}F$ P = 550 PSIA

H = 555 BTU/lb. mole

Cold Stream: T in = ? P = 60 PSIA

T out = $70^{o}F$ P = 50 PSIA H = 5120 BTU/lb.mole

H = 555 BTU/lb. mole

THEREFORE:

H in = 5675 BTU/lb. mole

T in = $5^{o}F$

Refrigeration Exchanger:

Cold Stream (Expander Exhaust)

T in = $-143^{o}F$; P = 70 PSIA ; H = 4427 BTU/lb.mole

T out = $5^{o}F$; P = 60 PSIA ; H = 5675 BTU/lb.mole

THEREFORE H Refrig. = 1248 BTU/lb. mole

@ 20MMscfd

FLOWRATE = 20MMscfd $\frac{\text{1 lb. mole}}{\text{386 SCF}}$ $\frac{1}{24}$ $\frac{\text{lb. mol}}{\text{Hr}}$

Refrigeration Capacity = 224 Tons @ $5^{o}F$

Expander Work = 906 Horsepower

The sizing and pricing of the major pieces of process equipment is the next step after the energy calculations for the system are completed. All the equipment is sized and priced for a flow rate of 20MMscfd with the intention of scaling up or down the costs for 50MM and 10MMscfd. The equipment to be sized and priced consists of the MEA unit, dehydrator, preheating or precooling equipment, and the expander generator system.

SIZING & PRICING

Precooling Core

H = 555 BTU/lb. mole

Assume Warm End $\Delta T = 10^oF = 80^oF - 70^oF$

Cold End $\Delta T = 20^oF - 5^oF = 15^oF$

The cooling curve for this exchanger is straight so the log mean temperature difference may be used.

$$\Delta Tlm = (15-10)/\ln (15/10) = 12.34^oF$$

For a Stewart-Warner brazed aluminum core (18 = x 29 = x 125 =) suitable for the pressures considered the UA value is 125,000 BTU/Hr ^oF. Using the equation;

$$Q = N (UA) \Delta Tlm$$

The number of basic cores (N) needed may now be determined.

$$N = (555) (2158)/(125,000) (12.34)$$

$$N = 0.769$$

$$\text{Number of IN}^3 \text{ of core} = (0.769) \left(65,250 \frac{IN^3}{Core}\right) = 50,177 \text{ IN}^3$$

From APCI Cost data it is now possible to determine the cost/IN^3. For 50,177 IN^3 of core the cost/IN^3 is \$.158/IN^3.

The total core cost is then,

$$\text{Total Cost} = (\$.158/IN^3)\ (50,177\ IN^3) = \$8,530.$$

MEA-CO_2 Removal Unit

The pricing of the MEA unit is handled with the use of cost data from previsous projects. For a natural gas stream containing 1/2 to 2 mole % CO_2 the concentration may be reduced to 50 PPM or less.

The unit cost for the system and utility requirements are given below.

Unit Cost = $168,300

Electric Power =	52 KW
Cooling Water =	532 GPM
Fuel Gas (1000 BTU/SCF) =	7,200 SCFH
MEA Make-up =	40#/Day

Cost Factors Used:

Power	- 0.8¢/KWH
Once thru cooling water	- 2¢/M Gal
Natural gas fuel	- 30¢/MM BTU
MEA	- 25¢/#

Annual Utility Costs:

Electric Power	=	$ 3,640/Yr
Cooling Water	=	5,600/Yr
Fuel Gas (1000 BTU/SCF)	=	18,920/Yr
MEA Make-up	=	3,650/Yr
Total Annual Costs		$31,810/Yr

Summary

Unit Cost = $168,300
Utilities = $ 31,810/Yr

Dehydrator:

The cost of this unit has been determined from data available on similarly sized unit.

Unit Cost = $50,000

Utilities = $ 5,000

EXPANDER-GENERATOR SYSTEM

Expander sizing requires that the enthalpy change per expander stage be less that 50 BTU/lb or 678 HP for a flow rate of 20MMscfd of natural gas and the actual volumetric flow rate meet specifications for the various expander turbine sizes available.

For the case being studied;

Horsepower = 906

ACFM Inlet = 362

ACFM Outlet = 1868

The horsepower exceeds 678 HP/stage. The expansion must therefore be split into two equal pressure ratio expansions (550 PSIA/196 PSIA = 196 PSIA/70 PSIA = 2.8).

For this case the operating conditions are:

1st Stage;	Inlet	T = 20°F	P = 550 PSIA
	Intermediate	T = -74°F	P = 196 PSIA
	Horsepower = 510		

2nd Stage; Intermediate T = $-74^{o}F$ P = 196 PSIA

T = $-144^{o}F$ P = 70 PSIA

HP = 396

ACFM Inlet = 362

ACFM Intermediate = 820

ACFM Outlet = 1868

The horsepower/stage requirement is then fulfilled and the volumetric flow rates satisfy the specifications for a 6" expander in series with a 9" expander. The inlet and outlet volumetric requirements for 6" and 9" expanders are given below:

6" Expander	Inlet ACFM	=	180-480
	Outlet ACFM	=	600-1500
9" Expander	Inlet ACFM	=	400-900
	Outlet ACFM	=	1200-3500

The first stage of the expansion is handled by a 6" while the second is handled by a 9" expander. Both expanders are coupled by means of a gear box to an electric generator.

Costs for such an expander system obtained from APCI data, are given in Table VI.

For the case studied:

Gear horsepower of unit chosen = 1500 HP

Unit cost = $135,000

INVESTMENT ANALYSIS:

The object is to obtain a value/ton figure for the case being studied and compare this with the value/ton figure for a conventional vapor compression refrigeration system.

Total Installed Cost = I = 5. (sum cost of all major equipment pieces)

Lange Factor = 5.0

A large lange factor is used because of the cryogenic nature of the plant.

Sum of Equipment Costs:

Precooling Core	-	$ 8,500
MEA Unit	-	168,300
Dehydrator	-	50,000
Expander-Generator	-	135,000
Total Installed Cost		$361,800

Utility costs are given below for a 365 day operating year.

Electric Power	-	$ 3,640/Yr
Cooling Water	-	5,600
Fuel Gas	-	18,920
Chemicals	-	3,650
Maintenance (3% I)	-	54,300
Dehydrator Costs	-	5,000
Total Direct Costs	=	$ 91,100/Yr

Capital costs include the following factors:

Depreciation	- 6.7% I
Tax and Insurance	- 2.0% I
6% Net Profit After Taxes	- 12.5% I
Capital Costs	= 21.2% I
	= $384,000/Yr

Power Credit

90% Efficient generator

0.8¢/KWH

$$(.90)(906HP)(.746 \frac{KW}{HP})(\frac{\$.008}{KW\text{-}HR})(8760 \frac{Hr}{Yr}) = \$42,900/Yr$$

Net Operating Costs

Direct Costs	= $ 91,100/Yr
Capital Charges	= 384,000
Gross Operating Costs	= $475,100
Power Credit	= 42,900
Net Annual Oper. Costs	= $432,200/Yr

Value/Ton of Refrigeration

$$\frac{\text{Value}}{\text{Ton}} = \frac{\text{Net Annual Operating Costs}}{\text{Tons of Refrigeration}} =$$

$$\frac{\$432,200/Yr}{224 \text{ tons @ } 5^{o}F} = \$1970/Yr. \text{ Ton @ } 5^{o}F$$

Scaling Up and Down

The cost factors for the flow rates of 50MMscfd and 10MMscfd are:

$$f_{50} = \left(\frac{50}{20}\right)^{.6} = 1.73$$

$$f_{10} = \left(\frac{10}{20}\right)^{.6} = 0.707$$

These factors are used to scale the total installed cost (I) and related costs (capital charges and maintenance) while factors of 2.5 and 0.5 are used to scale flow related costs (utilities and refrigeration capacity).

Results of the Scale-up

@ 50MMscfd

I = $3,120,000

Tons
Tons = 560 @ 5°F

HP = 2265

Value/Ton = $1330/Yr. Ton

@ 10MMscfd

I = $1,280,000

Tons = 112 @ 5°F

HP = 453

Value/Ton = $2,720/Yr. Ton

Total Installed Cost and Utility Cost

Data for conventional compressed vapor refrigeration systems is used to calculate value/ton data for the comparison with the expander data above.

@ 20MMscfd

I = $125,000

Tons = 224 tons @ $5^{o}F$

Value/Ton = $256/Yr.Ton

@ 50MMscfd

I = $285,000

Tons = 560 tons @ $5^{o}F$

Value/Ton = $245/Yr.Ton

@ 10MMscfd

I = $69,000

Tons = 112 tons @ $5^{o}F$

Value/Ton = $271/Yr.Ton

This concludes the calculations. This sample does not consider the ambient or preheating cases but is representative of most of the cases studied.

SOURCES OF INFORMATION

1 Herrin, J.P.; Hydrocarbon Proc.; June 1966, pp. 45-47.

2 Gardner, J.B. and Smith, K.C.; "Power Consumption in Low Temperature Refrigeration and Separation Processes", Advances in Cryogenic Engineering, Vol. 3, p. 32.

3 Bodle, W.W. and Proctor, R.C.; Cryogenic Engineering News, March 1968, pp. 22-25.

4 Over one hundred natural gas transmission and utility companies have been contacted in an attempt to collect availability data. The replies as of this writing are of sufficient scope to provide a general picture of natural gas availability in the United States and Canada.

5 Design information for the brazed aluminum cores has been provided by the Stewart-Warner Corporation and H.C. Rowles, APCI.

6 Aluminum core costs were made available from a study of that subject by J.W. Taverna, APCI.

7 Cost information for MEA equipment was available through correspondence with Graph Engineering, Inc. Prices for dehydrator equipment has been developed from previous APCI projects.

8 Expander sizing and pricing information has been provided by L.A. Ness, APCI.

9 Beters, D.L.; American Society of Cost Engineering, Paper A-2, 11-63.