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Offshore LNG Production

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Abstract
A natural gas liquefaction plant was designed for offshore production of LNG, using only N_2 and CO_2 as refrigerants in the cooling cycles to avoid potential hazards of mixed hydrocarbon refrigerants. The process was designed to accommodate 13,500 lb-mole/hr (roughly 1MMmtpa) of raw natural gas feed, and fits within all parameters required in the process specifications. Safety concerns, the start-up process, and other potential considerations are also included.

The Net Present Value of the project was found to be $37M at an internal rate of return (IRR) of 18.4%. Further analysis of the assumptions made in these calculations may be required before final project approval is made; however, estimates tend towards conservatism.

Disciplines
Chemical Engineering

Offshore LNG Production

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Dr. Raymond Gorte

April 14, 2009

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April 14, 2009

Dear Mr. Fabiano, Dr. Gorte, and Mr. Brostow,

Enclosed is our proposed process design for the offshore Natural Gas Liquefaction problem statement provided by Mr. Adam Brostow of Air Products and Chemicals. Our process is made up of seven main process blocks – the Liquefaction Process, Nitrogen Refrigeration Cycle, Cooling Water Supply, Power Generation, Steam Cycle, CO_2 Cooling, and Fractionation Train. The process has been designed free of any hydrocarbon refrigerants, reliant on only N_2 and CO_2 for cooling purposes. This process achieves the required 1MMmtpa capacity and purity rates specified by the problem statement.

The following report details the process, equipment needs and estimated costs, approximated power requirements, and a detailed economic analysis. The liquefied natural gas (LNG) Ship has been designed to operate for 40 years, being dry docked for repairs and maintenance after 20 years of operation.

Our proposed process design yields a NPV of $37 million with an IRR of 18.4%. Detailed economic analyses, including sensitivities to key input assumptions, have also been included and discussed. Energy prices and other potential risks to the long-term profitability of the process have also been addressed.

Sincerely,

Salma Al-Aidaroos	Nicholas Bass
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A natural gas liquefaction plant was designed for offshore production of LNG, using only N_2 and CO_2 as refrigerants in the cooling cycles to avoid potential hazards of mixed hydrocarbon refrigerants. The process was designed to accommodate 13,500 lb-mole/hr (roughly 1MMmtpa) of raw natural gas feed, and fits within all parameters required in the process specifications. Safety concerns, the start-up process, and other potential considerations are also included.

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Introduction and Background Information

Introduction

This project has been commissioned to explore the feasibility of an offshore FPSO (Floating, Production, Storage, and Offloading) natural gas liquefaction plant off the coast of Qatar, using only N_2 and CO_2 in the refrigeration cycles. Traditional on-land and initial offshore plants have included mixed refrigerants (hydrocarbons) in their refrigeration cycles; however, system leaks could lead to potentially hazardous accumulations near the ship's surface, considering many of these refrigerants have densities higher than air.

Project analysis will be based primarily on design & thermodynamic considerations, economic feasibility, and safety concerns. Also included are potential changes or additions to the presented base-case scenario, and sensitivities to the economic assumptions used to make final recommendations.

Background

The process designed for this project is an LNG production plant that is installed onboard a ship (FPSO i.e. Floating, Production, Storage and Offloading) that docks in remote areas of the sea (specifically, Qatar, the largest LNG distributor in 2007), extracting and liquefying natural gas. While the liquefaction process consumes a considerable amount of energy, its appeal lies in the reduction of required storage and transportation volumes – LNG takes up approximately 1/600th of the volume of its gas counterpart.

Turning natural gas into the liquid form facilitates its transportation worldwide, increasing its availability in areas where pipelines do not exist. After the liquefaction process, cryogenic sea vessels are used to transport the LNG to areas of the world where it is in high demand. These vessels are highly insulated and double-layered. The LNG is stored below its boiling point around atmospheric pressure. The LNG liquefaction ship is designed so that it is “docked” in the water for a period of 20 years, during which only the transportation vessels return back to shore with the LNG product. The liquefaction ship dry docks for complete maintenance following these 20 years, after which it is sent back for a second 20 year operational period.[1]

The need for such a facility is important to meet the increasing demand in natural gas worldwide. It will take advantage of the remote offshore natural gas reserves that would be impossible to reach without offshore technologies. In general, projects that are similar to this one are usually onshore and use mixed refrigerant cycle technology, normally with the propane pre-cooled cycle, or use cascade refrigeration. However, since the project is offshore, important process considerations differ, as the use of mixed refrigerants becomes more dangerous on FPSOs. Mixed refrigerants are designed to handle larger capacities at low operating temperatures; however, as a result, they need careful technical manufacturing and assembly solutions for proper and smooth operation. Therefore, it is very costly and not as safe as the nitrogen cycle that is suggested in this process.

One of the differentiating attributes of the suggested design is the nitrogen expander cycle. The first and most important advantage to this choice is that nitrogen is an inherently safe, inert refrigeration fluid. Of the utmost concern, explosion hazards are reduced. Additionally, should a large refrigeration cycle failure occur, leaks have very minimal environmental impact.

One recent example of a similar process is the Snohvit LNG Export Terminal in Melkoya Island, Norway. While this facility is not in the middle of the ocean, per se, its underlying concepts remain the same. Three significant gas fields are found near the area, but because Melkoya Island is in a rather remote area of the Barent Sea, a traditional dry-land LNG facility is not possible. Instead, liquefaction barges are used to avoid performing construction and

[1] (The California Energy Commission, 2008)

steelwork on the island itself. The LNG barges were constructed off-site in Spain, and the facility in Melkoya Island was constructed in a modular fashion, saving enormous costs. The project, commissioned in 2002, came on-line in late 2007. While it has encountered some cooling system problems since then, it currently has the capacity to process over 4 million tons of natural gas per year.[2]

[2] (Snohvit LNG Export Terminal, Melkoya Island, Hammerfest,)

Market and Competitive Analyses

As previously mentioned, natural gas is the second-leading energy source behind oil, accounting for 23% of global energy production. As shown in Figure 1, world consumption of natural gas in 2005 was approximately 104 trillion cubic feet, a figure that is expected to rise by approximately fifty percent by 2030.[3] In terms of liquefied natural gas, specifically, the Energy Information Administration reports that 8 trillion cubic feet of LNG was imported globally in 2006, and is expected to rise to 19 trillion cubic feet by 2030.

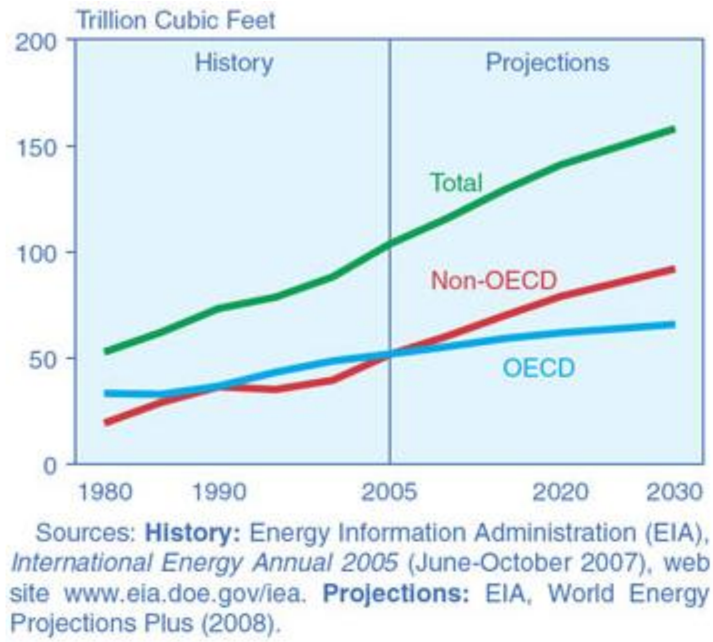

Figure 1: World Natural Gas Consumption (1980-2005 actual and 2006-2030 projections)

However, quantity (consumption) is only half of the revenue equation. Energy prices, which had been hovering at record levels in years past, have recently fallen sharply off their highs. This increased volatility has lead to some uncertainty with regard to natural gas projects, as large energy companies such as BP and Royal Dutch Shell have been taking financial hits due to the decreased values of their energy reserves. Figure 2 shows US Natural Gas Wellhead Prices, an estimate of the value of natural gas “at the well’s mouth.” [For simplicity, all values quoted in this report will be denominated in US dollars, and all financial projections will be

[3] (Energy Information Administration, 2008)

based on price quotes from US-traded energy futures, considered to be one of the most liquid energy pricing sources available.] Here, prices are quoted in $/(thousand cubic feet), a standard natural gas pricing unit. One standard cubic foot of natural gas has a heat of combustion of around 1000 BTU, so the units "thousand cubic feet" and "MMBTU" (million BTU) are often used interchangeably. Natural gas liquids (ethane, propane, butane, and other byproducts from the liquefaction process) will be addressed later, and are typically priced in $/gallon.

As will be discussed later in the economic analysis part of the report, energy prices play a key role in the determination of an LNG facility's profitability. If prices remain towards the lower end of recent values, commissioning new liquefaction capacity may not seem worthwhile. As prices rise, so does the project's attractiveness.

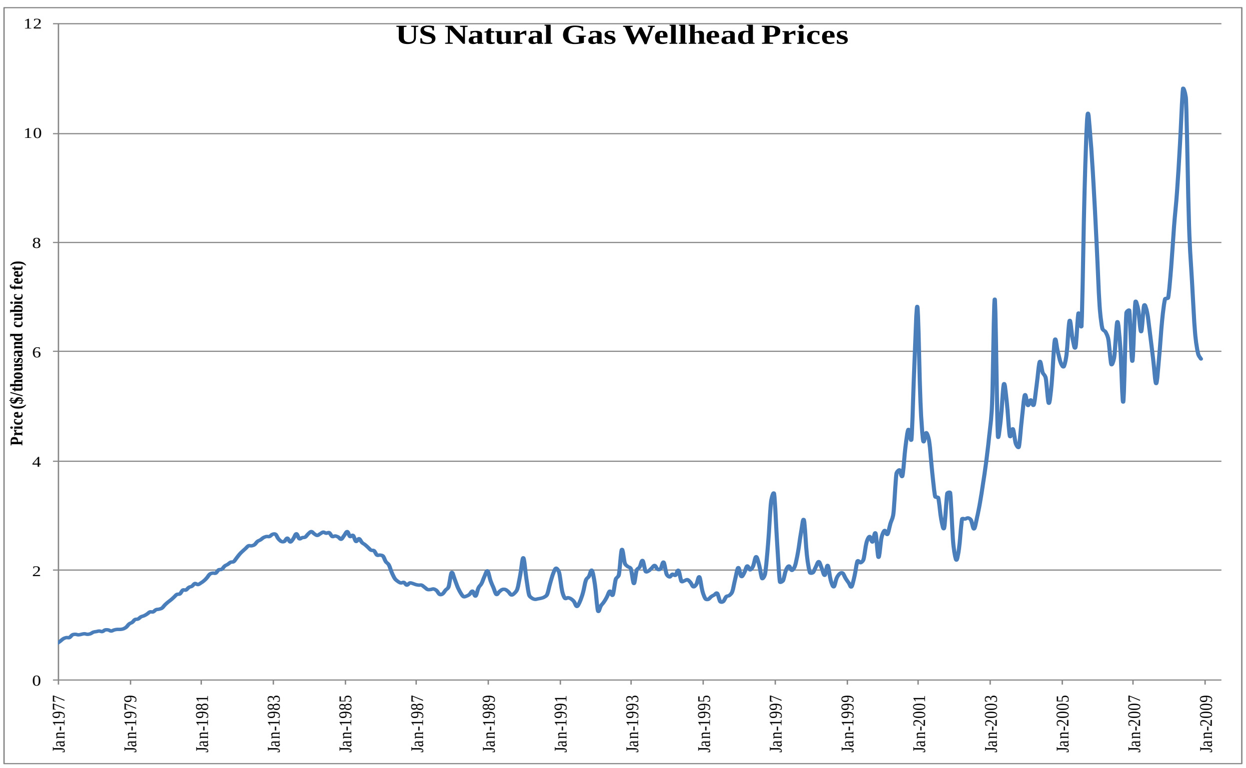

Figure 2: US Natural Gas Wellhead Prices[4]

[4] (Energy Information Administration, 2008)

LNG prices, as expected, are very highly correlated with natural gas prices (see Figure 2), with a slight premium added due to the ease of storage and transport. Further economic analysis and process sensitivities to macroeconomic factors will be discussed later.

Process Flow Diagrams and Material Balances

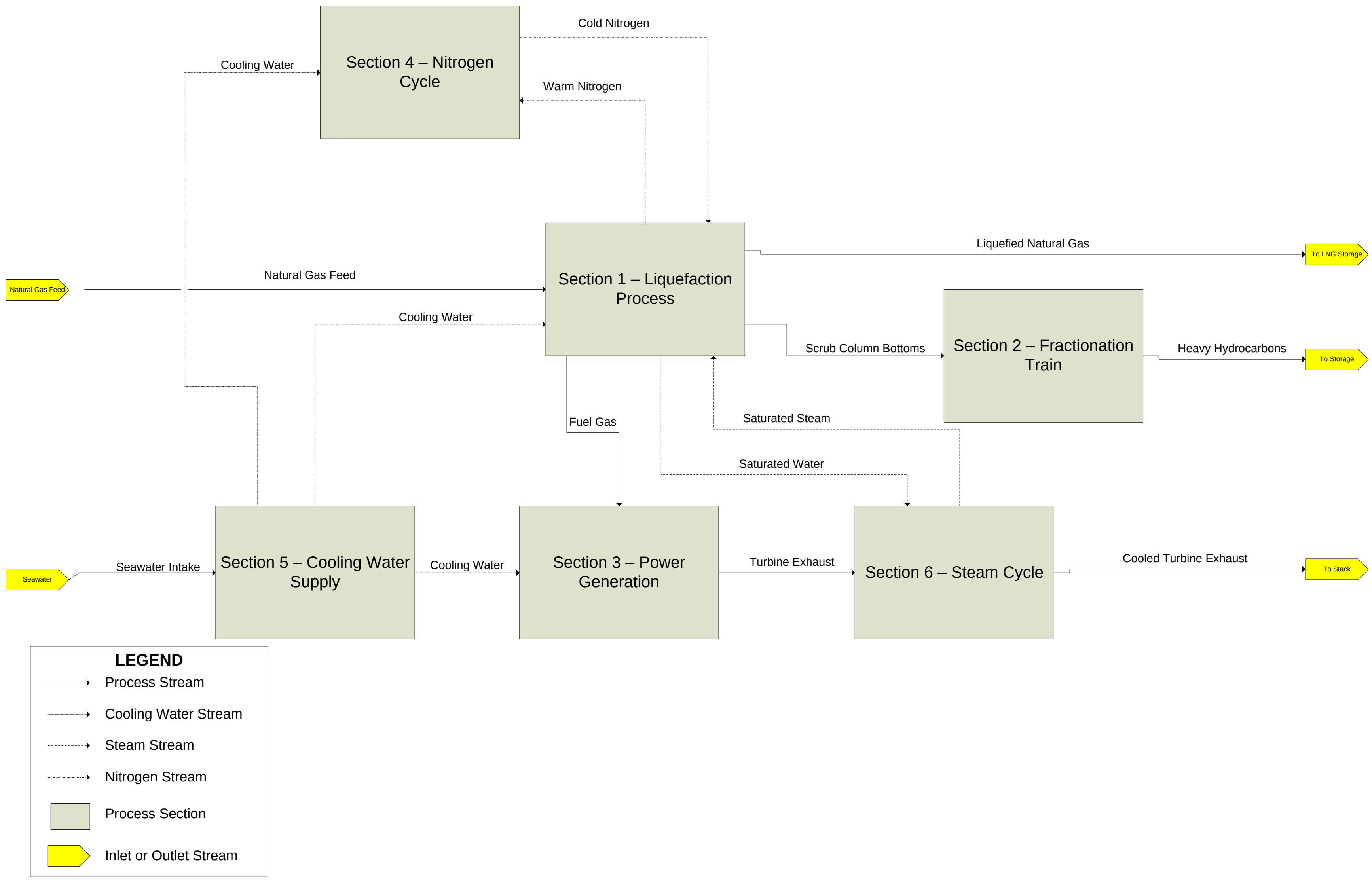

Figure 3: Process Block Flow Diagram

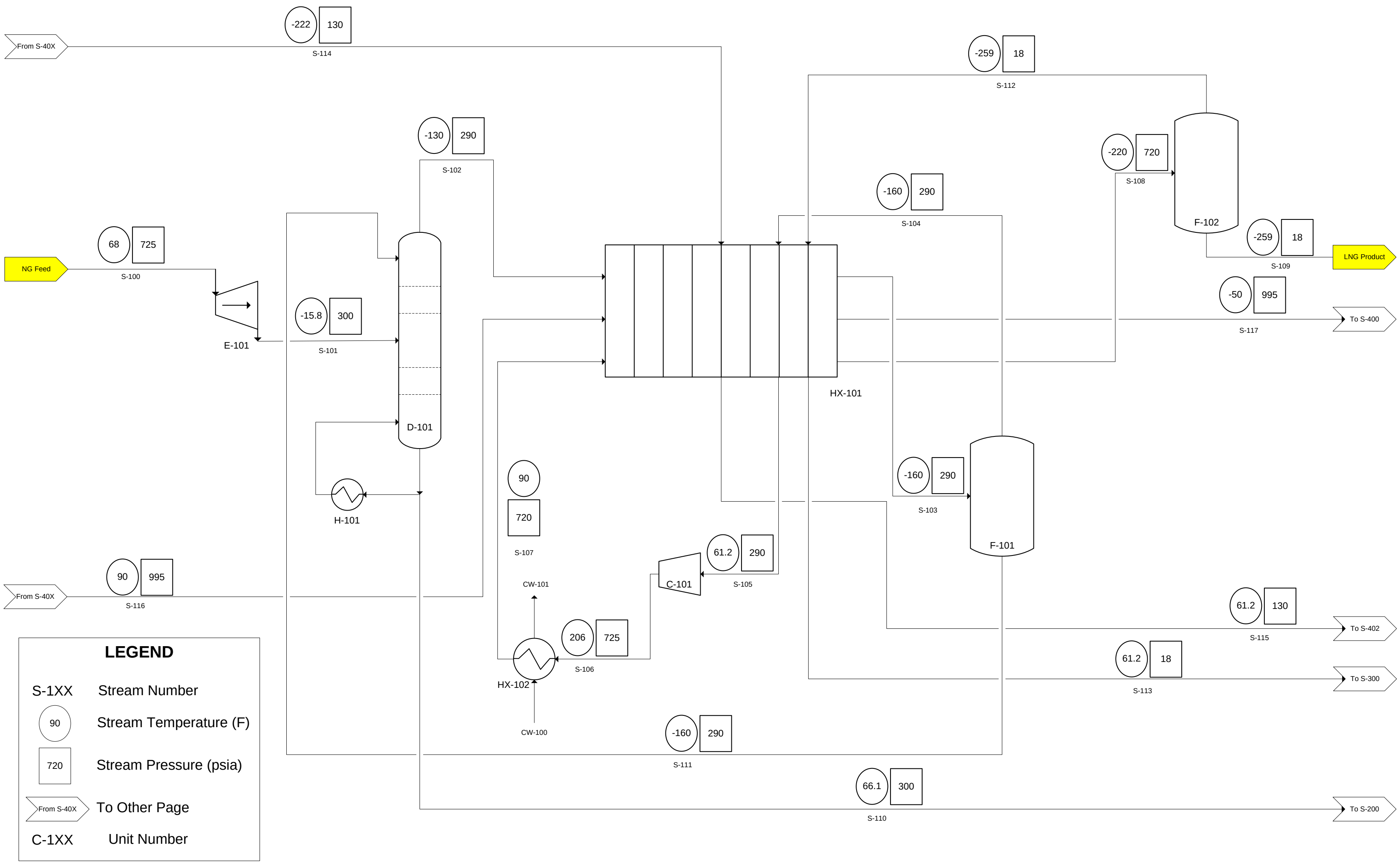

Figure 4: Process Flow Diagram Section 1 - Liquefaction Process

Table 1: Process Flow Diagram Section 1 - Stream Table

	S-100	S-101	S-102	S-103	S-104	S-105	S-106	S-107	S-108
Temperature (F)	68	-15.8	-129.8	-160	-160	61.2	206	90	-220
Pressure (psia)	725	300	290	290	290	290	725	720	720
Mole Flow (lb-mol/hr)	13500	13500	20832.69	20832.69	12319.97	12319.97	12319.97	12319.97	12319.97
Vapor Fraction	1	0.984	1	0.591	1	1	1	1	0
Enthalpy (Btu/hr)	-4.37E+08	-4.47E+08	-6.93E+08	-7.22E+08	-4.07E+08	-3.81E+08	-3.67E+08	-3.81E+08	-4.47E+08
Mole Flow (lb-mol/hr)									
Nitrogen	540	540	622.211	622.211	539.99	539.99	539.99	539.99	539.99
Carbon Dioxide	0	0	0	0	0	0	0	0	0
Methane	11745	11745	19635.72	19635.72	11734.24	11734.24	11734.24	11734.24	11734.24
Ethane	675	675	548.7273	548.7273	45.43008	45.43008	45.43008	45.43008	45.43008
Propane	270	270	24.40419	24.40419	0.307341	0.307341	0.307341	0.307341	0.307341
n-Butane	67.5	67.5	0.539389	0.539389	8.51E-04	8.51E-04	8.51E-04	8.51E-04	8.51E-04
Isobutane	67.5	67.5	1.015938	1.015938	3.11E-03	3.11E-03	3.11E-03	3.11E-03	3.11E-03
Isopentane	40.5	40.5	0.036013	0.036013	9.26E-06	9.26E-06	9.26E-06	9.26E-06	9.26E-06
n-Pentane	67.5	67.5	0.035745	0.035774	8.29E-06	8.29E-06	8.29E-06	8.29E-06	8.29E-06
n-Hexane	27	27	4.62E-04	4.62E-04	2.05E-08	2.05E-08	2.05E-08	2.05E-08	2.05E-08
Oxygen	0	0	0	0	0	0	0	0	0
Water	0	0	0	0	0	0	0	0	0
	S-109	**S-110**	**S-111**	**S-112**	**S-113**	**S-114**	**S-115**	**S-116**	**S-117**
Temperature (F)	-259	66.1	-160	-259	61.2	-222	61.2	90	-50
Pressure (psia)	18	300	290	18	18	130	130	995	995
Mole Flow (lb-mol/hr)	10118.17	1180	8512.687	2201.804	2201.804	69012.13	69012.13	69012.13	69012.13
Vapor Fraction	0	0	0	1	1	1	1	1	1
Enthalpy (Btu/hr)	-3.84E+08	-5.95E+07	-3.15E+08	-6.25E+07	-5.69E+07	-1.52E+08	-9.71E+04	-6.79E+06	-8.52E+07
Mole Flow (lb-mol/hr)									
Nitrogen	105.2375	9.94E-05	82.21112	434.7572	434.7572	69012.13	69012.13	69012.13	69012.13
Carbon Dioxide	0	0	0	0	0	0	0	0	0
Methane	9967.203	10.74426	7901.461	1767.032	1767.032	0	0	0	0
Ethane	45.41546	629.5665	503.2938	0.014622	0.014622	0	0	0	0
Propane	0.30734	269.693	24.09723	1.40E-06	1.40E-06	0	0	0	0
n-Butane	8.51E-04	67.49919	0.538575	3.72E-11	3.72E-11	0	0	0	0
Isobutane	3.11E-03	67.49695	1.012887	7.17E-10	7.17E-10	0	0	0	0
Isopentane	9.26E-06	40.49999	0.036007	7.39E-15	7.39E-15	0	0	0	0
n-Pentane	8.29E-06	67.5	0.03574	5.50E-15	5.50E-15	0	0	0	0
n-Hexane	2.05E-08	27	4.62E-04	4.67E-19	4.67E-19	0	0	0	0
Oxygen	0	0	0	0.00E+00	0.00E+00	0	0	0	0
Water	0	0	0	0.00E+00	0.00E+00	0	0	0	0

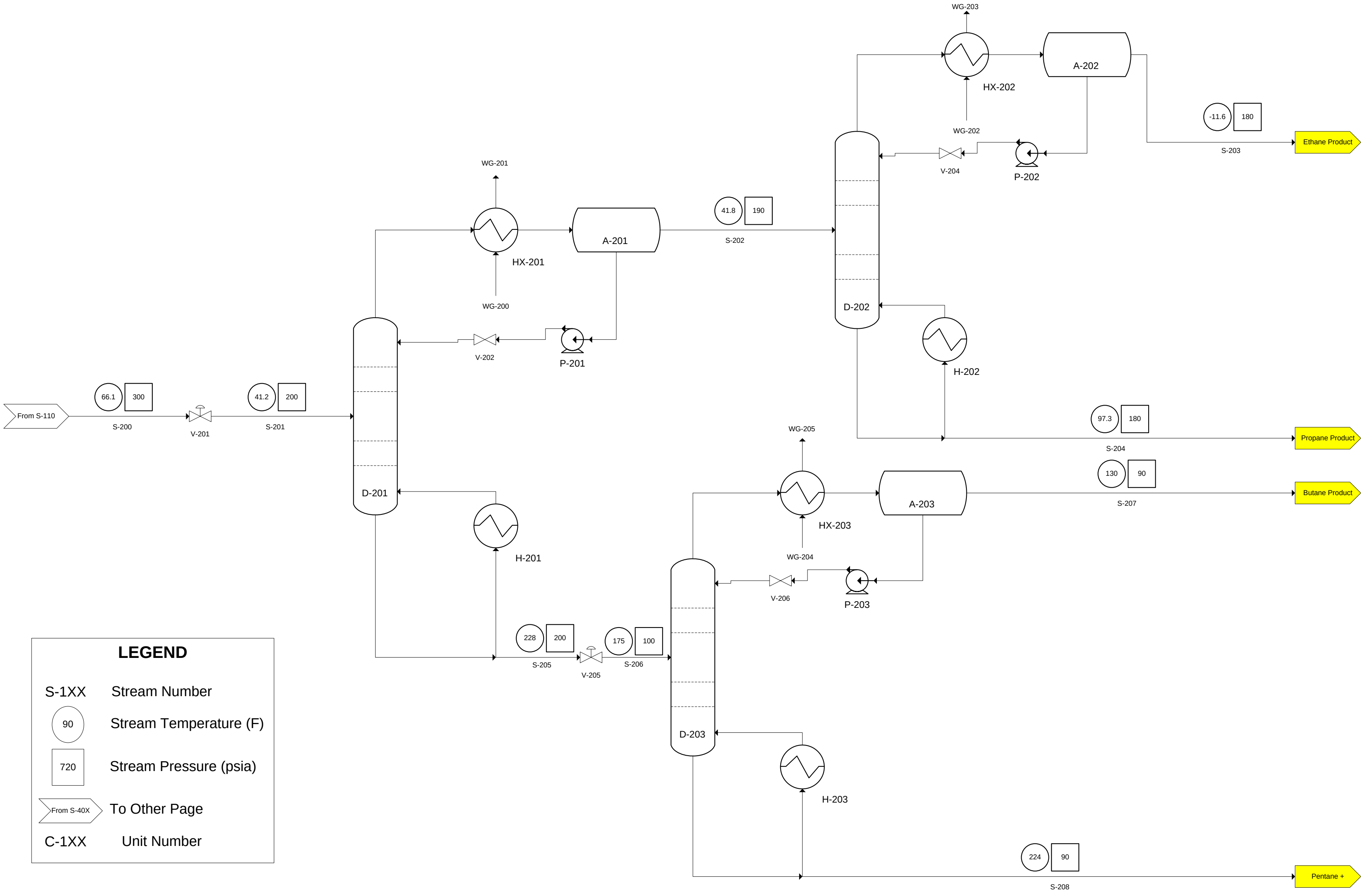

Figure 5: Process Flow Diagram Section 2 - Fractionation Train

Table 2: Process Flow Diagram Section 2 - Stream Table

	S-200	S-201	S-202	S-203	S-204	S-205	S-206	S-207	S-208
Temperature (F)	66.1	41.2	41.8	-11.3	97.3	228	175	130	224
Pressure (psia)	300	200	190	180	180	190	100	90	90
Mole Flow (lb-mol/hr)	1180	1180	910	640	270	270	270	135	135
Vapor Fraction	0	0.145639	1	1	0	0	0.304	1	0
Enthalpy (Btu/hr)	-5.95E+07	-5.95E+07	-3.61E+07	-2.40E+07	-1.38E+07	-1.74E+07	-1.74E+07	-7.40E+06	-9.49E+06
Mole Flow (lb-mol/hr)									
Nitrogen	9.94E-05	9.94E-05	9.94E-05	9.94E-05	3.57E-13	2.08E-14	2.08E-14	0	0
Carbon Dioxide	0	0	0	0	0	0	0	0	0
Methane	10.74426	10.74426	10.74426	10.74424	1.31E-05	4.43E-07	4.43E-07	4.43E-07	9.50E-17
Ethane	629.5665	629.5665	629.5234	622.0639	7.459518	0.0430383	0.0430383	0.0430383	2.33E-08
Propane	269.693	269.693	262.7133	7.191032	255.5223	6.979692	6.979692	6.978909	7.84E-04
n-Butane	67.49919	67.49919	1.224905	4.88E-05	1.224856	66.27428	66.27428	62.02797	4.246315
Isobutane	67.49695	67.49695	5.78232	6.47E-04	5.781673	61.71463	61.71463	60.73809	0.976542
Isopentane	40.49999	40.49999	6.64E-03	2.03E-09	6.64E-03	40.49335	40.49335	3.216224	37.27713
n-Pentane	67.5	67.5	4.99E-03	5.47E-10	4.99E-03	67.49501	67.49501	1.991153	65.50386
n-Hexane	27	27	4.20E-06	2.54E-16	4.20E-06	27	27	4.62E-03	26.99537
Oxygen	0	0	0	0	0	0	0	0	0
Water	0	0	0	0	0	0	0	0	0

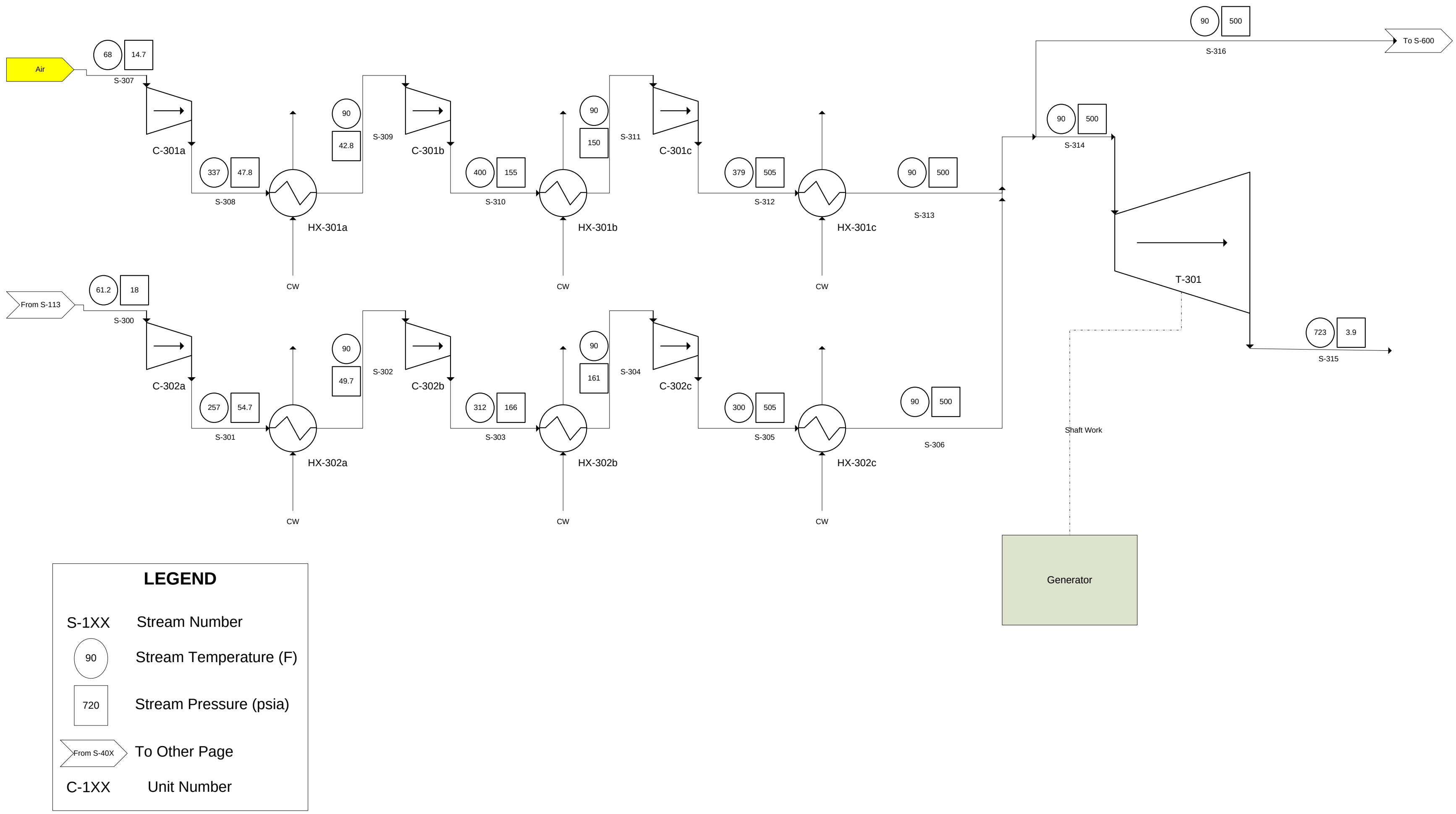

Figure 6: Process Flow Diagram Section 3 - Fuel Gas Turbine/Power Generation

Table 3: Process Flow Diagram Section 3 - Stream Table

	S-300	S-301	S-302	S-303	S-304	S-305	S-306	S-307	S-308	S-309
Temperature (F)	61.2	257	90	312	90	300	90	68	337	90
Pressure (psia)	18	54.7	49.7	166	161	505	500	14.7	47.8	42.8
Mole Flow (lb-mol/hr)	2201.804	2201.804	2201.804	2201.804	2201.804	2201.804	2201.804	30600	30600	30600
Vapor Fraction	1	1	1	1	1	1	1	1	1	1
Enthalpy (Btu/hr)	-5.69E+07							-2.00E+06		
Mole Flow (lb-mol/hr)										
Nitrogen	434.7572	434.7572	434.7572	434.7572	434.7572	434.7572	434.7572	24639	24639	24639
Carbon Dioxide	0	0	0	0	0	0	0	0	0	0
Methane	1767.032	1767.032	1767.032	1767.032	1767.032	1767.032	1767.032	0	0	0
Ethane	0.014622	0.014622	0.014622	0.014622	0.014622	0.014622	0.014622	0	0	0
Propane	1.40E-06	1.40E-06	1.40E-06	1.40E-06	1.40E-06	1.40E-06	1.40E-06	0	0	0
n-Butane	3.72E-11	3.72E-11	3.72E-11	3.72E-11	3.72E-11	3.72E-11	3.72E-11	0	0	0
Isobutane	7.17E-10	7.17E-10	7.17E-10	7.17E-10	7.17E-10	7.17E-10	7.17E-10	0	0	0
Isopentane	7.39E-15	7.39E-15	7.39E-15	7.39E-15	7.39E-15	7.39E-15	7.39E-15	0	0	0
n-Pentane	5.50E-15	5.50E-15	5.50E-15	5.50E-15	5.50E-15	5.50E-15	5.50E-15	0	0	0
n-Hexane	4.67E-19	4.67E-19	4.67E-19	4.67E-19	4.67E-19	4.67E-19	4.67E-19	0	0	0
Oxygen	0	0	0	0	0	0	0	6395.4	6395.4	6395.4
Water	0	0	0	0	0	0	0	0	0	0

	S-310	S-311	S-312	S-313	S-314	S-315	S-316
Temperature (F)	400	90	379	90	90	723	90
Pressure (psia)	155	150	505	500	500	3.9	500
Mole Flow (lb-mol/hr)	30600	30600	30600	30600	30854	30854	1948.1
Vapor Fraction	1	1	1	1	1	1	1
Enthalpy (Btu/hr)							
Mole Flow (lb-mol/hr)							
Nitrogen	24639	24639	24639	24639	23176	23176	1463
Carbon Dioxide	0	0	0	0	0	1662.12	0
Methane	0	0	0	0	1662.09	0	104.94
Ethane	0	0	0	0	1.38E-02	0	8.68E-04
Propane	0	0	0	0	TRACE	0	TRACE
n-Butane	0	0	0	0	TRACE	0	TRACE
Isobutane	0	0	0	0	TRACE	0	TRACE
Isopentane	0	0	0	0	TRACE	0	TRACE
n-Pentane	0	0	0	0	TRACE	0	TRACE
n-Hexane	0	0	0	0	TRACE	0	TRACE
Oxygen	6395.4	6395.4	6395.4	6395.4	6015.59	2691.36	379.81
Water	0	0	0	0	0	3324.23	0

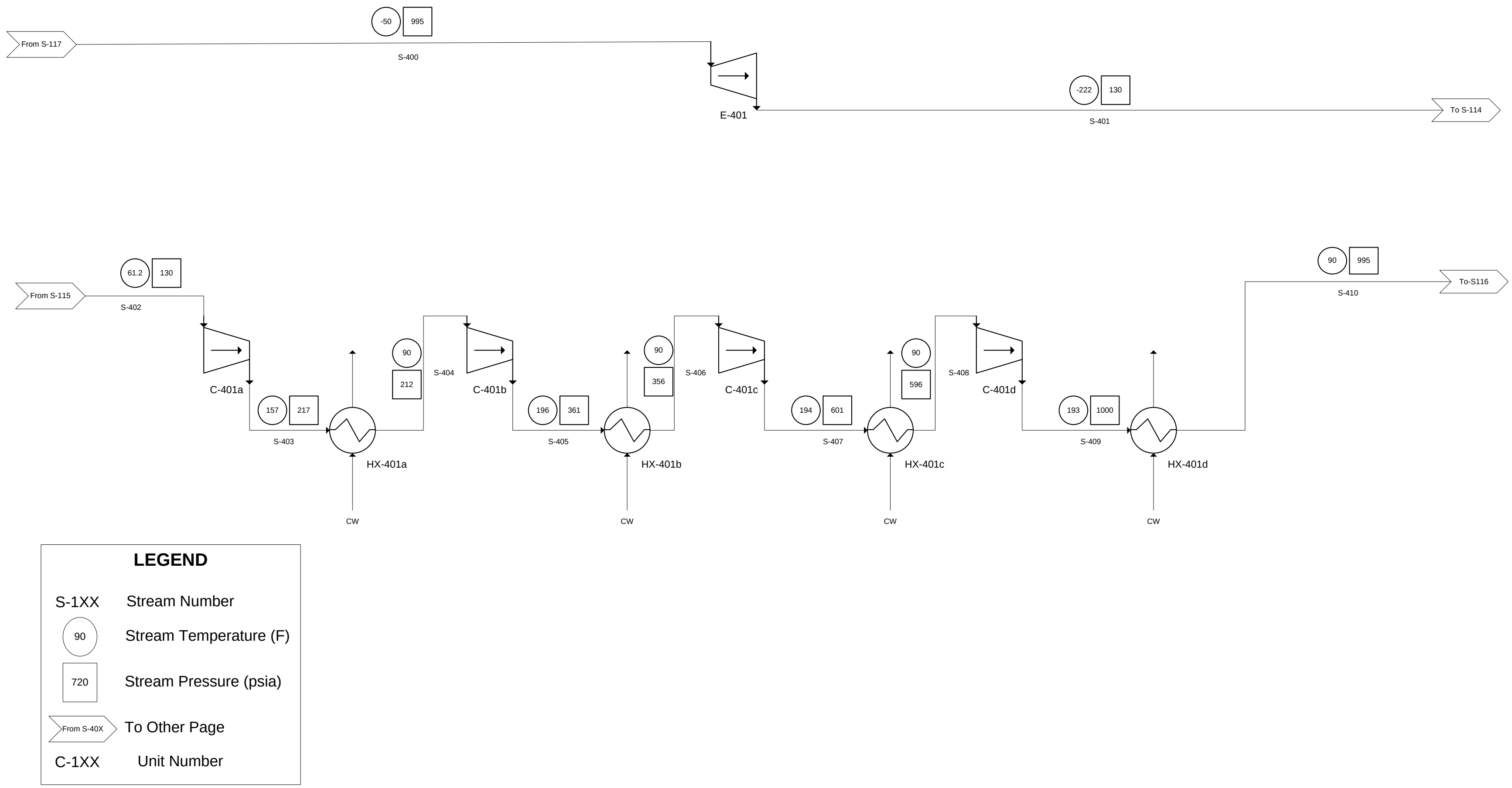

Figure 7: Process Flow Diagram Section 4 - Nitrogen Cycle

Table 4: Process Flow Diagram Section 4 - Stream Table

	S-400	S-401	S-402	S-403	S-404	S-405	S-406	S-407	S-408	S-409	S-410
Temperature (F)	-50	-222	61.2	157	90	196	90	194	90	193	90
Pressure (psia)	995	130	130	217	212	361	356	601	596	1000	995
Mole Flow (lb-mol/hr)	69012.13	69012.13	69012.13	69012.13	69012.13	69012.13	69012.13	69012.13	69012.13	69012.13	69012.13
Vapor Fraction	1	1	1	1	1	1	1	1	1	1	1
Enthalpy (Btu/hr)	-8.52E+07	-1.52E+08	-9.71E+04								-6.79E+06
Mole Flow (lb-mol/hr)											
Nitrogen	69012.13	69012.13	69012.13	69012.13	69012.13	69012.13	69012.13	69012.13	69012.13	69012.13	69012.13
Carbon Dioxide	0	0	0	0	0	0	0	0	0	0	0
Methane	0	0	0	0	0	0	0	0	0	0	0
Ethane	0	0	0	0	0	0	0	0	0	0	0
Propane	0	0	0	0	0	0	0	0	0	0	0
n-Butane	0	0	0	0	0	0	0	0	0	0	0
Isobutane	0	0	0	0	0	0	0	0	0	0	0
Isopentane	0	0	0	0	0	0	0	0	0	0	0
n-Pentane	0	0	0	0	0	0	0	0	0	0	0
n-Hexane	0	0	0	0	0	0	0	0	0	0	0
Oxygen	0	0	0	0	0	0	0	0	0	0	0
Water	0	0	0	0	0	0	0	0	0	0	0

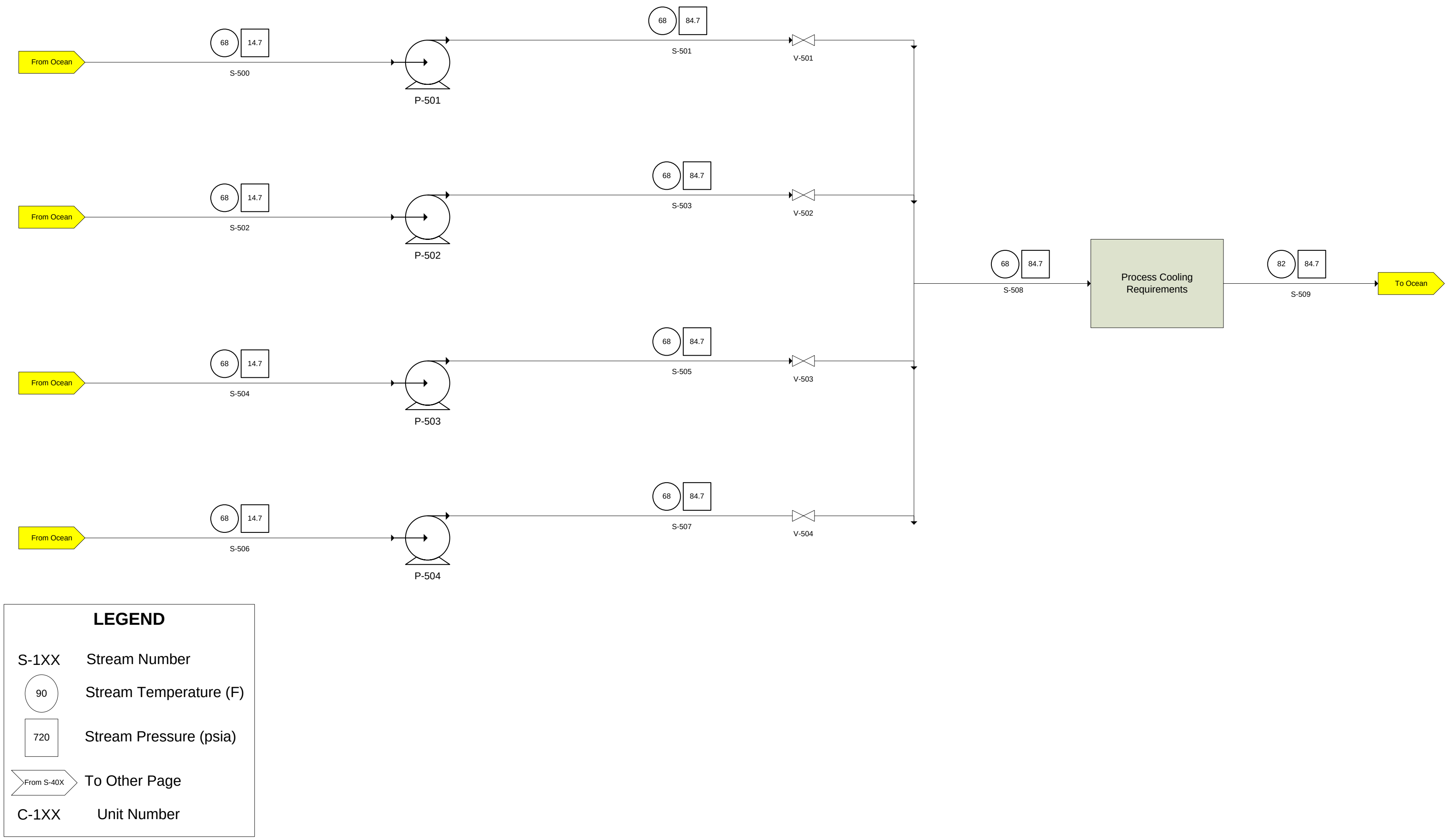

Figure 8: Process Flow Diagram Section 5 - Cooling Water Distribution

Table 5: Process Flow Diagram Section 5 - Stream Table

	S-500	S-501	S-502	S-503	S-504	S-505	S-506	S-507	S-508
Temperature (F)	68	68	68	68	68	68	68	68	68
Pressure (psia)	14.7	84.7	14.7	84.7	14.7	84.7	14.7	84.7	84.7
Mole Flow (lb-mol/hr)	563808	563808	563808	563808	563808	563808	563808	563808	1,691,424
Vapor Fraction	0	0	0	0	0	0	0	0	0
Enthalpy (Btu/hr)									
Mole Flow (lb-mol/hr)									
Nitrogen	0	0	0	0	0	0	0	0	0
Carbon Dioxide	0	0	0	0	0	0	0	0	0
Methane	0	0	0	0	0	0	0	0	0
Ethane	0	0	0	0	0	0	0	0	0
Propane	0	0	0	0	0	0	0	0	0
n-Butane	0	0	0	0	0	0	0	0	0
Isobutane	0	0	0	0	0	0	0	0	0
Isopentane	0	0	0	0	0	0	0	0	0
n-Pentane	0	0	0	0	0	0	0	0	0
n-Hexane	0	0	0	0	0	0	0	0	0
Oxygen	0	0	0	0	0	0	0	0	0
Water	563808	563808	563808	563808	563808	563808	563808	563808	1,691,424

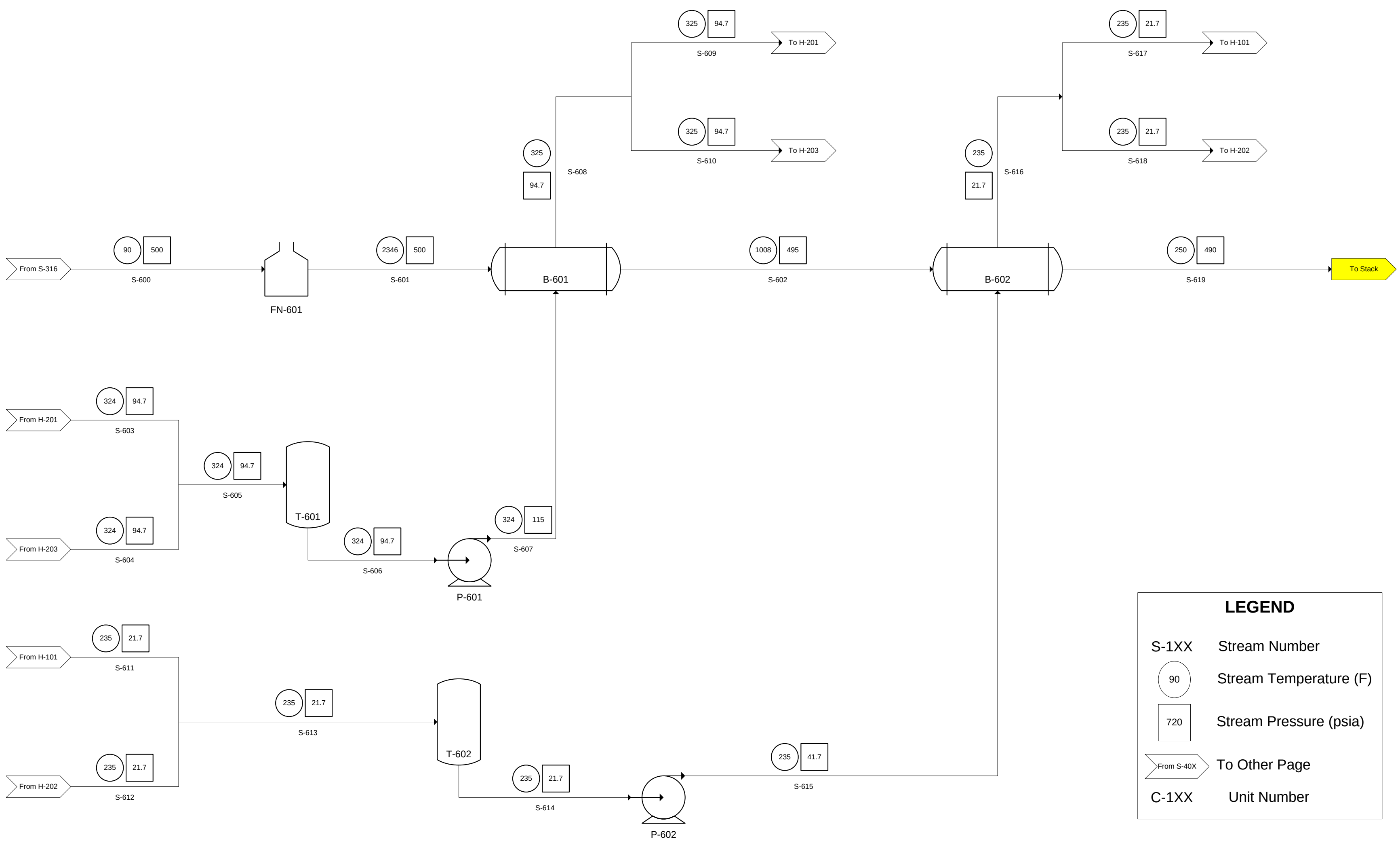

Figure 9: Process Flow Diagram Section 6 - Steam Cycle

Table 6: Process Flow Diagram Section 6 - Stream Tables

	S-600	S-601	S-602	S-603	S-604	S-605	S-606	S-607	S-608	S-609
Temperature (F)	90	2345.6	1008.25	324	324	324	324	324	325	325
Pressure (psia)	500	500	495	94.7	94.7	94.7	94.7	114.7	94.7	94.7
Mole Flow (lb-mol/hr)	1948.1	1948.01	1948.01	976.86	294.72	1271.58	1271.58	1271.58	1271.58	976.86
Vapor Fraction	1	1	1	0	0	0	0	0	1	1
Enthalpy (Btu/hr)		-1749.52	-1.33E+04	-1.16E+08	-3.50E+08	-1.51E+08	-1.51E+08	-1.51E+08	-1.30E+08	-1.30E+08
Mole Flow (lb-mol/hr)										
Nitrogen	1463	1463.27	1463.27	0	0	0	0	0	0	0
Carbon Dioxide	0	104.94	104.94	0	0	0	0	0	0	0
Methane	104.94	0	0	0	0	0	0	0	0	0
Ethane	8.68E-04	0	0	0	0	0	0	0	0	0
Propane	TRACE	0	0	0	0	0	0	0	0	0
n-Butane	TRACE	0	0	0	0	0	0	0	0	0
Isobutane	TRACE	0	0	0	0	0	0	0	0	0
Isopentane	TRACE	0	0	0	0	0	0	0	0	0
n-Pentane	TRACE	0	0	0	0	0	0	0	0	0
n-Hexane	TRACE	0	0	0	0	0	0	0	0	0
Oxygen	379.81	169.92	169.92	0	0	0	0	0	0	0
Water	0	209.88	209.88	976.86	294.72	1271.58	1271.58	1271.58	1271.58	976.86
	S-610	**S-611**	**S-612**	**S-613**	**S-614**	**S-615**	**S-616**	**S-617**	**S-618**	**S-619**
Temperature (F)	325	235	235	235	235	235	235	235	235	250
Pressure (psia)	94.7	21.7	21.7	21.7	21.7	41.7	21.7	21.7	21.7	490
Mole Flow (lb-mol/hr)	294.72	464.64	257.6	722.24	722.24	722.24	722.24	464.64	257.6	1948.01
Vapor Fraction	1	0	0	0	0	0	1	1	1	1
Enthalpy (Btu/hr)	-3.01E+07	-5.60E+07	-3.11E+07	-8.71E+07	-8.71E+07	-8.71E+07	-7.42E+07	-4.77E+07	-2.65E+07	-1.99E+04
Mole Flow (lb-mol/hr)										
Nitrogen	0	0	0	0	0	0	0	0	0	1463.27
Carbon Dioxide	0	0	0	0	0	0	0	0	0	104.94
Methane	0	0	0	0	0	0	0	0	0	0
Ethane	0	0	0	0	0	0	0	0	0	0
Propane	0	0	0	0	0	0	0	0	0	0
n-Butane	0	0	0	0	0	0	0	0	0	0
Isobutane	0	0	0	0	0	0	0	0	0	0
Isopentane	0	0	0	0	0	0	0	0	0	0
n-Pentane	0	0	0	0	0	0	0	0	0	0
n-Hexane	0	0	0	0	0	0	0	0	0	0
Oxygen	0	0	0	0	0	0	0	0	0	169.92
Water	294.72	464.64	257.6	722.24	722.24	722.24	722.24	464.64	257.6	209.88

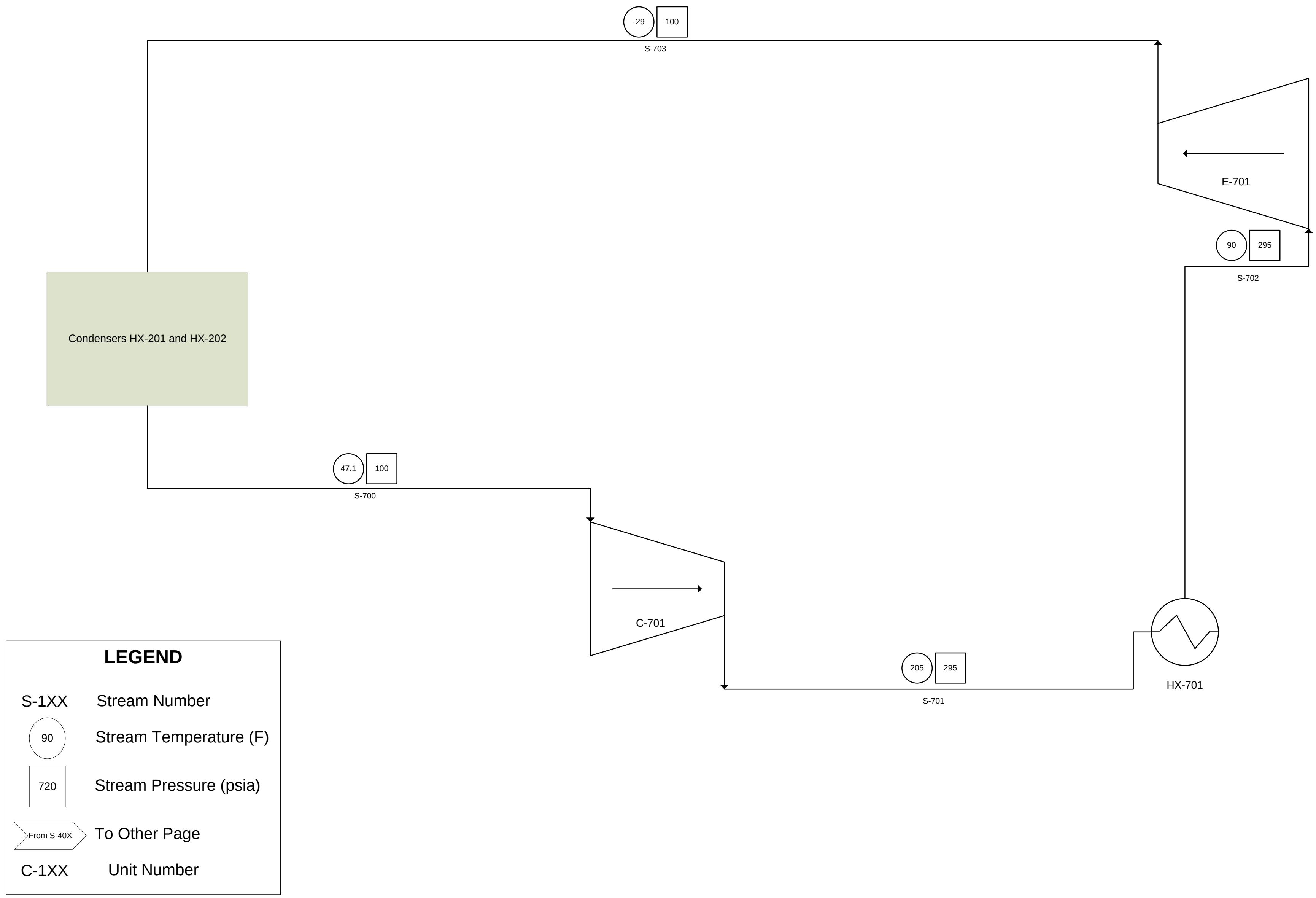

Figure 10: Process Flow Diagram Section 7 - CO_2 Condenser Cooling Loop

Table 7: Process Flow Diagram Section 7 - Stream Table

	S-700	S-701	S-702	S-703
Temperature (F)	47.1	205	90	-29
Pressure (psia)	100	295	295	100
Mole Flow (lb-mol/hr)	26,000	26,000	26,000	26,000
Vapor Fraction	1	1	1	1
Enthalpy (Btu/hr)	-4,409,000	-4,374,489	-4,405,396	-4,426,798
Mole Flow (lb-mol/hr)				
Nitrogen	0	0	0	0
Carbon Dioxide	26,000	26,000	26,000	26,000
Methane	0	0	0	0
Ethane	0	0	0	0
Propane	0	0	0	0
n-Butane	0	0	0	0
Isobutane	0	0	0	0
Isopentane	0	0	0	0
n-Pentane	0	0	0	0
n-Hexane	0	0	0	0
Oxygen	0	0	0	0
Water	0	0	0	0

Process Material Balance

Table 8: Process Material Balance

In Flows	Flow rate (lbmol/hr)	Out Flows	Flow rate (lbmol/hr)
S-100	13,500	S-109	10,118
S-307	30,600	S-203	640
S-500	563,808	S-204	270
S-502	563,808	S-207	135
S-504	563,808	S-208	135
		S-509	1,691,424
		S-619	32,802
Total	1,735,524	**Total**	1,735,524

Table 8 above provides the process material balance over all of the inlet streams and the outlet streams. S-100 is the natural gas feed stream, S-307 is the air feed stream, S-500, S-502, and S-504 are all cooling water inlet streams. Similarly, S-109 is the LNG product stream, S-203 is the ethane product stream, S-204 is the propane product stream, S-207 is the butanes product stream, S-208 is the pentanes plus product stream, S-509 is the cooling water outlet stream, and S-602 is the fuel gas turbine flue gas outlet stream. Since the total number of lbmol/hr in is the same as the total number of lbmol/hr out, the process satisfies the material balance.

Process Descriptions

Overall Process Outline

Figure 3 on page 16 is a block flow diagram of the process that shows the basic relationships between the various process sections. A natural gas feed stream enters Section 1 (Liquefaction Process). In this section, the heavy hydrocarbons are removed from the feed stream, and the resulting methane-rich stream is cooled down to -259F at 18 psia, where it emerges as a liquid and is sent to storage. The separated heavies are sent to Section 2 (Fractionation Train) to be separated further, via a train of separation columns. Additionally, a nitrogen-rich stream of methane is sent from Section 1 to provide fuel gas for Section 3 (Power Generation).

The utilities for the process are produced and delivered by Sections 3-6. Section 3 (Power Generation) provides the process with electrical power via a fuel gas turbine. Section 4 (Nitrogen Cycle) provides the process with the cold nitrogen necessary to liquefy the natural gas through a closed loop expansion-compression cycle. Section 5 (Cooling Water Supply) provides the process with cooling water via four seawater intake pumps. Section 6 (Steam Cycle) provides steam to power the reboilers for the separation columns in both Section 1 and Section 2. All of these sections will be described in detail below.

Section 1 – Liquefaction Process

The base case system considered for this plant is a single loop nitrogen cooling cycle. The first main section in the plant is the liquefaction process. In this section, the heavier hydrocarbons (C2+) are removed from the natural gas feed stream, and the resulting methane-rich stream is liquefied in the main heat exchanger. Figure 4 on page 17 shows the process flow diagram for this section, and Table 1 on page 18 provides detailed descriptions of the streams outlined in Figure 4.

The feed stream (S-100) enters on the left side of Figure 4, with a flow rate of 13,500 lbmol/hr at 725 psia and 68F. This stream is then expanded in the feed expander (E-101) to 300 psia and 45.1F. The feed expander shares a common shaft with the feed compressor, which offsets some of the required work of the compressor. The expanded stream, S-101, now enters the scrub column (D-101). The scrub column separates the feed stream into two streams, a methane-rich overhead product, and a bottoms product that contains almost all of the C3+ components. The overhead from the scrub column (S-102), which has a flow rate of 20,833 lbmol/hr and is made up of 94.2% methane, 2.6% ethane, 2.9% nitrogen, and 0.3% C3+, emerges from the top of the column at -130F, and 290 psia. The bottoms product from the scrub column (S-110), which has a flow rate of 1180 lbmol/hr and consists of 0.8% methane, 53.5% ethane, 22.9% propane, 11.4% butanes, and 11.4% C5+, emerges from the column at 66.1F and 300 psia. Stream S-110 is then sent to the fractionation train (Section 2, Figure 5 on page 19) for further separation.

The overhead from column D-101 (S-102) then proceeds to the main cryogenic heat exchanger (HX-101), where it is cooled to -160F using streams S-104, S-112, and S-114. The cooled stream (S-103) from the main heat exchanger now enters the reflux separator (F-101). The reflux separator is an isothermal flash vessel that provides additional methane/ethane separation, and provides reflux for column D-101 (in the form of the liquid flash product). Stream S-103 is isothermally flashed at -160F and 290 psia. The liquid product of the flash (S-111), which has a flow rate of 8513 lbmol/hr at -160F and 290 psia and consists of 92.8% methane, 5.9% ethane, 0.2% propane, and the balance C4+, is sent back to the scrub column, where it enters above the top stage and is used as reflux for the column. This eliminates the need for a separate condenser.

The vapor product from the reflux separator (S-104), which is comprised of 95.4% methane, 0.3% ethane, 4.2% nitrogen, and the balance C3+, is returned as a cold stream to the main heat exchanger (HX-101), where it is warmed to 61.2F at 290 psia before re-compression. As it enters HX-101 as a cold stream, stream S-104 provides cooling power that supplements that of the nitrogen in stream S-114.

The warmed reflux separator overhead (S-105) then enters the feed gas compressor (C-101). Here, the vapor is recompressed to 725 psia (S-106), in a single compression stage. The compressor has an isentropic efficiency of 86% and a mechanical efficiency of 100%. Stream S-106 is then cooled in the compressor cooling water heat exchanger (HX-102) using seawater to 90F at 720 psia. Recompressing the gas allows for less nitrogen to be used for cooling, which improves the overall power efficiency of the system. After recompression, the stream (S-107) re-enters the main heat exchanger as a hot stream, where it is cooled down to -220F at 720 psia by streams S-104, S-112, and S-114, and emerges as S-108.

This cooled methane-rich stream is then sent to the nitrogen rejection flash vessel (F-102). This vessel isothermally flashes the incoming stream at 18 psia to remove 80.5% of the nitrogen from the methane-rich stream. The liquid product of the flash (S-109) is the final liquefied natural gas product. It emerges at -259F and 18 psia, and is sent to the storage tanks. Stream S-109 has a flow rate of 10,119 lbmol/hr, and consists of 98.5% methane, 1.0% nitrogen, and the balance C2+.

The overhead product from the flash (S-112) is rich in nitrogen and will be used as fuel to power the fuel gas turbine. This stream has a flow rate of 2,201 lbmol/hr and consists of 19.7% nitrogen, 80.2% methane, and 0.1% C2+. Stream S-112 emerges from the flash vessel at -259F and 18 psia. It is then sent to the main heat exchanger as a cold stream, as it must be heated before it is compressed and fed to the turbine reaction chamber. This stream (S-113) emerges from the main exchanger at 61.2F and 18 psia. It then exits Section 1 and proceeds to Section 3 (Power Generation), which is shown in Figure 6 on page 21.

Stream S-114 is the cold nitrogen stream that provides the bulk of the cooling power for the main heat exchanger. This stream enters the heat exchanger at -222F and 130 psia, with a flow rate of 69,012 lbmol/hr of pure nitrogen. It emerges at the other end of the exchanger as stream S-115, with a temperature of 61.2F and a pressure of 130 psia. It then proceeds to Section 4 (Nitrogen Cycle), where it will be re-compressed, cooled, and expanded so that it is again at -222F.

Stream S-116 is the re-compressed nitrogen stream that must be pre-cooled before it can be expanded to provide the necessary cooling for the main heat exchanger. It enters HX-101 at 90F and 995 psia, with a flow rate of 69,012 lbmol/hr. It is cooled by streams S-112 and S-114 to -50F at 995 psia, when it emerges as stream S-117. Stream S-117 then proceeds from Section 1 to Section 4, where it will be expanded to provide cooling power.

Section 2 – Fractionation Train

Section 2 of the process is the fractionation train. The main purpose of this section is to further separate the heavy components (C2+) present in the bottoms product of D-101, stream S-110, so they can be sold for additional revenue. Figure 5, on page 19 above, shows the process flow diagram for this process section. Additionally, Table 2, on page 20 above, provides detailed information on the streams introduced in Figure 5.

The liquid stream (S-110 from Figure 4 or S-200 on Figure 5) coming from the bottom of the scrub column (D-101) contains 0.8% methane, 53.5% ethane, 22.9% propane, 11.4% butanes, and 11.4% C5+, at 66.1F and 300 psia. This stream is then throttled in a valve (V-201) to a temperature of 41.2F and a pressure of 200 psia (S-201). Decreasing the pressure of the stream increases the relative volatilities of the components, and thus makes separation easier.

Stream S-201 then proceeds to the first distillation column in the train (D-201). This column separates the components of S-201 into light (methane, ethane, and propane) components and heavy (butanes and higher) components. These streams are then further separated by other columns further along the train. The vapor product (S-202) of column D-201 has a flow rate of 910 lbmol/hr at 41.8F and 190 psia, and contains 1.2% methane, 69.2% ethane, 28.8% propane, 0.7% butanes, and the balance C5+. This stream (S-202) is then sent to the light distillation column (D-202), where it will be separated further. The liquid bottoms product (S-205) from column D-201 has a flow rate of 270 lbmol/hr at 228F and 190 psia. It contains <0.02% methane and ethane, 2.6% propane, 47.4% butanes, 40.0% pentanes, and the balance hexanes. Stream S-205 is then sent for further separation in the heavy distillation column D-203.

The overhead (S-202) of column D-201 then proceeds to the second distillation column, D-202. This distillation column further separates the light products, and produces a final ethane product as its overhead and a final butane product as its bottoms. Stream S-202 enters D-202 at 41.8F and 190 psia. The overhead product (S-203) has a flow rate of 640 lbmol/hr at -11.3F and 180 psia. It contains 1.6% methane, 97.2% ethane, 1.1% propane, and the balance C4+. This ethane rich stream is then sent to storage and will be sold as an ethane product. The bottoms product (S-204) of D-202 has a flow rate of 270 lbmol/hr at 97.3F and 180 psia. It contains essentially no methane, 2.8% ethane, 94.6% propane, 2.5% butanes, and the balance C5+. This propane rich stream is then sent to storage and will be sold as a propane product.

The bottoms product (S-205) of column D-201 is at 228F and 190 psia. It is throttled through a valve (V-202) to decrease the pressure of the stream from 190 psia to 100 psia in order to improve separation. The lower pressure stream (S-206) is then fed to the third distillation column, D-203. This column separates the butanes from the pentanes and higher. The overhead product from D-203 will be sold as a butanes product, and the bottoms product will be sold as a pentanes plus product. The overhead (S-207) from D-203 has a flow rate of 135 lbmol/hr at 130F and 90 psia. It contains 5.1% propane, 91.1% butanes, 3.8% C5+. This butane rich overhead product is then sent to storage. The bottoms product (S-208) from D-203 has a flow rate of 135 lbmol/hr at 224F and 90 psia. It contains 0.1% propane, 3.9% butanes, and 96.0% C5+. The stream is then sent to storage.

To arrive at this model for the fractionation train, two other models were also considered. The first model was to take the lightest component off in series in each distillation column. This method yielded a process that gave very similar purities, and required almost the same heat duty in the reboilers and condensers as the selected process; however the columns had more stages, and double the reflux ratios of the selected process. This would make those columns more expensive than the ones in the chosen process, hence it was decided that taking the lights off first would be less desirable.

The second model tested was where the heaviest component was taken off each time in a series of distillation columns. Again, this process yielded purities very similar to the chosen process and the columns were of similar size, however the reboiler and condenser duties of those columns were twice as large as the duties in the selected process. This means that the heavy process would be more expensive, like the light process, while yielding the same results as the selected process.

Section 3 – Power Generation

Section 3 of the process is the power generation process. This section takes the nitrogen rich stream (S-113) from the nitrogen rejection vessel (F-102) and burns it with air in order to power a turbine that in turn powers a generator that produces electrical power for the plant. Power generated from the fuel gas turbine is used to power all of the plant compressors and pumps. Figure 6 on page 21 above displays the process flow diagram for the power generation section. Additionally, Table 3 on page 22 gives specific information about the streams introduced in Figure 6.

The nitrogen-rich overhead product (S-113 in Figure 4, S-300 on Figure 6) from the nitrogen rejection flash vessel (F-102) emerges from the main heat exchanger (HX-101) at 61.2F and 18 psia. This fuel gas must be compressed to 500 psia before it can be combusted with excess air in the fuel gas turbine combustion chamber. The fuel gas, which has a flow rate of 2202 lbmol/hr and a composition of 19.7% nitrogen, 80.2% methane, and 0.1% C2+, is compressed to 500 psia in three stages (C-302a-c), with inter-cooling (HX-302a-c) in between stages to 90F with seawater. Each of the compressor stages has an isentropic efficiency of 78% and a mechanical efficiency of 100%. The compressed fuel gas (S-306) emerges after the last stage of the compressor at 500 psia and 90F.

Air is also needed to burn the fuel gas in the turbine combustion chamber. Air enters the system at 14.7 psia and 68F with a flow rate of 30600 lbmol/hr (S-307), and is compressed to 500 psia in a three stage compressor (C-301a-c), also with inter-cooling (HX-301a-c) between stages to 90F with seawater. Each stage of the compressor has an isentropic efficiency of 78% and a mechanical efficiency of 100%. The compressed air emerges from the final stage of the air compressor at 500 psia and 90F (S-313). 78% excess air is provided for combustion in order to keep the temperature in the combustion chamber below the required 2350F.

The compressed air stream (S-313) and compressed fuel gas stream (S-306) are combined. This stream is then split into two streams. Stream S-314 contains 94% of the original combined stream, and is fed into a combustion chamber. The methane is completely combusted, as are the residual heavier hydrocarbons. The product gas, consisting of nitrogen, oxygen, carbon dioxide, and water emerges from the combustion chamber at 2345F, and is fed into the turbine. The turbine converts the thermal energy of the product gas stream into shaft work, which then powers a generator that produces electrical power with 98% efficiency. This power is then distributed throughout the plant to power the process compressors and pumps. The flue gas (S-314) emerges from the turbine at 736F and 3.9 psia. This stream is vented to the atmosphere.

The remaining 6% of the combined air and fuel gas stream is sent as stream S-316 to the furnace in the steam generation cycle (Section 6 – Steam Generation).

Section 4 – Nitrogen Cycle

Section 4 of the process is the nitrogen cycle. In this process, nitrogen provides the cooling necessary to liquefy the methane. The nitrogen cycle is a closed-loop compression expansion cycle, in which the nitrogen is compressed 995 psia, cooled in the main exchanger to -50F, and expanded to 130 psia, where it emerges at -222F. Figure 7 above on page 23 provides the process flow diagram for the nitrogen cycle, and Table 4 on page 24 provides specific stream information for all of the streams in Figure 7.

Nitrogen is the main cooling fluid for the process, and it is circulated in a closed-loop system, so all nitrogen streams have the same flow rate of 69,012 lbmol/hr. Stream S-115 (Figure 4) or S-400 (Figure 7) emerges from the main heat exchanger at 61.2F and 130 psia. This stream must be recompressed to 995 psia and pre-cooled in the main heat exchanger before it can be expanded. The compression takes place in a four stage compressor (C-401a-d), with inter-cooling with seawater (HX-401a-d) to 90F between each stage. Each stage has an isentropic efficiency of 86% and a mechanical efficiency of 100%. The nitrogen emerges from the last stage of the compressor as stream S-410, at 90F and 995 psia. It then exits Section 4 and goes to Section 1 as stream S-114, to be cooled prior to expansion in the main heat exchanger to -50F.

After S-114 has been cooled in the main heat exchanger, it returns to Section 4 as S-402 (S-115 in Figure 4). Stream S-402 has been pre-cooled to -50F at 995 psia. It then enters the nitrogen expander (E-401), where it is isentropically expanded to 130 psia. After this expansion, the stream (S-401) is at -222F and 130 psia. The expander has an isentropic efficiency of 88% and a mechanical efficiency of 100%. The expander shares a common shaft with one of the stages of the nitrogen compressor (C-401a), so the shaft work produced is used to power that compressor stage. The cold nitrogen stream (S-401) is then sent to Section 1 to provide the main cooling power for HX-101. When it emerges from the heat exchanger as S-115, the nitrogen cycle begins anew.

Section 5 – Cooling Water Supply

Section 5 of the process is the cooling water supply. The process requires a significant amount of cooling water due to the heavy cooling load placed on the compressor inter-coolers. Since the process is offshore, cooling water cannot be purchased directly. Instead, seawater will be used for process cooling. The seawater is pumped directly from the ocean through a bank of pumps to a main manifold, from where it will be distributed to the required pieces of process equipment. Figure 8, on page 25, provides the process flow diagram for the cooling water supply. Table 5, on page 26 above, lists detailed information about the process streams associated with Figure 8.

The seawater is assumed to be available at 14.7 psia and 68F. The seawater is supplied to the process by three main seawater pumps (P-501, P-502, and P-503). These pumps pump the seawater through their associated streams (S-501, S-503, and S-505) at 68F and 84.7 psia to the main distribution manifold. An additional pump, P-504, is available should one of the main three pumps fail. Each pump is prepared to intake 563,808 lbmol/hr of water, and the total process requires 1,691,424 lbmol/hr of cooling water.

Four pumps are available in Section 5 to minimize the impact of a mechanical failure on the cooling water system. All four pumps have an identical capacity, and all of their outlet streams have an associated valve (V-501, V-502, V-503, V-504) that can be used to stop the flow to the main manifold. Should a single pump fail with this arrangement, the valve associated with the failed pump can be closed so that the pump can be repaired, and the spare pump can be brought online and have its valve activated. This will ensure that the process can continue to operate, because the spare pump will have the same capacity as the failed pump. Since multiple simultaneous failures are unlikely, this will ensure that the process is always supplied with cooling water. The electric power required to power the pumps is provided by the generator associated with the fuel gas turbine.

After distribution in the main cooling water manifold, the cooling water emerges from the various process heat exchangers at 82F and 84.7 psia. The cooling water is collected into a single main stream and is returned to the ocean at this temperature and pressure as S-609.

Section 6 – Steam Cycle

Section 6 of the process is the steam cycle. As with the cooling water, since the process is located offshore, steam cannot be purchased to power the distillation column reboilers, and must be generated onsite. A closed loop steam cycle is used for this process, with stream S-316 being burned in a furnace to provide the heat necessary to vaporize the steam. Figure 9, on page 27 above, displays the process flow diagram for the steam cycle. Table 6, on page 28, gives detailed information about all of the streams associated with Figure 9.

Two pressures of steam are used in the process. The first is MP steam, which is used to power the reboilers H-201 and H-203 (Section 2, Figure 5). This steam is available at 94.7 psia at its saturation temperature of 324F. The steam emerges from the reboilers in S-603 and S-604 as a saturated liquid. S-603 has a flow rate of 976.9 lbmol/hr, and S-604 has a flow rate of 294.7 lbmol/hr. Both streams are saturated water at 94.7 psia and 324F. The streams are combined as S-605 (also at 94.7 psia and 324F) and are sent to the MP steam condensate tank (T-601).

Stream S-606 is removed from the condensate tank at a flow rate of 1271.6 lbmol/hr at 94.7 psia and 324F, and is fed to the MP steam pump, P-601. This pump increases the pressure of the liquid stream to 114.7 psia, in order to counteract the pressure drop in the boiler. This stream (S-607) is then fed to the MP steam boiler (B-601), where the exhaust gas from the furnace, S-601, at 2346F and 500 psia is used to vaporize the saturated water. Stream S-601 has a flow rate of 1948.1 lbmol/hr, and is composed of nitrogen, oxygen, carbon dioxide, and water. This stream emerges at the other side of the exchanger as stream S-602, at 1008F and 495 psia. The saturated water is completely vaporized to saturated steam, and it emerges as S-608 at 325F and 94.7 psia. This stream is then split into S-609, which feeds H-201, and S-610, which feeds H-203. These streams have flow rates of 976.9 lbmol/hr and 294.72 lbmol/hr, respectively.

A similar process is used to supply the LP steam. This steam is used to power the reboilers H-101 (Section 1, Figure 4) and H-202 (Section 2, Figure 5), and is available at 235F and 21.7 psia. The steam emerges from reboilers H-101 and H-202 as saturated liquids as streams S-611 and S-612, respectively. Stream S-611 has a flow rate of 464.6 lbmol/hr and stream S-612 has a flow rate of 257.6 lbmol/hr. Both are saturated liquids at 235F and 21.7 psia.

These streams are then combined into S-613, which then feeds to the LP steam condensate tank (T-602). Stream S-614 is taken from T-602, with a flow rate of 722.2 lbmol/hr at 235F and 21.7 psia, and is fed to the LP steam pump, P-602. This pump increases the pressure of the liquid stream to 41.7 psia to counteract the 20 psia pressure drop in the LP steam boiler (B-602). The resulting stream (S-615), at 41.7 psia and 235F enters B-602, where it is vaporized by stream S-602. Stream S-602 emerges from the hot side of the boiler as stream S-619, and has been reduced in temperature to 250F. Stream S-619 is then vented to the atmosphere.

The LP steam emerges from the boiler as saturated steam in stream S-616, with a flow rate of 722.2 lbmol/hr at 235F and 21.7 psia. Stream S-616 is then split into streams S-617 and S-618, which power the reboilers H-101 and H-202 respectively. Stream S-617 has a flow rate of 464.6 lbmol/hr, and stream S-618 has a flow rate of 257.6 lbmol/hr.

The heat used to vaporize the steam is generated by a furnace that combusts stream S-600. This stream is combined air and fuel gas, and emerges from the furnace at a temperature of 2345F and a pressure of 500 psia. The fuel gas is completely combusted into carbon dioxide, and this hot gas stream is then used to vaporize both the medium pressure and low pressure steam streams in B-601 and B-602. After the heat has been used, the stream, now S-619, is vented to the atmosphere at 250F.

Section 7 – Carbon Dioxide Cycle

Section 7 of the process is the carbon dioxide cycle. In this process, the carbon dioxide is used to cool the condensers for distillation columns D-201 and D-202. Carbon dioxide was chosen over a chilled water/glycol mixture because the amounts of water and glycol required made the use of that process prohibitively expensive. In this cycle, carbon dioxide is cooled in a compression-expansion process similar to that of the nitrogen cycle. Figure 10 on page 29 shows the carbon dioxide cycle and Table 7 on page 30 gives more detail about the streams introduced there.

Cold carbon dioxide is used to cool the condensers HX-201 and HX-202 for the distillation columns D-201 and D-202. Stream S-703 enters the first condenser at -29F and 100 psia, with a flow rate of 26,000 lbmol/hr. The carbon dioxide exits from HX-202 at 47.1F and 100 psia as S-700.

Stream S-700 then proceeds to the carbon dioxide compressor, C-701, which compresses S-700 from 47.1F and 100 psia to 205F and 295 psia. The compressor has an isentropic efficiency of 85%. After compression, the new stream, S-701 is cooled using cooling water in HX-702 back down to 90F, as in the nitrogen cycle. After this cooling, the stream, S-702, is sent to the carbon dioxide expander to be expanded and cooled. The stream is expanded from 90F and 295 psia to -29F and 100 psia, and the cycle is completed. The expander has an efficiency of 88%, and shares a common shaft with the carbon dioxide compressor (C-701), which helps to offset the electric power requirement for the compressor.

Energy Balance and Utility Requirements

Process Energy Balance

The main challenge of this process is to provide enough cooling power through the nitrogen cycle to liquefy the natural gas and cool the other intermediate streams. Fortunately, due to the recompression of the reflux separator overhead (S-104) and the need to warm the fuel gas stream before compression (S-112), cooling power can also be provided by other streams. These three streams combine to provide all of the necessary cooling power for the process. All streams discussed in this section can be found in Figure 4, on page 17.

There are three streams that need to be cooled during the process. The first is the overhead from the scrub column (D-101), S-102, which is being cooled before it is sent to the reflux separator (F-101). The second is the recompressed nitrogen stream that is being cooled before it is expanded (S-116), and the last stream is the recompressed overhead from the reflux separator (S-107) that will emerge as the final LNG product.

There are a number of things that must be considered when determining the amount of nitrogen required in the system. The first is that efficiency in the cooling process is at its highest when there is a small temperature difference between the contacting streams. In the process that follows, the heating and cooling curves for the exchanger will be constructed. All of the critical streams in the process will be analyzed. These curves are the source of much valuable information. The curves show the amount of cooling power required to bring each stream down to the required temperature, and also where the cooling power is most needed in the system.

Figure 11 above shows the individual cooling curves for streams S-107, S-116, and S-102. These curves were generated in ASPEN by calculating the temperature associated with a particular cooling duty for the given stream in the main heat exchanger. The curves are created

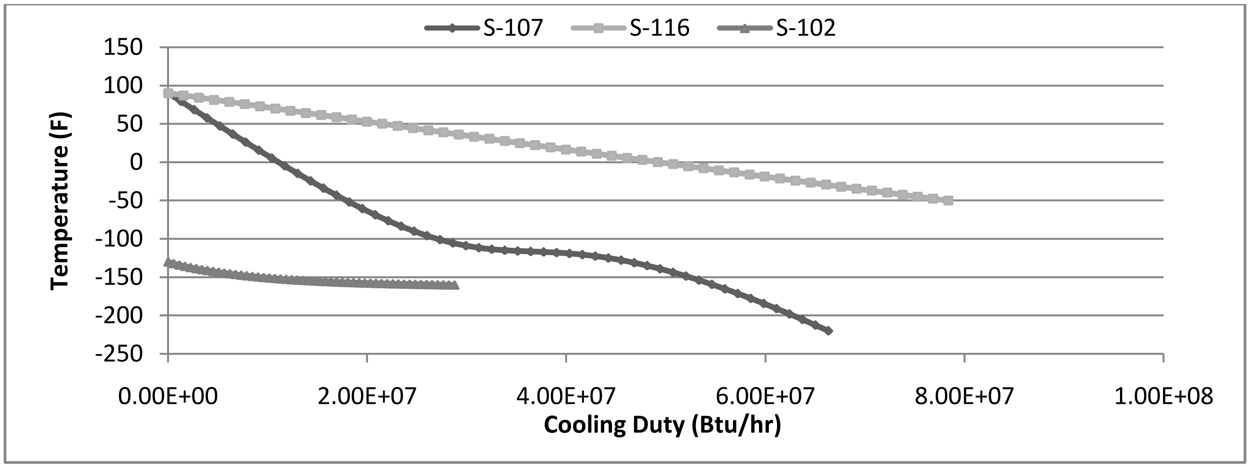

Figure 11: Individual Stream Cooling Curves

by plotting the temperature (F) on the y-axis and the cooling duty (Btu/hr) on the x-axis.

The shape of the individual cooling curves gives an indication of what the cooling duty is doing to the stream. For example, S-116 is the high pressure nitrogen stream that is being pre-cooled prior to expansion. The stream enters the exchanger at 90F, and exits at -50F, having been cooled with just under 80 MMBtu/hr. Since the stream remains at constant pressure, is pure nitrogen, and is a vapor the entire time that it is being cooled, the cooling curve generates a line with constant slope.

However, for a mixed composition fluid, such as the reflux separator overhead (S-107), a different picture emerges. S-107 is a vapor when it enters the main heat exchanger at 90F. As the stream is cooled, there is a significant flattening of the curve at about -120F. This section, where the curve becomes almost horizontal, is where a phase change of the stream takes place. Since the stream is not pure, the line is not perfectly horizontal, because different parts of the stream liquefy at different times. Once the entire stream has been liquefied, a straight section of the curve with constant slope is seen as the now liquid stream is cooled further.

In the same way that process cooling curves were generated for streams S-107, S-116, and S-102, process heating curves can be generated for the streams that are being heated in the main exchanger. Figure 12, below, shows the individual cooling curves for streams S-112, S-114, and S-104.

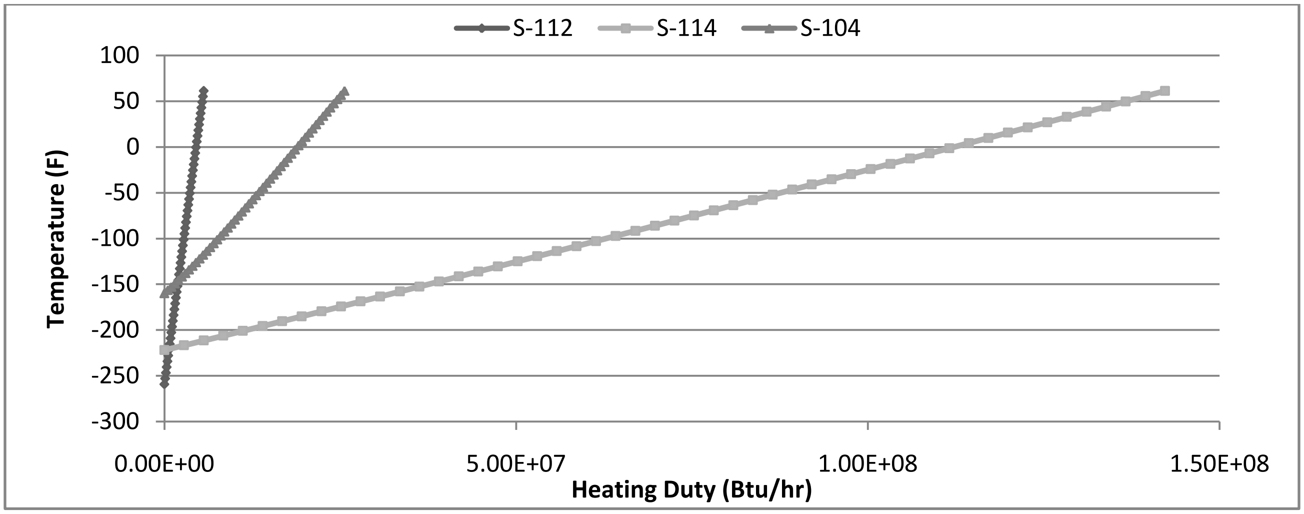

Figure 12: Individual Process Heating Curves

In Figure 12, we can see the process heating curves for streams S-112, S-114, and S-104. Stream S-112 corresponds to the overhead from the nitrogen rejection flash vessel (F-102). Stream S-114 corresponds to the cold nitrogen stream that is providing the bulk of the cooling for the process, and S-104 corresponds to the uncompressed overhead of the reflux separator (F-101) that is being heated before compression. All streams and process units associated with Figure 12 can be seen in Figure 4 on page 19.

The heating curves for these three streams were constructed in a similar manner to the cooling curves in Figure 11 above. Unlike the cooling curves above, all of the heating curves are straight lines with constant slope. This is because no phase changes occur for any of the streams that are being heated. S-114 has a significantly higher heating duty than the rest of the curves; this is because it has the greatest flow rate of the three, and encompasses a large temperature range.

The heating curve allows us to see which streams are available to cool the hot streams in the main exchanger. For example, S-104 enters the main exchanger at -160F. This means that it can be used to cool all streams that enter the exchanger at a higher temperature than it, but it can only cool them down to its inlet temperature of -160F. Therefore, while it can provide some assistance to S-112 and S-114 in cooling stream S-108, it no longer provides any cooling once the temperature of S-108 drops below -160F.

By adding the individual heating and cooling curves together, we can generate composite heating and cooling curves. Below, Figure 13 displays the composite cooling curve.

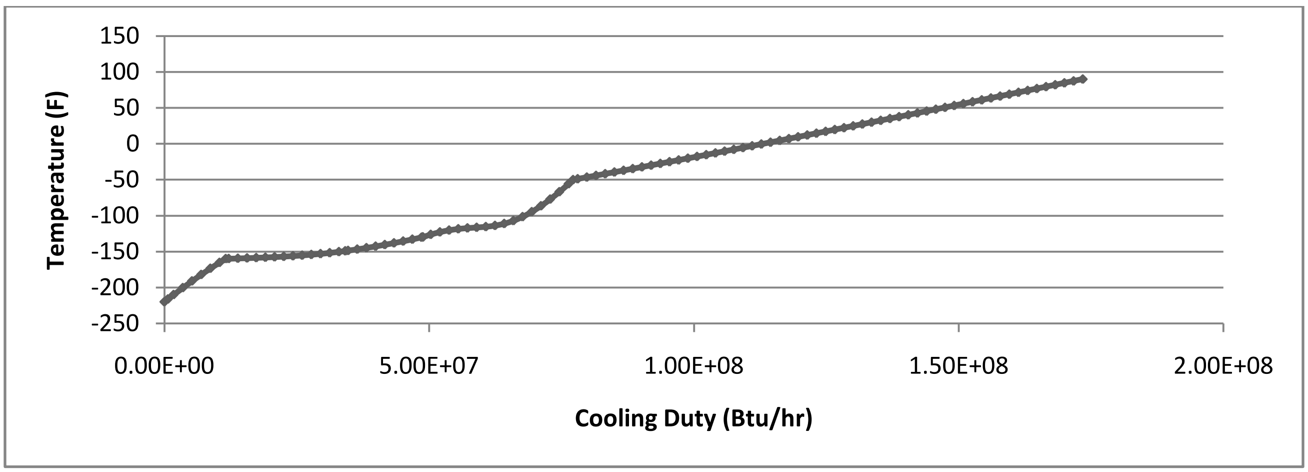

Figure 13: Composite Process Cooling Curve

This composite curve presents a summary view of what is happening in the cold streams of the main heat exchanger. The composite curve could be understood as the mean temperature of the cold streams within the main heat exchanger. Although the streams enter at different temperatures, since the exchanger is a plate-fin exchanger, the cold streams contact one another and exchange heat until they are the same temperature. It is assumed that this happens relatively quickly compared to the overall length of the exchanger, so the composite curve is a good representation of what is actually happening inside the exchanger.

In this case, it is clear that the compressed nitrogen stream (S-116) dominates the composite curve at high temperatures. However, once this stream exits the exchanger at -50F, S-107 is the only contributor until S-102 enters at -130F. From there until -160F, both curves contribute to the composite curve, leading to the nearly horizontal section of the curve seen at about 25 MMBtu/hr. This portion is where both of the streams are being liquefied. Once S-102 is removed at -160F, the remainder of the horizontal portion is from S-107, and the liquefaction finishes. Once liquefaction is complete, a straight line of constant slope results until S-107 is removed from the exchanger at -220F.

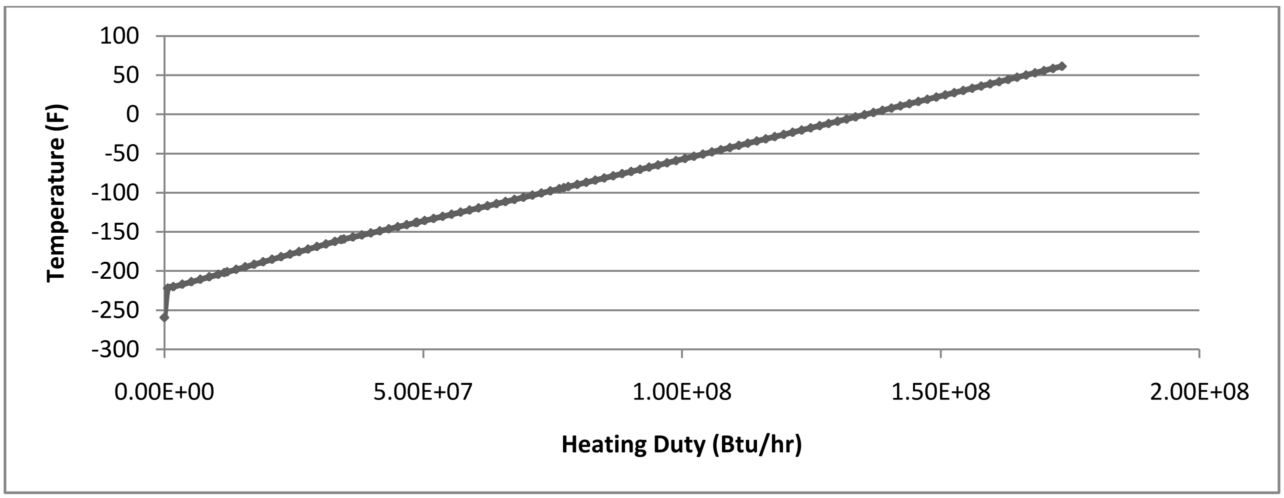

Figure 14: Composite Process Heating Curve

Figure 14 above shows the composite process heating curve. Clearly, the composite curve is dominated by S-114, which enters the exchanger at -222F. This is to be expected, however, since S-114 has the greatest flow rate, and can therefore accept the greatest amount of heat among the three cooling streams. The small, steep section at the beginning of the composite curve corresponds to the region where S-112 is the only stream contributing to the composite curve. The initial slope of the heating curve in that region is the same as the slope of S-112 on the individual heating curve shown in Figure 12 above. The addition of S-104 at -160F is barely discernable in the composite curve, which demonstrates that compared to the main nitrogen stream (S-112), it provides a relatively small amount of cooling power.

Once both the composite heating and the composite cooling curves have been generated, a combined composite curve can be generated to observe the efficiency of the heat exchanger. Figure 15 below displays the combined composite curves for the main heat exchanger:

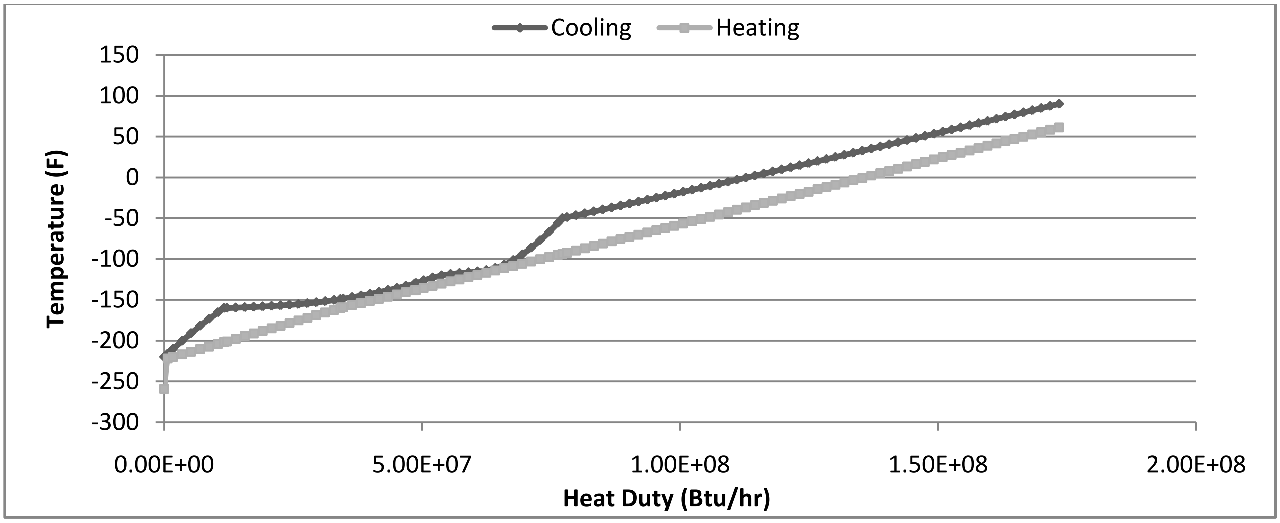

Figure 15: Combined Composite Process Heating and Cooling Curves

Both the composite heating and the composite cooling curve have been plotted on the same axes in Figure 15. A plot such as this allows for easy comparison of the curves. The area between the curves is indicative of the efficiency of the heat exchanger; the more area between the composite curves, the less efficient the process is.

For each exchanger, there is a characteristic internal temperature approach, this is the closest that the composite cooling and composite heating streams can approach in the exchanger. In the exchanger used in this process, the minimum internal temperature approach is 3F. As is apparent from Figure 15, the minimum temperature approach occurs at a heat duty of approximately 65 MMBtu/hr, where the composite cooling curve has a temperature of -114F and the composite heating curve has a temperature of -111F. This is the pinch point in this process, and since the heating curve has a constant slope, a closer temperature approach at the warm end of the exchanger cannot be achieved.

In order to determine the amount of nitrogen required for liquefaction, the composite heating and cooling curves are generated as in Figure 15. The flow rate of S-114 is then manipulated so that the minimum internal temperature difference is achieved. The analysis for this particular process found that 5.11 lbmol of nitrogen is required per lbmol feed of natural gas. Since the initial natural gas feed (S-100) has a flow rate of 13,500 lbmol/hr, this amounts to 69012 lbmol/hr of nitrogen.

Utilities Requirements

Cooling Water

Cooling water is one of the main utilities required for the process. Cooling water is required to reduce the temperatures of the intermediate streams in the multi-stage compressors that are used in the process. Specifically, cooling water is required for the air compressor (HX-301a-c), the feed gas compressor (HX-302a-c), the nitrogen compressor (HX-401a-d), and the feed compressor. The cooling water is used to cool the internal streams in these compressors to 90F in between compression stages. Seawater is used to cool all of these required loads.

Current environmental regulations dictate that in order to minimize thermal pollution, seawater must be discharged no more than 14F above the temperature that it was removed from the ocean. Additionally, no water discharged should exceed 95F in temperature.[5] The heat capacity of seawater was determined to be 0.953 Btu/lb-F, and the seawater is assumed to be available at 68F. Incorporating the maximum possible temperature difference (14F) gives an outlet temperature of 82F, which meets the appropriate environmental regulations. Table 9 below

[5] (Bin Mahfouz, El-Halwagi, & Abdel-Wahab, 2006)

summarizes the cooling duties required by various pieces of process equipment, and provides the total amount of cooling water required for the overall process. Sample cooling water requirement calculations can be found in Appendix IX. Individual cooling duties were calculated for each heat exchanger in a compressor unit, and the final values were added together to produce a net load for the compressor unit.

Table 9: Process Cooling Water Requirements

Process Unit	Required Cooling Duty	Cooling Water Requirement	
	Btu/year	MMlb/year	MMgal/year
HX-301a-c	1.57485×10^{12}	118,036.68	13,760.40
HX-302a-c	1.18982×10^{11}	8,917.84	1,039.62
HX-401a-d	1.63028×10^{12}	7,598.84	885.85
HX-102	1.01384×10^{11}	2,867.65	334.30
HX-203	3.82603×10^{10}	122,191.95	14,244.81
TOTAL	3.46376×10^{12}	259,612.96	30,264.98

Table 9 shows that 30.26 billion gallons of cooling water are required per year to cool the process. The largest loads come from the nitrogen and the air compressors, which is to be expected. The cooling water will be supplied to the process units via a bank of pumps, as outlined in the process descriptions above on page 25. Multiple pumps will be used to ensure that the system will remain operational should one of the pumps fail. A process flow diagram of the cooling water distribution system can be seen in Figure 8 on page 25.

Since the process is located offshore, it is impossible to purchase or deliver cooling water to the vessel while the process is in operation. Because of this, seawater is to be used whenever possible for process cooling. Cooling with seawater introduces a number of new challenges. From an equipment standpoint, seawater cooling requires that any equipment exposed to the seawater must be made of corrosion-resistant materials, and must be cleaned regularly. Additionally, an effective and environmentally friendly biocide must be injected into the seawater at intake in order to minimize the effects of fouling due to microscopic sea life, and strict environmental regulations concerning the discharge temperature of the seawater must be followed.

A closed loop freshwater cooling system was discussed as a possible alternative way of delivering the required cooling power. In this process, the seawater would be used to cool the fresh water after it had been heated by the process equipment. However, such a system would come with increased costs; instead of simply purchasing corrosion resistant heat exchangers for the intercoolers, regular heat exchangers could be purchased, but additional corrosion resistant heat exchangers would need to be purchased to cool the fresh water which would not otherwise be required. Fresh water storage and production would also need to be taken into consideration. Additionally, the minimum temperature of the fresh water would be higher than the assumed inlet temperature of the seawater (the effective fresh water temperature would be 73F, assuming a 5F temperature difference in the seawater/fresh water exchangers), thereby increasing the amount of fresh water required for cooling. Due to these concerns, the current process uses an open-loop seawater cooling system.

Electric Power Generation

Since the ship will be in operation offshore, it will be impossible to purchase the electrical power necessary to power the process compressors and pumps. The process contains four main compressors; one for air, one for nitrogen, one for fuel gas, and one for the feed gas. Some of the power requirements of these compressors are offset by the use of a coupled expander-compressor, in which a compressor and an expander share a common shaft. The process contains two such units, the feed expander (E-101) is coupled with the feed compressor (C-101), and the nitrogen expander (E-401) is coupled with the first stage of the nitrogen compressor (C-401a). Although this generated power is helpful in offsetting the power requirements of the compressors, it is not enough to meet the total process requirements.

In order to meet the process power requirements, the process includes a fuel gas turbine and generator system that uses the nitrogen-rich overhead stream (S-113) from the nitrogen rejection vessel (F-102) as fuel. The fuel gas turbine and generator system provides enough electrical power to power all of the compressors and pumps in the process, and any excess power can be used to provide electricity for living quarters, storage units, and other assorted onboard power requirements. The generator has an efficiency of 98%.

Table 10: Process Power Requirements

Process Unit	Power Consumed	Power Generated
	Hp	Hp
C-101	5,697	---
C-301a-c	73,235	---
C-302a-c	4,699	---
C-401a-d	76,347	---
C-701	13,595	---
P-501 – P-505	3,830	---
P-601 – P-602	3	---
E-101	---	3,415
E-401	---	26,243
E-701	---	8,411
Generator	---	163,647
TOTALS	177,406	201,716

Table 10 provides a summary of the process power requirements. The power requirements for all of the process units listed above were taken directly from the ASPEN simulation results, which are catalogued in Appendix X. The power generated value for the generator was calculated by multiplying the generator efficiency (98%) by the amount of work generated by the fuel gas turbine, as determined in ASPEN.

Table 10 clearly shows that more than enough power is produced by the fuel gas turbine and generator to meet the power requirements of the process. It is also important to note that the power generated by E-101 and E-401 is applied directly to the power load of C-101 and C-401a respectively. This gives an effective power requirement for C-101 of 2,282 Hp, and an effective power requirement for C-401a-d of 50,104 Hp. With these new numbers, the total power required from the power generation system is 147,748 Hp, which is significantly less than the 163,647 Hp generated by the power generation system.

The feed gas for the fuel gas turbine (S-306) contains 1,767 lbmol/hr of methane. The natural gas feed to the process contains 11,745 lbmol/hr of methane. This means that 15% of the methane taken into the system is used to power the system, giving 85% recovery of the inlet methane in the liquefaction process.

Steam

Steam is required in the process to power the reboilers (H-201, H-202, and H-203) for the distillation columns in the fractionation train (Figure 5, page 19) and the reboiler for the scrub column (H-101). Two pressure of steam are needed, MP steam at 94.7 psia that powers H-201 and H-203, and LP steam at 21.7 psia that powers H-101 and H-202. Both pressures of steam are provided at their saturation temperature, 325F for MP steam, and 235F for HP steam.

Table 11: Process Steam Requirements

Process Unit	Heat Duty (Btu/hr)	MP Steam Req. (lbmol/hr)	LP Steam Req. (lbmol/hr)
H-101	8,306,066	---	464.64
H-201	17,483,308	976.86	---
H-202	4,716,273	---	257.6
H-203	4,980,650	294.72	---
TOTAL	35,486,297	1271.58	722.24

Table 11 provides a summary of the steam loads required to power the four reboilers in the process. The heat duty for each process unit was determined from the ASPEN simulations, which can be found in Appendix X. The amount of steam required was determined by setting a design specification in ASPEN so that the steam flow rate was varied until the liquid fraction of the combined streams after the reboilers (S-603 and S-604 for MP, S-611 and S-612 for LP) was found to be 1.

The steam is in a closed loop cycle, and it is regenerated by using the available heat in the flue gas from the fuel gas turbine. The specific process for steam generation is outlined in the process descriptions on page 27 above, and the process flow diagram for the steam cycle can be found in Figure 9 on page 27 above. The amount of heat available from the flue gas stream is sufficient to supply both the MP (B-601) and LP (B-602) boilers with enough heat to completely vaporize the condensate.

Unit Descriptions

Feed Expander E-101

The feed gas expander (E-101) is a carbon steel turbine that is used to expand the natural gas feed, S-100, from 725 psia and 68F to 300 psia and -15.8F prior to its entry into distillation column D-101. The turbine operates at an isentropic efficiency of 88%, and it shares a common shaft with the feed compressor (C-101), which allows its generated shaft work to be used to offset the work load required by the feed compressor.

The expansion of S-100 produces 3,415 hp, which is used to power the shaft of the feed compressor (C-101). The purchase cost of the expander is $459,481, and the bare module cost is $1,474,935. For additional specific information on the feed expander, see the specification sheet on page 78.

Scrub Column D-101

The scrub column, D-101, is the first distillation column in the process. It takes the feed stream (S-101) at 300 psia and 13,500 lbmol/hr and separates the methane and nitrogen from the rest of the hydrocarbons in the feed. The overhead product, stream S-102, flows at 20,833lbmol/hr and has a composition of 94.5% methane, 3% nitrogen and 2.5% ethane. The bottoms product, stream S-110, flows at 1180lbmol/hr and its composition is 53.3% ethane, 22.9% propane, 11.4% butane, 9.2% pentane, 2% hexane and 1.2% methane. This stream will be sent to the fractionation train for further distillation.

The column has no condenser and its reflux is instead supplied by a stream that enters at the top which flows at 8512 lbmol/hr and is mostly methane with ethane. The column is 12’7” tall and has diameter of 12’10”. The column runs at 300 psia and has 6 stages, which are comprised of stainless steel sieve trays. The overall pressure drop in the column is assumed to be

10 Psi. Stainless steel will be the primary material in the column and it will cost $342,050. For more information on column D-101, see the specification sheet on page 79.

Reboiler H-101

Reboiler H-101 provides the boil-up for the scrub column (D-101). The inlet temperature is 30F and the outlet is 66F. The reboiler is a shell and tube heat exchanger made out of carbon steel for the shell and stainless steel for the tubes. The reboiler has a heat duty of 8,306,070 Btu/hr which will be supplied by 464.64 lbmol/hr of 7 Psi steam. The cost of the reboiler will be $77,078. For more information about reboiler H-101, see the specification sheet on page 80.

Plate-Fin Heat Exchanger HX-101

The plate-fin heat exchanger, HX-101, is the main heat exchanger for the process. It is a brazed aluminum plate-fin heat exchanger that handles multiple hot and cold streams in a single unit. Specifically, this unit handles three hot streams (S-102, S-107, and S-116) and three cold streams (S-104, S-112, and S-114).

Stream S-102 is the overhead from the scrub column. It is comprised of 94.2% methane, 2.6% ethane, 2.9% nitrogen, and 0.3% C3+, and enters the exchanger at -130F and 290 psia. The stream is cooled and partially liquefied to -160F, where it emerges as S-103 and proceeds to the reflux separator (F-101). Stream S-107 is the compressed overhead of the reflux separator. It enters the exchanger at 90F and 720 psia, exits as S-108, at -220F. The last hot stream is the compressed nitrogen stream, S-116, which enters the exchanger at 90F and 995 psia, and is cooled to -50F before it is expanded to provide the bulk of the cooling power for the exchanger as S-114.

S-104 is the lower pressure overhead of the reflux separator. It is at -160F and 290 psia, and is warmed to 61.2F so that it can be recompressed once it comes out as stream S-105. S-112 is the cold nitrogen stream, and supplies the bulk of the cooling power to the main exchanger. It enters the exchanger at -222F and 130 psia, and emerges at 61.2F. Lastly, S-112 is the overhead from the nitrogen rejection column. It enters the exchanger at -259F and 18 psia, and emerges at 61.2F, as S-113, prior to combustion in the fuel gas turbine.

The heat exchanger consists of two assemblies, which are further divided into two pieces. The top part of one assembly services the hot fluids. It contains four aluminum cores that measure 1065 x 1575 x 4000 mm (W x L x H). The bottom section of the exchanger services the cold fluids, and contains two aluminum cores that measure 1220 x 1525 x 2400 mm. Both of these pieces constitute one assembly, and two assemblies make up the entire exchanger. Figure 24 in Appendix VII shows precisely what streams are contacting which other streams in a particular section of the exchanger.

The quoted price for the exchanger from the Applied UA consultant was $5,500,000 as the purchase cost, and the bare module cost is $16,500,000. A sample calculation was undertaken to confirm the calculations of the consultant, as a design exercise. That calculation can be found in Appendix VII, on page 182. For more information on the main heat exchanger, HX-101, see the specification sheet on pages 81 and 82.

Reflux Separator F-101

The reflux separator, F-101, takes the cooled overhead product (S-103), which flows at 20,833lbmol/hr, from column D-101 and flashes the stream at -160F and 290 psia. The bottoms product from the drum, S-111, has a flow rate of 8,513 lbmol/hr and is comprised of 1.0% nitrogen, 92.8% methane, and 5.9% ethane, with the balance C3+. It is at -160F and 290 psia, and is sent back to D-101 to be used as reflux. The overhead of F-102, S-104, has a flow rate of 12,320 lbmol/hr, and is comprised of 4.3% nitrogen, 95.2% methane, 0.4% ethane, and the balance C3+. The stream is at -160F and 290 psia, and is sent back to the main heat exchanger to be warmed prior to compression.

The vessel is 14’11” tall and 7’6” wide. It is being run with a hold up time of 3 minutes, and has a purchase cost of $94,187 and a bare module cost of $431,780. The vessel will be made of stainless steel. For more information about flash vessel F-101, see the specification sheet on page 83.

Feed Compressor C-101 and HX-102

The feed gas compressor (C-101) is a single stage carbon steel compressor that is used to recompress the warmed overhead from the reflux separator (S-105) from 61.2F and 290 psia to 206F and 725 psia. The compressor operates at an efficiency of 85%. This compressor also shares a common shaft with the feed expander (E-101), which helps to offset its electrical power requirement. After emerging from the compression stage, stream S-106 is cooled using seawater in HX-102 to 90F. The heat exchanger requires a cooling duty of 13,964,538 Btu/hr. The seawater required for cooling is provided by the cooling water distribution system.

In order to compress S-105, 5,697.5 hp is required. Of this, 3415.4 hp is provided by the shared shaft with the feed expander, leaving 2,282.1 hp to be provided via electricity from the power generation system. The purchase cost of the compressor is $2,080,294, calculated from correlations in Seider et al. The bare module cost of the compressor was calculated to be $6,677,743. For more details on the feed gas compressor, see the specification sheet on page 84. For more details on the feed gas compressor cooler, HX-102, see the specification sheet on page 85.

Nitrogen Rejection Vessel F-102

The nitrogen rejection vessel, F-102, removes some of the nitrogen from the final liquefied gas stream before it is sent to storage. The feed stream to the vessel, S-108, flows at 12,320 lbmol/hr, -220F and 390 psia. It consists of 4.3% nitrogen, 95.2% methane, 0.4% ethane, and the balance C3+. The stream is then flashed at 18 psia. The resulting overhead stream has a flow rate of 2,202 lbmol/hr and is composed of 19.7% nitrogen, 80.2% methane, and the balance C2+. This stream is then sent back through the main heat exchanger (HX-101) to be warmed before it is compressed and fed to the fuel gas turbine (T-301). The bottoms product of this flash (S-109) is the final LNG product. It has a flow rate of 10,118 lbmol/hr, and is comprised of 1.0% nitrogen, 98.5% methane, and 4.4% ethane, with the remaining 0.1% as C3+. This stream is sent to LNG storage.

The flash vessel is 17'6" tall and has a diameter of 8'9". The vessel will be made out of stainless steel and will have a purchase cost of $4,908 and a bare module cost of $14,969. For more information about the nitrogen rejection flash vessel, F-102, see the specification sheet on page 86.

Distillation Column D-201

This distillation column is the first column in the fractionation train, and it is used to separate the light hydrocarbons (ethane and propane) from the heavier hydrocarbons (butanes and higher) in stream S-201. This stream flows into the column at 41.2F and 200 psia. It has a flow rate of 1180 lbmol/hr, and contains 0.8% methane, 53.5% ethane, 22.9% propane, 11.4% butanes, and 11.4% C5+.

The overhead product of the column, S-202, emerges from the column at 41.8F and 190 psia. It has a flow rate of 910 lbmol/hr, and is comprised of 1.2% methane, 69.2% ethane, 28.8% propane, 0.7% butanes, and the balance C5+. This stream is sent to the second distillation column in the series, D-202, to be further separated. The bottoms product, S-205, emerges at 226F and 180 psia. It contains <0.02% methane and ethane, 2.6% propane, 47.4% butanes, 40.0% pentanes, and the balance hexanes, and is sent to the third distillation column in the sequence, D-203, to be further separated.

Column D-201 is 14’ 9” tall with a diameter of 4’ 1” and contains 10 stages in the form of stainless steel sieve trays. The column is run at 200 psia and has a reflux ratio of 2. Stainless steel is the primary material in the column. Column D-201 has a purchase cost of $52,830 and a bare module cost of $219,773. For more information about distillation column D-201, see the specification sheet on page 87.

Reboiler H-201

Reboiler H-201 provides the boil up for the first fractionation column D-201. The inlet temperature is 206F and the outlet temperature is 229F. It is a shell and tube heat exchanger made out of carbon steel for the shell and stainless steel for the tubes. The heat duty of the reboiler is 16,301,100 Btu/hr which will be supplied by 976.9 lbmol/hr of 80 Psi steam. The cost of the reboiler will be $116,580. For more information on reboiler H-201, see the specification sheet on page 88.

Condenser HX-201

Condenser HX-201 condenses the vapor stream from the top of the first fractionation column D-201. The inlet temperature is 70F and the outlet temperature is 45F. The condenser is a shell and tube heat exchanger, made with carbon steel for the shell and stainless steel for the tubes. The condenser has a heat duty of – 11,437,600 Btu/hr which will be cooled by 26,000 lbmol/hr of a chilled carbon dioxide stream. The purchase cost of HX-201 is $178,305, and its bare module cost is $565,225. For more information about condenser HX-201, see the specification sheet on page 89.

Reflux Accumulator A-201

Accumulator A-201 separates the vapor overhead product from the liquid that will be used as a reflux stream for the first fractionation column, D-201. The height of the drum is 5'8" and its width is 2'10". The holdup time for the drum is 3 minutes, it is made of stainless steel and it will cost $20,479. For more information about reflux accumulator A-201, see the specification sheet on page 90.

Distillation Column D-202

Distillation column D-202 separates stream S-202 into ethane and propane products. The feed stream is the overhead stream from column D-201, and consists of 1.2% methane, 69.2% ethane, 28.8% propane, 0.7% butanes, and the balance C5+. It is at 41.8F and 190 psia. The overhead product (S-203) from D-202 is the final ethane product. It comes out of the column at -11.6F and 180 psia, with the flow rate of 640 lbmol/hr, and is comprised of 0.6% methane, 97.2% ethane, 1.1% propane, and the balance C4+. The bottoms product (S-204) emerges from the column at 97.3F and 180 psia. It contains essentially no methane, 2.8% ethane, 94.6% propane, 2.5% butanes, and the balance C5+, and has a flow rate of 270 lbmol/hr.

The column is 14'9" tall and 3'3" in diameter and contains 10 stages in the form of stainless steel sieve trays. The column is being run at 200 Psia and a reflux ratio of 2. The column is made out of stainless steel and has a purchase cost of $45,564 and a bare module cost of $189,546. For more information on distillation column D-202, see the specification sheet on page 91.

Reboiler H-202

Reboiler H-202 provides the boil up for fractionation column D-202. The inlet temperature is 95F and the outlet temperature is 101F. It is a shell and tube heat exchanger made out of carbon steel for the shell and stainless steel for the tubes. The heat duty of the reboiler is 4,604,670 Btu/hr which will be supplied by 257.6 lbmol/hr of 7 Psi steam. The cost of the reboiler will be $72,499. For more information about reboiler H-202, see the specification sheet on page 92.

Condenser HX-202

Condenser HX-202 condenses the vapor stream from the top of the fractionation column D-202. The inlet temperature is 1F and the outlet temperature is -4F. The condenser is a shell and tube heat exchanger, made with carbon steel for the shell and stainless steel for the tubes. The condenser has a heat duty of -6,332,100Btu/hr which will be cooled by 26,000 lbmol/hr of a chilled carbon dioxide stream. The purchase cost of HX-202 is $260,631, and the bare module cost is $826,201. For more information about condenser HX-202, see the specification sheet on page 93.

Reflux Accumulator A-202

Accumulator A-202 separates the vapor overhead product from the liquid that will be used as a reflux stream in the light hydrocarbons fractionation column, D-202. The calculated height of the drum is 6’1” and its width is 3’. The holdup time for the drum is 3 minutes, it is made of stainless steel and it will cost $21,767. For more information on reflux accumulator A-202, see the specification sheet on page 94.

Distillation Column D-203

Distillation column D-203 is the third distillation column in the fractionation train. It takes as its feed the expanded bottoms product (S-206) from distillation column D-201, and separates the butanes from the pentanes plus so that both can be sold as final products. The feed stream has a flow rate of 270 lbmol/hr and is at 175F and 100 psia. It contains <0.02% methane and ethane, 2.6% propane, 47.4% butanes, 40.0% pentanes, and the balance hexanes.

The overhead product of D-203 is stream S-207. It emerges from the column at 130F and 90 psia, with a flow rate of 135 lbmol/hr. It contains 5.1% propane, 91.1% butanes, 3.8% C5+. The bottoms product, S-208, emerges from the column at 224F and 90 psia, with a flow rate of 135 lbmol/hr. It consists of 0.1% propane, 3.9% butanes, and 96.0% C5+.

The height of the column is 14’9” and its diameter is 2’3”. The column has 10 stages in the form of stainless steel sieve trays, is being run at 100 psia and has a reflux ratio of 4. It is made out of stainless steel and has a purchase cost of $37,715 and a bare module cost of $156,892. For more information about distillation column D-203, see the specification sheet on page 95.

Reboiler H-203

Reboiler H-203 provides the boil up for fractionation column D-203. The inlet temperature is 215F and the outlet temperature is 225F. It is a shell and tube heat exchanger made out of carbon steel for the shell and stainless steel for the tubes. The heat duty of the reboiler is 4,918,390 Btu/hr which will be supplied by 294.72 lbmol/hr of 80 Psi steam. The cost of the reboiler will be $69,345. For more information about reboiler H-203, see the specification sheet on page 96.

Condenser HX-203

Condenser HX-203 condenses the vapor stream from the top of the fractionation column D-203. The inlet temperature is 145F and the outlet temperature is 137F. The condenser is a shell and tube heat exchanger, made with carbon steel for the shell and stainless steel for the tubes. The condenser has a heat duty of -4,408,330Btu/hr which will be cooled by 940,000 lb/hr of cooling water. The purchase cost of HX-203 is $98,966, and the bare module cost is $313,724. For more information about HX-203, see the specification sheet on page 97.

Reflux Accumulator A-203

Accumulator A-203 separates the vapor overhead product from the liquid that will be used as a reflux stream in the heavy hydrocarbons fractionation column, D-203. The calculated height of the drum is 9' and its width is 2'3". The holdup time for the drum is 3 minutes, it is made of stainless steel and it will cost $16,561. For more information on reflux accumulator A-203, see the specification sheet on page 98.

Air Compressor C-301a-c

The air compressor (C-301a-c) is a multistage centrifugal compressor made of stainless steel that compresses an incoming air stream (S-307) to 500 psia in preparation for combustion in the fuel gas turbine combustion chamber. The compressor consists of three compression stages, each with an efficiency of 78%, and has a total inlet capacity of 30100 lbmol/hr of air. In between compression stages, the intermediate streams are cooled using seawater to 90F (in HX-301a, HX-301b, and HX-301c). Additionally, after the last compression stage, the outlet stream is also cooled to 90F with seawater. When the compressed air emerges from the final stage (S-313), it is sent to be burned in the fuel gas turbine combustion chamber.

The compressor requires 72,039 hp of energy in order to produce the outlet stream at the required pressure. The compressor also requires 181,820,257 Btu/hr of cooling duty, which is supplied by the cooling water distribution system. The purchase cost and bare module cost of the air compressor are included in the purchase cost and bare module cost of the fuel gas turbine and generator (T-301). For additional details about the air compressor, see its specification sheet on page 100.

Fuel Gas Compressor C-302a-c

The fuel gas compressor (C-302a-c) is a multi-stage centrifugal compressor made of stainless steel that compresses the warmed overhead from the nitrogen rejection vessel (S-113) to 500 psia in preparation for combustion in the fuel gas turbine combustion chamber. The compressor consists of three compression stages, each with an efficiency of 78%, and has a total inlet capacity of 2201.8 lbmol/hr of fuel gas. In between compression stages, as with the air compressor, the intermediate streams are cooled to 90F using seawater in HX-302a, HX-302b, and HX-302c. When the compressed fuel gas emerges from the third stage of the compressor as S-308, it is sent to the combustion chamber of the fuel gas turbine to be burned with the air stream.

The compressor is significantly smaller than the air compressor, and only requires 4699 hp of electric power to compress the fuel gas to the required pressure. The intercoolers in turn require 11,899,493 Btu/hr of cooling duty, which is supplied via the cooling water distribution system to the individual intercoolers. The purchase cost and the bare module cost of the fuel gas compressor are included in the purchase cost and the bare module cost of the fuel gas turbine and generator (T-301). For more detailed information on the fuel gas compressor, see the specification sheet on page 99.

Fuel Gas Turbine & Generator T-301

The fuel gas turbine (T-301) is used to produce the electrical power required to power all of the equipment in the process. The feed stream to the turbine is S-113, which is the warmed overhead from the nitrogen rejection vessel (F-102). This stream has a flow rate of 2,163 lbmol/hr, and is at 61.3F and 18 psia. The stream is comprised of 0.7% nitrogen, 80.2% methane, and 0.1% C2+. This stream is compressed in the fuel gas compressor (C-302a-c) to 500 psia, and is mixed with air also compressed to 500 psia in the air compressor (C-301a-c). These are then burned in the fuel gas turbine combustion chamber.

The specific model chosen for the process was the 179.9 MW ALSTOM GT13E2 – 50Hz turbine/generator combination from ALSTOM. The turbine spins at 30,000 rpm to produce the electrical energy needed. The generator produces 173,398 hp of electric power, which is then used to power the compressors and pumps in the process. The outlet stream from the turbine, S-314, flows at 32,302 lbmol/hr with a composition of 75% nitrogen, 11% water vapor, 8.5% oxygen, and 5.5% carbon dioxide. The pressure of S-113 is 3.9 psia, and its temperature is 738F.

The purchase cost of the turbine will be $48,000,000, and the bare module cost will be $86,619,004. These costs include the costs of both the air compressor (C-301a-c) and the fuel gas compressor (C-302a-c), in addition to the cost of the turbine and the generator. For more information about T-301, see the specification sheet on page 100.

Nitrogen Expander E-401

Expander E-401 lowers the temperature and pressure of the pre-cooled, compressed nitrogen stream (S-400 in Figure 7 above) from -50F and 995 psia to -222F and 130 psia when it exits the expander as S-401. Both of these streams have flow rates of 69,012 lbmol/hr. Expander E-401 produces 26,024 hp from the expansion of the nitrogen, and shares a shaft with the first stage of the nitrogen compressor (C-401a), which allows this produced work to help offset the power requirements of the nitrogen compressor. The efficiency of the expander is 88%.

The expander will be made of stainless steel, and the purchase cost and bare module cost will be $4,000,000 and $12,840,000 respectively. For more detailed information on expander E-401, see the specification sheet on page 101.

Nitrogen Compressor C-401a-d

Compressor C-401a-d is a multi-stage centrifugal compressor used to compress the warmed nitrogen stream that emerges from the main heat exchanger (HX-101) before its subsequent expansion. The compressor consists of four compression stages (C-401a-d), each with inter-cooling (HX-401a-d) to 90F. The efficiency of each compression stage is 86%.

The warm nitrogen stream (S-402 in Figure 7 above) enters the first compression stage at 61F and 130 psia. It emerges from the last stage of the compressor as stream S-410, at 90F and 995 psia. The compression requires 76,347 hp, 26,024 hp of which is provided by the shared shaft of the nitrogen expander (E-401), and 50,323 hp of which is supplied by the power generation system (T-301). Each compression stage is made out of carbon-steel, the total purchase cost for the entire system is $20,984,883, and the bare module cost is $67,361,474. For more information about compressor C-401a-d, see the specification sheet on page 102.

Cooling Water Pumps P-501 – P-504

The cooling water pumps, P-501 through P-504, are used to provide the process with seawater that will be used for cooling. Although only three pumps are sufficient to provide the process with cooling water, a spare pump is available in case one of the other pumps fails. The pumps to be used are centrifugal pumps that affect a pressure change of 70 psi. They pump cooling water to the system from 14.7 psia to 84.7 psia, each at a flow rate of 559,473 lbmol/hr.

Each pump requires 714 kW of electrical power, which is provided by the power generation system. The pumps are made of Inconel-600, which is corrosion resistant. The purchase cost of all four of the pumps is $476,368 and the bare module cost is $1,530,830. For more information on pumps P-501 to P-504, see the specification sheet on page 103.

MP Steam Boiler B-601

The MP steam boiler, B-601, is used to vaporize the saturated water to be used as steam to power reboilers H-201 and H-203. The hot gas emerging from the furnace (S-601) is used to provide the heat of vaporization. This stream emerges from the furnace at 2345F and 500 psia, and leaves the boiler at 1008F and 495 psia. The saturated water stream (S-607) enters at 324F and 114.7 psia. It is vaporized, and emerges as saturated steam (S-608) at 325F and 94.7 psia.

Boiler B-601 is a shell and tube heat exchanger that is made of 304 stainless steel for both the shell and the tube portions of the exchanger. It has a total heat duty of 22,460,492 Btu/hr and a total heat transfer area of 395.1 square feet. The boiler has a purchase cost of $48,476 and a bare module cost of $153,669. For more information about steam boiler B-601, see the specification sheet on page 104.

MP Steam Condensate Tank T-601

The MP steam condensate tank, T-601, is a vessel that is used to hold the condensed steam that has just been used to heat the MP steam boiler (B-601). Stream S-605 empties the saturated water into tank T-601 at a flow rate of 1,271.68 lbmol/hr at 324F and 94.7 psia. The tank empties into stream S-606. The tank is designed to have five minutes of liquid hold-up time. It is constructed of 304 stainless steel, and is 6’1” in height and 3’ in diameter. It has a purchase cost of $26,930 and a bare module cost of $82,136. For more information on tank T-601, see the specification sheet on page 105.

MP Steam Pump P-601

The MP steam pump, P-601, is used to increase the pressure of the saturated water stream once it leaves the condensate tank (T-601) before it is vaporized in B-601. The pump to be used is a centrifugal pump that affects a pressure change of 20 psi. It pumps stream S-606 from 94.7 psia to 114.7 psia. The pump requires 1.047 kW of electrical power, and has an efficiency of 45.2%. The purchase cost of the pump is $12,117 and the bare module cost is $38,887. For more information on pump P-601, see the specification sheet on page 106.

LP Steam Boiler B-602

The LP steam boiler, B-602, is used to vaporize the saturated water to be used as steam to power reboilers H-101 and H-202. The gas from the furnace, after it has vaporized the MP steam, is used to provide the heat of vaporization. This stream, S-602, enters at 1008F and 495 psia. This stream emerges from the boiler at 250F, and is vented to the atmosphere. The saturated water stream (S-615) enters at 235F and 41.7 psia. It is vaporized, and emerges as saturated steam (S-616) at 235F and 21.7 psia.

Boiler B-602 is a shell and tube heat exchanger that is made of 304 stainless steel for both the shell and the tube portions of the exchanger. It has a total heat duty of 13,019,950 Btu/hr and a total heat transfer area of 223.4 square feet. The boiler has a purchase cost of $42,759 and a bare module cost of $135,546. For more information about steam boiler B-602, see the specification sheet on page 107.

LP Steam Condensate Tank T-602

The LP steam condensate tank, T-602, is a vessel that is used to hold the condensed steam that has just been used to heat the LP steam boiler (B-602). Stream S-613 empties the saturated water into tank T-602 at a flow rate of 722.24 lbmol/hr at 235F and 21.7 psia. The tank in turn empties into stream S-614. The tank is designed to have five minutes of liquid hold-up time. It is constructed of 304 stainless steel, and is 5'10" in height and 2'11" in diameter. The tank has a purchase cost of $20,369 and a bare module cost of $62,124. For more information about tank T-602, see the specification sheet on page 108.

LP Steam Pump P-602

The LP steam pump, P-602, is used to increase the pressure of the saturated water stream once it leaves the condensate tank (T-602) before it is vaporized in B-602. The pump to be used is a centrifugal pump that affects a pressure change of 20 psi. It pumps stream S-614 from 21.7 psia to 41.7 psia. The pump requires 0.703 kW of electrical power, and has an efficiency of 35.5%. The purchase cost of the pump is $12,016 and the bare module cost is $38,802. For more information on pump P-602, see the specification sheet on page 109.

Steam Generation Furnace FN-601

The Steam Generation furnace FN-601 is used to increase the temperature of the flue gas before it heads to the steam generation cycle. The furnace takes 7% of the fuel gas turbine feed and burns it to generate the 36,219,100 Btu/hr required to heat stream S-600. This heats the 1948lbmol/hr of 73% methane stream from 90F to 2346F. The furnace will be made out of stainless steel at a purchase cost of $2,177,474 and a bare module cost of $6,902,594. For more information on furnace FN-601 see the specification sheet on page 110.

Carbon Dioxide Compressor C-701 and HX-702

The carbon dioxide compressor (C-701) is a single stage carbon steel compressor that is used to compress the warmed carbon dioxide stream (S-700) from 47.1F and 100 psia to 90F and

295 psia after being cooled to 90F in the intercooler heat exchanger, HX-702. The compressor operates at an efficiency of 85%. The seawater required for cooling is provided by the cooling water distribution system.

In order to compress the carbon dioxide stream, 13,595 hp is required, 8,411 hp of which is provided by the shared shaft with the carbon dioxide expander, and the rest of which is provided via electricity from the power generation system. The purchase cost of the compressor is $4,251,719 calculated from correlations in Seider et al. The bare module cost of the compressor was calculated to be $13,648,018. Relative to equipment costs, the purchase cost of carbon dioxide is minimal, and is included in the bare module cost. For more details on the carbon dioxide compressor, see the specification sheet on page 111.

Carbon Dioxide Expander E-701

Expander E-701 lowers the temperature and pressure of the compressed carbon dioxide stream (S-702 in Figure 10 above) from 90F and 295 psia to -29F and 100 psia when it exits the expander as S-703. Both of these streams have flow rates of 26,000 lbmol/hr. Expander E-701 produces 8,411 hp from the expansion of the carbon dioxide, and shares a shaft with carbon dioxide compressor (C-701), which allows this produced work to help offset the power requirements of that compressor. The efficiency of the expander is 88%.

The expander will be made of stainless steel, and the purchase cost and bare module cost will be $953,560 and $3,060,928 respectively. For more detailed information on expander E-701, see the specification sheet on page 112.

Unit Specification Sheets

Feed Expander

Identification:	Item:	Turbine		Date: 04/14/2009
	Item No.	E-101		
	Quantity:	1		
Function:				
Operation:	Continuous			

Materials Handled:	Inlet Stream	Outlet Stream
Stream ID:	S-100	S-101
Quantity (lbmol/hr):	13500.00	13500.00
Composition (lbmol/hr):	Vapor	Vapor
Nitrogen	540.00	540.00
Methane	11745.00	11745.00
Ethane	675.00	675.00
Propane	270.00	270.00
Isobutane	67.50	67.50
n-butane	67.50	67.50
Isopentane	40.50	40.50
n-Pentane	67.50	67.50
Hexanes	27.00	27.00
Temperatu *hexanes*	68	-15.84
pressure (in psia) :	725	300

Design Data:

Type:	
Net work generated:	3415 hp
Isentropic Efficiency:	0.88

Comments:

Shares shaft with C-101

Scrub Column

Identification:	Item:	Distillation Column		Date: 04/14/2009	
	Item No.	D-101			
	Quantity:	1			

Function: To separate the methane from the other heavier components in the feed

Operation: Continuous

Materials Handled:	Inlet Feed	Inlet Reflux	Top Out	Bottom Out
Stream ID:	S-101	S-111	S-102	S-110
Quantity (lbmol/hr):	13500.00	8512.69	20833.00	1180.00
Composition (lbmol/hr):	Mixed	Liquid	Vapor	Vapor
Methane	11745.00	7901.46	19636.00	10.74
Ethane	675.00	503.29	548.73	629.57
Propane	270.00	24.10	24.40	269.69
Isobutane	67.50	0.54	0.54	67.50
n-Butane	67.50	1.01	1.02	67.50
Isopentane	40.50	3.60E-02	3.60E-02	40.50
n-Pentane	67.50	3.57E-02	3.57E-02	67.50
Hexanes	27.00	4.62E-04	4.62E-04	27.00
Nitrogen	540.00	82.21	622.21	0.00
Temperature (in F):	-15.8406	-160	-129.806	66.1381
pressure (in psia) :	300	290	290	300

Design Data:

Material:	304 Stainless Steel
Stages:	6
Pressure:	290 psi
Pressure drop per stage:	2 psi
Diameter:	12'10"
Height:	12'7"
Tray spacing:	0.5ft
Tray Type:	Sieve

Comments:

Reboiler

Identification:	Item:	Heat Exchanger	Date: 04/14/2009
	Item No.	H-101	
	Quantity:	1	

Function: To provide vapor boil-up for the liquid product in D-101 column

Operation: Continuous

Materials Handled:

	Bottom Stage of D-101
Stream ID:	S-110
Quantity (lbmol/hr):	1180.00
Composition (lbmol/hr):	Vapor
Methane	10.74
Ethane	629.57
Propane	269.69
Isobutane	67.50
n-butane	67.50
iIsopentane	40.50
n-pentane	67.50
hexanes	27.00
Inlet Temperature (in F):	29.872
Outlet Temperature (in F)	66.138

Design Data: Shell and Tube Heat Exchanger

Material of Construction:	Carbon Steel/ 304 Stainless Steel (Shell/Tube)
Heat duty:	8306070 Btu/hr
Utilities:	8380.22 lb/hr
Type:	7.107 psi steam

Comments:

Brazed Aluminium Plate-Fin Heat Exchanger

Identification: Item: Brazed Aluminum Plate-Fin Exchanger   Date: 04/14/2009
Item No. HX-101
Quantity: 1

Function: To exchange heat between process streams to pre-cool nitrogen, liquefy natural gas, and cool compressed fluids

Operation: Continuous

Materials Handled:	Cold In	Cold In	Cold In	Cold Out	Cold Out	Cold Out
Stream ID:	S-104	S-112	S-114	S-105	S-113	S-115
Quantity (lbmol/hr):	12319.97	2201.8	69012.13	12319.97	2201.8	69012.13
Composition (lbmol/hr):						
Nitrogen	539.99	434.76	69012.13	539.99	434.76	69012.13
Carbon Dioxide	0	0	0	0	0	0
Methane	11734.24	1767.03	0	11734.24	1767.03	0
Ethane	45.43	0.014622	0	45.43	0.014622	0
Propane	0.0307	1.40E-06	0	0.0307	1.40E-06	0
n-Butane	8.51E-04	3.72E-11	0	8.51E-04	3.72E-11	0
Isobutane	3.11E-03	7.17E-10	0	3.11E-03	7.17E-10	0
n-Pentane	9.26E-06	TRACE	0	9.26E-06	TRACE	0
Isopentane	8.29E-06	TRACE	0	8.29E-06	TRACE	0
Hexanes	2.05E-06	TRACE	0	2.05E-06	TRACE	0
Temperature (F)	-160	-259	-222	61.2	61.2	61.2
Pressure (psia)	290	18	130	290	18	130

Materials Handled:	Hot In	Hot In	Hot In	Hot Out	Hot Out	Hot Out
Stream ID:	S-102	S-107	S-116	S-103	S-108	S-117
Quantity (lbmol/hr):	20832.69	12319.97	69012.13	20832.69	12319.97	69012.13
Composition (lbmol/hr):						
Nitrogen	622.211	539.99	69012.13	622.211	539.99	69012.13
Carbon Dioxide	0	0	0	0	0	0
Methane	19635.72	11734.24	0	19635.72	11734.24	0
Ethane	548.73	45.43	0	548.73	45.43	0
Propane	24.4	0.0307	0	24.4	0.0307	0
n-Butane	0.539	8.51E-04	0	0.539	8.51E-04	0
Isobutane	1.016	3.11E-03	0	1.016	3.11E-03	0
n-Pentane	0.036	9.26E-06	0	0.036	9.26E-06	0
Isopentane	0.035	8.29E-06	0	0.035	8.29E-06	0
Hexanes	4.62E-04	2.05E-06	0	4.62E-04	2.05E-06	0
Temperature (F)	-130	90	90	-160	-220	-50
Pressure (psia)	290	720	995	290	720	995

Design Data:	Brazed Aluminum Plate-Fin Heat Exchanger	
Material of Construction:	Aluminum	
Heat duty:	173438547	Btu/hr
Number of Assemblies:	2	
Assembly 1		
No. of Cores:	4	
Core Height:	1065	mm
Core Width:	1575	mm
Core Length:	4000	mm
Assembly 2		
No. of Cores:	2	
Core Height:	1220	mm
Core Width:	1525	mm
Core Length:	2400	mm
Fin type(s):	1/8 Serrated, 4/8 Serrated, 6/8 Serrated 5% Perforated	
Purchase Cost:	$5,500,000	

Reflux Flash Vessel

Identification:	Item:	Flash Vessel		Date: 04/14/2009
	Item No.	F-101		
	Quantity:	1		

Function: To further separate the methane from the other components and send the bottoms liquid stream back to the scrub column as reflux

Operation: Continuous

Materials Handled:

	Inlet Feed	Top Out	Bottom Out
Stream ID:	S-103	S-104	S-111
Quantity (lbmol/hr):	20833.00	12320.00	8512.6868
Composition (lbmol/hr):	Mixed	Vapor	Liquid
Methane	19636.00	11734.00	7901.46
Ethane	548.73	45.43	503.29
Propane	24.40	0.31	24.10
Isobutane	0.54	8.51E-04	0.54
n-Butane	1.02	3.11E-03	1.01
Isopentane	0.04	9.26E-06	0.04
n-Pentane	0.04	8.29E-06	0.04
Hexanes	4.62E-04	2.05E-08	4.62E-04
Nitrogen	622.21	539.99	82.21
Temperature (in F):	-159.9992	-160	-160

Design Data:

Material:	304 Stainless Steel
Pressure:	290 psi
Diameter:	7'6"
Height:	14'11"
Vapor fraction:	.59138
Hold-up time:	3 min

Comments:

Feed Compressor

Identification:	Item: Compressor		Date: 04/14/2009
	Item No. C-101		
	Quantity: 1		
Function:			
Operation:	Continuous		

Materials Handled:	Inlet Stream	Outlet Stream
Stream ID:	S-105	S-107
Quantity (lbmol/hr):	12320	12320
Composition (lbmol/hr):	Vapor	Vapor
Nitrogen	540	540
Methane	11734	11734
Ethane	45	45
Propane	0.31	0.31
Temperature (in F):	61.23	90
pressure (in psia) :	290	725

Design Data:

Type:	Centrifugal
Net work required:	5698 hp
Isentropic Efficiency:	0.85

Comments:

Shared shaft with E-101

Intercooler

Identification:	Item:	Heat Exchanger	Date: 04/14/2009
	Item No.	HX-102	
	Quantity:	1	

Function: To cool down the stream from the compressor C-101 before entering the main heat exchanger

Operation: Continuous

Materials Handled:	Inlet Stream	Outlet Stream
Stream ID:	S-106	S-107
Quantity (lbmol/hr):	12319.97	12319.97
Composition (lbmol/hr):	Vapor	Vapor
Methane	11734.24	11734.24
Ethane	45.43	45.43
Propane	0.31	0.31
Nitrogen	539.99	539.99
Temperature (in F):	205.85	90
pressure (in psia) :	725	720

Design Data: Shell and Tube Heat exchanger

Material of Construction:	Carbon Steel/ 304 Stainless Steel (Shell/Tube)
Heat duty:	-13965000 Btu/hr
Pressure drop:	5
Utilities:	940000 lb/hr
Type:	Cooling water

Comments:

Nitrogen Rejection Vessel

Identification: Item: Flash Vessel — Date: 04/14/2009

Item No. F-102

Quantity: 1

Function: To remove the nitrogen from the LNG product stream before sending it to LNG storage as final product

Operation: Continuous

Materials Handled:

	Inlet Feed	Top Out	Bottom Out
Stream ID:	S-108	S-112	S-109
Quantity (lbmol/hr):	12320.00	2201.80	10118.00
Composition (lbmol/hr):	Mixed	Vapor	Liquid
Methane	11734.00	1767.03	9967.20
Ethane	45.43	1.46E-02	45.42
Propane	0.31	1.40E-06	0.31
Isobutane	8.51E-04	3.72E-11	8.51E-04
n-Butane	3.11E-03	7.17E-10	3.11E-03
Isopentane	9.26E-03	7.39E-15	9.26E-06
n-Pentane	3.29E-06	5.50E-15	8.29E-06
Hexanes	2.05E-08	4.67E-19	2.05E-08
Nitrogen	539.99	434.76	105.24
Temperature (in F):	-220	-259.209	-259.2085

Design Data:

Material:	304 Stainless Steel
Pressure:	18 psi
Diameter:	8'9"
Height:	17'6"
Vapor fraction:	.17872
Hold-up time:	5 min

Comments:

Fractionation Column

Identification: Item: Distillation Column — Date: 04/14/2009
Item No. D-201
Quantity: 1

Function: To separate the heavy and the light components coming from the bottoms product stream of the scrub column.

Operation: Continuous

Materials Handled:

	Inlet Feed	Top Out	Bottom Out
Stream ID:	S-201	S-202	S-205
Quantity (lbmol/hr):	1179.97	909.97	270
Composition (lbmol/hr):	Mixed	Vapor	Liquid
Methane	7.67	7.67	2.14E-07
Ethane	632.7	632.6616	3.84E-02
Propane	269.6	262.5463	7.0537
Isobutane	67.5	1.258	66.242
n-Butane	67.5	5.8213	61.6787
Isopentane	40.5	7.26E-03	40.4927
n-Pentane	67.5	5.49E-03	67.4945
Hexanes	27	4.73E-06	27
Temperature (in F):	51	45.2958	228.606

Design Data:

Material:	304 Stainless Steel
Stages:	10
Pressure:	200 psi
Molar Reflux Ratio:	2
Diameter:	4'1"
Height:	14'9"
Tray spacing:	0.5 ft
Tray Type:	Sieve

Comments:

Reboiler

Identification:	Item:	Heat Exchanger	Date: 04/14/2009
	Item No.	H-201	
	Quantity:	1	

Function: To create vapor boil-up of the liquid product in column D-201

Operation: Continuous

Materials Handled:

	Bottom Stage of D-201
Stream ID:	S-205
Quantity (lbmol/hr):	270
Composition (lbmol/hr):	Liquid
Methane	2.14E-07
Ethane	3.84E-02
Propane	7.0537
Isobutane	66.242
n-Butane	61.6787
Isopentane	40.4927
n-Pentane	67.4945
Hexanes	27
Inlet Temperature (in F):	206.18
Outlet Temperature (in F)	228.606

Design Data: Shell and Tube Heat Exchanger

Material of Construction:	Carbon Steel/ 304 Stainless Steel (Shell/Tube)
Heat duty:	0.163011E08 Btu/hr
Utilities:	18090.7 lb/hr
Type:	79.77 psi steam

Comments:

Condenser

Identification:	Item:	Heat Exchanger	Date: 04/14/2009
	Item No.	HX-201	
	Quantity:	1	

Function: To create liquid flow for reflux and liquid product in D-201 column

Operation: Continuous

Materials Handled:

	Top Stage of D-201
Stream ID:	S-202
Quantity (lbmol/hr):	909.97
Composition (lbmol/hr):	Vapor
Methane	7.67
Ethane	632.6616
Propane	262.5463
Isobutane	1.258
n-butane	5.8213
iIsopentane	7.26E-03
n-pentane	5.49E-03
hexanes	4.73E-06

Inlet Temperature (in F): 70.466
Outlet Temperature (in F) 45.296

Design Data: Shell and Tube Heat Exchanger

Material of Construction:	304 Stainless Steel/ 304 Stainless Steel (Shell/Tube)
Heat duty:	-0.114376E8 Btu/hr
Utilities:	26,000lbmol/hr
Type:	CO2

Comments:

Reflux Accumulator

Identification:	Item:	Drum Vessel	Date: 04/14/2009
	Item No.	A-201	
	Quantity:	1	

Function: To separate the liquid and the vapor coming from the condenser where the liquid returns as reflux into column D-201

Operation: Continuous

Materials Handled:	Inlet Feed	Reflux Bottom Out	Top Out
Stream ID:	From condenser	To pump	S-202
Quantity (lbmol/hr):	1516.62	606.65	909.97
Composition (lbmol/hr):	Mixed	Liquid	Vapor
Methane	12.78	5.11	7.67
Ethane	1054.44	421.77	632.6616
Propane	437.58	175.03	262.5463
Isobutane	2.10	0.84	1.258
n-Butane	9.70	3.88	5.8213
Isopentane	1.21E-02	4.84E-03	7.26E-03
n-Pentane	9.15E-03	3.66E-03	5.49E-03
Hexanes	7.88E-06	3.15E-06	4.73E-06
Temperature (in F):	45.296	45.296	45.296

Design Data:

Material:	304 Stainless Steel
Pressure:	200 psia
Molar Reflux Ratio:	2
Diameter:	2' 10"
Height:	5' 8"
Vapor fraction:	0.333
Hold-up time:	3 min

Comments:

Fractionation Column

Identification:	Item:	Distillation Column	Date: 04/14/2009
	Item No.	D-202	
	Quantity:	1	

Function: To separate the ethane and the propane coming from the top vapor product stream of the fractionation column D-201

Operation: Continuous

Materials Handled:

	Inlet Feed	Top Out	Bottom Out
Stream ID:	S-202	S-203	S-204
Quantity (lbmol/hr):	909.97	639.97	270
Composition (lbmol/hr):	Mixed	Vapor	Liquid
Methane	7.67	7.67	9.37E-06
Ethane	632.6616	625.0975	7.56
Propane	262.5463	7.2018	255.3445
Isobutane	1.258	5.06E-05	1.258
n-butane	5.8213	6.57E-04	5.8207
iIsopentane	7.26E-03	2.24E-09	7.26E-03
n-pentane	5.49E-03	6.09E-10	5.49E-03
hexanes	4.73E-06	2.91E-16	4.73E-06
Temperature (in F):	45.2958	-4.2486	100.9358

Design Data:

Material:	304 Stainless Steel
Stages:	10
Pressure:	200 psi
Molar Reflux Ratio:	2
Diameter:	3'3"
Height:	14'9"
Tray spacing:	.5 ft
Tray Type:	Sieve

Comments:

Reboiler

Identification:	Item:	Heat Exchanger	Date: 04/14/2009
	Item No.	H-202	
	Quantity:	1	

Function: To create vapor boil-up of the liquid product in column D-202

Operation: Continuous

Materials Handled:	Bottom Stage of D-202
Stream ID:	S-204
Quantity (lbmol/hr):	270
Composition (lbmol/hr):	Liquid
Methane	9.37E-06
Ethane	7.56
Propane	255.3445
Isobutane	1.258
n-Butane	5.8207
Isopentane	7.26E-03
n-Pentane	5.49E-03
Hexanes	4.73E-06
Inlet Temperature (in F):	94.694
Outlet Temperature (in F)	100.94

Design Data: Shell and Tube Heat Exchanger

Material of Construction:	Carbon Steel/ 304 Stainless Steel (Shell/Tube)
Heat duty:	4604670 Btu/hr
Utilities:	4645.77 lb/hr
Type:	7.107 psi steam

Comments:

Condenser

Identification:	Item:	Heat Exchanger	Date: 04/14/2009
	Item No.	HX-202	
	Quantity:	1	

Function: To create liquid flow for reflux and liquid product in D-202 column

Operation: Continuous

Materials Handled:

	Top Stage of D-202
Stream ID:	S-203
Quantity (lbmol/hr):	639.97
Composition (lbmol/hr):	Vapor
Methane	7.67
Ethane	625.0975
Propane	7.2018
Isobutane	5.06E-05
n-Butane	6.57E-04
Isopentane	2.24E-09
n-Pentane	6.09E-10
Hexanes	2.91E-16
Inlet Temperature (in F):	1.0863
Outlet Temperature (in F)	-4.2486

Design Data: Shell and Tube Heat Exchanger

Material of Construction:	304 Stainless Steel/ 304 Stainless Steel (Shell/Tube)
Heat duty:	-6332100 Btu/hr
Utilities:	26,000 lbmol/hr
Type:	CO2

Comments:

Reflux Accumulator

Identification:	Item: Drum Vessel		Date: 04/14/2009
	Item No. A-202		
	Quantity: 1		

Function:	To separate the liquid and the vapor coming from the condenser where the liquid returns as reflux into column D-202
Operation:	Continuous

Materials Handled:	Inlet Feed	Reflux Bottom Out	Top Out
Stream ID:	From condenser	To pump	S-203
Quantity (lbmol/hr):	1066.62	426.65	639.97
Composition (lbmol/hr):	Mixed	Liquid	Vapor
Methane	12.78	5.11	7.67
Ethane	1041.83	416.73	625.0975
Propane	12.00	4.80	7.2018
Isobutane	8.43E-05	3.37E-05	5.06E-05
n-Butane	1.09E-03	4.38E-04	6.57E-04
Isopentane	3.74E-09	1.49E-09	2.24E-09
n-Pentane	1.01E-09	4.06E-10	6.09E-10
Hexanes	4.85E-16	1.94E-16	2.91E-16
Temperature (in F):	-4.2486	-4.2486	-4.2486

Design Data:

Material:	304 Stainless Steel
Pressure:	200 psi
Diameter:	3'
Height:	6'1"
Vapor fraction:	0.333
Hold-up time:	3 min

Comments:

Fractionation Column

Identification: Item: Distillation Column — Date: 04/14/2009
Item No. D-203
Quantity: 1

Function: To separate the butanes and the pentane coming from the bottom liquid product stream of the fractionation column D-201

Operation: Continuous

Materials Handled:

	Inlet Feed	Top Out	Bottom Out
Stream ID:	S-206	S-207	S-208
Quantity (lbmol/hr):	270	135	135
Composition (lbmol/hr):	Mixed	Vapor	Liquid
Methane	2.14E-07	2.14E-07	4.58E-17
Ethane	3.84E-02	3.84E-02	2.08E-08
Propane	7.0537	7.0529	7.92E-04
Isobutane	66.24	62.00	4.2449
n-Butane	61.68	60.70	0.976
Isopentane	40.49	3.21	37.28
n-Pentane	67.49	1.99	65.50
Hexanes	27.00	4.62E-03	27.00
Temperature (in F):	175.7994	137.4423	224.8653

Design Data:

Material:	304 Stainless Steel
Stages:	10
Pressure:	100 psi
Molar Reflux Ratio:	4
Diameter:	2'3"
Height:	14'9"
Tray spacing:	.5 ft
Tray Type:	Sieve

Comments:

Reboiler

Identification:	Item:	Heat Exchanger	Date: 04/14/2009
	Item No.	H-203	
	Quantity:	1	

Function: To create vapor boil-up of the liquid product in column D-203

Operation: Continuous

Materials Handled:	Bottom Stage of D-203
Stream ID:	S-208
Quantity (lbmol/hr):	135
Composition (lbmol/hr):	Liquid
Methane	4.58E-17
Ethane	2.08E-08
Propane	7.92E-04
Isobutane	4.2449
n-butane	0.976
iIsopentane	37.28
n-pentane	65.50
hexanes	27.00
Inlet Temperature (in F):	215.14
Outlet Temperature (in F)	224.87

Design Data: Shell and Tube Heat Exchanger

Material of Construction:	Carbon Steel/ 304 Stainless Steel (Shell/Tube)
Heat duty:	4918390 Btu/hr
Utilities:	5458.36 lb/hr
Type:	79.77 psi steam

Comments:

Condenser

Identification:	Item:	Heat Exchanger	Date: 04/14/2009
	Item No.	HX-203	
	Quantity:	1	

Function: To create liquid flow for reflux and liquid product in D-203 column

Operation: Continuous

Materials Handled:

	Top Stage of D-203
Stream ID:	S-207
Quantity (lbmol/hr):	135
Composition (lbmol/hr):	Vapor
Methane	2.14E-07
Ethane	3.84E-02
Propane	7.0529
Isobutane	62.00
n-Butane	60.70
Isopentane	3.21
n-Pentane	1.99
Hexanes	4.62E-03
Inlet Temperature (in F):	144.54
Outlet Temperature (in F)	137.44

Design Data: Shell and Tube Heat Exchanger

Material of Construction:	304 Stainless Steel/ 304 Stainless Steel (Shell/Tube)
Heat duty:	-4408330 Btu/hr
Utilities:	940000 lb/hr
Type:	Cooling water

Comments:

Reflux Accumulator

Identification:	Item:	Drum Vessel		Date: 04/14/2009
	Item No.	A-203		
	Quantity:	1		

Function: To separate the liquid and the vapor coming from the condenser where the liquid returns as reflux into column D-203

Operation: Continuous

Materials Handled:	Inlet Feed	Reflux Bottom Out	Top Out
Stream ID:	From condenser	To pump	S-207
Quantity (lbmol/hr):	243.00	108.00	135
Composition (lbmol/hr):	Mixed	Liquid	Vapor
Methane	3.85E-07	1.71E-07	2.14E-07
Ethane	0.07	0.03	3.84E-02
Propane	12.70	5.64	7.0529
Isobutane	111.59	49.60	62.00
n-Butane	109.26	48.56	60.70
Isopentane	5.79	2.57	3.21
n-Pentane	3.58	1.59	1.99
Hexanes	0.01	3.69E-03	4.62E-03
Temperature (in F):	137.44	137.44	137.44

Design Data:

Material:	304 Stainless Steel
Pressure:	100 psi
Diameter:	2'3"
Height:	9'
Vapor fraction:	.333
Hold-up time:	3 min

Comments:

Fuel Gas Compressor

Identification:	Item: Compressor		Date: 04/14/2009
	Item No. C-302		
	Quantity: 1		

Function: Increases the pressure of the gas inlet stream to produce pressure energy

Operation: Continuous

Materials Handled:	Inlet Stream	Outlet Stream
Stream ID:	S-300	S-306
Quantity (lbmol/hr):	2201.80	2201.80
Composition (lbmol/hr):	Vapor	Vapor
Nitrogen	434.76	434.76
Methane	1.77E+03	1.77E+03
Ethane	0.01	0.01
Temperature (in F):	61.22	90
pressure (in psia) :	18	505

Design Data:

Type:

Net work required: 4698.928 hp

No. of stages: 3

Comments:

Natural Gas Turbine Power Generation System

Identification:	Item: Gas Turbine & Generator		Date: 04/14/2009
	Item No. T-301 and Power Generator		
	Quantity: 1		
Function:	To generate electricity from burning natural gas		
Operation:	Continuous		

Materials Handled:

	Inlet Stream	Outlet Stream
Stream ID:	S-314	S-315
Quantity (lbmol/hr):	30854.00	30854.00
Composition (lbmol/hr):	Vapor	Vapor
Nitrogen	23176.00	23176.00
Carbon Dioxide	0.00	1662.12
Oxygen	6015.59	2691.36
Water	0	3.32E+03
Methane	1.66E+03	0
Ethane	1.38E-02	0
Temperature (in F):	90	732
Pressure (in psia) :	500	3.9293

Design Data:

Type:	179.9 MW ALSTOM GT13E2	
Turbine Stage Power Production:	166,987	hp
Generator Efficiency:	98%	
Generated Electric Power:	163,647	hp
Total Available Electric Output:	163,647	hp
Air Compressor Power Requirement:	73,235	hp
Combustion Chamber Inlet Temp.:	90	F
Combustion Chamber Temp.:	2345	F
Turbine Stage Inlet Tem.:	2345	F
Turbine Stage Outlet Temp.:	723	F
Purchase Cost:	$48,000,000	

Comments:

Nitrogen Expander

Identification:	Item:	Turbine	Date: 04/14/2009
	Item No.	E-401	
	Quantity:	1	

Function: Uses the pressure energy from the nitrogen inlet stream to produce electrical energy

Operation: Continuous

Materials Handled:	Inlet Stream	Outlet Stream
Stream ID:	S-400	S-401
Quantity (lbmol/hr):	69012.00	69012.00
Composition (lbmol/hr):	Vapor	Vapor
Nitrogen	69005	69005
Temperature (in F):	-50	-222
pressure (in psia) :	995	130

Design Data:

Materials:	Stainless steel
Net work generated:	26243 hp
Isentropic Efficiency:	.88

Comments:

Shares common shaft with C-401a

Nitrogen Compressor

Identification:	Item:	Compressor	Date: 04/14/2009
	Item No.	C-401(a-d)	
	Quantity:	4	

Function: Increases the pressure of the nitrogen stream to produce pressure energy

Operation: Continuous

Materials Handled:	Inlet Stream	Outlet Stream
Stream ID:	S-402	S-410
Quantity (lbmol/hr):	69012.13	69012.13
Composition (lbmol/hr):	Vapor	Vapor
Nitrogen	69012.00	69012.00
Temperature (in F):	61.22	90
pressure (in psia) :	130	995

Design Data:

Materials:	Carbon Steel	
Type:	Centrifugal	
Net work required:	76,347	hp
(a)	17,870	hp
(b)	19,810	hp
(c)	19,430	hp
(d)	19,247	hp
Total Cooling Load:	1.91E+08	Btu/hr

Comments:

4 stage compression

Pump

Identification:	Item:	Pump		Date: 04/14/2009
	Item No.	P-501	(same for P502-P504)	
	Quantity:	4		
Function:	To pump and distribute the cooling water in the process			
Operation:	Continuous			

Materials Handled:	Inlet Stream	Outlet Stream
Stream ID:	1	2
Composition (lbmol/hr):	Liquid	Liquid
Water	563808.00	563808.00
Temperature (in F):	68	68.0409
pressure (in psia) :	14.7	84.6959

Design Data:

Material of Construction:	Inconel-600
Type:	Centrifugal
Pressure Change:	70 psi
Electricity Required:	957.6 hp

Comments:

The pumps have one in line spare pump

Pumps are corrosion resistant

High Pressure Boiler

Identification: Item: Heat Exchanger — Date: 04/14/2009
Item No. B-601
Quantity: 1

Function: To generate steam for the column reboilers

Operation: Continuous

Materials Handled:	Cold In	Hot In	Hot Out	Cold Out
Stream ID:	S-607	S-600	S-601	S-608
Quantity (lbmol/hr):	1307.17	1948.01	1948.01	1307.17
Composition (lbmol/hr):	Vapor	Liquid	Liquid	Liquid
Nitrogen	0.00	1463.27	1463.27	0.00
Carbon Dioxide	0.00	104.94	104.94	0.00
Oxygen	0.00	169.92	169.92	0.00
Water	1307.17	209.88	209.88	1307.17
Temperature (in F):	300.01	2345.618	1008.251	325.34
pressure (in psia) :	114.7	500	500	94.7

Design Data: Shell and Tube Heat Exchanger

Material of Construction:	304 Stainless Steel/ 304 Stainless Steel (Shell/Tube)
Heat duty:	22460389.1 Btu/hr
Heat Transfer Area:	119.9 ft^2
Heat Transfer Coefficient:	149.7 BTU/hr-ft^2-R

Comments:

Medium Pressure Tank

Identification:	Item: Drum Vessel		Date: 04/14/2009
	Item No. T-601		
	Quantity: 1		

Function: To hold the water coming from the column reboilers after being used as steam to be transported to the boilers for steam re-generation

Operation: Continuous

Materials Handled:	Inlet	Bottom
Stream ID:	S-605	S-606
Quantity (lbmol/hr):	1307.17	1307.17
Composition (lbmol/hr):	Liquid	Liquid
Water	1307.17	1307.17
Temperature (in F):	299.92	299.92

Design Data:

Material:	304 Stainless Steel
Pressure:	94.70 psia
Diameter:	3'
Height:	6' 1"
Hold-up time:	5 min

Comments:

High Pressure Pump

Identification:	Item: Pump		Date: 04/14/2009
	Item No. P-601		
	Quantity: 1		
Function:	To pump the water from the Tank to the boiler		
Operation:	Continuous		

Materials Handled:	Inlet Stream	Outlet Stream
Stream ID:	S-606	S-607
Quantity (lbmol/hr):	1307.17	1307.17
Composition (lbmol/hr):	Liquid	Liquid
Water	1307.17	1307.17
Temperature (in F):	299.91	300.01
pressure (in psia) :	94.7	114.7

Design Data:

Type:	Centrifugal
Pressure Change:	20 psi
Electricity Required:	1.047 KW
Efficiency:	0.452

Comments:

Low Pressure Boiler

Identification:	Item:	Heat Exchanger		Date: 04/14/2009
	Item No.	B-602		
	Quantity:	1		

Function: To generate steam for the column reboilers

Operation: Continuous

Materials Handled:	Cold In	Hot In	Hot Out	Cold Out
Stream ID:	S-615	S-602	S-619	S-616
Quantity (lbmol/hr):	723.05	1948.01	1948.01	723.05
Composition (lbmol/hr):	Liquid	Vapor	Vapor	Vapor
Nitrogen	0.00	1463.27	1463.27	0.00
Carbon Dioxide	0.00	104.94	104.94	0.00
Oxygen	0.00	169.92	169.92	0.00
Water	723.05	209.88	209.88	723.05
Temperature (in F):	228.68	1008.251	250.0015	235.36
pressure (in psia) :	41.7	500	500	21.7

Design Data: Shell and Tube Heat Exchanger

Material of Construction:	304 Stainless Steel/ 304 Stainless Steel (Shell/Tube)
Heat duty:	13019959 Btu/hr
Heat Transfer Area:	1197.85 ft^2
Heat Transfer Coefficient:	149.7 BTU/hr-ft^2-R

Comments:

Low Pressure Tank

Identification:	Item: Drum Vessel		Date: 04/14/2009
	Item No. T-602		
	Quantity: 1		

Function: To hold the water coming from the column reboilers after being used as steam to be transported to the boilers for steam re-generation

Operation: Continuous

Materials Handled:	Inlet	Bottom
Stream ID:	S-613	S-614
Quantity (lbmol/hr):	723.05	723.05
Composition (lbmol/hr):	Liquid	Liquid
Water	723.05	723.05
Temperature (in F):	228.55	228.55

Design Data:

Material:	304 Stainless Steel
Pressure:	21.70 psia
Diameter:	2' 11"
Height:	5' 10"
Hold-up time:	5 min

Comments:

Low Pressure Pump

Identification:	Item: Pump		Date: 04/14/2009
	Item No. P-602		
	Quantity: 1		
Function:	To pump the water from the Tank to the boiler		
Operation:	Continuous		

Materials Handled:	Inlet Stream	Outlet Stream
Stream ID:	S-614	S-615
Quantity (lbmol/hr):	723.05	723.05
Composition (lbmol/hr):	Liquid	Liquid
Water	723.05	723.05
Temperature (in F):	228.55	228.55
pressure (in psia) :	21.7	41.7

Design Data:

Type:	Centrifugal
Pressure Change:	20 psi
Electricity Required:	0.703 KW
Efficiency:	0.355

Comments:

Furnace

Identification:	Item: Furnace		Date: 04/14/2009
	Item No. FN-601		
	Quantity: 1		

Function:	To increase the temperature of the stream used to heat the boilers for steam generation
Operation:	Continuous

Materials Handled:	Inlet Stream	Outlet Stream
Stream ID:	S-600	S_601
Quantity (lbmol/hr):	1948.01	1948.01
Composition (lbmol/hr):	Vapor	Vapor
Methane	104.94	0.00
Nitrogen	1463.27	1463.27
Oxygen	379.81	169.92
Water	0.00	209.88
Temperature (in F):	89.77609	2345.618
pressure (in psia) :	500	500

Design Data:

Material of Construction:	304 Stainless Steel
Heat of combustion:	-36219100 Btu/hr

Comments:

CO_2 Compressor

Identification:	Item: Compressor		Date: 04/07/2009
	Item No. C-701		
	Quantity: 1		
Function:	To compress the warm CO_2 in the condensor cooling loop		
Operation:	Continuous		

Materials Handled:	Inlet Stream	Outlet Stream
Stream ID:	S-703	S-704
Quantity (lbmol/hr):	26000.00	26000.00
Composition (lbmol/hr):	Vapor	Vapor
Nitrogen	26000.00	26000.00
Temperature (in F):	47.1	205
pressure (in psia) :	100	295

Design Data:

Type:	Centrifugal
Net work required:	13595 hp
Isentropic Efficiency:	0.86

Comments:

CO2 Expander

Identification:	Item: Turbine		Date: 04/14/2009
	Item No. E-701		
	Quantity: 1		
Function:	To expand the cool, compressed CO3		
Operation:	Continuous		

Materials Handled:	Inlet Stream	Outlet Stream
Stream ID:	S-705	S-701
Quantity (lbmol/hr):	26000.00	26000.00
Composition (lbmol/hr):	Vapor	Vapor
Nitrogen	26000.00	26000.00
Temperature (in F):	90	-29
Pressure (in psia) :	295	100

Design Data:

Type:	Stainless Steel
Net work generated:	8411 hp
Isentropic Efficiency:	0.88

Comments:

Equipment Cost Summary

Table 12 on the following page shows all of the equipment to be used in the LNG Liquefaction Process. The first column shows the unit numbers, as referenced on the Process Flow Diagrams, with the Equipment description, Purchase Cost (in adjusted 2009 $), Bare Module Factor and Bare Module Cost used for each unit. The final column shows the source used for each cost estimation.

Total process machinery amounted to a bare module cost of $224,573,334, with an additional amount for the Ship's cost and needed start-up gases, for a total equipment cost of $399,573,334.

Table 12: Equipment Cost Summary

Item No.	Equipment	Cp (Adj)	Bare-Module Factor	C_{Bm}	Source
D-101	Scrub Column	$514,117	4.16	$2,138,729	Seider et al Correlation
H-101	Scrub col reboiler	$115,852	3.17	$367,252	Seider et al Correlation
F-101	Reflux flash drum	$141,567	3.05	$431,780	Seider et al Correlation
F-102	N2 rejection flash	$4,908	3.05	$14,969	Seider et al Correlation
HX-102	Intercooler	$116,474	3.17	$369,224	Seider et al Correlation
D-201	Fractionation column 1	$52,830	4.16	$219,773	Seider et al Correlation
H-201	Col 1 reboiler	$175,225	3.17	$555,464	Seider et al Correlation
HX-201	Col 1 condenser	$178,305	3.17	$565,225	Seider et al Correlation
A-201	Col 1 accumulator	$30,781	3.05	$93,883	Seider et al Correlation
P-201	Col 1 pump & motor	$42,533	2.51	$106,880	Seider et al Correlation
D-202	Fractionation column 2	$45,564	4.16	$189,546	Seider et al Correlation
H-202	Col 2 reboiler	$108,969	3.17	$345,433	Seider et al Correlation
HX-202	Col 2 condenser	$260,631	3.17	$826,201	Seider et al Correlation
A-202	Col 2 accumulator	$32,717	3.05	$99,788	Seider et al Correlation
P-202	Col 2 pump & motor	$30,943	2.55	$78,905	Seider et al Correlation
D-203	Fractionation column 3	$37,715	4.16	$156,892	Seider et al Correlation
H-203	Col 3 reboiler	$104,229	3.17	$330,407	Seider et al Correlation
HX-203	Col 3 condenser	$98,966	3.17	$313,724	Seider et al Correlation
A-203	Col 3 accumulator	$24,892	3.05	$75,919	Seider et al Correlation
P-203	Col 3 pump & motor	$13,169	2.72	$35,787	Seider et al Correlation
B1-B4	Water pumps (4)	$476,368	3.21	$1,530,830	Seider et al Correlation
C-401	N2 compressor	$20,984,883	3.21	$67,361,474	Couper et al Correlation
E-401	N2 expander	$4,000,000	3.21	$12,840,000	Seider Correlation/Atlas Copco Estimate
HX-101	Main Heat Exchanger (BAHX)	$5,500,000	3.00	$16,500,000	Applied UA Price Quote
C-101	Feed gas compressor	$2,080,294	3.21	$6,677,743	Seider et al Correlation
E-101	Feed gas expander	$459,481	3.21	$1,474,935	Seider et al Correlation
T-301	Fuel Gas Turbine	$43,309,502	2.00	$86,619,004	Alstom Comparable Turbine
B-601	Boiler 1	$40,128	3.17	$127,207	Seider et al Correlation
B-602	Boiler 2	$49,170	3.17	$155,869	Seider et al Correlation
T-601	Tank 1	$26,930	3.05	$82,136	Seider et al Correlation
T-602	Tank 2	$20,369	3.05	$62,124	Seider et al Correlation
P-601	Pump 1	$12,117	3.21	$38,887	Seider et al Correlation
P-602	Pump 2	$12,016	3.23	$38,802	Seider et al Correlation
FN-601	Furnace	$2,177,474	3.17	$6,902,594	Seider et al Correlation
C-701	CO2 Compressor	$4,251,719	3.21	$13,648,018	Couper et al Correlation
E-701	CO2 Expander	$953,560	3.21	$3,060,928	Seider et al Correlation
SHIP	LNG Ship		-	$175,000,000	SBM Gas estimate
TANK	Liquid Nitrogen Tank		-	$137,000	Assumed $0.10/L
	Total			***$399,573,334***	

Fixed-Capital Investment Summary

The fixed costs for the LNG project encompass equipment and ship costs (shown in the previous section), as well as contingency costs, contractor fees, and start-up costs. Table 13 and Table 14 show the Total Permanent Investment input assumptions and the Fixed-Capital Investment Summary, respectively, for the project.

Table 13: Total Permanent Investment Input Assumptions

Total Permanent Investment		
Cost of Site Preparations:	0%	of Total Bare Module Costs
Cost of Service Facilities:	0%	of Total Bare Module Costs
Costs of Contingencies and Contractor Fees:	25%	of Direct Permanent Investment*
Cost of Plant Start-Up:	10%	of Depreciable Capital*
		*excluding cost of ship

Table 14: Fixed-Capital Investment Summary

Fixed-Capital Investment Summary			
Bare Module Costs			
	Process Machinery	$ 224,573,334	
	LNG Ship	$ 175,000,000	
	Direct Permanent Investment		**$ 399,573,334**
Additional Depreciable Capital			
	Cost of Contingencies & Contractor Fees	$ 56,143,334	
	Total Depreciable Capital		**$ 455,716,668**
Additional Permanent Investment			
	Cost of Plant Start-Up	$ 28,071,667	
Total Capital Investment			***$ 483,788,334***

From above, it can be seen that the total process machinery cost was estimated at $224,573,334, and a total Direct Permanent Investment of $399,573,334 with the price of the LNG ship included. Since the process will be located in an offshore setting, no additional site preparation or service facility costs were assumed in addition to those accounted for in the price of the ship and the bare module factors of the individual equipment pieces.

As suggested by industry consultants, 25% of the Direct Permanent Investment, excluding the cost of the ship (aka the process machinery cost) was used as the estimate for the costs of contingencies and internal/external contractor fees. Plant start-up was estimated at 10% of depreciable capital, again excluding the cost of the ship.

This yields a Total Capital Investment of $483,788,334 to be incurred in 2010.

Operating Cost and Economic Analysis

Economic Summary

After careful economic analysis of the offshore liquefied natural gas process, the following have been identified as key factors affecting the potential profitability of the project:

- Energy prices in the future, especially natural gas
- The cost of the LNG ship
- Cost of key process equipment, namely the turbine and compressors

The assumptions made in this analysis attempt to quantify these uncertainties; however, more detailed analyses of these factors may be warranted before final investment is made. The base-case economic analysis of the LNG Liquefaction Process proved profitable at a cost of capital of 17%. Given the estimates, the Net Present Value (NPV) of the project was $37,275,885 with an internal rate of return (IRR) of 18.4% and third-year return on investment (ROI) of 7.3%.

Table 15 summarizes the revenues, variable costs, and fixed costs attributed to LNG in each year of operation in $/MMBTU (fractionation train revenues and expenses eliminated for this table, for “LNG-only” comparison). This *excludes* up-front capital cost considerations.

Table 15: Revenue, Variable Cost, Fixed Cost, and Margin Overview for LNG in $/MMBTU

Year	Revenue	Variable Cost	Fixed Cost	Margin
2011	$6.62	$0.662	$3.852	$2.102
2012	$6.75	$0.675	$2.568	$3.510
2013	$6.76	$0.676	$1.926	$4.155
2014	$6.82	$0.682	$1.984	$4.158
2015	$6.90	$0.690	$2.044	$4.171
2016	$7.02	$0.702	$2.105	$4.210
2017	$7.18	$0.718	$2.168	$4.292
2018	$7.38	$0.738	$2.233	$4.407
2019	$7.56	$0.756	$2.300	$4.507
2020	$7.43	$0.743	$2.369	$4.318
2021	$7.22	$0.722	$2.440	$4.057
2022	$7.29	$0.729	$2.513	$4.047
2023	$7.40	$0.740	$2.589	$4.073
2024	$7.77	$0.777	$2.666	$4.325
2025	$8.08	$0.808	$2.746	$4.521
2026	$8.38	$0.838	$2.829	$4.711
2027	$8.67	$0.867	$2.914	$4.893
2028	$8.92	$0.892	$3.001	$5.023
2029	$9.09	$0.909	$3.091	$5.089
2030	$9.25	$0.925	$3.184	$5.145

Table 16: General Information, Process Information, and Chronology

General Information							
	Process Title:	**Offshore Liquefied Natural Gas**					
	Product:	**LNG**					
	Plant Site Location:	**Qatar (offshore)**					
	Site Factor:	**1.00**					
	Operating Hours per Year:	**8497**					
	Operating Days Per Year:	**354**					
	Operating Factor:	**0.9700**					
Product Information							
This Process will Yield							
		10,118	lb-mol of LNG per hour				
		242,832	lb-mol of LNG per day				
		85,974,670	lb-mol of LNG per year				
	Price	**$ 5.50**	/mmBTU	or	**$ 2.10**	/lb-mol	
Chronology							
		Distribution of			**Production**		
Year	**Action**	**Permanent Investment**			**Capacity**		**Depreciation**
2009	Design	0%			0.0%		
2010	Construction	100%			0.0%		
2011	Production				45.0%		14.29%
2012	Production				67.5%		24.49%
2013	Production				90.0%		17.49%
2014	Production				90.0%		12.49%
2015	Production				90.0%		8.93%
2016	Production				90.0%		8.92%
2017	Production				90.0%		8.93%
2018	Production				90.0%		4.46%
2019	Production				90.0%		
2020	Production				90.0%		
2021	Production				90.0%		
2022	Production				90.0%		
2023	Production				90.0%		
2024	Production				90.0%		
2025	Production				90.0%		
2026	Production				90.0%		
2027	Production				90.0%		
2028	Production				90.0%		
2029	Production				90.0%		
2030	Production				90.0%		

Table 17: Sales, Costs, and Cash Flow Estimates for the Liquefied Natural Gas Process

Year	LNG Price ($/MMBTU)	Sales	Capital Costs	Var Costs	Fixed Costs	7-year MACRS	Depreciation	EBIT	Net Earnings	Cash Flow	ROI
2009											
2010			483,788,334							(483,788,334)	
2011	$6.62	105,717,849		10,571,785	59,398,657	14.29	65,121,912	(29,374,519)	(18,505,947)	46,615,965	-3.8%
2012	$6.75	161,996,420		16,199,642	59,398,657	24.49	111,605,012	(25,206,916)	(15,880,357)	95,724,655	-3.3%
2013	$6.76	216,668,497		21,666,850	59,398,657	17.49	79,704,845	55,898,127	35,215,820	114,920,665	7.3%
2014	$6.82	219,246,320		21,924,632	61,180,617	12.49	56,919,012	79,222,047	49,909,890	106,828,901	10.3%
2015	$6.90	222,187,164		22,218,716	63,016,035	8.93	40,695,498	96,256,905	60,641,850	101,337,349	12.5%
2016	$7.02	226,087,544		22,608,754	64,906,517	8.92	40,649,927	97,922,337	61,691,072	102,340,999	12.8%
2017	$7.18	231,437,203		23,143,720	66,853,712	8.93	40,695,498	100,744,263	63,468,886	104,164,384	13.1%
2018	$7.38	237,921,628		23,792,163	68,859,323	4.46	20,324,963	124,945,174	78,715,459	99,040,423	16.3%
2019	$7.56	244,015,405		24,401,541	70,925,103			148,688,761	93,673,920	93,673,920	19.4%
2020	$7.43	240,873,879		24,087,388	73,052,856			143,733,635	90,552,190	90,552,190	18.7%
2021	$7.22	235,479,427		23,547,943	75,244,442			136,687,043	86,112,837	86,112,837	17.8%
2022	$7.29	238,263,880		23,826,388	77,501,775			136,935,717	86,269,501	86,269,501	17.8%
2023	$7.40	242,350,068		24,235,007	79,826,828			138,288,233	87,121,587	87,121,587	18.0%
2024	$7.77	253,788,329		25,378,833	82,221,633			146,187,863	92,098,354	92,098,354	19.0%
2025	$8.08	263,566,694		26,356,669	84,688,282			152,521,743	96,088,698	96,088,698	19.9%
2026	$8.38	273,208,024		27,320,802	87,228,931			158,658,291	99,954,723	99,954,723	20.7%
2027	$8.67	282,747,729		28,274,773	89,845,799			164,627,157	103,715,109	103,715,109	21.4%
2028	$8.92	290,659,425		29,065,942	92,541,173			169,052,310	106,502,955	106,502,955	22.0%
2029	$9.09	296,663,535		29,666,354	95,317,408			171,679,774	108,158,258	108,158,258	22.4%
2030	$9.25	302,437,170	(35,000,000)	30,243,717	98,176,930			209,016,523	131,680,410	131,680,410	27.2%
									NPV	$37,275,885	
									IRR	18.4%	

Table 16 shows the General Information, Product Information, and Chronology of the Offshore Liquefied Natural Gas project. As mentioned previously, the project will be taking place off the coast of Qatar. The process will be continuous, with an operating factor of 0.9700 (suggested by Mr. Adam Brostow of Air Products), which equates to approximately 354 days or 8497 hours per year. At 100% capacity this process yields 85.97M lb-mol of LNG per year, at a 2009 price of $5.50/MMBTU (roughly $2.10 per lb-mol). Forecasting prices in future years will be discussed in a subsequent section.

The Chronology section shows that 2009 is reserved as the design year, with 2010 for construction and the point of capital investment. Production will start at a reduced scale in 2011 and will reach 90% of its normal approximated operating scale by 2013. Depreciation expenses taken in each year are shown in the last column. These types of LNG ships are usually designed to last for 40 years, but are dry-docked for repairs and maintenance after 20. Therefore, instead of trying to model the economics of the dry-dock process and needed maintenance (figures that are very uncertain this far in advance) all economic analysis will be done given a 20 year project cycle with a salvage value taken for the ship in 2030.

Table 17 on the previous page shows a calculation of cash flows for the LNG process, with a net present value (NPV) of and an internal rate of return (IRR) of 18.4%. All input variables and assumptions are discussed in subsequent pages.

Variable Costs

Table 18 below shows the Variable Cost assumptions used for this process.

Table 18: Variable Cost Input Assumptions

Variable Costs							
		Transfer Expenses:	8%	of Sales			
		Administrative Expenses:	2%	of Sales			

Since the energy prices used for revenue analysis reflect *at port* spot prices, 8% of sales was used to estimate the costs of bringing the LNG and other products from the offshore ship to port. Conversely, this could also be seen as the reduction in the LNG's value if this step of the process were outsourced to another company – either way, only 92% of the estimated energy prices will be recovered. An additional 2% was added for miscellaneous administrative expenses, so variable costs equate to 10% of total sales in each year, as reflected in the cash flow table.

Fixed Costs

Table 19, below, shows the Fixed Cost assumptions used for this process. For operations, it was assumed that three total operators were needed at all times, receiving an hourly wage of $30. In reality, this number could be higher or lower depending on the labor source. Typically, if an energy company were to bring employees over from the United State or Europe, higher wages will need to be offered as an incentive to work offshore. However, if local labor is employed, rates may be lower. Figure 16, below, shows a sensitivity analysis based on the labor rate used. Total wages were calculated by assuming three operators on-shift at all times at $30/hr rate, 24 hours a day for 365 days/year.

Table 19: Fixed Cost Input Assumptions

Fixed Costs			
	Operations		
	Operators per Shift:	3	(assuming 4 shifts)
	Direct Wages and Benefits:	$30	/hour
	Direct Salaries and Benefits:	15%	of Direct Wages and Benefits
	Technical Assistance/Engineering:	33.3%	of Direct Wages and Benefits
	Maintenance		
	Wages and Benefits:	4.50%	of Total Depreciable Capital
	Salaries and Benefits:	25%	of Maintenance Wages and Benefits
	Materials and Services:	100%	of Maintenance Wages and Benefits
	Maintenance Overhead:	5%	of Maintenance Wages and Benefits
	Operating Overhead		
	General Plant Overhead:	7.10%	of Maintenance Wages and Benefits
	Mechanical Department Services:	2.40%	of Maintenance Wages and Benefits
	Insurance		
	Insurance:	2%	of Total Depreciable Capital

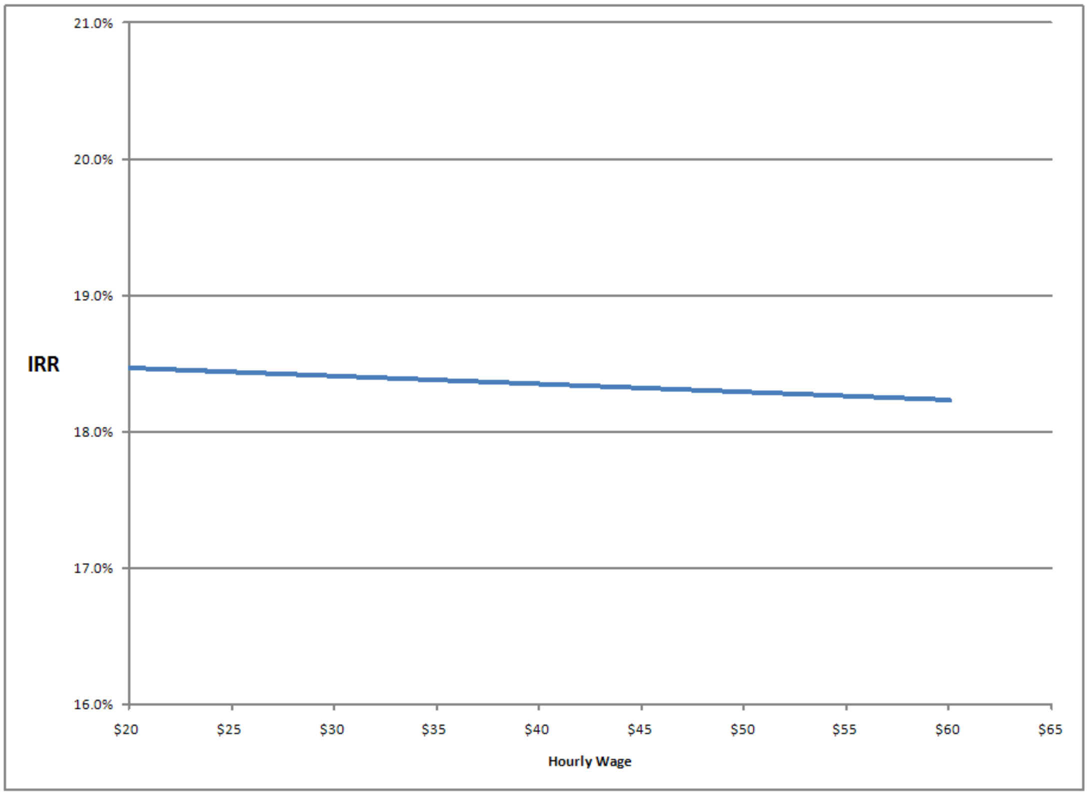

Figure 16: Sensitivity of the IRR to a Change in the Operator Hourly Wage

As can be seen, while higher wages obviously lower profitability, a doubling of the wage rate does not have a drastic effect on the internal rate of return.

Technical assistant and engineering personnel were estimated to be hired at a ratio of 1 engineer per 3 operators, so their total wages are approximately a third of the direct wages and benefits paid to the operators. Maintenance wages were assumed to be 4.5% of total depreciable capital, with an additional factor of 1.3 times this figure for maintenance salaries and benefits, materials and services, and overhead. 7.1% and 2.4% of maintenance wages were also added for general plant overhead and mechanical department services, respectively. Lastly, insurance was estimated at 2% of total depreciable capital.

Assumptions and Economic Uncertainties

Tax Rate – The tax rate for this process has been assumed to be 37%.

"No Carry-Over Losses" – All economic modeling was done based on the assumption that the parent company was a profitable entity – that is, accounting losses in the first two years of operation would be used to offset taxable earnings in other divisions of the company, and not carried over to reduce taxes in subsequent years.

Cost of Capital – For the purpose of discounting (the "time-value of money"), the cost of capital was assumed to be 17% (annually compounded). This number was estimated from the 2008 Annual Reports of BP and Royal Dutch Shell. However, considering that a company's cost of capital can be affected by its capital structure (debt vs. equity) and access to credit markets at the time of investment, sensitivity to this figure has been included in Figure 17. Here, the x-axis shows varying possible rates to use as the cost of capital, and their affects on the overall NPV. Notice that this graph's x-intercept is at 18.4%, the IRR of the project.

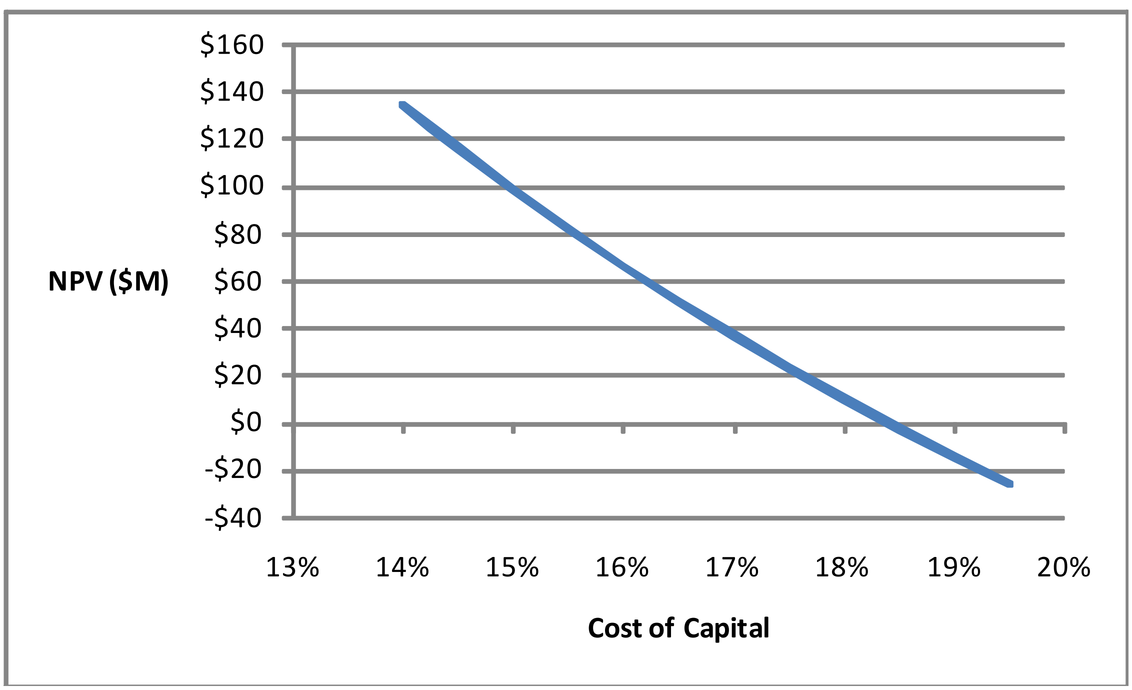

Figure 17: Net Present Value's Sensitivity to the Cost of Capital discounting rate

All cash flows have been discounted to 2010 as the base year, as this will be the time of initial capital investment.

Depreciation – Any equipment used in the collection or processing of natural gas falls within the 49.23 IRS Asset Class, and is thus subject to a 7 year modified accelerated cost recovery system (MACRS) for depreciation purposes. Because MACRS uses a half-year depreciation convention, depreciation expenses are actually taken year-end during the first 8 years of production (14.29, 24.49, 17.49, 12.49, 8.93, 8.92, 8.93, and 4.46% of total capital costs each year, respectively).[6]

Working Capital – It has been assumed that no working capital is required in this process. Because natural gas has no tangible "cost of goods," (that is, drilling fees are not a function of the amount of LNG onboard in inventory) and no additional raw materials need to be stored, the working capital requirement is not the same as for a traditional product or service.

Process Ship/Barge – For the purposes of this process, the physical ship on which the LNG process is being built has been estimated at a cost of $175M. This is by far the largest capital cost incurred in the process, and thus a large driver of the net present value (NPV) and internal rate of return (IRR) calculations. An article by SBM Gas suggests using $100,000/day as an approximated lease rate for a barge with a 1.4MMTON capacity.[7] Assuming annual lease payments, this equates to an equivalent (discounted at 17%) up-front capital cost of $215M. Because this process only produces 1MMTON, the "six-tenths" rule was applied to find a value of $175M. This figured seems reasonable, as Samsung Heavy Industries barges tend to run in the $200-300M range for double the capacity. Refer to Table 20 for sensitivity analysis on this assumption. The rows of this figure represent varying assumed costs for the process ship (values

[6] (Internal Revenue Service, 2007)
[7] (SBM Gas & Power, Kivi Niria, 2006)

shown on the left-hand side) while the columns represent percent deviation from the assumed energy prices (values up top). The matrix in the center shows the IRR for each ship-price cost pair, with those values greater than 17% (the assumed cost of capital, indicating a positive NPV) highlighted.

Table 20: Sensitivity of IRR to Changes in Energy Prices and Ship Price

		Change in energy Prices from assumptions										
		-25%	**-20%**	**-15%**	**-10%**	**-5%**	**0%**	**5%**	**10%**	**15%**	**20%**	**25%**
	$100	15.6%	17.1%	18.6%	20.0%	21.3%	22.7%	24.0%	25.2%	26.5%	27.7%	28.9%
	$125	14.2%	15.7%	17.1%	18.5%	19.8%	21.1%	22.4%	23.6%	24.8%	26.0%	27.2%
	$150	12.9%	14.4%	15.8%	17.2%	18.5%	19.7%	20.9%	22.1%	23.3%	24.5%	25.6%
	$175	11.7%	13.2%	14.6%	15.9%	17.2%	18.4%	19.6%	20.8%	21.9%	23.0%	24.1%
Price of	**$200**	10.6%	12.0%	13.4%	14.7%	16.0%	17.2%	18.4%	19.5%	20.6%	21.7%	22.8%
Ship ($M)	**$225**	9.5%	11.0%	12.3%	13.6%	14.9%	16.1%	17.2%	18.3%	19.4%	20.5%	21.5%
	$250	8.5%	9.9%	11.3%	12.6%	13.8%	15.0%	16.1%	17.2%	18.3%	19.3%	20.4%
	$275	7.5%	8.9%	10.3%	11.6%	12.8%	14.0%	15.1%	16.2%	17.2%	18.3%	19.3%
	$300	6.5%	8.0%	9.4%	10.6%	11.9%	13.0%	14.1%	15.2%	16.2%	17.3%	18.2%

Salvage Value – Ships for these processes typically have a lifetime of 40 years, but are dry-docked for repairs and maintenance after a 20 year span. Therefore, since it would be near impossible to project the specifics of these costs, a salvage value has been taken instead. This has been estimated as 20% ($35M) of the original cost of the ship. Because the ship has been fully depreciated by 2030, proceeds from the salvage value are subject to the normal tax of 37%. Because of the relatively large discount rate and the 20 year difference from the beginning of the project, the NPV of the project is relatively insensitive to the salvage value taken (Figure 18).

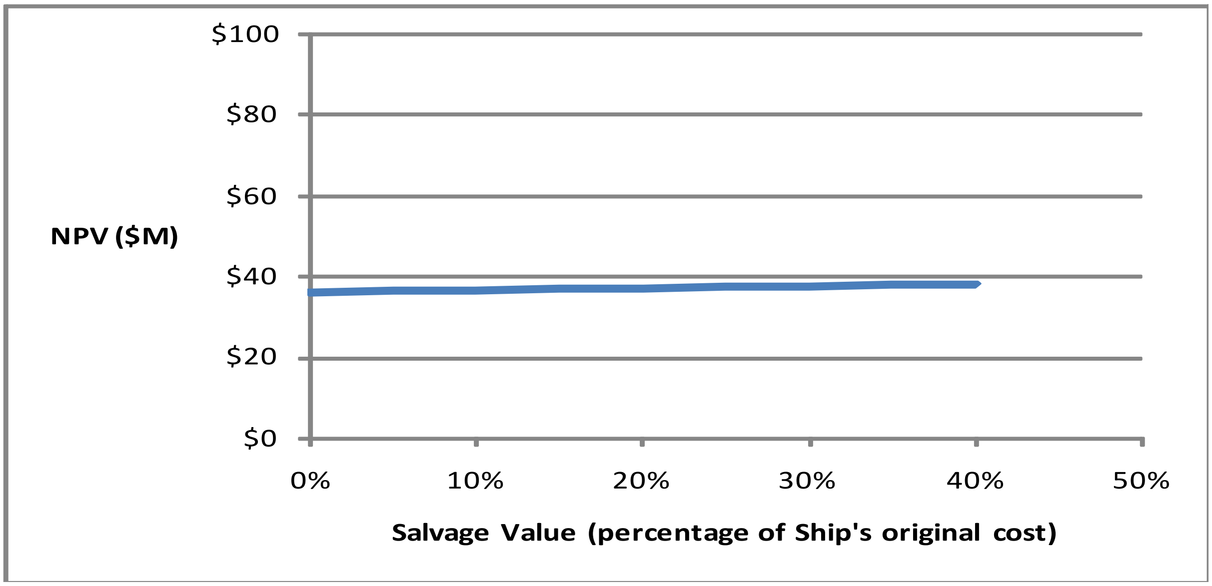

Figure 18: Sensitivity of NPV to Salvage Value

Energy Prices –Table 21: Energy Prices below shows the current energy prices for revenue purposes:

Table 21: Energy Prices

	Price	Unit
LNG	$5.50	MMBTU
Propane	$0.67	gallon
Ethane	$0.40	gallon
Butane +	$0.62	gallon

All figures were estimated from New York Mercantile Exchange futures contract data from late March, 2009.[8] Natural Gas price forecasts from the Energy Information Administration were used as the basis for revenue projections in the cash flow analysis, with assumed prices (in $/MMBTU) for each year shown in the second column of Table 17. These forecasts are based on many complex factors, such as projections of natural gas supply and demand, forecasts of oil supply and demand, volatility estimates, and inflation (more information on these forecasts can be found on the EIA's 2009 Annual Energy Outlook webpage). The price drop in 2020 is based

[8] (New York Mercantile Exchange)

on the anticipated opening of the new Alaskan Pipeline – an effective supply surge.[9] The presented estimates represent a compound annual growth rate (CAGR) of 1.78%, "overly-conservative" according to an interviewed equity associate from Sanford C. Bernstein's Oil & Gas Exploration and Production team.[10] Therefore, cash flows may, in reality, be higher based on LNG prices – Figure 19 shows the process' NPV as a function of the deviation from predicted energy prices. Again, the x-axis shows the average deviation from the assumed energy prices in the model, while the y-axis shows the new NPV. For instance, if prices drop by 15%, the new NPV is approximately -$50M.

For propane, ethane, and "butane+" (butane, pentane and hexane were lumped together for pricing purposes, as these higher hydrocarbons are often prices together as "NGL" – natural gas liquids – for futures quotes), the current spot prices, adjusted annually at 3% inflation, were used for revenue estimations.

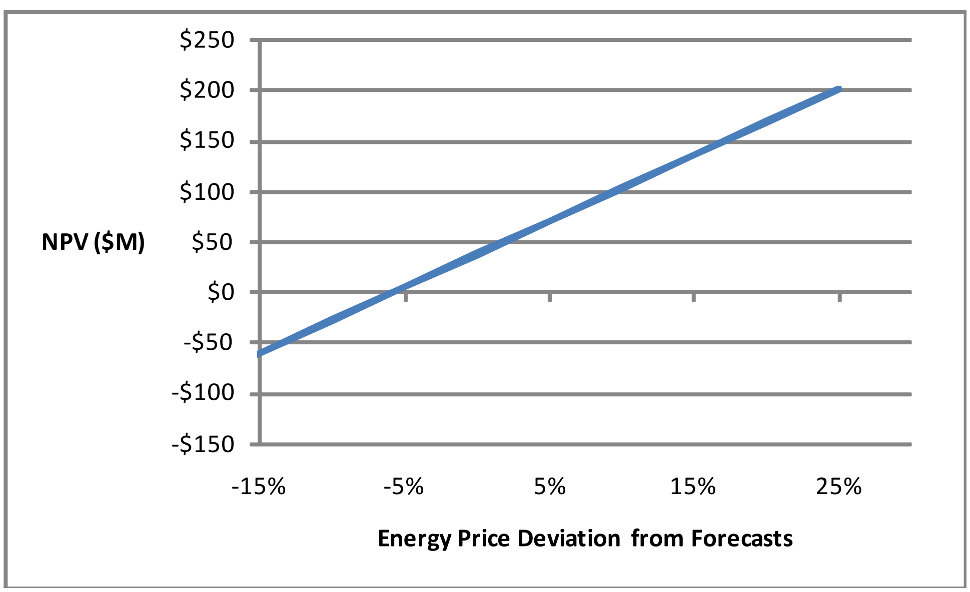

Figure 19: Sensitivity of NPV to Deviation of Energy Prices from Forecasts

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[9] (Energy Information Administration, 2009)
[10] (Lockshin, 2009)

Compressors – Aside from the LNG ship, the nitrogen cycle compressors (4 stage compression) were one of the highest capital costs, estimated at $67,361,474. Some uncertainty in this price and design exists, so an analysis of the NPV's sensitivity to this figure is included below (Figure 20). Because this is an up-front capital cost, the NPV is heavily influenced by this value.

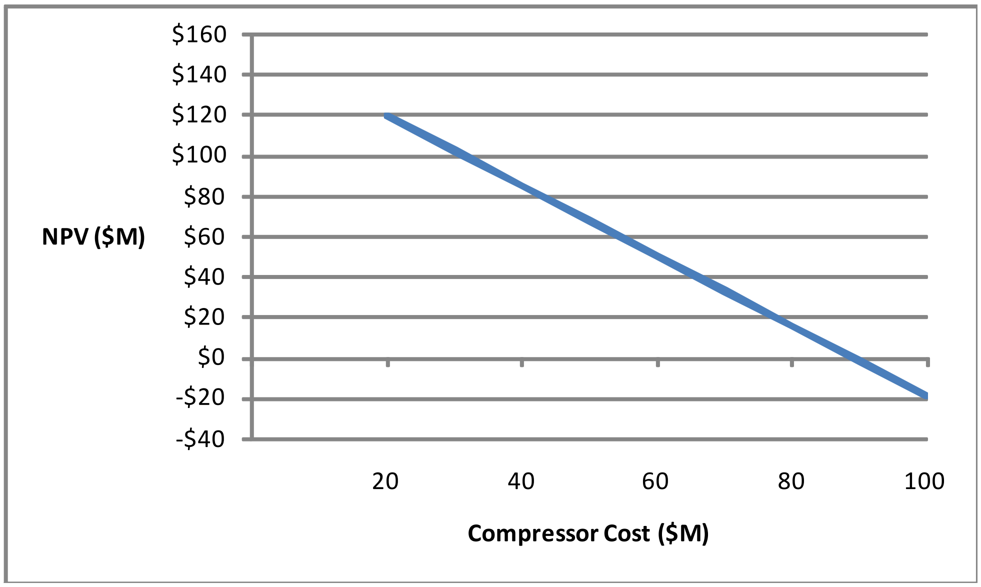

Figure 20: Sensitivity of NPV to Compressor Costs

Overall Sensitivity Analysis: Figure 21, on Page 135, shows a normalized sensitivity of each factor's affect, ceteris paribus, on the Net Present Value of the project. Each factor (Energy Prices, Nitrogen Compressor Price, Ship Cost, Ship Salvage Value, Wages, and Contingencies and Contractor Fees) has been scaled by *percent change*, or the percent difference from the assumed values mentioned in the report. All costs have *negative* slopes, signifying that an increase in a cost decreases the NPV, while effects on revenues have *positive* slopes. As can be seen – by the absolute value of the slopes – this process is most sensitive to the Energy Prices (revenue, represented by the triangles), followed by Ship Cost (X's), Compressor Price (squares), and Contingencies and Contractor Fees (circles).

By contrast, the process is relatively insensitive to the assumed operator wages and benefits (*, discussed in the following section) and Ship Salvage Value (diamonds). In the case of wages, their affect on costs is approximately an order-of-magnitude less than some of the capital expenditures. The ship's salvage value is realized in 2030, and is thus small when discounted to current value.

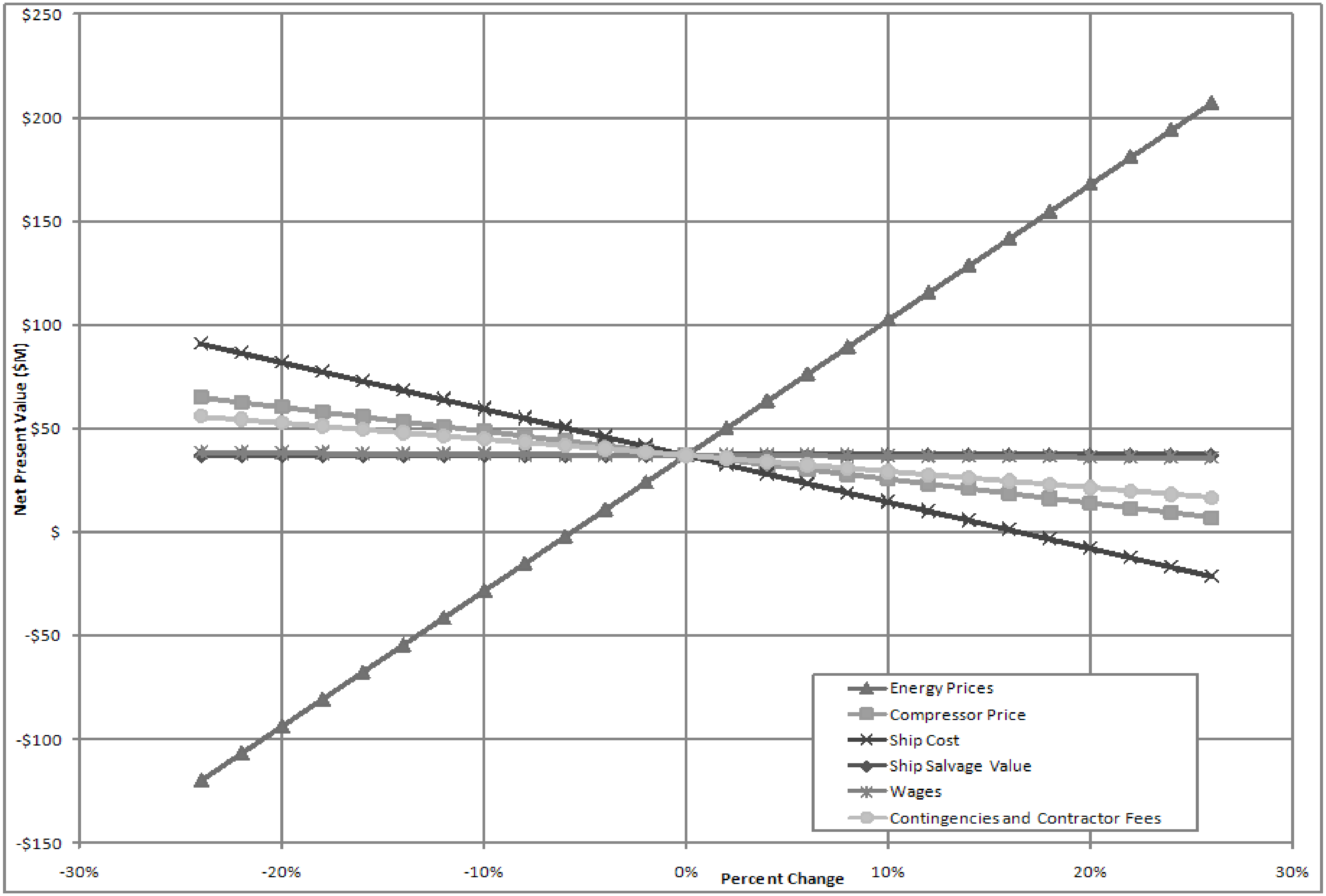

Figure 21: Multi-Variable Analysis of Sensitivity to NPV

Other Considerations

Eliminating the Fractionation Process

Table 22, on the following page, shows estimated cash flows with the fractionation train eliminated. Here, all process equipment required for capturing the ethane, propane, butane, and heavier hydrocarbons was eliminated, as were revenues for these products.

The net present value of this new process is approximately M; however, caution should be taken when comparing this with the previously-quote NPV, as the size of initial investment will be smaller. Instead, a comparison should be based on IRR -- 17.4% for the new process vs. 18.4% for the base-case. Therefore, the fractionation chain *should* be included. Also note the higher ROI figures on the right-most column in Table 17, (Page 123) compared to those in the new scenario.

Table 22: Cash Flow of the LNG Process without Fractionation Train

Year	LNG Price ($/MMBTU)	Sales	Capital Costs	Var Costs	Fixed Costs	7-year MACRS	Depreciation	EBIT	Net Earnings	Cash Flow	ROI
2009											
2010			454,980,703							(454,980,703)	
2011	$6.62	96,269,001		9,626,900	56,052,389	14.29	61,379,539	(30,789,840)	(19,397,600)	41,981,939	-4.3%
2012	$6.75	147,397,951		14,739,795	56,052,389	24.49	105,191,386	(28,585,643)	(18,008,955)	87,182,430	-4.0%
2013	$6.76	196,619,932		19,661,993	56,052,389	17.49	75,124,432	45,781,101	28,842,094	103,966,525	6.3%
2014	$6.82	198,596,299		19,859,630	57,733,961	12.49	53,648,036	67,354,660	42,433,436	96,081,472	9.3%
2015	$6.90	200,917,642		20,091,764	59,465,979	8.93	38,356,843	83,003,047	52,291,920	90,648,762	11.5%
2016	$7.02	204,179,936		20,417,994	61,249,959	8.92	38,313,890	84,198,085	53,044,794	91,358,683	11.7%
2017	$7.18	208,872,367		20,887,237	63,087,458	8.93	38,356,843	86,540,822	54,520,718	92,877,560	12.0%
2018	$7.38	214,679,847		21,467,985	64,980,081	4.46	19,156,945	109,074,831	68,717,144	87,874,089	15.1%
2019	$7.56	220,076,371		22,007,637	66,929,484			131,139,250	82,617,727	82,617,727	18.2%
2020	$7.43	216,216,674		21,621,667	68,937,368			125,657,638	79,164,312	79,164,312	17.4%
2021	$7.22	210,082,506		21,008,251	71,005,489			118,068,766	74,383,322	74,383,322	16.3%
2022	$7.29	212,105,050		21,210,505	73,135,654			117,758,891	74,188,102	74,188,102	16.3%
2023	$7.40	215,406,474		21,540,647	75,329,724			118,536,103	74,677,745	74,677,745	16.4%
2024	$7.77	226,036,427		22,603,643	77,589,615			125,843,169	79,281,196	79,281,196	17.4%
2025	$8.08	234,982,235		23,498,224	79,917,304			131,566,708	82,887,026	82,887,026	18.2%
2026	$8.38	243,766,031		24,376,603	82,314,823			137,074,605	86,357,001	86,357,001	19.0%
2027	$8.67	252,422,476		25,242,248	84,784,268			142,395,961	89,709,456	89,709,456	19.7%
2028	$8.92	259,424,415		25,942,441	87,327,796			146,154,178	92,077,132	92,077,132	20.2%
2029	$9.09	264,491,475		26,449,147	89,947,629			148,094,698	93,299,660	93,299,660	20.5%
2030	$9.25	269,299,948	(35,000,000)	26,929,995	92,646,058			184,723,895	116,376,054	116,376,054	25.6%
									NPV	$9,593,641	
									IRR	17.4%	

Doubling (2.0MMmtpa) and Halving (0.5MMmtpa) Capacity

The base-case process was also re-simulated and the economic analysis re-run using 2.0MMmtpa and 0.5MMmtpa feed gas flow rates. When the process was doubled, the project returned an internal rate of return of 24.7% ($344M NPV) with a third-year ROI of 14.4% -- a more appealing project than the base-case. When the capacity was doubled, the IRR dropped to 12.4% (-$71M NPV) with an ROI of 1.6% -- creating a loss. Modeled cash flows under these scenarios can be found in Table 23 and Table 24 on the following pages.

However, these results are somewhat expected. All of the revenues in this process, in the form of sellable LNG and fractionation products, are linearly dependent on the process flow rate. On the other hand, all equipment costs are modeled as increasing with capacity (in terms of flow rates, heat duty, power requirements, etc) to a power of less than 1. Therefore, the "net gain" of increasing the capacity is inherently positive.

Scale is therefore affected by factors somewhat outside the scope of this project – practical size limits, risk, and so forth. For instance, a ship can only be so big before assembly and transportation become too difficult. While economies of scale dictate that a larger plant is more efficient, having smaller facilities can be seen as "safer," both in a personal and economic sense. In the case of a mishap, less physical and environmental harm is likely with a smaller plant, and less total capacity is lost if a shutdown is required. Additionally, the natural gas deposits themselves may not be able to supply the increased flow rates for a sustainable amount of time.

Recent plant proposals and existing facilities with mixed refrigerants seem to regard 1.4 – 2.0MMmtpa as the ideal capacity range for this type of process. However, this may be another issue to consider before a final decision is made.

Table 23: Cash Flows when Plant Capacity Doubled

Year	LNG Price ($/MMBTU)	Sales	Capital Costs	Var Costs	Fixed Costs	7-year MACRS	Depreciation	EBIT	Net Earnings	Cash Flow	ROI
2009											
2010			758,281,842							(758,281,842)	
2011	$6.62	211,435,698		21,143,570	92,331,914	14.29	101,953,549	(3,993,349)	(2,515,810)	99,437,739	-0.3%
2012	$6.75	323,992,840		32,399,284	92,331,914	24.49	174,726,550	24,535,067	15,457,092	190,183,642	2.0%
2013	$6.76	433,336,993		43,333,699	92,331,914	17.49	124,784,294	172,887,068	108,918,853	233,703,147	14.4%
2014	$6.82	438,492,640		43,849,264	95,101,871	12.49	89,111,254	210,430,238	132,571,050	221,682,304	17.5%
2015	$6.90	444,374,328		44,437,433	97,954,927	8.93	63,712,050	238,269,910	150,110,043	213,822,093	19.8%
2016	$7.02	452,175,088		45,217,509	100,893,575	8.92	63,640,704	242,423,291	152,726,673	216,367,377	20.1%
2017	$7.18	462,874,406		46,287,441	103,920,383	8.93	63,712,050	248,954,524	156,841,350	220,553,400	20.7%
2018	$7.38	475,843,255		47,584,326	107,037,994	4.46	31,820,352	289,400,579	182,322,365	214,142,717	24.0%
2019	$7.56	488,030,810		48,803,081	110,249,134			328,978,595	207,256,515	207,256,515	27.3%
2020	$7.43	481,747,758		48,174,776	113,556,608			320,016,375	201,610,316	201,610,316	26.6%
2021	$7.22	470,958,854		47,095,885	116,963,306			306,899,663	193,346,788	193,346,788	25.5%
2022	$7.29	476,527,759		47,652,776	120,472,205			308,402,778	194,293,750	194,293,750	25.6%
2023	$7.40	484,700,136		48,470,014	124,086,371			312,143,751	196,650,563	196,650,563	25.9%
2024	$7.77	507,576,658		50,757,666	127,808,963			329,010,029	207,276,319	207,276,319	27.3%
2025	$8.08	527,133,388		52,713,339	131,643,231			342,776,818	215,949,395	215,949,395	28.5%
2026	$8.38	546,416,048		54,641,605	135,592,528			356,181,915	224,394,606	224,394,606	29.6%
2027	$8.67	565,495,458		56,549,546	139,660,304			369,285,608	232,649,933	232,649,933	30.7%
2028	$8.92	581,318,850		58,131,885	143,850,113			379,336,851	238,982,216	238,982,216	31.5%
2029	$9.09	593,327,070		59,332,707	148,165,617			385,828,747	243,072,110	243,072,110	32.1%
2030	$9.25	604,874,341	(53,050,080)	60,487,434	152,610,585			444,826,401	280,240,633	280,240,633	37.0%
									NPV	$344,467,128	
									IRR	24.7%	

Table 24: Cash Flows when Plant Capacity Halved

Year	LNG Price ($/MMBTU)	Sales	Capital Costs	Var Costs	Fixed Costs	7-year MACRS	Depreciation	EBIT	Net Earnings	Cash Flow	ROI
2009											
2010			310,640,551							(310,640,551)	
2011	$6.62	52,858,924		5,285,892	38,594,321	14.29	41,854,922	(32,876,226)	(20,712,022)	21,142,900	-6.7%
2012	$6.75	80,998,210		8,099,821	38,594,321	24.49	71,730,374	(37,426,331)	(23,578,588)	48,151,786	-7.6%
2013	$6.76	108,334,248		10,833,425	38,594,321	17.49	51,227,613	7,678,872	4,837,689	56,065,302	1.6%
2014	$6.82	109,623,160		10,962,316	39,752,151	12.49	36,582,784	22,325,897	14,065,315	50,648,099	4.5%
2015	$6.90	111,093,582		11,109,358	40,944,715	8.93	26,155,665	32,883,834	20,716,816	46,872,481	6.7%
2016	$7.02	113,043,772		11,304,377	42,173,057	8.92	26,126,376	33,439,953	21,067,171	47,193,546	6.8%
2017	$7.18	115,718,602		11,571,860	43,438,249	8.93	26,155,665	34,552,819	21,768,276	47,923,941	7.0%
2018	$7.38	118,960,814		11,896,081	44,741,396	4.46	13,063,188	49,260,144	31,033,891	44,097,079	10.0%
2019	$7.56	122,007,703		12,200,770	46,083,638			63,723,294	40,145,675	40,145,675	12.9%
2020	$7.43	120,436,940		12,043,694	47,466,147			60,927,098	38,384,072	38,384,072	12.4%
2021	$7.22	117,739,714		11,773,971	48,890,132			57,075,611	35,957,635	35,957,635	11.6%
2022	$7.29	119,131,940		11,913,194	50,356,836			56,861,910	35,823,003	35,823,003	11.5%
2023	$7.40	121,175,034		12,117,503	51,867,541			57,189,990	36,029,694	36,029,694	11.6%
2024	$7.77	126,894,164		12,689,416	53,423,567			60,781,181	38,292,144	38,292,144	12.3%
2025	$8.08	131,783,347		13,178,335	55,026,274			63,578,739	40,054,605	40,054,605	12.9%
2026	$8.38	136,604,012		13,660,401	56,677,062			66,266,549	41,747,926	41,747,926	13.4%
2027	$8.67	141,373,864		14,137,386	58,377,374			68,859,104	43,381,236	43,381,236	14.0%
2028	$8.92	145,329,712		14,532,971	60,128,695			70,668,046	44,520,869	44,520,869	14.3%
2029	$9.09	148,331,768		14,833,177	61,932,556			71,566,035	45,086,602	45,086,602	14.5%
2030	$9.25	151,218,585	(23,091,388)	15,121,859	63,790,533			95,397,582	60,100,477	60,100,477	19.3%
									NPV	($70,815,986)	
									IRR	12.4%	

Risks

The following have been identified as major economic uncertainties inherent in any liquefied natural gas process: (1) changes in macroeconomic climate/growth, (2) changes in energy prices, (3) restrictions on carbon emissions, (4) demand-side technological advances, (5) supply-side technological advances, and (6) geology & access. These are described below.

Macroeconomic climate/growth – Traditionally, energy consumption has been highly correlated with economic growth. Economic success, often measured using metrics such as gross domestic product (GDP) or gross national product (GNP), requires construction, production, transportation, and so forth – steps that inherently require energy. The emergence of significant global growth, such as China's economic boom of the late 1990s and early 2000s, could boost energy and natural gas demand, raising prices. However, global stagnation or even recession could conversely make the energy industry less attractive.

Energy Prices – Energy consumption is only half of the revenue battle – prices also play an important role. Forces outside an individual company's control, such as global supply, global demand, and future expectations, all play a key role in determining the price of oil, gas, and electricity prices. Unfortunately, due to the volatility of prices and the long-term scope of the project (20 years), no reliable projections of energy prices can be made that far in the future.

However, shorter-term risk can be mitigated via *hedging*. Financial instruments such as (short) natural gas futures, (short) forward contracts, and "put" options can be used to lock-in energy prices or reduce downside. When entering a "short" contract, the contract's issuer agrees to buy a specified quantity of the commodity at a specified priced and future date (or, put differently, the company agrees to *sell* the commodity at those terms). A "put" option awards money if the spot price of natural gas falls below a specified "strike price," in order to offset the decreased revenue from the lower price. These instruments can be traded on such exchanges as the New York Mercantile Exchange, Chicago Board of Trade, and European Energy Exchange. Some hedges are cash-settled (no products exchange hands, just the cash equivalent between the agreed rate and the spot price), while some require actual delivery of natural gas. Again, these types of derivatives cannot be used for longer-term energy fluctuations, but contracts can typically be found for time frames up to 18 – 24 months.

Carbon Emissions – Any changes in carbon emissions policies in the US or worldwide – from cap-and-trade restrictions to mandated reductions – have the potential to change demand for hydrocarbon-based energy sources. If carbon emissions become restricted or more expensive, oil and gas become less desirable, driving down prices.

Demand-side Technological Advances – Advances that reduce or eliminate hydrocarbon-based energy needs (hydrogen fuel cells, hybrid vehicles, more efficient natural gas home appliances, etc) have a negative impact on natural gas demand and prices. However, many advances are actually *substituting* natural gas for less-cleaner burning fuels. One such recent innovation is compressed natural gas (CNG) buses, now used in many public transit systems.[11] While natural gas is a cleaner fuel than some of its petroleum counterparts and can be seen as a short-term fix to reduced greenhouse effects, an ideal situation would be complete energy independence from non-renewable sources.

Supply-side Technological Advances – Supply affects energy prices to a similar degree as demand – in many cases; supply is a *greater* contributor, due to the relative inelasticity of energy demand. Usually changes in supply are affected by short-term “shocks.” These can take the form of natural disasters, disruptions on a gas pipeline, or a large production facility going off-line. However, supply *advances* can also play a role. If new processes are discovered to

[11] (Office of Energy Efficiency and Renewable Energy, DOE, 2000)

make natural gas production or transportation easier (such as LNG), there exists the potential for supply to be affected. Too much supply, or offerings at a low price, has the potential to create negative effects on revenue.

Geology and Access – While beyond the scope of this project, geological concerns also come into play. While gas fields are well-analyzed and mapped before a project is commissioned, some uncertainty exists in the actual size and lifetime of natural gas sites. Economic analysis has been done assuming 1 million metric tons of product per year; however, this figure is far from a certainty.

Process Extensions and Additional Considerations

Process Extensions

Carbon Dioxide Pre-Cooling of Feed and Nitrogen

Introduction

The major process extension that was considered for this project was the use of carbon dioxide as a pre-coolant. The re-compressed stream from the reflux separator (F-101) overhead, S-107, and the compressed nitrogen stream, S-116, would be pre-cooled by the carbon dioxide in a separate plate-fin heat exchanger. In theory, the pre-cooling would decrease the amount of nitrogen required for the main cooling cycle, because it would decrease the cooling load of the nitrogen at higher temperatures, enabling less nitrogen to be used in the cooling cycle.

Process Description

Figure 22 below shows the added equipment required if the carbon dioxide pre-cooling system were to be added to the system. Table 25 below gives specific information on the streams that are introduced in Figure 22. In the pre-cooling cycle, a closed-loop carbon dioxide loop is used to pre-cool the compressed overhead from the reflux separator (S-107) and the compressed nitrogen stream (S-116), before both are further cooled in the main heat exchanger (HX-101).

The carbon dioxide loop is similar to the nitrogen loop described above, except that the process creates liquid carbon dioxide through Joule-Thompson expansion. A cold, mixed liquid and vapor carbon dioxide stream (S-808) is fed into the cold end of the heat exchanger (HX-801) to provide the required cooling duty. It has a flow rate of 33,335 lbmol/hr and is at -41F and 140 psia. This stream cools both of the inlet streams and emerges from the exchanger as stream S-805 at 87F and 140 psia. It is then sent to the carbon dioxide compressor, C-801, where it is recompressed to 965 psia, at 410F and emerges as stream S-806. This stream is then cooled in

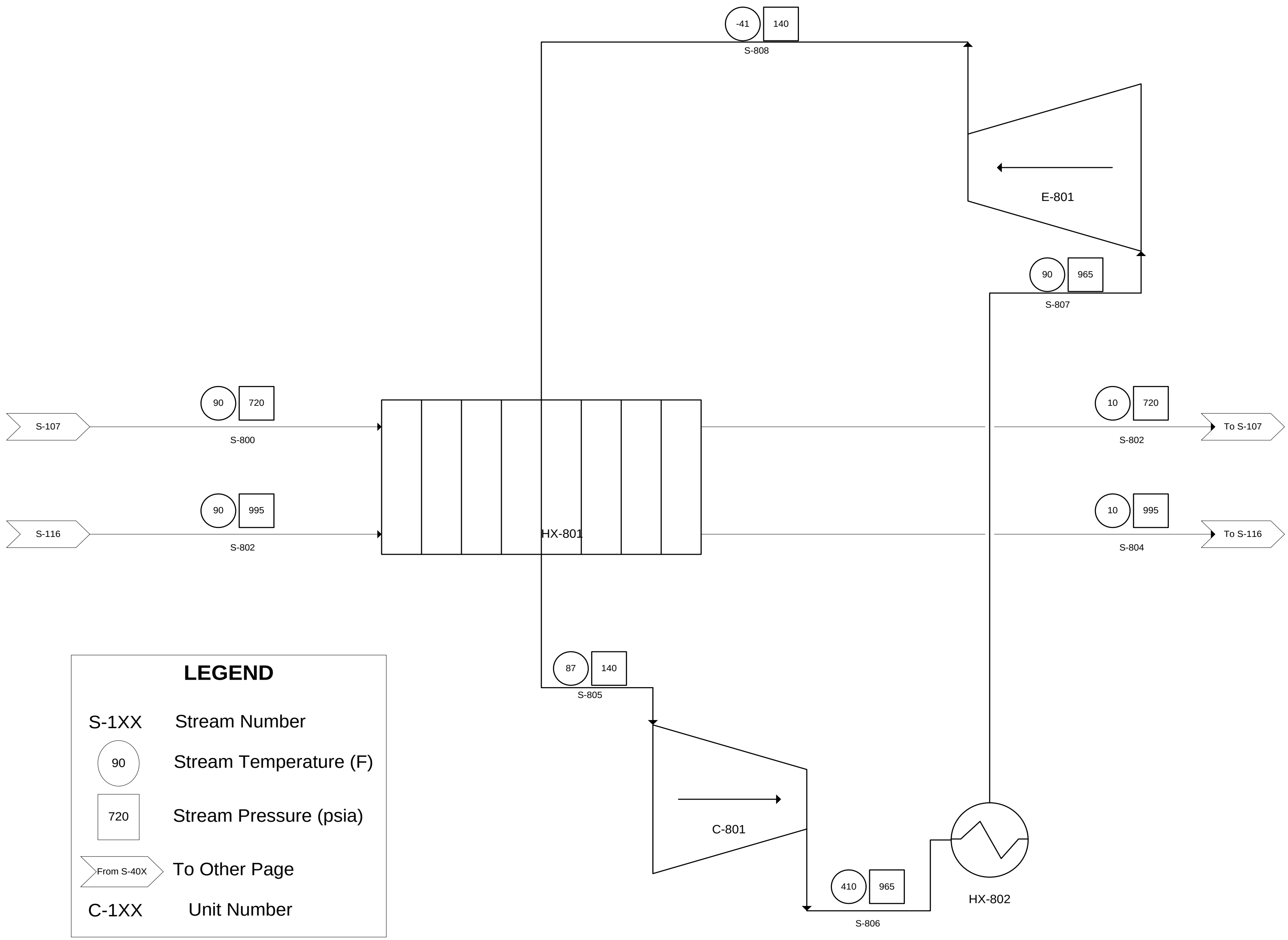

Figure 22: Additions to Process with CO_2 Pre-Cooling

Table 25: Additions to Process with CO_2 - Stream Table

	S-801	S-802	S-803	S-804	S-805	S-806	S-807	S-808
Temperature (F)	90	10	90	10	87	410	90	-41
Pressure (psia)	720	720	995	995	140	965	965	140
Mole Flow (lb-mol/hr)	12319.97	12319.97	69012.13	69012.13	33334.54	33334.54	33334.54	33334.54
Vapor Fraction	1	1	1	1	1	1	1	0.94
Enthalpy (Btu/hr)	-3.81E+08	-3.91E+08	-6.79E+06	-4.89E+07	-5.64E+09	-5.55E+09	-5.69E+09	-5.70E+09
Mole Flow (lb-mol/hr)								
Nitrogen	539.99	539.99	69012.13	69012.13	0	0	0	0
Carbon Dioxide	0	0	0	0	33334.54	33334.54	33334.54	33334.54
Methane	11734.24	11734.24	0	0	0	0	0	0
Ethane	45.43008	45.43008	0	0	0	0	0	0
Propane	0.307341	0.307341	0	0	0	0	0	0
n-Butane	8.51E-04	8.51E-04	0	0	0	0	0	0
Isobutane	3.11E-03	3.11E-03	0	0	0	0	0	0
Isopentane	9.26E-06	9.26E-06	0	0	0	0	0	0
n-Pentane	8.29E-06	8.29E-06	0	0	0	0	0	0
n-Hexane	2.05E-08	2.05E-08	0	0	0	0	0	0
Oxygen	0	0	0	0	0	0	0	0
Water	0	0	0	0	0	0	0	0

HX-802 using cooling water back to 90F. From there, the stream, S-807, is sent to the CO_2 expander, where it is expanded isentropically in a Joule-Thompson valve to 140 psia, causing some of the carbon dioxide to liquefy. This stream then becomes S-808, completing the cycle.

The carbon dioxide is used to cool the two incoming streams, S-801 and S-803, which correspond to S-107 and S-116 in Figure 4, above. S-801 is the compressed overhead from the reflux separator, and enters the carbon dioxide exchanger (HX-801) at 90F, 720 psia. It has the same composition as S-107 described above. It is cooled with the carbon dioxide to 10F at the same pressure (S-802), and then proceeds to the main heat exchanger (HX-101) to complete liquefaction.

Stream S-803 is the recompressed nitrogen stream. It enters the carbon dioxide exchanger at a flow rate of 69,012 lbmol/hr, at 90F and 995 psia. It is cooled using carbon dioxide to 10F at 995 psia (S-804), and proceeds from there to the main heat exchanger, where it will be cooled to -50F before being expanded to provide the bulk of the cooling power for the main exchanger.

Benefits and Drawbacks

The purpose of including carbon dioxide pre-cooling is to reduce the cooling power required in the main exchanger by taking off some of the load at the warm end of the cooling process. In this respect, the carbon dioxide pre-cooling succeeds admirably, as it reduces the overall heat duty for H-101 from 173,438,547 Btu/hr to 118,960,295 Btu/hr. This in turn reduces the size of the main exchanger, because less length is needed in each passage in order to get the proper cooling. The reduction in size comes with a corresponding reduction in purchase and installation costs for HX-101.

However, there is no corresponding decrease in the power required by the main nitrogen compressor. This is due to the fact that the limiting factor in the amount of nitrogen that can be used for cooling exists below the threshold of carbon dioxide cooling. In other words, while the carbon dioxide takes on the cooling load of the nitrogen at the warm end of the exchanger, the pinch point that dictates the minimum amount of nitrogen that can be used is at a much lower temperature, around -111F, which is not affected by the carbon dioxide cooling.

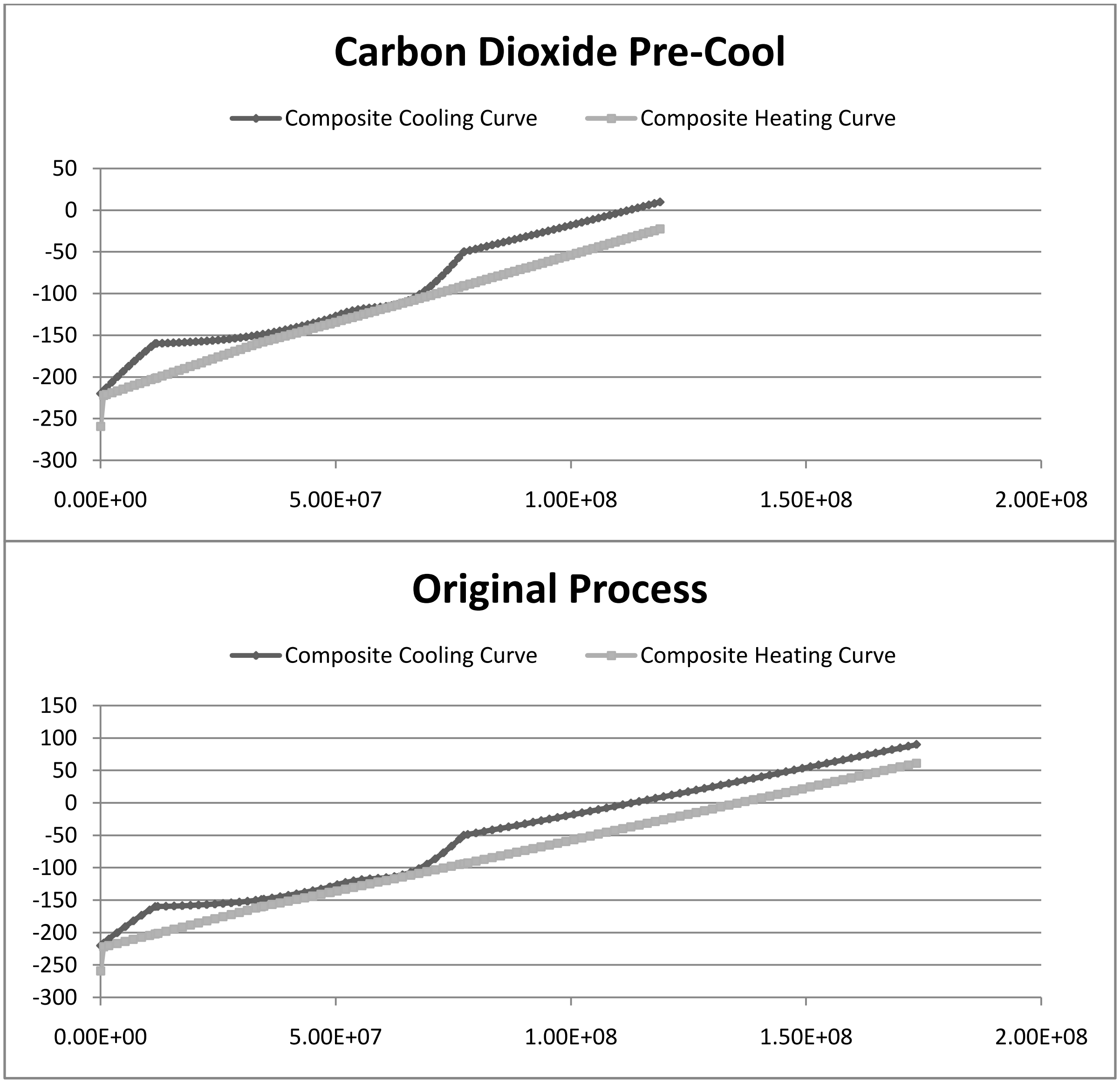

Figure 23: Combined Composite Heat Duty Curves, with and without Carbon Dioxide Pre-Cooling

Figure 23 above shows the combined composite heating and cooling curves for the main heat exchanger, HX-101, when pre-cooling is used (top graph), and when no pre-cooling is used (bottom graph). These graphs were generated using the process described in the Energy Balances section above; from ASPEN output reports of the main heat exchanger. The bottom graph of Figure 23 is the same as Figure 15 above.

The reduction in heat duty at the hot end of the main exchanger can be clearly seen when the graphs are compared. The top graph, the process with carbon dioxide pre-cooling, requires that the nitrogen stream only cool the input streams from 10F, rather than 90F, as is the case in the base case. The cooling down to 10F is provided by the addition of the carbon dioxide cooling exchanger (HX-801).

Close examination of Figure 23 shows why the introduction of carbon dioxide pre-cooling has no effect on the power requirements of the nitrogen compressor. Since the carbon dioxide can only be cooled to a certain point before it freezes, there is an effective limit on the cooling power of the carbon dioxide. As discussed above in the energy balance section, the minimum temperature approach inside the main heat exchanger dictates the minimum amount of nitrogen that can be used for the cooling. The minimum acceptable temperature approach for the exchanger used in this process is 3F, which occurs at -111F on the hot side and -114F on the cold side of the exchanger.

Since carbon dioxide freezes at approximately -70F, it is impossible to use carbon dioxide to offset some of the nitrogen required in that section of the exchanger. Therefore, regardless of whether or not carbon dioxide is used for pre-cooling, the required amount of nitrogen in the main exchanger remains the same. If the amount of nitrogen is decreased, cross-over occurs in the main heat exchanger.

Conclusions and Recommendations

Because the inclusion of a carbon dioxide pre-cooling loop does not decrease the amount of nitrogen in the system, its inclusion in the process is not recommended, as the power required by the nitrogen compressor remains the same. Additionally, the inclusion of a carbon dioxide loop would require the purchase of additional equipment, including another compressor train (C-801), a liquid-expander (E-801), and a second high-capacity heat exchanger, which would significantly increase equipment costs.

Furthermore, the power requirements of the system would increase, due to the need to power the carbon dioxide compressors. ASPEN simulations indicate that an additional 37,375 hp would be required to power the compressor, which increases the power requirements of the system to beyond what the initially installed fuel gas turbine (T-301) can supply. This would necessitate the purchase of a large fuel gas turbine system, as well as the need to increase the amount of methane being fed to the fuel gas turbine to supply the extra power. Cooling water requirements would also increase, as cooling water is required for the intercoolers (HX-802). The inclusion of the carbon dioxide pre-cooling loop if the existing base case remains unchanged is not recommended at this time.

Additional Considerations

Mercury Removal for the Main Heat Exchanger

The main heat exchanger in the process is a brazed aluminum plate-fin heat exchanger. Because the main heat exchanger is constructed of aluminum, liquid metal embrittlement caused by contact with mercury in one of the internal streams is a concern. Although mercury generally occurs in very low levels in natural gas reservoirs, it can accumulate over the operation of a plant and eventually achieve quantities that can result in the degradation and failure of the main heat exchanger.

Liquid metal embrittlement occurs when a corrosive liquid metal, such as mercury, find its way into small cracks in the aluminum body of the heat exchanger. The metal penetrates deep into the cracks, ensuring that the tip of the crack is always in contact with the mercury. This in turn creates more cracking, even at low stresses. Over time, the integrity of the aluminum heat exchanger may be compromised.[12]

Although there is currently no mercury present in the process specifications, it is recommended that future teams consider the possibility that trace amounts of mercury may accumulate in the exchanger over the course of operation. If the situation is deemed likely, then the system should be designed with a mercury removal process, such as a static guard bed, before the natural gas stream enters the main exchanger, in order to prevent degradation in the main heat exchanger.

Onsite Nitrogen Generation

The current process uses nitrogen in a closed loop as the cooling fluid to liquefy the natural gas. However, over time, nitrogen will be lost from the system due to imperfect seals and other system losses. In the base case design, nitrogen is purchased as a liquid from off-site, and shipped to the plant at necessary intervals, where it is stored in a specially designed nitrogen tank until it is required to replenish the system. The tank also contains enough space to store the nitrogen required for plant start-up.

Over time, however, the cost of purchasing and transporting nitrogen offshore to the ship may be prohibitive. The addition of an on-site nitrogen generation unit, while increasing initial capital costs, would eliminate the purchase and transport cost of the required nitrogen. On-site nitrogen generation units, such as those sold by Air Products Norway, are readily available, have a relatively small footprint, and could be powered by the extra power generated from the fuel gas turbine. These units use membrane-separation to produce nitrogen from intake air.[13] Future

[12] (Coade & Coldham, 2006)
[13] (Air Products Norway, 2008)

teams looking into the viability of this process should consider the purchase of a nitrogen generation system to help offset the recurring purchasing costs associated with ensuring that the system is adequately supplied with nitrogen.

Onsite Fresh Water Generation and Steam Co-generation

The current base case design includes a closed-loop steam system that supplies the heat necessary to power the reboilers for all of the process distillation columns. The steam is used to power the reboilers, and then the condensate is collected in a large vessel and split into two streams to be pumped to either 7 psig or 80 psig. These streams are then heated to their appropriate temperatures using the excess heat available from the exhaust of the fuel gas turbine.

Currently, the fresh water required to produce this steam will be purchased and stored aboard the ship, with enough in reserve to replenish the system as necessary. Similar to the situation described above with nitrogen, it may be beneficial to have an onboard source of fresh water, so that the shipping and purchase costs of replacement water can be eliminated. The system could also potentially be used to supply fresh water for the crew and other onboard applications. Reverse osmosis systems, such as the Vantage M86 Reverse Osmosis System sold by Siemens, are pre-engineered, pre-assembled, and have a compact footprint. They can be used to generate high-purity fresh water from readily available seawater.[14] Purchasing a system such as this would eliminate the need to purchase offsite water for steam generation.

Nitrogen Rejection Column

The current base case design includes a flash vessel, F-102, that is used to removed excess nitrogen from the final natural gas stream, S-108. This vessel removes most of the nitrogen in the incoming stream, leaving the final LNG product, S-109, with only 1% nitrogen

[14] (Siemens Water Technologies, 2008)

content. However, the use of a flash vessel causes a significant amount of methane to be taken off in the overhead as well. Even though this methane is eventually used to power the fuel gas turbine that provides the electrical power for the plant, more methane than is necessary for this power generation is provided.

Future teams may wish to consider the addition of a distillation column in place of F-102 in the process. The addition of such a column would allow for the system to be tweaked so that only the amount of methane required to power the fuel gas turbine is removed from the system, while maintaining the required nitrogen levels in the final product. This would allow for a larger percentage of the methane to be recovered in the final product, increasing the profitability of the process.

Important Considerations

Environmental Concerns

In any industry, one of the most important concerns that need to be taken into consideration is the environmental issues and problems that may arise from operation. In this design project, we are concerned with air pollution, minimizing waste and supporting marine life.

In this process design, the impact that the process will have on air pollution is minimal. However, in Qatar, some technologies have already been implemented to improve the quality of air. To maintain clean air, emissions are reduced using a common condensate VOC control system that is used to burn the vapors that result in the formation of smog.[15]To control pollution, smokeless flaring is used to burn off excess gas while reducing the pollution that flaring causes.[16]

To protect sea life, the amount of chlorine used to control sea organisms from clogging the seawater cooling system is reduced using the Pulse-Chlorination technology. Environmentally friendly silicon-based anti-fouling paints that have slippery surface characteristic can be used to protect the hull of the ship from adhesion by undesirable marine growth. The greatest and most dangerous risk to marine wildlife would be the result of leaks and spills which can be minimized by taking all the necessary precautions with regards to the equipment, maintenance and operation parameters. It is also extremely necessary to make sure that the pipeline used to extract and transport natural gas are in optimal conditions.

As mentioned earlier, another major environmental plus is the use of nitrogen as a refrigerant instead of hydrocarbons.

Safety Considerations

[15] (Environment, 2009)
[16] (Environment, 2009)

Even though LNG is non-toxic, asphyxiation can occur due to lack of oxygen in confined, unventilated areas. In the case of spills or leaks, the LNG will vaporize, creating a possible explosion. Therefore, strict industrial standards and safeguards must be followed, including regular inspection and maintenance of all piping, equipment, and storage tanks, along with appropriate process control technologies.[17]

Another safety concern that needs to be taken into consideration is the high pressure in the distillation columns. Here, pressure valves can be placed at the top of the columns to relieve the pressure in the columns in the case of unexpected pressure build-up where the valve can vent the natural gas into the air and it will vaporize. Control valves are to be placed in necessary areas in the process to constantly monitor the process and control unexpected deviations that may occur. Most refrigeration cycles in LNG processes use gas turbine-driven compressors to reach the necessary cryogenic temperatures. In these cases, the compression units must be efficient, robust and very reliable for environmental and safety reasons.

Two factors that allow for the maximum efficiency of a plant to be attained are the chemical construction of the plant and the plant maintenance.[18] Corrosion is one of the main causes of plant and equipment breakdown as a result the selection of the most suitable material of construction of the equipments, depending on the chemical it's processing, is very crucial.[19] The material of construction should be selected such that it is able to withstand unpredicted changes in conditions or chemical composition. Corrosion resistant materials such as stainless steel are chosen for the units that will be processing corrosive chemicals like hydrocarbons and sea water. This will make the equipments last longer and stay in good conditions so as to function properly and avoid potential operational problems that occur due to build up of

[17] (Importing LNG, 2005)
[18] (Chemical Plant Design, 2009)
[19] (Corrosion/selection of materials, 2006)

sediments, fouling or corrosion in the process units. Other selected material of construction is carbon steel that can be used with units operating on cooler chemicals such as fresh water. Highly corrosion resistant Inconel-600 will be used for the centrifugal pumps. All the equipments have to be of high-quality and purchased from reliable sources that have performed rigorous testing on the equipments.

Regular inspection and maintenance will be performed on the equipment to make sure that no corrosion results in leakage of chemicals during operation. The frequency of inspection of process units that handle hazardous material will be high initially until a history of its performance has been made such that its conditions can be predicted. Careful design of the pipelines is also very important since certain fluid velocities in pipes especially around bends can result in the early erosion of the pipes coating in these areas. The natural gas in the pipes will also be routinely sampled to make sure that the composition is correct and that the pipes are in good conditions inside and out. Preventative maintenance can also be scheduled to check on the valves and any build up of materials on surfaces that could result in the clogging of pipelines.[20]

Plant Start-Up

The plant start-up process is standard in every plant where the first step is to dry up the entire system. Dry nitrogen is used to purge all the equipments so that the system is dry before operation. With cryogenic process units, the surfaces of the units must be cooled slowly and gradually before contacting the cryogenic liquids. LNG is usually used in this case. Also, all the process equipments must be tested and checked before operation to make sure they are in top conditions. Refrigerants such as mixed cycle refrigerants may take hours before they reach a

[20] (The Transportation of Natural Gas, 2004)

stable state. However, in this project, the refrigerant is the nitrogen expander cycle which starts-up quickly in no more than an hour and can easily and rapidly shutdown in all conditions.

Conclusion and Recommendations

Based on the process design, economic, and safety analysis presented herein, the Offshore LNG Production project is hereby recommended as a feasible and profitable project, under the assumed economic conditions. The process has been successfully designed within the parameters required in the process specifications. All cooling loops are free of hydrocarbons, using only N_2 in the main loop and CO_2 in the condenser cooling cycle, with both the natural gas production rate and LNG purity within the specifications.

The Net Present Value of the project, including the fractionation train for recovery of higher hydrocarbons, was found to be $37M at an internal rate of return (IRR) of 18.4%. Further analysis of the assumptions made in these calculations may be required before final project approval is made due to the volatile nature of energy prices and related costs; however, estimates tend towards conservatism.

Other considerations potentially affecting the finances of the project have been presented, with no major deterrents to investment currently existing. All supplementary documentation is included in the following appendices, as referenced in the body of this report.

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Appendix I: Column Sizing Calculations

Sample Calculation for Distillation Column Size and Price:

(Yoko Kawashima, 2001)

D-201:

From the Aspen stream report:

G: 31137 lb/hr L: 17865 lb/hr

ρ_G: 1.5675 lb/ft^3 ρ_L : 30.5067 lb/ft^3

$$equation\ 1: U_f = \left[\frac{4d_p g}{3C_D}\right]^{1/2} \sqrt{\left[\frac{\rho_L - \rho_G}{\rho_G}\right]}$$

$$equation\ 2: C = \left[\frac{4d_p g}{3C_D}\right]^{1/2}$$

$$equation\ 3: C = C_{SB} F_{ST} F_F F_{HA}$$

$$equation\ 4: F_{st} = \left(\frac{\sigma}{20}\right)^{.2}$$

$$equation\ 5: F_{LG} = \left(\frac{L}{G}\right)\left(\frac{\rho_G}{\rho_L}\right)^{.5}$$

Solve for equation 5 from the given. F_{LG} = 0.1301. Find C_{SB} using the flooding correlation graph in Seider *et al.* C_{SB} vs. F_{LG}. C_{SB} = 0.15 ft/s. solve equation 4, σ = 10 dyne/cm (estimate), F_{ST} =0.87. F_F =1 for nonfoaming. F_{HA} =1, hole area for sieve trays. Next solve for equation 3 to find C. C = 0.131 ft/s. from equation 1, find U_f and solve for D_T from equation 6 where *f* = 0.85, and for .1<=F_{LG}<=1, $\left(\frac{A_d}{A_T}\right)$ = 0.1033.

$$equation\ 6: D_T = \left[\frac{4G}{(fU_f)\pi\left(1-\left(\frac{A_d}{A_T}\right)\right)\rho_G}\right]^{1/2}$$

D_T = 4.053 ft

Find height of column from the number of stages and column heuristics. L= 14.71 ft
P=200 psia and the Material of Construction is 304 Stainless Steel.

$$equation\ 7: P_d = \exp\{0.6068 + 0.91615[\ln(P)] + 0.0015655[\ln(P)]^2\}$$

$$equation\ 8: t_p = \frac{P_d D}{2SE - 1.2P_d}$$

$$equation\ 9: W = \pi(D + t_s)(L + 0.8D)t_s\rho$$

Solve for P_d from equation 7. P_d = 245.7 psia. Then solve for t_p where S=19000. In this case, t_p=t_S. Next, we can find the weight from equation 9. W= 505.5 lb

$$equation\ 10: C_V = \exp\{7.0132 + 0.18255[\ln(W)] + 0.02297[\ln(W)]^2\}$$

$$equation\ 11: C_{PL} = 361.8(\mathrm{D})^{0.73960}(\mathrm{L})^{0.70684}$$

$$equation\ 12: C_P = F_M C_V + C_{PL}$$

$$equation\ 13: C_T = N_T F_{NT} F_{TT} F_{TM} C_{BT}$$

$$equation\ 14: F_{NT} = \left(\frac{2.25}{1.0414^{N_T}}\right) for\ N_T < 20$$

$$equation\ 15: F_{TM} = 1.189 + 0.0577D\ for\ 304\ stainless\ steel$$

$$equation\ 15: C_{BT} = 468\exp(0.1739D)$$

Next, solve for C_V from equation 10. C_V =\$ 3056.3. From equation 11, find the added cost of platform and ladders. C_{PL} = \$ 6814. Find C_P where the bare-module factor is 1.7, C_P=\$12010 Next, solve for the cost of trays from equation 13. N_T = number of trays.
$F_{TT} = 1$ for sieve trays. And solve for the other parameters using equations 13, 14 and 15. C_T = \$23139. Total C_p= C_P + C_T = \$35149.

Appendix II: Reboiler, Condenser, and Heater Calculations

Sample Calculation for a Reboiler (H-201):

$$equation\ 16: A = \frac{Q}{U\Delta T_{\mathrm{LM}}}$$

$$equation\ 17: C_B = \exp\{11.967 - 0.8709[\ln(A)] + 0.09005[\ln(A)]^2\}$$

$$equation\ 18: C_P = F_P F_M F_L C_B$$

$$equation\ 19: F_P = 0.9803 + 0.018\left(\frac{P}{100}\right) + \ 0.0017\left(\frac{P}{100}\right)^2$$

From the Aspen report, the heat duty is given. Guess U and find the LMTD from the four stream of the reboiler. Using equation 16, find the area. A=1734 ft^2. Solve for C_B from equation 17. C_B = $35620. Next find C_P using equation 18 where $F_M = 1.75 + \left(\frac{A}{100}\right)^{0.13} for\ stainless\ steel$ F_L=1 and equation 19 for P=200. C_P = $116580

The calculations for a **condenser** and a **heater** are the same as the reboiler, the only difference is that:

$$equation\ 20(a): C_B = \exp\{11.667 - 0.8709[\ln(A)] + 0.09005[\ln(A)]^2\}$$

$$equation\ 20(b): C_B = \exp\{11.0545 - 0.9228[\ln(A)] + 0.09861[\ln(A)]^2\}$$

Where equation 20(a) is for the condenser and equation 20(b) is for the heater.

Appendix III: Pump and Motor Calculations

Sample Calculations for a Pump and its Motor:

Pump (P-201):

From the Aspen report, the pressure drop, density and flowrate, Q, are given. Find the pump head of the fluid flowing (which is pressure rise over liquid density) and solve for S in equation 21. Next, solve for C_B and then C_P where F_T= 1 in this case and F_M= 1.35 for the material of construction chosen. Therefore C_P= $8929.

$$equation\ 21: S = Q(H)^{0.5}$$

$$equation\ 22: C_B = \exp\{9.2951 - 0.6019[\ln(S)] + 0.0519[\ln(S)]^2\}$$

$$equation\ 23: C_P = F_T F_M C_B$$

Pump motor:

Using the flowrate, Q, from the Aspen report, find η_P using equation 24. Next, find P_B using equation 25 and then find η_M from equation 26. After that, we can solve for P_C and C_B from equations 27 and 28 and finally we can find C_P where F_T=1. C_P = $19369.

$$equation\ 24: \eta_P = -0.316 + 0.24015[\ln(Q)] - 0.01199[\ln(Q\}]^2$$

$$equation\ 25: P_B = \frac{QH\rho}{33000\eta_P}$$

$$equation\ 26: \eta_M = 0.8 + 0.0319[\ln(P_B)] - 0.00182[\ln(P_B)]^2$$

$$equation\ 27: P_C = P_B/\eta_M$$

$$equation\ 28: \\ = \exp\{5.4866 + 0.13141[\ln(P_C)] + 0.053255[\ln(P_C)]^2 + 0.028628[\ln(P_C)]^3 \\ - 0.0035549[\ln(P_C)]^4\}$$

$$equation\ 29: C_P = F_T C_B$$

Total C_P of pump ad motor = $28298.

Appendix IV: Flash Vessel Calculations

Sample Calculation for a Flash Drum (A-201):

L/D = 2, hold-up time =3 min. The mass flowrate, density and vapor fraction were provided by the Aspen report. Find the velocity and solve for the diameter using equation 30. D=2.84 ft, L=5.7 ft. Next find P_d, t_p and W using equations 7, 8 and 9. Then solve for C_V using equation 31 and find C_{PL} using equation 32 and finally use equation 12 to find C_P. C_P =$20479

$$equation\ 30: D = \left[\frac{4\nu\tau}{\pi}\right]^{1/3}$$

$$equation\ 31: C_V = \exp\{8.9552 - 0.2330[\ln(W)] + 0.04333[\ln(W)]^2\}$$

$$equation\ 32: C_{PL} = 2005(D)^{0.20294}$$

Appendix V: Compressor and Expander Sample Calculations

Feed Expander
Unit: E-101

Assumptions:
Carbon Steel Material
Pressure Discharge
20-5,000 HP

Feed Expander

C(p) Adj	$	459,481.41	=420*(HP)^0.81*CE/394

592.2 2009 CE Index
394 2000 CE Index

3415 Power (HP)

3.21 Bare Module Factor

Bare Module Cost
$ 1,474,935

Compressor Costing
Unit: C-401(a-d)

	HP	Purchase Cost
(a)	17,868	$3,419,119
(b)	19,809	$3,644,869
(c)	19,433	$3,601,818
(d)	19,237	$3,579,252
	Total	$14,245,057

Purchase Cost = 7900*HP^(0.62)

Correlation From:
Chemical Process Equipment: Selection and Design
James R. Couper, W. Roy Penney, James R. Fair, Stanley M. Walas
2nd Edition, Gulf Professional Publishing, 2004

Valid up to 30,000 HP
Figure in 2003 dollars
(No material specified)

Purchase Cost	$14,245,057	
Bare Module Factor	3.21	Seider et al
2009 CE Index	592.2	
2003 CE Index	402	
Bare Module Cost	$67,361,474	

Appendix VI: Furnace Sample Calculations

Sample Calculation for Furnace (FN-601):

$$equation\ 33: C_B = \exp\{0.32325 + 0.766[\ln(Q)]\}$$

$$equation\ 34: C_P = F_P F_M C_B$$

$$equation\ 35: F_P = 0.986 - 0.0035\left(\frac{P}{100}\right) + \ 0.0175\left(\frac{P}{100}\right)^2$$

The material factor for the furnace is 1.7 since stainless steel is used. The Q is provided from the Aspen files to solve for equation 33. Using equation 33 and 35, the purchase cost is found from equation 34.

Appendix VII: HX-101 Confirmation Sizing Calculations and Contacting Streams Diagram

HX-101 Sizing Confirmation Calculations

The brazed aluminum plate-fin heat exchanger that will be used in the process was sized by a consultant at Applied UA. The equations here reproduce one of the calculations in order to give a general idea of how they were produced. The nitrogen rejection stream in zones 3 & 4 was the stream that was chosen to be reproduced. Zones 3 & 4 cover the temperature range of -256.3F to -135.2F for the cold streams and -132.1F to -220F for the hot streams.

Sizing was done using the method outlined in Stewart-Warner's Product Information brochure[21]. Some important assumptions were required before sizing began. First, confirmation sizing was done using lanced fins (15 fins/inch, 0.008" thickness, 0.375" height), because these were the fins that most closely resembled those used in the actual sizing. The nitrogen rejection stream used 6/8 serrated fins (15 fins/inch, 0.0079" thickness, 0.3791" height), so the results can be expected to differ, perhaps significantly, depending on the effectiveness of serrated fins vs. that of lanced fins.

STEP 1: Calculate Stream Reynold's Number

$$\mathrm{Re} = \frac{4r_h G}{\mu} = \frac{4 \times .002107\ ft \times 12113.1 \dfrac{lb}{ft^2 \cdot hr}}{.0145145\ \dfrac{lb}{ft \cdot hr}} = 7033.6$$

Here, r_h is the hydraulic radius, G is the mass velocity of the stream, and μ is the kinematic viscosity at the midpoint temperature.

$$G = \frac{\dot{m}}{A_c} = \frac{9972.75\ lb/hr}{0.8233\ ft^2} = 12113.1 \frac{lb}{hr \cdot ft^2}$$

Here, m is the mass flow rate, and A_c is the free stream area.

$$A_c = A_c^{'} N W_e = 0.002242 \times 8 \times 45.9 = 0.8233\, ft^2$$

Here, A_c' is the free stream area factor, N is the number of passages per core, and W_e is the assumed width of the passage.

STEP 2: Use Correlation Graph to Find Value for 'j'

After calculating the Re, Figure 2 in the Stewart-Warner brochure is used to determine the value of j. The value of j as determined from reading the graph is 0.008.

STEP 3: Solve for the Heat Transfer Coefficient

---

[21] (Stewart Warner)

$$\Pr = \frac{Cp \cdot \mu}{k} = \frac{0.428 \dfrac{Btu}{lb \cdot F} \times 0.0145145 \dfrac{lb}{ft^2 \cdot hr}}{0.009 \dfrac{Btu}{ft^2 \cdot hr \cdot F}} = 0.69025$$

Here, Cp is the constant pressure heat capacity, and k is the thermal conductivity of the aluminum.

$$h = \frac{jGCp}{\Pr^{2/3}} = \frac{0.008 \cdot 12113.1 \dfrac{lb}{hr \cdot ft^2} \cdot 0.428 \dfrac{Btu}{lb \cdot F}}{0.6903^{2/3}} = 53.1 \frac{Btu}{hr \cdot ft^2 \cdot F}$$

STEP 4: Determine the Weighted Log-Mean Temperature Difference

The weighted log mean temperature difference takes into account the actual heat duties of each section of the exchanger. The exchanger is split into small pieces, and the log-mean temperature difference is calculated for each section. Then, the total heat duty for the exchanger is divided by the sum of the other heat duties divided by their log-mean temperature differences, yielding the weighted log-mean temperature difference.

$$WTDLMTD = \frac{Q_T}{\sum \dfrac{Q_i}{LMTD_i}} = 20.71F$$

STEP 5: Determine the Required Area

$$A = \frac{Q}{h \times WTDLMTD} = \frac{2160000 \dfrac{Btu}{hr}}{53.1 \dfrac{Btu}{hr \cdot ft^2 \cdot F} \times 20.71F} = 1964.17\, ft^2$$

This area is per core, so multiply by the number of cores (4) to get the total heat transfer area, 7856.7 ft^2. Given that different correlations were used, this compares very favorably to the consultant's calculated area of 7238 ft^2, as it is within 10%.

Contacting Streams Diagram

Because HX-101 is a heat exchanger with multiple hot and cold streams, it can be difficult to understand exactly which streams are contacting which other streams in various parts of the exchanger. The following diagram, Figure 24, provides a general overview of where the individual streams are contacting one another.

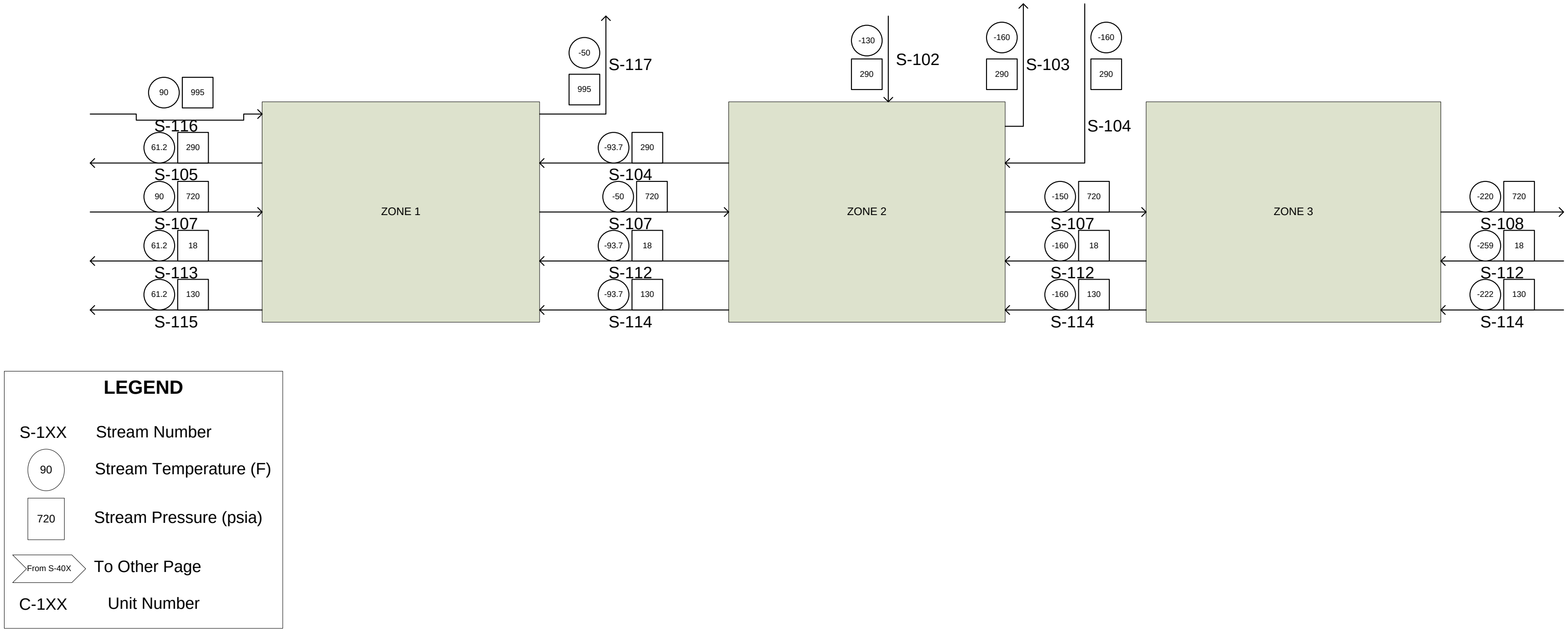


Figure 24: HX-101 Contacting Streams Diagram

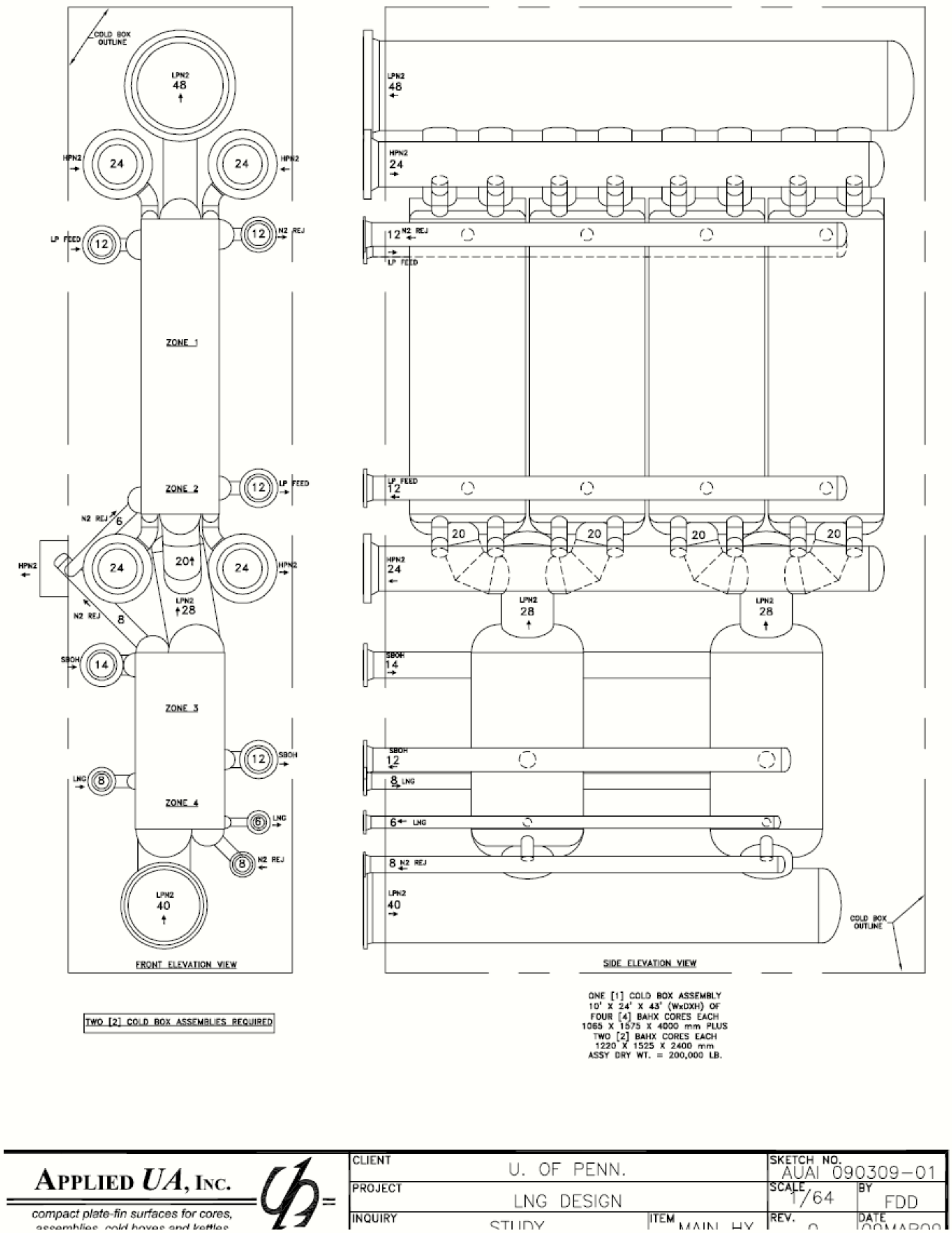
The figure above shows all of the streams entering and exiting the main heat exchanger, HX-101. The right side of the diagram is the cold end of the exchanger, and the left side of the diagram is the warm end. Streams are labeled as they were in the unit description for the main heat exchanger.

Streams S-112 and S-114 enter at the far right side of the diagram. These streams are not at equal temperatures when they enter, but they are assumed to quickly equilibrate, rendering them an equal temperature after an insignificant portion of the main heat exchanger. This is an assumption that will be used for the rest of this discussion. All of the hot streams at the endpoint of a given zone are assumed to be the same temperature, as are all cold streams at the same endpoint.

Streams S-112 and S-114 cool stream S-107 to its final temperature, -220F in zone 3. When they exit zone 3, S-112 and S-114 are at -160F, and the entering stream, S-107, is at -150F. At the beginning of zone 2, stream S-104 is added to the cold end of the exchanger. It enters at -160F. An additional hot stream is also added in this zone, S-102. S-102 enters at -130F and is cooled to -160F by contacting S-104, S-112, and S-114.

In zone 1, S-116 is cooled from 90F by contacting S-104, S-112, and S-114. Stream S-107 is cooled by all three streams as well. Stream S-104 exits as S-105 at 68.2F, as do S-112 and S-114. S-116 is removed at -50F.

Appendix VIII: BAHX Specification Sheet from Applied UA



COLD BOX OUTLINE
LPN2 48
HPN2 24
LP FEED 12
N2 REJ 12
ZONE 1
ZONE 2
N2 REJ 6
20
LPN2 28
N2 REJ 8
SBOH 14
ZONE 3
SBOH 12
LNG 8
ZONE 4
LNG 6
LPN2 40
FRONT ELEVATION VIEW
SIDE ELEVATION VIEW
TWO [2] COLD BOX ASSEMBLIES REQUIRED
ONE [1] COLD BOX ASSEMBLY
10' X 24' X 43' (WxDXH) OF
FOUR [4] BAHX CORES EACH
1065 X 1575 X 4000 mm PLUS
TWO [2] BAHX CORES EACH
1220 X 1525 X 2400 mm
ASSY DRY WT. = 200,000 LB.
APPLIED UA, INC.
compact plate-fin surfaces for cores,
CLIENT
U. OF PENN.
PROJECT
LNG DESIGN
INQUIRY
ITEM
SKETCH NO.
AUAI 090309-01
SCALE
1/64
BY
FDD
REV.
DATE

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fives cryogenie

HEAT EXCHANGER SPECIFICATION SHEET N° 5390-1

CUSTOMER: U. OF PENN	PROJECT: LNG DESIGN	N.C. ORDER N°: FOR QUOTATION
ITEM N°: ZONES 1 & 2	LOCATION: STUDY	CUST. JOB N°:
CASE: DESIGN	PLANT SERVICE: MAIN HX	CONSTRUCTION CODE: ASME VIII, DIV. 1

01	FLUID		A. HPN2	B. LP FEED	C. LPN2	D. N2 REJ
02	TOTAL FLOWRATE	lb/hr	2,458,183	253,718	2,458,183	39,891
03	VAPOR FLOWRATE IN	lb/hr	2,458,183	253,718	2,458,183	39,891
04	VAPOR FLOWRATE OUT	lb/hr	2,458,183	229,485	2,458,183	39,891
05	- MOL. WGT. IN/OUT		28.013 / 28.013	18.794 / 17.712	28.013 / 28.013	18.437 / 18.437
06	LIQUID FLOWRATE IN	lb/hr	0	0		
07	LIQUID FLOWRATE OUT	lb/hr	0	24,233	0	0
08	- MOL. WGT. IN/OUT		--- / ---	--- / 44.587	--- / ---	--- / ---
09	TEMPERATURE IN	°F	90.0	45.1	-135.2	-135.2
10	TEMPERATURE OUT	°F	-85.0	-50.0	85.95	85.95
11	DEW POINT / BUBBLE POINT					
12	OPERATING PRESSURE	psig	1005.0	400.0	166.1	18.0
13	ALLOWABLE PRESSURE DROP	psi	10.0	4.0	3.0	2.0
14	TOTAL HEAT TRANSFERRED	MMBTU/hr	127.390	16.487	140.101	3.776
15	CORRECTED MTD (GLOBAL)	°F	11.55 / 44	11.552	44 / 11.55	44 / 11.55
16	FOULING FACTOR	°F-hr-ft²/Btu	NIL	NIL	NIL	NIL
17	DESIGN TEMPERATURE	°F	150 / -275	150 / -275	150 / -275	150 / -275
18	DESIGN PRESSURE	psig	1,180	500	200	50
19	TEST PRESSURE HYDRO/PNEU	psig	per ASME	per ASME	per ASME	per ASME
20	NUMBER OF ASSEMBLIES : TWO [2]		WIDTH : **1,065** mm		TYPE OF HEAT EXCHANGER : COUNTERFLOW	
21	NR OF CORES/ASSEMBLY : FOUR [4]		HEIGHT : **1,575** mm		TOTAL NR OF LAYERS / CORE : 157 + 2	
22	TOTAL NR OF CORES : EIGHT [8]		LENGTH : **4,000** mm		PARTING SHEETS(EXT. 6 mm) : 2.5 mm	
23	NR OF PASSAGES / CORE		66	12	73	6
24	EFFECTIVE PASSAGE WIDTH	inches	39.6	39.6	39.6	39.6
25	TYPE OF FINS		4/8 SERR.	5% PERF	4/8 SERR.	6/8 SERR.
26	FIN HEIGHT	inches	0.2008	0.2008	0.3791	0.3791
27	FIN THICKNESS	inches	0.0138	0.0079	0.0079	0.0079
28	NUMBER OF FINS PER INCH		18.8	25.0	21.0	15
29	EFFECTIVE PASSAGE LENGTH	inches	125	116	125	125
30	TOTAL HEAT TRANSFER AREA	ft²	154,396	34,428	346,223	21,268
31	TOTAL FREE FLOW AREA	ft²	20.10	4.09	49.72	4.32
32	CALCULATED PRESSURE DROP	psi	7.0	3.5	2.6	1.8
33	HEADER SIZE IN/OUT	NPS(")	9[2] 9[2]	16 12	22 22	14 16
34	NOZZLE SIZE IN/OUT	NPS(")	8[4] 8[4]	6 6	20 18[2]	6 6
35	SUBMANIFOLD SIZE IN/OUT	NPS(")				
36	MANIFOLD SIZE IN/OUT	NPS(")	24[2] 24[2]	12 12	--- 48	--- 12
37	CONNECTIONS IN/OUT	NPS(")	24[2] 24[2]	12 12	--- 48	--- 12
38	RFWN AL. FLANGE RATING		600# 600#	300# 300#	--- 150#	--- 150#
39						
40			FIN 2934	FIN 2216	FIN 2945	FIN 2946

41 NOTES :
1. FOR INSTALLATION IN SERIES WITH ZONES 3 & 4
2. ESTIMATED ASSEMBLY DRY WEIGHT = 200,000 LB.
3. TWO [2] COLD BOX ASSEMBLIES REQUIRED
4. EACH COLD BOX ASSEMBLY 10' X 24' 43' (WxDxH)

REV.	DATE	ISSUED BY	APPROVED BY
0	9-Mar-09	AUAI / FDD	

DOC No. SECTION4 / 004-10 (28/01/00)

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fives cryogenie

HEAT EXCHANGER SPECIFICATION SHEET N° 5390-1

CUSTOMER: U. OF PENN	PROJECT: LNG DESIGN	N.C. ORDER N°: FOR QUOTATION
ITEM N°: ZONES 3 & 4	LOCATION: STUDY	CUST. JOB N°:
CASE: DESIGN	PLANT SERVICE: MAIN HX	CONSTRUCTION CODE: ASME VIII, DIV. 1

01	FLUID		A. SBOH	B. LNG	C. LPN2	D. N2 REJ		
02	TOTAL FLOWRATE	lb/hr	392,155	204,719	2,458,183	39,891		
03	VAPOR FLOWRATE IN	lb/hr	392,155	204,719	2,458,183	39,891		
04	VAPOR FLOWRATE OUT	lb/hr	204,662	0	2,458,183	39,891		
05	- MOL. WGT. IN/OUT		16.656 / 16.617	16.616 / ---	28.013 / 28.013	18.437 / 18.437		
06	LIQUID FLOWRATE IN	lb/hr	0					
07	LIQUID FLOWRATE OUT	lb/hr	187,493	204,719	0	0		
08	- MOL. WGT. IN/OUT		35.44 / 37.58	--- / 16.616	--- / ---	--- / ---		
09	TEMPERATURE IN	°F	-132.1	-147.0	-231.5	-259.3		
10	TEMPERATURE OUT	°F	-147.0	-220.0	-135.19	135.19		
11	DEW POINT / BUBBLE POINT							
12	OPERATING PRESSURE	psig	390.0	390.0	166.1	18.0		
13	ALLOWABLE PRESSURE DROP	psi	2.0	1.5	2.5	1.5		
14	TOTAL HEAT TRANSFERRED	MMBTU/hr	28.339	39.490	65.669	2.160		
15	CORRECTED MTD (GLOBAL)	°F	8.4	38.0	38 / 8.4	38 / 8.4		
16	FOULING FACTOR	°F-hr-ft²/Btu	NIL	NIL	NIL	NIL		
17	DESIGN TEMPERATURE	°F	150 / -275	150 / -275	150 / -275	150 / -275		
18	DESIGN PRESSURE	psig	500	500	200	50		
19	TEST PRESSURE HYDRO/PNEU	psig	per ASME	per ASME	per ASME	per ASME		
20	NUMBER OF ASSEMBLIES :	TWO [2]	WIDTH :	1,220 mm		TYPE OF HEAT EXCHANGER : COUNTERFLOW		
21	NR OF CORES/ASSEMBLY :	TWO [2]	HEIGHT :	1,525 mm		TOTAL NR OF LAYERS / CORE : 169 + 2		
22	TOTAL NR OF CORES :	FOUR [4]	LENGTH :	2,400 mm		PARTING SHEETS(EXT. 6 mm) : 1.5 mm		
23	NR OF PASSAGES / CORE		84	84	77	8		
24	EFFECTIVE PASSAGE WIDTH	inches	45.9	45.9	45.9	45.9		
25	TYPE OF FINS		1/8 SERR.	1/8 SERR.	4/8 SERR.	6/8 SERR.		
26	FIN HEIGHT	inches	0.2008	0.2008	0.3791	0.3791		
27	FIN THICKNESS	inches	0.0079	0.0079	0.0079	0.0079		
28	NUMBER OF FINS PER INCH		26.0	26.0	21.0	15		
29	EFFECTIVE PASSAGE LENGTH	inches	40	15	55	55		
30	TOTAL HEAT TRANSFER AREA	ft²	49,794	18,673	93,213	7,238		
31	TOTAL FREE FLOW AREA	ft²	16.43	16.43	30.43	3.34		
32	CALCULATED PRESSURE DROP	psi	1.5	1.1	2.2	1.5		
33	HEADER SIZE IN/OUT	NPS(")	12 10	8 6	30 30	18 18		
34	NOZZLE SIZE IN/OUT	NPS(")	10 8	6 4	28 28	6 8		
35	SUBMANIFOLD SIZE IN/OUT	NPS(")						
36	MANIFOLD SIZE IN/OUT	NPS(")	14 12	8 6	40 ---	8 ---		
37	CONNECTIONS IN/OUT	NPS(")	14 12	8 6	40 ---	8 ---		
38	RFWN AL. FLANGE RATING		300# 300#	300# 300#	150# ---	150# ---		
39								
40			FIN 2912	FIN 2912	FIN 2945	FIN 2946		

41 NOTES :

1. FOR INSTALLATION IN SERIES WITH ZONES 1 & 2
2. ESTIMATED ASSEMBLY DRY WEIGHT = 200,000 LB.
3. TWO [2] COLD BOX ASSEMBLIES REQUIRED
4. EACH COLD BOX ASSEMBLY 10' X 24' 43' (WxDxH)

REV.	DATE	ISSUED BY	APPROVED BY
0	9-Mar-09	AUAI / FDD	

DOC No. SECTION4 / 004-10 (28/01/00)

Appendix IX: Cooling Water Requirement Calculations

Cooling Water Requirement Calculations

The amount of cooling water required is calculated from the individual heat duties of all of the pieces of equipment that require cooling water. The required cooling water calculation will be provided for one unit here, and the rest will be summarized afterwards.

STEP 1: Determine the water flow in lb/hr

From references, the constant pressure heat capacity of seawater was found to be 0.953 Btu/lb-F. The maximum allowable temperature difference is 14F. The air compressor (HX-301a-c) requires a cooling duty of 0.18182E+09 Btu/hr.

$$\frac{lb\ H_2O}{hr} = \frac{Cooling\ duty}{Cp_{cw} \times \Delta T} = \frac{0.18182 \times 10^9\ Btu/hr}{0.953\ Btu/lb \cdot F \times 14\ F} = 1.36276 \times 10^7\ \frac{lb\ H_2O}{hr}$$

STEP 2: Convert the flow to yearly values

The plant is assumed to operate 24 hours a day, for 355 days per year.

$$\frac{lb\ H_2O}{year} = \frac{lb\ H_2O}{hr} \times \frac{24\ hr}{day} \times \frac{355\ day}{year} = 1.36276 \times 10^7 \times 24 \times 355 = 116{,}101{,}124{,}269 \frac{lb}{year}$$

STEP 3: Convert to gallons

The density of seawater is such that one gallon is approximately 8.33 lbs.

$$\frac{gal\ H_2O}{year} = \frac{lb\ H_2O}{year} \times \frac{gal}{8.33\ lb} = 116{,}101{,}124{,}269\ \frac{lb}{year} \times \frac{gal}{8.33\ lb} = 13{,}534{,}754{,}520 \frac{gal}{year}$$

STEP 4: Tabulate results

Process Unit	Cooling Duty	Cooling Water Requirement	
	Btu/year	MMlb/year	MMgal/year
HX-301a-c	1.57485×10^{12}	118,036.68	13,760.40
HX-302a-c	1.18982×10^{11}	8,917.84	1,039.62
HX-401a-d	1.63028×10^{12}	7,598.84	885.85
HX-102	1.01384×10^{11}	2,867.65	334.30
HX-203	3.82603×10^{10}	122,191.95	14,244.81
TOTALS	3.46376×10^{12}	259,612.96	30,264.98

Appendix X: ASPEN Files

```
BLOCK:  HX101     MODEL: MHEATX
------------------------------

  HOT SIDE:     INLET STREAM     OUTLET STREAM
                ------------     -------------
                COMPCOOL         N2-LNG
                N2COMP           HP-N2-C
                SCRUBOH          REFFEED

  COLD SIDE:    INLET STREAM     OUTLET STREAM
                ------------     -------------
                N2-PURGE         FUELGAS
                COLDN2           WARMN2
                REFOH            PRECOMP

  PROPERTIES FOR STREAM COMPCOOL
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

  PROPERTIES FOR STREAM N2COMP
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

  PROPERTIES FOR STREAM SCRUBOH
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

  PROPERTIES FOR STREAM N2-PURGE
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

  PROPERTIES FOR STREAM COLDN2
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

  PROPERTIES FOR STREAM REFOH
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                    ***  MASS AND ENERGY BALANCE  ***
                                IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)          185699.         185699.         0.00000
      MASS(LB/HR   )        0.466669E+07    0.466669E+07      0.00000
      ENTHALPY(BTU/HR  )   -0.170219E+10   -0.170219E+10    0.423718E-07

                        ***  INPUT DATA  ***

  MAXIMUM NO. ITERATIONS                                   30
  CONVERGENCE TOLERANCE                                     0.000100000

  SPECIFICATIONS FOR STREAM COMPCOOL:
  TWO    PHASE  TP  FLASH
  SPECIFIED TEMPERATURE              F                       -220.000
  PRESSURE DROP                      PSI                        0.0
  MAXIMUM NO. ITERATIONS                                       30
  CONVERGENCE TOLERANCE                                         0.000100000

  SPECIFICATIONS FOR STREAM N2COMP  :
  TWO    PHASE  TP  FLASH
  SPECIFIED TEMPERATURE              F                        -50.0000
  PRESSURE DROP                      PSI                        0.0
```

```
 MAXIMUM NO. ITERATIONS                                              30
 CONVERGENCE TOLERANCE                                                0.000100000

 SPECIFICATIONS FOR STREAM SCRUBOH :
 TWO    PHASE   TP  FLASH
 SPECIFIED TEMPERATURE                F                            -160.000
 PRESSURE DROP                        PSI                             0.0
 MAXIMUM NO. ITERATIONS                                              30
 CONVERGENCE TOLERANCE                                                0.000100000

 SPECIFICATIONS FOR STREAM N2-PURGE:
 TWO    PHASE       FLASH
 PRESSURE DROP                        PSI                             0.0
 MAXIMUM NO. ITERATIONS                                              30
 CONVERGENCE TOLERANCE                                                0.000100000

 SPECIFICATIONS FOR STREAM COLDN2  :
 TWO    PHASE       FLASH
 PRESSURE DROP                        PSI                             0.0
 MAXIMUM NO. ITERATIONS                                              30
 CONVERGENCE TOLERANCE                                                0.000100000

 SPECIFICATIONS FOR STREAM REFOH   :
 TWO    PHASE       FLASH
 PRESSURE DROP                        PSI                             0.0
 MAXIMUM NO. ITERATIONS                                              30
 CONVERGENCE TOLERANCE                                                0.000100000

                        ***  RESULTS  ***

INLET                    OUTLET       OUTLET       OUTLET
STREAM     DUTY          TEMPERATURE  PRESSURE     VAPOR FRAC
           BTU/HR        F            PSIA

COMPCOOL  -0.66309E+08   -220.00       720.00        0.0000
N2COMP    -0.78371E+08    -50.00       995.00        1.0000
SCRUBOH   -0.28758E+08   -160.00       290.00        0.5914
N2-PURGE   0.55944E+07     61.22       18.000        1.0000
COLDN2     0.14223E+09     61.22       130.00        1.0000
REFOH      0.25611E+08     61.22       290.00        1.0000


          ------------------------------------------
          |                                        |
COMPCOOL  |                                        | N2-LNG
--------->|         12320.     LBMOL/HR            |--------->
   90.00  |                                        |  -220.00
          |                                        |
N2COMP    |                                        | HP-N2-C
--------->|         69012.     LBMOL/HR            |--------->
   90.00  |                                        |   -50.00
          |                                        |
SCRUBOH   |                                        | REFFEED
--------->|         20833.     LBMOL/HR            |--------->
 -129.81  |                                        |  -160.00
          |                                        |
FUELGAS   |                                        | N2-PURGE
```

```
 <---------|          2201.8     LBMOL/HR          |<---------
     61.22 |                                       |  -259.21
           |                                       |
  WARMN2   |                                       | COLDN2
 <---------|          69012.     LBMOL/HR          |<---------
     61.22 |                                       |  -221.91
           |                                       |
  PRECOMP  |                                       | REFOH
 <---------|          12320.     LBMOL/HR          |<---------
     61.22 |                                       |  -160.00
           |                                       |
           -----------------------------------------
```

```
                       ***  INTERNAL ANALYSIS  ***

  FLOW IS COUNTERCURRENT.
  DUTY                             0.17344E+09 BTU/HR
  UA                               0.10032E+08 BTU/HR-R
  AVERAGE LMTD (DUTY/UA)            17.289     F
  MIN TEMP APPROACH                 3.0093     F
  HOT-SIDE TEMP APPROACH            28.779     F
  COLD-SIDE TEMP APPROACH           39.209     F
  HOT-SIDE NTU                      17.930
  COLD-SIDE NTU                     18.534

  DUTY          T HOT     T COLD    DELTA T   LMTD    UA ZONE     Q ZONE
UA           PINCH  STREAM IN/OUT/DEW/

POINT  BUBBLE POINT
  BTU/HR          F         F         F         F     BTU/HR-R    BTU/HR
BTU/HR-R

   0.000       -220.00   -259.21     39.21
  0.6526E+06   -216.23   -221.91      5.67     17.35  0.3762E+05  0.6526E+06
0.3762E+05  LOC    IN  COLDN2
  0.1734E+07   -210.07   -219.96      9.89      7.59  0.1426E+06  0.1082E+07
0.1802E+06
  0.3469E+07   -200.42   -216.84     16.42     12.88  0.1346E+06  0.1734E+07
0.3148E+06
  0.5203E+07   -191.08   -213.71     22.63     19.36  0.8959E+05  0.1734E+07
0.4044E+06
  0.6938E+07   -182.08   -210.56     28.48     25.44  0.6817E+05  0.1734E+07
0.4726E+06
  0.8672E+07   -173.46   -207.40     33.94     31.13  0.5572E+05  0.1734E+07
0.5283E+06
  0.1041E+08   -165.27   -204.23     38.96     36.39  0.4766E+05  0.1734E+07
0.5760E+06
  0.1158E+08   -160.00   -202.08     42.08     40.50  0.2894E+05  0.1172E+07
0.6049E+06         OUT SCRUBOH
  0.1214E+08   -159.88   -201.05     41.17     41.63  0.1350E+05  0.5621E+06
0.6184E+06
  0.1388E+08   -159.50   -197.86     38.37     39.75  0.4363E+05  0.1734E+07
0.6621E+06
  0.1561E+08   -159.08   -194.67     35.58     36.96  0.4693E+05  0.1734E+07
0.7090E+06
```

```
  0.1734E+08   -158.62   -191.46     32.84     34.19  0.5072E+05  0.1734E+07
0.7597E+06
  0.1908E+08   -158.11   -188.25     30.14     31.47  0.5512E+05  0.1734E+07
0.8148E+06
  0.2081E+08   -157.53   -185.02     27.49     28.79  0.6024E+05  0.1734E+07
0.8751E+06
  0.2255E+08   -156.88   -181.79     24.92     26.18  0.6625E+05  0.1734E+07
0.9413E+06
  0.2428E+08   -156.12   -178.55     22.43     23.65  0.7333E+05  0.1734E+07
0.1015E+07
  0.2602E+08   -155.26   -175.31     20.05     21.22  0.8174E+05  0.1734E+07
0.1096E+07
  0.2775E+08   -154.25   -172.06     17.80     18.90  0.9174E+05  0.1734E+07
0.1188E+07
  0.2948E+08   -153.10   -168.80     15.70     16.73  0.1037E+06  0.1734E+07
0.1292E+07
  0.3122E+08   -151.78   -165.53     13.75     14.70  0.1180E+06  0.1734E+07
0.1410E+07
  0.3295E+08   -150.28   -162.26     11.98     12.85  0.1350E+06  0.1734E+07
0.1545E+07
  0.3415E+08   -149.15   -160.00     10.85     11.40  0.1050E+06  0.1198E+07
0.1650E+07          IN  REFOH
  0.3469E+08   -148.62   -159.21     10.59     10.72  0.5007E+05  0.5365E+06
0.1700E+07
  0.3642E+08   -146.79   -156.62      9.84     10.21  0.1699E+06  0.1734E+07
0.1870E+07
  0.3816E+08   -144.80   -154.03      9.23      9.53  0.1820E+06  0.1734E+07
0.2052E+07
  0.3989E+08   -142.69   -151.43      8.74      8.98  0.1930E+06  0.1734E+07
0.2245E+07
  0.4163E+08   -140.45   -148.82      8.37      8.56  0.2027E+06  0.1734E+07
0.2447E+07
  0.4336E+08   -138.07   -146.20      8.12      8.25  0.2103E+06  0.1734E+07
0.2658E+07
  0.4509E+08   -135.57   -143.57      7.99      8.06  0.2153E+06  0.1734E+07
0.2873E+07
  0.4683E+08   -132.94   -140.92      7.98      7.99  0.2172E+06  0.1734E+07
0.3090E+07  LOC
  0.4856E+08   -130.20   -138.28      8.07      8.03  0.2161E+06  0.1734E+07
0.3306E+07
  0.4881E+08   -129.81   -137.91      8.10      8.09  0.3002E+05  0.2428E+06
0.3336E+07          DP  SCRUBOH
  0.4881E+08   -129.81   -137.91      8.10      8.10   135.8       1100.
0.3336E+07          IN  SCRUBOH
  0.5030E+08   -126.24   -135.62      9.38      8.72  0.1708E+06  0.1490E+07
0.3507E+07
  0.5203E+08   -122.85   -132.95     10.11      9.74  0.1781E+06  0.1734E+07
0.3685E+07
  0.5377E+08   -120.28   -130.28     10.00     10.05  0.1725E+06  0.1734E+07
0.3858E+07
  0.5550E+08   -118.49   -127.61      9.12      9.55  0.1816E+06  0.1734E+07
0.4039E+07
  0.5723E+08   -117.30   -124.92      7.62      8.35  0.2077E+06  0.1734E+07
0.4247E+07

  DUTY          T HOT     T COLD    DELTA T   LMTD    UA ZONE     Q ZONE
UA          PINCH  STREAM IN/OUT/DEW/
```

```
POINT  BUBBLE POINT
  BTU/HR          F          F          F          F     BTU/HR-R    BTU/HR
BTU/HR-R

  0.5897E+08   -116.42   -122.23      5.81       6.68  0.2598E+06  0.1734E+07
0.4507E+07
  0.6070E+08   -115.43   -119.54      4.11       4.91  0.3531E+06  0.1734E+07
0.4860E+07
  0.6244E+08   -113.81   -116.84      3.03       3.54  0.4895E+06  0.1734E+07
0.5350E+07
  0.6417E+08   -111.12   -114.13      3.01       3.02  0.5744E+06  0.1734E+07
0.5924E+07  GBL
  0.6591E+08   -107.08   -111.42      4.34       3.64  0.4771E+06  0.1734E+07
0.6401E+07
  0.6764E+08   -101.58   -108.71      7.13       5.62  0.3085E+06  0.1734E+07
0.6709E+07
  0.6938E+08    -94.66   -105.99     11.33       9.07  0.1913E+06  0.1734E+07
0.6901E+07
  0.7111E+08    -86.46   -103.26     16.80      13.88  0.1249E+06  0.1734E+07
0.7026E+07
  0.7284E+08    -77.15   -100.54     23.39      19.91  0.8710E+05  0.1734E+07
0.7113E+07
  0.7458E+08    -66.88    -97.80     30.93      26.98  0.6428E+05  0.1734E+07
0.7177E+07
  0.7631E+08    -55.80    -95.07     39.27      34.93  0.4965E+05  0.1734E+07
0.7227E+07
  0.7718E+08    -50.00    -93.70     43.70      41.45  0.2090E+05  0.8664E+06
0.7248E+07          OUT N2COMP
  0.7805E+08    -48.83    -92.33     43.50      43.60  0.1991E+05  0.8680E+06
0.7268E+07
  0.7978E+08    -46.50    -89.59     43.08      43.29  0.4006E+05  0.1734E+07
0.7308E+07
  0.8152E+08    -44.16    -86.84     42.68      42.88  0.4045E+05  0.1734E+07
0.7348E+07
  0.8325E+08    -41.81    -84.09     42.28      42.48  0.4083E+05  0.1734E+07
0.7389E+07
  0.8498E+08    -39.45    -81.34     41.89      42.09  0.4121E+05  0.1734E+07
0.7430E+07
  0.8672E+08    -37.08    -78.59     41.51      41.70  0.4159E+05  0.1734E+07
0.7472E+07
  0.8845E+08    -34.69    -75.83     41.13      41.32  0.4197E+05  0.1734E+07
0.7514E+07
  0.9019E+08    -32.30    -73.07     40.76      40.95  0.4236E+05  0.1734E+07
0.7556E+07
  0.9192E+08    -29.90    -70.30     40.40      40.58  0.4274E+05  0.1734E+07
0.7599E+07
  0.9366E+08    -27.49    -67.54     40.04      40.22  0.4312E+05  0.1734E+07
0.7642E+07
  0.9539E+08    -25.08    -64.77     39.69      39.87  0.4350E+05  0.1734E+07
0.7685E+07
  0.9713E+08    -22.65    -62.00     39.35      39.52  0.4388E+05  0.1734E+07
0.7729E+07
  0.9886E+08    -20.22    -59.23     39.01      39.18  0.4427E+05  0.1734E+07
0.7773E+07
  0.1006E+09    -17.77    -56.45     38.68      38.84  0.4465E+05  0.1734E+07
0.7818E+07
```

0.1023E+09	-15.32	-53.68	38.35	38.51	0.4503E+05	0.1734E+07	0.7863E+07
0.1041E+09	-12.87	-50.90	38.03	38.19	0.4541E+05	0.1734E+07	0.7909E+07
0.1058E+09	-10.40	-48.12	37.71	37.87	0.4580E+05	0.1734E+07	0.7954E+07
0.1075E+09	-7.93	-45.33	37.40	37.56	0.4618E+05	0.1734E+07	0.8001E+07
0.1093E+09	-5.45	-42.55	37.10	37.25	0.4656E+05	0.1734E+07	0.8047E+07
0.1110E+09	-2.96	-39.77	36.80	36.95	0.4694E+05	0.1734E+07	0.8094E+07
0.1127E+09	-0.47	-36.98	36.51	36.66	0.4731E+05	0.1734E+07	0.8141E+07
0.1145E+09	2.03	-34.19	36.21	36.36	0.4770E+05	0.1734E+07	0.8189E+07
0.1162E+09	4.54	-31.41	35.95	36.08	0.4807E+05	0.1734E+07	0.8237E+07
0.1179E+09	7.05	-28.60	35.65	35.80	0.4845E+05	0.1734E+07	0.8286E+07
0.1197E+09	9.57	-25.81	35.37	35.51	0.4884E+05	0.1734E+07	0.8334E+07
0.1214E+09	12.09	-23.01	35.10	35.24	0.4922E+05	0.1734E+07	0.8384E+07
0.1231E+09	14.62	-20.22	34.84	34.97	0.4960E+05	0.1734E+07	0.8433E+07
0.1249E+09	17.16	-17.42	34.58	34.71	0.4997E+05	0.1734E+07	0.8483E+07
0.1266E+09	19.70	-14.62	34.32	34.45	0.5035E+05	0.1734E+07	0.8534E+07
0.1283E+09	22.24	-11.82	34.06	34.19	0.5073E+05	0.1734E+07	0.8584E+07
0.1301E+09	24.80	-9.02	33.82	33.94	0.5110E+05	0.1734E+07	0.8635E+07
0.1318E+09	27.35	-6.22	33.57	33.69	0.5148E+05	0.1734E+07	0.8687E+07
0.1335E+09	29.92	-3.41	33.33	33.45	0.5185E+05	0.1734E+07	0.8739E+07
0.1353E+09	32.48	-0.61	33.09	33.21	0.5222E+05	0.1734E+07	0.8791E+07
0.1370E+09	35.06	2.19	32.86	32.98	0.5259E+05	0.1734E+07	0.8844E+07
0.1388E+09	37.63	5.00	32.63	32.75	0.5296E+05	0.1734E+07	0.8896E+07
0.1405E+09	40.21	7.81	32.41	32.52	0.5333E+05	0.1734E+07	0.8950E+07
0.1422E+09	42.80	10.61	32.19	32.30	0.5370E+05	0.1734E+07	0.9004E+07

```
  DUTY          T HOT     T COLD    DELTA T   LMTD    UA ZONE     Q ZONE
UA          PINCH  STREAM IN/OUT/DEW/

POINT  BUBBLE POINT
  BTU/HR          F          F          F          F     BTU/HR-R    BTU/HR
BTU/HR-R
```

0.1440E+09	45.39	13.42	31.97	32.08	0.5407E+05	0.1734E+07	0.9058E+07
0.1457E+09	47.98	16.23	31.75	31.86	0.5444E+05	0.1734E+07	0.9112E+07
0.1474E+09	50.58	19.04	31.54	31.65	0.5480E+05	0.1734E+07	0.9167E+07
0.1492E+09	53.18	21.85	31.34	31.44	0.5517E+05	0.1734E+07	0.9222E+07
0.1509E+09	55.79	24.66	31.13	31.23	0.5553E+05	0.1734E+07	0.9278E+07
0.1526E+09	58.40	27.47	30.93	31.03	0.5589E+05	0.1734E+07	0.9333E+07
0.1544E+09	61.02	30.28	30.74	30.83	0.5625E+05	0.1734E+07	0.9390E+07
0.1561E+09	63.63	33.09	30.54	30.64	0.5661E+05	0.1734E+07	0.9446E+07
0.1578E+09	66.26	35.90	30.35	30.45	0.5696E+05	0.1734E+07	0.9503E+07
0.1596E+09	68.88	38.72	30.17	30.26	0.5732E+05	0.1734E+07	0.9561E+07
0.1613E+09	71.51	41.53	29.98	30.07	0.5767E+05	0.1734E+07	0.9618E+07
0.1630E+09	74.14	44.34	29.80	29.89	0.5802E+05	0.1734E+07	0.9676E+07
0.1648E+09	76.78	47.15	29.62	29.71	0.5837E+05	0.1734E+07	0.9735E+07
0.1665E+09	79.42	49.97	29.45	29.54	0.5872E+05	0.1734E+07	0.9793E+07
0.1682E+09	82.06	52.78	29.28	29.36	0.5907E+05	0.1734E+07	0.9852E+07
0.1700E+09	84.70	55.59	29.11	29.19	0.5941E+05	0.1734E+07	0.9912E+07
0.1717E+09	87.35	58.41	28.94	29.03	0.5975E+05	0.1734E+07	0.9972E+07
0.1734E+09	90.00	61.22	28.78	28.86	0.6010E+05	0.1734E+07	0.1003E+08

```
 GBL = GLOBAL     LOC = LOCAL     DP  = DEW POINT     BP  = BUBBLE POINT

BLOCK:  B-601    MODEL: HEATX
-----------------------------
  HOT SIDE:
  ---------
  INLET STREAM:           FURNEXH
  OUTLET STREAM:          EXHINT
  PROPERTY OPTION SET:    PENG-ROB  STANDARD PR EQUATION OF STATE
  COLD SIDE:
  ----------
  INLET STREAM:           36
  OUTLET STREAM:          HPSTEAMH
  PROPERTY OPTION SET:    PENG-ROB  STANDARD PR EQUATION OF STATE

                      ***  MASS AND ENERGY BALANCE  ***
                                 IN                OUT         RELATIVE DIFF.
   TOTAL BALANCE
```

```
    MOLE(LBMOL/HR)            3255.18         3255.18          0.00000
    MASS(LB/HR   )            78377.1         78377.1          0.00000
    ENTHALPY(BTU/HR  )      -0.159339E+09   -0.159339E+09    0.187037E-15

                       ***  INPUT DATA  ***

FLASH SPECS FOR HOT SIDE:
TWO    PHASE      FLASH
MAXIMUM NO. ITERATIONS                                  30
CONVERGENCE TOLERANCE                                    0.000100000

FLASH SPECS FOR COLD SIDE:
TWO    PHASE      FLASH
MAXIMUM NO. ITERATIONS                                  30
CONVERGENCE TOLERANCE                                    0.000100000

FLOW DIRECTION AND SPECIFICATION:
  COUNTERCURRENT   HEAT EXCHANGER
  SPECIFIED COLD VAPOR FRACTION
  SPECIFIED VALUE                                        1.0000
  LMTD CORRECTION FACTOR                                 1.00000

PRESSURE SPECIFICATION:
  HOT  SIDE PRESSURE DROP        PSI                     0.0000
  COLD SIDE OUTLET PRESSURE      PSIA                   94.6959

HEAT TRANSFER COEFFICIENT SPECIFICATION:
  HOT LIQUID    COLD LIQUID      BTU/HR-SQFT-R         149.6937
  HOT 2-PHASE   COLD LIQUID      BTU/HR-SQFT-R         149.6937
  HOT VAPOR     COLD LIQUID      BTU/HR-SQFT-R         149.6937
  HOT LIQUID    COLD 2-PHASE     BTU/HR-SQFT-R         149.6937
  HOT 2-PHASE   COLD 2-PHASE     BTU/HR-SQFT-R         149.6937
  HOT VAPOR     COLD 2-PHASE     BTU/HR-SQFT-R         149.6937
  HOT LIQUID    COLD VAPOR       BTU/HR-SQFT-R         149.6937
  HOT 2-PHASE   COLD VAPOR       BTU/HR-SQFT-R         149.6937
  HOT VAPOR     COLD VAPOR       BTU/HR-SQFT-R         149.6937

                     ***  OVERALL RESULTS  ***

STREAMS:
                --------------------------------------
                |                                    |
FURNEXH   ----->|                HOT                 |-----> EXHINT
T=  2.3456D+03  |                                    |       T=
1.0083D+03
P=  5.0000D+02  |                                    |       P=
5.0000D+02
V=  1.0000D+00  |                                    |       V=
1.0000D+00
                |                                    |
HPSTEAMH  <-----|                COLD                |<----- 36
T=  3.2534D+02  |                                    |       T=
3.0002D+02
P=  9.4696D+01  |                                    |       P=
1.1470D+02
V=  1.0000D+00  |                                    |       V=
0.0000D+00
```

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                    --------------------------------------

    DUTY AND AREA:
      CALCULATED HEAT DUTY           BTU/HR              22460389.1061
      CALCULATED (REQUIRED) AREA     SQFT                     119.8716
      ACTUAL EXCHANGER AREA          SQFT                     119.8716
      PER CENT OVER-DESIGN                                      0.0000

    HEAT TRANSFER COEFFICIENT:
      AVERAGE COEFFICIENT (DIRTY)    BTU/HR-SQFT-R            149.6937
      UA (DIRTY)                     BTU/HR-R               17944.0236

    LOG-MEAN TEMPERATURE DIFFERENCE:
      LMTD CORRECTION FACTOR                                    1.0000
      LMTD (CORRECTED)               F                       1251.6919
      NUMBER OF SHELLS IN SERIES                                  1

    PRESSURE DROP:
      HOTSIDE, TOTAL                 PSI                        0.0000
      COLDSIDE, TOTAL                PSI                       20.0000

    PRESSURE DROP PARAMETER:
      HOT SIDE:                                               0.0000
      COLD SIDE:                                            0.10517E+06

                          ***  ZONE RESULTS  ***

    TEMPERATURE LEAVING EACH ZONE:

                                   HOT
          -------------------------------------------------------------
          |                                                           |
  FURNEXH |                        VAP                                | EXHINT
  ------> |                                                           |------>
  2345.6  |                                                           | 1008.3
          |                                                           |
  HPSTEAMH|                        BOIL                               | 36
  <------ |                                                           |<------
   325.3  |                                                           |  300.0
          |                                                           |
          -------------------------------------------------------------
                                   COLD

    ZONE HEAT TRANSFER AND AREA:

    ZONE       HEAT DUTY       AREA       DTLM       AVERAGE U       UA
               BTU/HR          SQFT        F         BTU/HR-SQFT-R   BTU/HR-R
      1   22460389.106       119.8716    1251.6919   149.6937       17944.0236

  BLOCK:  B-602    MODEL: HEATX
  -----------------------------
    HOT SIDE:
    ---------
    INLET STREAM:           EXHINT
    OUTLET STREAM:          EXHOUT
    PROPERTY OPTION SET:    PENG-ROB  STANDARD PR EQUATION OF STATE
    COLD SIDE:
```

```
----------
INLET STREAM:           34
OUTLET STREAM:          LPSTEAMH
PROPERTY OPTION SET:    PENG-ROB  STANDARD PR EQUATION OF STATE

                   ***  MASS AND ENERGY BALANCE  ***
                            IN                OUT        RELATIVE DIFF.
 TOTAL BALANCE
    MOLE(LBMOL/HR)            2671.07         2671.07         0.00000
    MASS(LB/HR   )            67854.1         67854.1         0.00000
    ENTHALPY(BTU/HR  )      -0.113164E+09   -0.113164E+09    0.131677E-15

                       ***  INPUT DATA  ***

FLASH SPECS FOR HOT SIDE:
TWO    PHASE      FLASH
MAXIMUM NO. ITERATIONS                                     30
CONVERGENCE TOLERANCE                                       0.000100000

FLASH SPECS FOR COLD SIDE:
TWO    PHASE      FLASH
MAXIMUM NO. ITERATIONS                                     30
CONVERGENCE TOLERANCE                                       0.000100000

FLOW DIRECTION AND SPECIFICATION:
  COUNTERCURRENT   HEAT EXCHANGER
  SPECIFIED COLD VAPOR FRACTION
  SPECIFIED VALUE                                           1.0000
  LMTD CORRECTION FACTOR                                    1.00000

PRESSURE SPECIFICATION:
  HOT  SIDE PRESSURE DROP        PSI                        0.0000
  COLD SIDE OUTLET PRESSURE      PSIA                      21.6959

HEAT TRANSFER COEFFICIENT SPECIFICATION:
  HOT LIQUID    COLD LIQUID      BTU/HR-SQFT-R           149.6937
  HOT 2-PHASE   COLD LIQUID      BTU/HR-SQFT-R           149.6937
  HOT VAPOR     COLD LIQUID      BTU/HR-SQFT-R           149.6937
  HOT LIQUID    COLD 2-PHASE     BTU/HR-SQFT-R           149.6937
  HOT 2-PHASE   COLD 2-PHASE     BTU/HR-SQFT-R           149.6937
  HOT VAPOR     COLD 2-PHASE     BTU/HR-SQFT-R           149.6937
  HOT LIQUID    COLD VAPOR       BTU/HR-SQFT-R           149.6937
  HOT 2-PHASE   COLD VAPOR       BTU/HR-SQFT-R           149.6937
  HOT VAPOR     COLD VAPOR       BTU/HR-SQFT-R           149.6937

                     ***  OVERALL RESULTS  ***

STREAMS:
                -------------------------------------
                |                                    |
EXHINT    ----->|                HOT                 |-----> EXHOUT
T=  1.0083D+03  |                                    |       T=
2.5000D+02
P=  5.0000D+02  |                                    |       P=
5.0000D+02
V=  1.0000D+00  |                                    |       V=  9.5459D-
01
```

```
                |                                     |
   LPSTEAMH  <-----|               COLD               |<----- 34
   T=  2.3536D+02  |                                  |       T=
2.2868D+02
   P=  2.1696D+01  |                                  |       P=
4.1696D+01
   V=  1.0000D+00  |                                  |       V=
0.0000D+00
                   --------------------------------------

   DUTY AND AREA:
     CALCULATED HEAT DUTY            BTU/HR          13019958.7672
     CALCULATED (REQUIRED) AREA      SQFT                1197.8505
     ACTUAL EXCHANGER AREA           SQFT                1197.8505
     PER CENT OVER-DESIGN                                   0.0000

   HEAT TRANSFER COEFFICIENT:
     AVERAGE COEFFICIENT (DIRTY)     BTU/HR-SQFT-R        149.6937
     UA (DIRTY)                      BTU/HR-R           179310.6242

   LOG-MEAN TEMPERATURE DIFFERENCE:
     LMTD CORRECTION FACTOR                                 1.0000
     LMTD (CORRECTED)                F                     72.6112
     NUMBER OF SHELLS IN SERIES                           1

   PRESSURE DROP:
     HOTSIDE, TOTAL                  PSI                    0.0000
     COLDSIDE, TOTAL                 PSI                   20.0000

   PRESSURE DROP PARAMETER:
     HOT SIDE:                                            0.0000
     COLD SIDE:                                           86883.

                        ***  ZONE RESULTS  ***

   TEMPERATURE LEAVING EACH ZONE:

                                  HOT
          -------------------------------------------------------------
          |                            |                              |
 EXHINT   |            VAP             |             COND             | EXHOUT
 ------>  |                            |                              |------>
 1008.3   |                       280.1|                              |  250.0
          |                            |                              |
 LPSTEAMH|            BOIL             |             BOIL             | 34
 <------  |                            |                              |<------
  235.4   |                       267.7|                              |  228.7
          |                            |                              |
          -------------------------------------------------------------
                                  COLD

   ZONE HEAT TRANSFER AND AREA:

   ZONE       HEAT DUTY        AREA        DTLM       AVERAGE U        UA
              BTU/HR           SQFT         F         BTU/HR-SQFT-R   BTU/HR-R
     1   11056171.270       401.2574     184.0680    149.6937       60065.6919
     2    1963787.497       796.5931      16.4685    149.6937      119244.9322
```

```
BLOCK:  C-101    MODEL: COMPR
-----------------------------
  INLET STREAM:          PRECOMP
  OUTLET STREAM:         COMP
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                     ***  MASS AND ENERGY BALANCE  ***
                                   IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)           12320.0         12320.0         0.00000
      MASS(LB/HR   )           204756.         204756.         0.00000
      ENTHALPY(BTU/HR  )     -0.381200E+09   -0.366703E+09   -0.380297E-01

                          ***  INPUT DATA  ***

  ISENTROPIC CENTRIFUGAL COMPRESSOR
   OUTLET PRESSURE  PSIA                                    725.000
   ISENTROPIC EFFICIENCY                                      0.86000
   MECHANICAL EFFICIENCY                                      1.00000

                          ***  RESULTS  ***

   INDICATED  HORSEPOWER REQUIREMENT  HP                  5,697.50
   BRAKE      HORSEPOWER REQUIREMENT  HP                  5,697.50
   NET WORK REQUIRED                  HP                  5,697.50
   POWER LOSSES                       HP                      0.0
   ISENTROPIC HORSEPOWER REQUIREMENT  HP                  4,899.85
   CALCULATED OUTLET TEMP  F                                205.853
   ISENTROPIC TEMPERATURE  F                                189.243
   EFFICIENCY (POLYTR/ISENTR) USED                            0.86000
   OUTLET VAPOR FRACTION                                      1.00000
   HEAD DEVELOPED,       FT-LBF/LB                       47,381.7
   MECHANICAL EFFICIENCY USED                                 1.00000
   INLET HEAT CAPACITY RATIO                                  1.37939
   INLET VOLUMETRIC FLOW RATE , CUFT/HR                 226,648.
   OUTLET VOLUMETRIC FLOW RATE, CUFT/HR                 116,451.
   INLET  COMPRESSIBILITY FACTOR                              0.95442
   OUTLET COMPRESSIBILITY FACTOR                              0.95952
   AV. ISENT. VOL. EXPONENT                                   1.31612
   AV. ISENT. TEMP EXPONENT                                   1.31550
   AV. ACTUAL VOL. EXPONENT                                   1.37596
   AV. ACTUAL TEMP EXPONENT                                   1.36503

BLOCK:  C-301A-C MODEL: MCOMPR
------------------------------
  INLET STREAMS:    IN-AIR       TO STAGE   1
  OUTLET STREAMS:   AIR        FROM STAGE   3
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                     ***  MASS AND ENERGY BALANCE  ***
                                   IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)           30600.0         30600.0         0.00000
      MASS(LB/HR   )           882700.         882700.         0.00000
      ENTHALPY(BTU/HR  )     -0.202955E+07    -528206.       -0.739742
```

*** INPUT DATA ***

ISENTROPIC CENTRIFUGAL COMPRESSOR

NUMBER OF STAGES	3
FINAL PRESSURE, PSIA	505.000

COMPRESSOR SPECIFICATIONS PER STAGE

STAGE NUMBER	MECHANICAL EFFICIENCY	ISENTROPIC EFFICIENCY
1	1.000	0.7800
2	1.000	0.7800
3	1.000	0.7800

COOLER SPECIFICATIONS PER STAGE

STAGE NUMBER	PRESSURE DROP PSI	TEMPERATURE F
1	5.000	90.00
2	5.000	90.00
3	5.000	90.00

*** RESULTS ***

FINAL PRESSURE, PSIA	500.000
TOTAL WORK REQUIRED, HP	73,235.1
TOTAL COOLING DUTY , BTU/HR	-0.184841+09

*** PROFILE ***

COMPRESSOR PROFILE

STAGE NUMBER	OUTLET PRESSURE PSIA	PRESSURE RATIO	OUTLET TEMPERATURE F
1	47.79	3.251	336.6
2	155.3	3.631	400.4
3	505.0	3.359	379.0

STAGE NUMBER	INDICATED HORSEPOWER HP	BRAKE HORSEPOWER HP
1	0.2263E+05	0.2263E+05
2	0.2622E+05	0.2622E+05
3	0.2438E+05	0.2438E+05

COOLER PROFILE

STAGE NUMBER	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	COOLING LOAD BTU/HR	VAPOR FRACTION
1	90.00	42.79	-.5307E+08	1.000

2	90.00	150.3	-.6745E+08	1.000
3	90.00	500.0	-.6432E+08	1.000

BLOCK: C-302A-C MODEL: MCOMPR

INLET STREAMS: FUELGAS TO STAGE 1
OUTLET STREAMS: FG FROM STAGE 3
PROPERTY OPTION SET: PENG-ROB STANDARD PR EQUATION OF STATE

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(LBMOL/HR)	2201.80	2201.80	0.00000
MASS(LB/HR)	40527.6	40527.6	0.00000
ENTHALPY(BTU/HR)	-0.569159E+08	-0.568593E+08	-0.994775E-03

*** INPUT DATA ***

ISENTROPIC CENTRIFUGAL COMPRESSOR
NUMBER OF STAGES 3
FINAL PRESSURE, PSIA 505.000

COMPRESSOR SPECIFICATIONS PER STAGE

STAGE NUMBER	MECHANICAL EFFICIENCY	ISENTROPIC EFFICIENCY
1	1.000	0.7800
2	1.000	0.7800
3	1.000	0.7800

COOLER SPECIFICATIONS PER STAGE

STAGE NUMBER	PRESSURE DROP PSI	TEMPERATURE F
1	5.000	90.00
2	5.000	90.00
3	5.000	90.00

*** RESULTS ***

FINAL PRESSURE, PSIA 500.000
TOTAL WORK REQUIRED, HP 4,698.93
TOTAL COOLING DUTY , BTU/HR -0.118995+08

*** PROFILE ***

COMPRESSOR PROFILE

STAGE NUMBER	OUTLET PRESSURE PSIA	PRESSURE RATIO	OUTLET TEMPERATURE F
1	54.69	3.039	257.4
2	166.2	3.344	311.6

3	505.0	3.133	300.4

STAGE NUMBER	INDICATED HORSEPOWER HP	BRAKE HORSEPOWER HP
1	1456.	1456.
2	1680.	1680.
3	1562.	1562.

COOLER PROFILE

STAGE NUMBER	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	COOLING LOAD BTU/HR	VAPOR FRACTION
1	90.00	49.69	-.3215E+07	1.000
2	90.00	161.2	-.4384E+07	1.000
3	90.00	500.0	-.4301E+07	1.000

BLOCK: C-401A-D MODEL: MCOMPR

INLET STREAMS: WARMN2 TO STAGE 1
OUTLET STREAMS: N2COMP FROM STAGE 4
PROPERTY OPTION SET: PENG-ROB STANDARD PR EQUATION OF STATE

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(LBMOL/HR)	69012.1	69012.1	0.00000
MASS(LB/HR)	0.193327E+07	0.193327E+07	0.00000
ENTHALPY(BTU/HR)	-0.970642E+07	-0.679489E+07	-0.299959

*** INPUT DATA ***

ISENTROPIC CENTRIFUGAL COMPRESSOR

NUMBER OF STAGES	4
FINAL PRESSURE, PSIA	1,000.00

COMPRESSOR SPECIFICATIONS PER STAGE

STAGE NUMBER	MECHANICAL EFFICIENCY	ISENTROPIC EFFICIENCY
1	1.000	0.8600
2	1.000	0.8600
3	1.000	0.8600
4	1.000	0.8600

COOLER SPECIFICATIONS PER STAGE

STAGE NUMBER	PRESSURE DROP PSI	TEMPERATURE F
1	5.000	90.00
2	5.000	90.00
3	5.000	90.00

4	5.000	90.00

*** RESULTS ***

FINAL PRESSURE, PSIA	995.000
TOTAL WORK REQUIRED, HP	76,346.9
TOTAL COOLING DUTY , BTU/HR	-0.191348+09

*** PROFILE ***

COMPRESSOR PROFILE

STAGE NUMBER	OUTLET PRESSURE PSIA	PRESSURE RATIO	OUTLET TEMPERATURE F
1	216.5	1.665	156.5
2	360.6	1.705	195.5
3	600.5	1.689	193.7
4	1000.	1.679	192.7

STAGE NUMBER	INDICATED HORSEPOWER HP	BRAKE HORSEPOWER HP
1	0.1787E+05	0.1787E+05
2	0.1981E+05	0.1981E+05
3	0.1943E+05	0.1943E+05
4	0.1924E+05	0.1924E+05

COOLER PROFILE

STAGE NUMBER	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	COOLING LOAD BTU/HR	VAPOR FRACTION
1	90.00	211.5	-.3258E+08	1.000
2	90.00	355.6	-.5240E+08	1.000
3	90.00	595.5	-.5261E+08	1.000
4	90.00	995.0	-.5376E+08	1.000

BLOCK: C-701 MODEL: COMPR

INLET STREAM: CO2WARM
OUTLET STREAM: TOCO2COO
PROPERTY OPTION SET: PENG-ROB STANDARD PR EQUATION OF STATE

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(LBMOL/HR)	26000.0	26000.0	0.00000
MASS(LB/HR)	0.114425E+07	0.114425E+07	0.00000
ENTHALPY(BTU/HR)	-0.440908E+10	-0.437449E+10	-0.784544E-02

*** INPUT DATA ***

ISENTROPIC CENTRIFUGAL COMPRESSOR
OUTLET PRESSURE PSIA 295.000

```
   ISENTROPIC EFFICIENCY                                          0.88000
   MECHANICAL EFFICIENCY                                          1.00000

                       ***  RESULTS  ***

   INDICATED  HORSEPOWER REQUIREMENT  HP                  13,594.8
   BRAKE      HORSEPOWER REQUIREMENT  HP                  13,594.8
   NET WORK REQUIRED                  HP                  13,594.8
   POWER LOSSES                       HP                       0.0
   ISENTROPIC HORSEPOWER REQUIREMENT  HP                  11,963.5
   CALCULATED OUTLET TEMP  F                                 204.966
   ISENTROPIC TEMPERATURE  F                                 189.621
   EFFICIENCY (POLYTR/ISENTR) USED                            0.88000
   OUTLET VAPOR FRACTION                                      1.00000
   HEAD DEVELOPED,       FT-LBF/LB                        20,701.4
   MECHANICAL EFFICIENCY USED                                 1.00000
   INLET HEAT CAPACITY RATIO                                  1.35316
   INLET VOLUMETRIC FLOW RATE , CUFT/HR                1,348,980.
   OUTLET VOLUMETRIC FLOW RATE, CUFT/HR                  593,408.
   INLET  COMPRESSIBILITY FACTOR                              0.95403
   OUTLET COMPRESSIBILITY FACTOR                              0.94398
   AV. ISENT. VOL. EXPONENT                                   1.27299
   AV. ISENT. TEMP EXPONENT                                   1.29715
   AV. ACTUAL VOL. EXPONENT                                   1.31731
   AV. ACTUAL TEMP EXPONENT                                   1.33453

BLOCK:  D-101    MODEL: RADFRAC
-------------------------------
   INLETS    - LPFEED   STAGE   2
               REFLUX   STAGE   1
   OUTLETS   - SCRUBOH  STAGE   1
               TOFRAC   STAGE   6
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                  ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)          22012.7         22012.7         0.00000
      MASS(LB/HR   )          399074.         399074.        0.437570E-15
      ENTHALPY(BTU/HR  )    -0.761239E+09   -0.752932E+09   -0.109121E-01


                        *********************
                        ****  INPUT DATA  ****
                        *********************

  ****   INPUT PARAMETERS   ****

   NUMBER OF STAGES                                            6
   ALGORITHM OPTION                                         STANDARD
   ABSORBER OPTION                                          NO
   INITIALIZATION OPTION                                    STANDARD
   HYDRAULIC PARAMETER CALCULATIONS                         NO
   INSIDE LOOP CONVERGENCE METHOD                           BROYDEN
   DESIGN SPECIFICATION METHOD                              NESTED
   MAXIMUM NO. OF OUTSIDE LOOP ITERATIONS                     25
```

MAXIMUM NO. OF INSIDE LOOP ITERATIONS		10
MAXIMUM NUMBER OF FLASH ITERATIONS		50
FLASH TOLERANCE		0.000100000
OUTSIDE LOOP CONVERGENCE TOLERANCE		0.000100000

**** COL-SPECS ****

MOLAR VAPOR DIST / TOTAL DIST		1.00000
MOLAR BOTTOMS RATE	LBMOL/HR	1,180.00
CONDENSER DUTY (W/O SUBCOOL)	BTU/HR	0.0

**** PROFILES ****

P-SPEC	STAGE 1	PRES, PSIA	290.000

****

**** RESULTS ****

****

*** COMPONENT SPLIT FRACTIONS ***

	OUTLET STREAMS	
	SCRUBOH	TOFRAC
COMPONENT:		
NITRO-01	1.0000	.15969E-06
METHA-01	.99945	.54688E-03
ETHAN-01	.46570	.53430
PROPA-01	.82980E-01	.91702
N-BUT-01	.79277E-02	.99207
ISOBU-01	.14828E-01	.98517
2-MET-01	.88842E-03	.99911
N-PEN-01	.52927E-03	.99947
N-HEX-01	.17095E-04	.99998

*** SUMMARY OF KEY RESULTS ***

TOP STAGE TEMPERATURE	F	-129.806
BOTTOM STAGE TEMPERATURE	F	66.1381
TOP STAGE LIQUID FLOW	LBMOL/HR	2,268.73
BOTTOM STAGE LIQUID FLOW	LBMOL/HR	1,180.00
TOP STAGE VAPOR FLOW	LBMOL/HR	20,832.7
BOTTOM STAGE VAPOR FLOW	LBMOL/HR	1,441.68
MOLAR BOILUP RATIO		1.22177
CONDENSER DUTY (W/O SUBCOOL)	BTU/HR	0.0
REBOILER DUTY	BTU/HR	8,306,070.

**** MAXIMUM FINAL RELATIVE ERRORS ****

DEW POINT	0.47363E-03	STAGE= 3	
BUBBLE POINT	0.14440E-03	STAGE= 4	
COMPONENT MASS BALANCE	0.11527E-05	STAGE= 2	COMP=PROPA-01
ENERGY BALANCE	0.12792E-04	STAGE= 4	

```
 ****    PROFILES   ****

 **NOTE** REPORTED VALUES FOR STAGE LIQUID AND VAPOR RATES ARE THE FLOWS
          FROM THE STAGE EXCLUDING ANY SIDE PRODUCT.  FOR THE FIRST STAGE,
          THE REPORTED VAPOR FLOW IS THE VAPOR DISTILLATE FLOW.  FOR THE
          LAST STAGE, THE REPORTED LIQUID FLOW IS THE LIQUID BOTTOMS FLOW.

                                              ENTHALPY
STAGE TEMPERATURE   PRESSURE                 BTU/LBMOL          HEAT DUTY
      F             PSIA            LIQUID        VAPOR         BTU/HR

  1  -129.81        290.00        -43291.      -33288.
  2  -110.39        292.00        -46552.      -34023.
  3  -69.591        294.00        -47278.      -34363.
  4  -9.9378        296.00        -46903.      -35649.
  5   29.872        298.00        -46989.      -37174.
  6   66.138        300.00        -50391.      -38442.       .83061+07

STAGE     FLOW RATE                   FEED RATE                 PRODUCT RATE
           LBMOL/HR                    LBMOL/HR                   LBMOL/HR
      LIQUID     VAPOR       LIQUID    VAPOR    MIXED      LIQUID    VAPOR
  1  2269.     0.2083E+05  8512.5658 .13291+05
.20833+05
  2  2233.      1298.       209.2363
  3  2210.      1053.
  4  2457.      1030.
  5  2622.      1277.
  6  1180.      1442.                                     1180.0000

   ****  MASS FLOW PROFILES  ****

STAGE     FLOW RATE                   FEED RATE                 PRODUCT RATE
           LB/HR                       LB/HR                      LB/HR
      LIQUID     VAPOR       LIQUID    VAPOR    MIXED      LIQUID    VAPOR
  1 0.5906E+05 0.3501E+06  .14535+06 .24196+06
.35011+06
  2 0.6830E+05 0.2186E+05  .11757+05
  3 0.7302E+05 0.1934E+05
  4 0.8528E+05 0.2406E+05
  5 0.9503E+05 0.3632E+05
  6 0.4896E+05 0.4607E+05                                 .48961+05

                         ****   MOLE-X-PROFILE     ****
  STAGE     NITRO-01      METHA-01      ETHAN-01      PROPA-01      N-BUT-01
     1    0.28555E-02   0.53204       0.29742       0.10887       0.19694E-
01
     2    0.38798E-03   0.40252       0.35175       0.12406       0.30364E-
01
     3    0.50826E-04   0.24364       0.49861       0.13374       0.31184E-
01
     4    0.65196E-05   0.10532       0.64006       0.13961       0.29279E-
01
     5    0.79640E-06   0.35522E-01   0.67508       0.16901       0.31686E-
01
     6    0.84206E-07   0.91053E-02   0.53353       0.22855       0.57203E-
01
```

**** MOLE-X-PROFILE ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.22024E-01	0.69031E-02	0.91558E-02	0.10356E-02
2	0.30424E-01	0.18157E-01	0.30252E-01	0.12092E-01
3	0.31380E-01	0.18455E-01	0.30710E-01	0.12231E-01
4	0.29777E-01	0.16889E-01	0.28018E-01	0.11046E-01
5	0.32979E-01	0.17090E-01	0.28042E-01	0.10592E-01
6	0.57201E-01	0.34322E-01	0.57203E-01	0.22881E-01

**** MOLE-Y-PROFILE ****

STAGE	NITRO-01	METHA-01	ETHAN-01	PROPA-01	N-BUT-01
1	0.29867E-01	0.94254	0.26340E-01	0.11714E-02	0.25891E-04
2	0.52420E-02	0.94521	0.47086E-01	0.22476E-02	0.74575E-04
3	0.82256E-03	0.84329	0.14808	0.69789E-02	0.29507E-03
4	0.10895E-03	0.51227	0.45862	0.25141E-01	0.13820E-02
5	0.12468E-04	0.19426	0.73852	0.57392E-01	0.34687E-02
6	0.13793E-05	0.57144E-01	0.79094	0.12027	0.10801E-01

**** MOLE-Y-PROFILE ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.48767E-04	0.17287E-05	0.17158E-05	0.22156E-07
2	0.11896E-03	0.96314E-05	0.11489E-04	0.55137E-06
3	0.42356E-03	0.46953E-04	0.57378E-04	0.34118E-05
4	0.18048E-02	0.28054E-03	0.36468E-03	0.30916E-04
5	0.44282E-02	0.77439E-03	0.10410E-02	0.10637E-03
6	0.13154E-01	0.29852E-02	0.41742E-02	0.53306E-03

**** K-VALUES ****

STAGE	NITRO-01	METHA-01	ETHAN-01	PROPA-01	N-BUT-01
1	10.459	1.7716	0.88565E-01	0.10761E-01	0.13148E-02
2	13.509	2.3482	0.13390	0.18126E-01	0.24576E-02
3	16.173	3.4604	0.29712	0.52229E-01	0.94750E-02
4	16.698	4.8618	0.71661	0.18017	0.47240E-01
5	15.650	5.4676	1.0940	0.33963	0.10950
6	16.379	6.2754	1.4824	0.52623	0.18883

**** K-VALUES ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.22145E-02	0.25045E-03	0.18743E-03	0.21399E-04
2	0.39122E-02	0.53084E-03	0.38006E-03	0.45642E-04
3	0.13514E-01	0.25485E-02	0.18716E-02	0.27958E-03
4	0.60657E-01	0.16629E-01	0.13031E-01	0.28032E-02
5	0.13430	0.45330E-01	0.37137E-01	0.10048E-01
6	0.22998	0.86984E-01	0.72979E-01	0.23300E-01

**** MASS-X-PROFILE ****

STAGE	NITRO-01	METHA-01	ETHAN-01	PROPA-01	N-BUT-01

1	0.30728E-02	0.32788	0.34355	0.18441	0.43972E-01
2	0.35537E-03	0.21114	0.34583	0.17887	0.57706E-01
3	0.43096E-04	0.11830	0.45381	0.17850	0.54861E-01
4	0.52608E-05	0.48669E-01	0.55438	0.17733	0.49020E-01
5	0.61545E-06	0.15721E-01	0.55999	0.20559	0.50806E-01
6	0.56851E-07	0.35205E-02	0.38665	0.24290	0.80130E-01

**** MASS-X-PROFILE ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.49174E-01	0.19133E-01	0.25376E-01	0.34281E-02
2	0.57819E-01	0.42835E-01	0.71369E-01	0.34071E-01
3	0.55206E-01	0.40303E-01	0.67066E-01	0.31902E-01
4	0.49853E-01	0.35099E-01	0.58229E-01	0.27420E-01
5	0.52880E-01	0.34015E-01	0.55815E-01	0.25180E-01
6	0.80128E-01	0.59682E-01	0.99469E-01	0.47523E-01

**** MASS-Y-PROFILE ****

STAGE	NITRO-01	METHA-01	ETHAN-01	PROPA-01	N-BUT-01
1	0.49785E-01	0.89974	0.47128E-01	0.30737E-02	0.89546E-04
2	0.87210E-02	0.90055	0.84085E-01	0.58860E-02	0.25742E-03
3	0.12549E-02	0.73679	0.24250	0.16760E-01	0.93404E-03
4	0.13068E-03	0.35189	0.59048	0.47468E-01	0.34394E-02
5	0.12275E-04	0.10953	0.78046	0.88943E-01	0.70857E-02
6	0.12091E-05	0.28686E-01	0.74420	0.16595	0.19644E-01

**** MASS-Y-PROFILE ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.16866E-03	0.74215E-05	0.73662E-05	0.11361E-06
2	0.41062E-03	0.41269E-04	0.49228E-04	0.28219E-05
3	0.13408E-02	0.18450E-03	0.22546E-03	0.16013E-04
4	0.44917E-02	0.86667E-03	0.11266E-02	0.11408E-03
5	0.90456E-02	0.19636E-02	0.26397E-02	0.32217E-03
6	0.23924E-01	0.67395E-02	0.94239E-02	0.14374E-02

```
BLOCK:  D-201    MODEL: RADFRAC
-------------------------------
   INLETS   - 22       STAGE   5
   OUTLETS  - LIGHT    STAGE   1
              HEAVY    STAGE  10
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                    ***  MASS AND ENERGY BALANCE  ***
                                IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)         1180.00          1180.00          0.00000
```

```
   MASS(LB/HR   )             48961.2        48961.2       0.386377E-14
   ENTHALPY(BTU/HR  )      -0.594620E+08   -0.535768E+08   -0.989744E-01
```

```
                    *********************
                    ****  INPUT DATA  ****
                    *********************
```

**** INPUT PARAMETERS ****

NUMBER OF STAGES	10
ALGORITHM OPTION	STANDARD
ABSORBER OPTION	NO
INITIALIZATION OPTION	STANDARD
HYDRAULIC PARAMETER CALCULATIONS	NO
INSIDE LOOP CONVERGENCE METHOD	BROYDEN
DESIGN SPECIFICATION METHOD	NESTED
MAXIMUM NO. OF OUTSIDE LOOP ITERATIONS	25
MAXIMUM NO. OF INSIDE LOOP ITERATIONS	10
MAXIMUM NUMBER OF FLASH ITERATIONS	50
FLASH TOLERANCE	0.000100000
OUTSIDE LOOP CONVERGENCE TOLERANCE	0.000100000

**** COL-SPECS ****

MOLAR VAPOR DIST / TOTAL DIST		1.00000
MOLAR REFLUX RATIO		2.00000
MOLAR BOTTOMS RATE	LBMOL/HR	270.000

**** PROFILES ****

P-SPEC	STAGE 1 PRES, PSIA	190.000

```
                     ******************
                     ****  RESULTS  ****
                     ******************
```

*** COMPONENT SPLIT FRACTIONS ***

	OUTLET STREAMS	
	LIGHT	HEAVY
COMPONENT:		
NITRO-01	1.0000	.16604E-09
METHA-01	1.0000	.34669E-07
ETHAN-01	.99994	.62352E-04
PROPA-01	.97510	.24903E-01
N-BUT-01	.17064E-01	.98294
ISOBU-01	.82811E-01	.91719
2-MET-01	.14485E-03	.99986
N-PEN-01	.64440E-04	.99994
N-HEX-01	.12401E-06	1.0000

*** SUMMARY OF KEY RESULTS ***

TOP STAGE TEMPERATURE	F	41.8504
BOTTOM STAGE TEMPERATURE	F	228.297
TOP STAGE LIQUID FLOW	LBMOL/HR	1,820.00
BOTTOM STAGE LIQUID FLOW	LBMOL/HR	270.000
TOP STAGE VAPOR FLOW	LBMOL/HR	910.000
BOTTOM STAGE VAPOR FLOW	LBMOL/HR	2,244.42
MOLAR REFLUX RATIO		2.00000
MOLAR BOILUP RATIO		8.31265
CONDENSER DUTY (W/O SUBCOOL)	BTU/HR	-0.115982+08
REBOILER DUTY	BTU/HR	0.174833+08

**** MAXIMUM FINAL RELATIVE ERRORS ****

DEW POINT	0.22014E-04	STAGE= 6	
BUBBLE POINT	0.96172E-05	STAGE= 7	
COMPONENT MASS BALANCE	0.54354E-07	STAGE= 5	COMP=N-HEX-01
ENERGY BALANCE	0.43367E-04	STAGE= 7	

**** PROFILES ****

NOTE REPORTED VALUES FOR STAGE LIQUID AND VAPOR RATES ARE THE FLOWS FROM THE STAGE EXCLUDING ANY SIDE PRODUCT. FOR THE FIRST STAGE, THE REPORTED VAPOR FLOW IS THE VAPOR DISTILLATE FLOW. FOR THE LAST STAGE, THE REPORTED LIQUID FLOW IS THE LIQUID BOTTOMS FLOW.

STAGE	TEMPERATURE F	PRESSURE PSIA	ENTHALPY BTU/LBMOL LIQUID	ENTHALPY BTU/LBMOL VAPOR	HEAT DUTY BTU/HR
1	41.850	190.00	-48826.	-39739.	-.11598+08
2	67.619	191.00	-50979.	-41549.	
3	84.073	192.00	-52700.	-42799.	
4	97.604	193.00	-54562.	-43723.	
5	116.95	194.00	-57025.	-45056.	
6	148.11	195.00	-59065.	-48577.	
7	169.68	196.00	-60512.	-51037.	
8	187.25	197.00	-61709.	-52703.	
9	205.42	198.00	-62970.	-53915.	
10	228.30	199.00	-64498.	-54996.	.17483+08

STAGE	FLOW RATE LBMOL/HR LIQUID	FLOW RATE LBMOL/HR VAPOR	FEED RATE LBMOL/HR LIQUID	FEED RATE LBMOL/HR VAPOR	FEED RATE LBMOL/HR MIXED	PRODUCT RATE LBMOL/HR LIQUID	PRODUCT RATE LBMOL/HR VAPOR
1	1820.	910.0					
910.0000							
2	1758.	2730.					
3	1696.	2668.					
4	1604.	2606.		182.4713			
5	2578.	2332.	997.5286				
6	2631.	2308.					
7	2647.	2361.					
8	2610.	2377.					
9	2514.	2340.					
10	270.0	2244.				270.0000	

**** MASS FLOW PROFILES ****

STAGE	FLOW RATE LB/HR LIQUID	FLOW RATE LB/HR VAPOR	FEED RATE LB/HR LIQUID	FEED RATE LB/HR VAPOR	FEED RATE LB/HR MIXED	PRODUCT RATE LB/HR LIQUID	PRODUCT RATE LB/HR VAPOR
1	0.7101E+05	0.3109E+05					.31091+05
2	0.7434E+05	0.1021E+06					
3	0.7581E+05	0.1054E+06					
4	0.7573E+05	0.1069E+06		5863.3934			
5	0.1310E+06	0.1010E+06	.43098+05				
6	0.1431E+06	0.1131E+06					
7	0.1512E+06	0.1253E+06					
8	0.1556E+06	0.1334E+06					
9	0.1570E+06	0.1377E+06					
10	0.1787E+05	0.1391E+06				.17870+05	

**** MOLE-X-PROFILE ****

STAGE	NITRO-01	METHA-01	ETHAN-01	PROPA-01	N-BUT-01
1	0.40822E-08	0.12994E-02	0.39868	0.56118	0.78417E-02
2	0.14506E-08	0.48472E-03	0.23050	0.66818	0.23439E-01
3	0.14196E-08	0.41972E-03	0.15976	0.64163	0.52399E-01
4	0.14457E-08	0.40936E-03	0.12998	0.54043	0.96249E-01
5	0.28425E-09	0.17560E-03	0.98604E-01	0.41300	0.14663
6	0.12430E-10	0.17066E-04	0.31444E-01	0.31028	0.21361
7	0.56798E-12	0.16482E-05	0.91527E-02	0.19972	0.27075
8	0.26667E-13	0.15815E-06	0.24874E-02	0.11399	0.30341
9	0.12718E-14	0.14979E-07	0.63018E-03	0.57497E-01	0.29722
10	0.0000	0.13796E-08	0.14539E-03	0.24875E-01	0.24573

**** MOLE-X-PROFILE ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.30828E-01	0.91714E-04	0.79372E-04	0.24734E-06
2	0.76285E-01	0.55545E-03	0.54723E-03	0.59579E-05
3	0.14043	0.24811E-02	0.27886E-02	0.98985E-04
4	0.21094	0.90477E-02	0.11681E-01	0.12605E-02
5	0.26406	0.26402E-01	0.39316E-01	0.11809E-01
6	0.34799	0.35075E-01	0.49244E-01	0.12336E-01
7	0.38638	0.50842E-01	0.69250E-01	0.13903E-01
8	0.37532	0.77280E-01	0.10777	0.19746E-01
9	0.31871	0.11415	0.17131	0.40477E-01
10	0.22929	0.14998	0.24998	0.10000

**** MOLE-Y-PROFILE ****

STAGE	NITRO-01	METHA-01	ETHAN-01	PROPA-01	N-BUT-01
1	0.10919E-06	0.11807E-01	0.69179	0.28899	0.12657E-02
2	0.39118E-07	0.48019E-02	0.49639	0.47045	0.56497E-02
3	0.38195E-07	0.43461E-02	0.38782	0.53886	0.15877E-01
4	0.39054E-07	0.43962E-02	0.34555	0.51848	0.34543E-01

5	0.77315E-08	0.19779E-02	0.29583	0.47500	0.65958E-01
6	0.31750E-09	0.19614E-03	0.11012	0.45840	0.13504
7	0.13851E-10	0.19018E-04	0.35023E-01	0.34293	0.20993
8	0.63249E-12	0.18353E-05	0.10176E-01	0.21958	0.27359
9	0.29737E-13	0.17624E-06	0.27576E-02	0.12428	0.31006
10	0.14175E-14	0.16615E-07	0.68850E-03	0.61421E-01	0.30341

**** MOLE-Y-PROFILE ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.61423E-02	0.64468E-05	0.47799E-05	0.36795E-08
2	0.22599E-01	0.63292E-04	0.54508E-04	0.16612E-06
3	0.52363E-01	0.36821E-03	0.36223E-03	0.39272E-05
4	0.93533E-01	0.16169E-02	0.18164E-02	0.64420E-04
5	0.14657	0.60419E-02	0.77883E-02	0.84115E-03
6	0.26812	0.11947E-01	0.14673E-01	0.14926E-02
7	0.36157	0.21932E-01	0.26283E-01	0.23094E-02
8	0.40423	0.39581E-01	0.48720E-01	0.41230E-02
9	0.39217	0.68892E-01	0.91359E-01	0.10486E-01
10	0.32947	0.10984	0.16185	0.33317E-01

**** K-VALUES ****

STAGE	NITRO-01	METHA-01	ETHAN-01	PROPA-01	N-BUT-01
1	26.748	9.0864	1.7352	0.51496	0.16140
2	26.969	9.9068	2.1535	0.70406	0.24103
3	26.908	10.355	2.4276	0.83983	0.30299
4	27.016	10.740	2.6586	0.95938	0.35888
5	27.204	11.265	3.0003	1.1501	0.44979
6	25.550	11.495	3.5025	1.4774	0.63217
7	24.398	11.542	3.8272	1.7171	0.77536
8	23.715	11.604	4.0908	1.9263	0.90173
9	23.380	11.765	4.3757	2.1614	1.0432
10	23.196	12.043	4.7355	2.4692	1.2347

**** K-VALUES ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.19924	0.70292E-01	0.60222E-01	0.14876E-01
2	0.29624	0.11394	0.99605E-01	0.27881E-01
3	0.37287	0.14840	0.12989	0.39672E-01
4	0.44339	0.17870	0.15550	0.51100E-01
5	0.55504	0.22883	0.19808	0.71221E-01
6	0.77046	0.34058	0.29794	0.12097
7	0.93578	0.43132	0.37948	0.16607
8	1.0770	0.51218	0.45208	0.20881
9	1.2305	0.60353	0.53330	0.25907
10	1.4369	0.73238	0.64744	0.33317

**** MASS-X-PROFILE ****

STAGE	NITRO-01	METHA-01	ETHAN-01	PROPA-01	N-BUT-01
1	0.29311E-08	0.53431E-03	0.30727	0.63427	0.11682E-01
2	0.96110E-09	0.18392E-03	0.16393	0.69689	0.32223E-01
3	0.88963E-09	0.15063E-03	0.10747	0.63296	0.68134E-01
4	0.85796E-09	0.13912E-03	0.82795E-01	0.50484	0.11851
5	0.15672E-09	0.55443E-04	0.58355E-01	0.35844	0.16774

6	0.63986E-11	0.50312E-05	0.17375E-01	0.25144	0.22816
7	0.27846E-12	0.46277E-06	0.48166E-02	0.15413	0.27541
8	0.12533E-13	0.42566E-07	0.12548E-02	0.84334E-01	0.29587
9	0.57061E-15	0.38486E-08	0.30348E-03	0.40606E-01	0.27668
10	0.0000	0.33440E-09	0.66052E-04	0.16573E-01	0.21580

**** MASS-X-PROFILE ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.45927E-01	0.16961E-03	0.14678E-03	0.54633E-06
2	0.10487	0.94787E-03	0.93385E-03	0.12144E-04
3	0.18259	0.40047E-02	0.45009E-02	0.19083E-03
4	0.25973	0.13829E-01	0.17853E-01	0.23012E-02
5	0.30207	0.37491E-01	0.55829E-01	0.20028E-01
6	0.37169	0.46505E-01	0.65291E-01	0.19536E-01
7	0.39304	0.64198E-01	0.87442E-01	0.20969E-01
8	0.36599	0.93547E-01	0.13045	0.28550E-01
9	0.29669	0.13190	0.19796	0.55866E-01
10	0.20136	0.16349	0.27251	0.13020

**** MASS-Y-PROFILE ****

STAGE	NITRO-01	METHA-01	ETHAN-01	PROPA-01	N-BUT-01
1	0.89528E-07	0.55440E-02	0.60885	0.37298	0.21532E-02
2	0.29302E-07	0.20599E-02	0.39911	0.55470	0.87805E-02
3	0.27079E-07	0.17646E-02	0.29514	0.60137	0.23355E-01
4	0.26669E-07	0.17193E-02	0.25329	0.55735	0.48944E-01
5	0.50024E-08	0.73289E-03	0.20546	0.48378	0.88546E-01
6	0.18147E-09	0.64201E-04	0.67562E-01	0.41244	0.16015
7	0.73114E-11	0.57488E-05	0.19844E-01	0.28494	0.22992
8	0.31577E-12	0.52473E-06	0.54531E-02	0.17256	0.28339
9	0.14157E-13	0.48047E-07	0.14091E-02	0.93128E-01	0.30626
10	0.64058E-15	0.42999E-08	0.33398E-03	0.43693E-01	0.28450

**** MASS-Y-PROFILE ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.10449E-01	0.13614E-04	0.10094E-04	0.92809E-08
2	0.35123E-01	0.12210E-03	0.10516E-03	0.38279E-06
3	0.77027E-01	0.67236E-03	0.66143E-03	0.85653E-05
4	0.13253	0.28439E-02	0.31948E-02	0.13533E-03
5	0.19677	0.10068E-01	0.12979E-01	0.16742E-02
6	0.31797	0.17587E-01	0.21601E-01	0.26245E-02
7	0.39599	0.29817E-01	0.35732E-01	0.37501E-02
8	0.41872	0.50894E-01	0.62645E-01	0.63321E-02
9	0.38736	0.84469E-01	0.11202	0.15356E-01
10	0.30893	0.12785	0.18838	0.46318E-01

```
BLOCK:  D-202    MODEL: RADFRAC
-------------------------------
   INLETS   - LIGHT    STAGE   6
   OUTLETS  - ETH      STAGE   1
              PROP     STAGE  10
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE
```

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(LBMOL/HR)	910.000	910.000	0.00000
MASS(LB/HR)	31091.0	31091.0	-0.271781E-11
ENTHALPY(BTU/HR)	-0.361622E+08	-0.379251E+08	0.464836E-01

****
**** INPUT DATA ****
****

**** INPUT PARAMETERS ****

NUMBER OF STAGES	10
ALGORITHM OPTION	STANDARD
ABSORBER OPTION	NO
INITIALIZATION OPTION	STANDARD
HYDRAULIC PARAMETER CALCULATIONS	NO
INSIDE LOOP CONVERGENCE METHOD	BROYDEN
DESIGN SPECIFICATION METHOD	NESTED
MAXIMUM NO. OF OUTSIDE LOOP ITERATIONS	25
MAXIMUM NO. OF INSIDE LOOP ITERATIONS	10
MAXIMUM NUMBER OF FLASH ITERATIONS	50
FLASH TOLERANCE	0.000100000
OUTSIDE LOOP CONVERGENCE TOLERANCE	0.000100000

**** COL-SPECS ****

MOLAR VAPOR DIST / TOTAL DIST		1.00000
MOLAR REFLUX RATIO		2.00000
MOLAR BOTTOMS RATE	LBMOL/HR	270.000

**** PROFILES ****

P-SPEC STAGE 1 PRES, PSIA 180.000

****
**** RESULTS ****
****

*** COMPONENT SPLIT FRACTIONS ***

	OUTLET STREAMS	
	ETH	PROP
COMPONENT:		
NITRO-01	1.0000	.20364E-08
METHA-01	1.0000	.79360E-06
ETHAN-01	.99013	.98666E-02
PROPA-01	.22585E-01	.97742
N-BUT-01	.25083E-04	.99997
ISOBU-01	.73514E-04	.99993
2-MET-01	.15989E-06	1.0000
N-PEN-01	.54707E-07	1.0000

N-HEX-01   .22606E-10   1.0000

***   SUMMARY OF KEY RESULTS   ***

TOP STAGE TEMPERATURE	F	-11.5930
BOTTOM STAGE TEMPERATURE	F	97.3295
TOP STAGE LIQUID FLOW	LBMOL/HR	1,280.00
BOTTOM STAGE LIQUID FLOW	LBMOL/HR	270.000
TOP STAGE VAPOR FLOW	LBMOL/HR	640.000
BOTTOM STAGE VAPOR FLOW	LBMOL/HR	773.733
MOLAR REFLUX RATIO		2.00000
MOLAR BOILUP RATIO		2.86568
CONDENSER DUTY (W/O SUBCOOL)	BTU/HR	-6,479,160.
REBOILER DUTY	BTU/HR	4,716,270.

****   MAXIMUM FINAL RELATIVE ERRORS   ****

DEW POINT	0.24815E-04	STAGE= 5	
BUBBLE POINT	0.38571E-05	STAGE= 5	
COMPONENT MASS BALANCE	0.29411E-05	STAGE= 5	COMP=METHA-01
ENERGY BALANCE	0.79410E-04	STAGE= 7	

****   PROFILES   ****

NOTE REPORTED VALUES FOR STAGE LIQUID AND VAPOR RATES ARE THE FLOWS FROM THE STAGE EXCLUDING ANY SIDE PRODUCT. FOR THE FIRST STAGE, THE REPORTED VAPOR FLOW IS THE VAPOR DISTILLATE FLOW. FOR THE LAST STAGE, THE REPORTED LIQUID FLOW IS THE LIQUID BOTTOMS FLOW.

STAGE	TEMPERATURE F	PRESSURE PSIA	ENTHALPY BTU/LBMOL LIQUID	ENTHALPY BTU/LBMOL VAPOR	HEAT DUTY BTU/HR
1	-11.593	180.00	-42873.	-37567.	-.64792+07
2	-6.0850	181.00	-43587.	-37730.	
3	3.7383	182.00	-44934.	-38041.	
4	18.448	183.00	-46631.	-38639.	
5	33.566	184.00	-48047.	-39424.	
6	50.512	185.00	-49193.	-40558.	
9	91.428	188.00	-51124.	-44318.	
10	97.329	189.00	-51415.	-44927.	.47163+07

STAGE	FLOW RATE LBMOL/HR LIQUID	FLOW RATE LBMOL/HR VAPOR	FEED RATE LBMOL/HR LIQUID	FEED RATE LBMOL/HR VAPOR	FEED RATE LBMOL/HR MIXED	PRODUCT RATE LBMOL/HR LIQUID	PRODUCT RATE LBMOL/HR VAPOR
1	1280.	640.0					
640.0000							
2	1223.	1920.					
3	1138.	1863.					
4	1064.	1778.					
5	1021.	1704.		910.0000			
6	1009.	751.0					
9	1044.	760.1					
10	270.0	773.7				270.0000	

```
   ****  MASS FLOW PROFILES  ****

 STAGE     FLOW RATE                   FEED RATE                 PRODUCT RATE
            LB/HR                       LB/HR                      LB/HR
       LIQUID     VAPOR       LIQUID    VAPOR    MIXED      LIQUID    VAPOR
   1 0.3913E+05 0.1918E+05
.19177+05
   2 0.3859E+05 0.5830E+05
   3 0.3801E+05 0.5776E+05
   4 0.3810E+05 0.5719E+05
   5 0.3872E+05 0.5728E+05          .31091+05
   6 0.4020E+05 0.2681E+05
   9 0.4549E+05 0.3218E+05
  10 0.1191E+05 0.3357E+05                                  .11914+05

                         ****   MOLE-X-PROFILE     ****
   STAGE     NITRO-01      METHA-01      ETHAN-01      PROPA-01      N-BUT-01
      1     0.66604E-08   0.24434E-02   0.95964       0.37912E-01   0.69829E-
06
      2     0.23479E-08   0.10119E-02   0.89237       0.10654       0.67256E-
05
      3     0.21778E-08   0.83855E-03   0.76159       0.23711       0.52142E-
04
      4     0.21540E-08   0.78336E-03   0.59136       0.40566       0.30309E-
03
      5     0.21781E-08   0.75170E-03   0.44641       0.54479       0.13248E-
02
      6     0.10738E-09   0.10643E-03   0.31243       0.67875       0.14157E-
02
      9     0.14372E-13   0.25054E-06   0.51199E-01   0.93434       0.22769E-
02
     10      0.0000       0.31580E-07   0.23005E-01   0.95199       0.42657E-
02

                         ****   MOLE-X-PROFILE     ****
   STAGE     ISOBU-01      2-MET-01      N-PEN-01      N-HEX-01
      1     0.81054E-05   0.68950E-10   0.22699E-10    0.0000
      2     0.64008E-04   0.19034E-08   0.79826E-09   0.63137E-14
      3     0.40292E-03   0.40409E-07   0.21056E-07   0.79983E-12
      4     0.18955E-02   0.59479E-06   0.37272E-06   0.64374E-10
      5     0.67161E-02   0.61470E-05   0.45170E-05   0.33265E-08
      6     0.72924E-02   0.63836E-05   0.46815E-05   0.33899E-08
      9     0.12165E-01   0.87490E-05   0.62794E-05   0.37152E-08
     10     0.20700E-01   0.21728E-04   0.16110E-04   0.12401E-07

                         ****   MOLE-Y-PROFILE     ****
   STAGE     NITRO-01      METHA-01      ETHAN-01      PROPA-01      N-BUT-01
      1     0.15525E-06   0.16788E-01   0.97393       0.92801E-02   0.45141E-
07
      2     0.56192E-07   0.72249E-02   0.96440       0.28368E-01   0.48057E-
06
      3     0.54878E-07   0.64316E-02   0.92039       0.73130E-01   0.44306E-
05
      4     0.57276E-07   0.65793E-02   0.83802       0.15511       0.33390E-
04
      5     0.59658E-07   0.67946E-02   0.73505       0.25678       0.18927E-
03
```

6	0.29613E-08	0.10219E-02	0.59864	0.39838	0.26744E-03
9	0.37329E-12	0.25897E-05	0.13154	0.86265	0.78330E-03
10	0.19125E-13	0.32694E-06	0.61037E-01	0.92819	0.15829E-02

**** MOLE-Y-PROFILE ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.64204E-06	0.14656E-11	0.37181E-12	0.0000
2	0.56176E-05	0.46455E-10	0.15257E-10	0.0000
3	0.42239E-04	0.12500E-08	0.52415E-09	0.41447E-14
4	0.25813E-03	0.25865E-07	0.13477E-07	0.51194E-12
5	0.11838E-02	0.37139E-06	0.23273E-06	0.40195E-10
6	0.16883E-02	0.54514E-06	0.34896E-06	0.63815E-10
9	0.50175E-02	0.15514E-05	0.10111E-05	0.18131E-09
10	0.91859E-02	0.42199E-05	0.28490E-05	0.68407E-09

**** K-VALUES ****

STAGE	NITRO-01	METHA-01	ETHAN-01	PROPA-01	N-BUT-01
1	23.310	6.8706	1.0149	0.24478	0.64646E-01
2	23.933	7.1401	1.0807	0.26626	0.71455E-01
3	25.198	7.6698	1.2085	0.30842	0.84972E-01
4	26.591	8.3991	1.4171	0.38235	0.11016
5	27.395	9.0404	1.6466	0.47131	0.14284
6	27.582	9.6039	1.9162	0.58691	0.18888
9	25.976	10.337	2.5693	0.92327	0.34401
10	25.520	10.353	2.6532	0.97499	0.37106

**** K-VALUES ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.79212E-01	0.21257E-01	0.16380E-01	0.32437E-02
2	0.87765E-01	0.24407E-01	0.19113E-01	0.38563E-02
3	0.10483	0.30934E-01	0.24894E-01	0.51823E-02
4	0.13617	0.43484E-01	0.36157E-01	0.79519E-02
5	0.17624	0.60403E-01	0.51508E-01	0.12079E-01
6	0.23149	0.85379E-01	0.74523E-01	0.18819E-01
9	0.41246	0.17731	0.16101	0.48798E-01
10	0.44376	0.19421	0.17684	0.55160E-01

**** MASS-X-PROFILE ****

STAGE	NITRO-01	METHA-01	ETHAN-01	PROPA-01	N-BUT-01
1	0.61040E-08	0.12824E-02	0.94401	0.54692E-01	0.13278E-05
2	0.20845E-08	0.51449E-03	0.85045	0.14890	0.12390E-04
3	0.18268E-08	0.40282E-03	0.68572	0.31308	0.90748E-04
4	0.16850E-08	0.35094E-03	0.49655	0.49952	0.49193E-03
5	0.16088E-08	0.31796E-03	0.35393	0.63341	0.20303E-02
6	0.75519E-10	0.42862E-04	0.23584	0.75139	0.20658E-02

9	0.92379E-14	0.92225E-07	0.35325E-01	0.94539	0.30366E-02
10	0.0000	0.11482E-07	0.15677E-01	0.95137	0.56190E-02

**** MASS-X-PROFILE ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.15412E-04	0.16275E-09	0.53578E-10	0.0000
2	0.11791E-03	0.43524E-08	0.18254E-08	0.17245E-13
3	0.70125E-03	0.87301E-07	0.45490E-07	0.20639E-11
4	0.30766E-02	0.11984E-05	0.75095E-06	0.15491E-09
5	0.10292E-01	0.11694E-04	0.85930E-05	0.75585E-08
6	0.10641E-01	0.11563E-04	0.84796E-05	0.73338E-08
9	0.16224E-01	0.14484E-04	0.10396E-04	0.73463E-08
10	0.27267E-01	0.35528E-04	0.26342E-04	0.24220E-07

**** MASS-Y-PROFILE ****

STAGE	NITRO-01	METHA-01	ETHAN-01	PROPA-01	N-BUT-01
1	0.14515E-06	0.89881E-02	0.97735	0.13657E-01	0.87563E-07
2	0.51838E-07	0.38170E-02	0.95498	0.41194E-01	0.91985E-06
3	0.49581E-07	0.33277E-02	0.89258	0.10400	0.83053E-05
4	0.49890E-07	0.32819E-02	0.78352	0.21267	0.60345E-04
5	0.49717E-07	0.32428E-02	0.65753	0.33685	0.32726E-03
6	0.23238E-08	0.45926E-03	0.50425	0.49210	0.43545E-03
9	0.24702E-12	0.98140E-06	0.93434E-01	0.89860	0.10755E-02
10	0.12347E-13	0.12088E-06	0.42298E-01	0.94327	0.21203E-02

**** MASS-Y-PROFILE ****

STAGE	ISOBU-01	2-MET-01	N-PEN-01	N-HEX-01
1	0.12454E-05	0.35291E-11	0.89528E-12	0.0000
2	0.10752E-04	0.11038E-09	0.36250E-10	0.0000
3	0.79179E-04	0.29086E-08	0.12197E-08	0.11520E-13
4	0.46650E-03	0.58025E-07	0.30235E-07	0.13718E-11
5	0.20469E-02	0.79714E-06	0.49953E-06	0.10305E-09
6	0.27488E-02	0.11018E-05	0.70529E-06	0.15405E-09
9	0.68891E-02	0.26441E-05	0.17233E-05	0.36909E-09
10	0.12305E-01	0.70167E-05	0.47372E-05	0.13586E-08

```
BLOCK:  D-203    MODEL: RADFRAC
-------------------------------
   INLETS    - 23       STAGE   5
   OUTLETS   - BUTS     STAGE   1
               PENTANE  STAGE  10
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                    ***  MASS AND ENERGY BALANCE  ***
                                   IN              OUT         RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)             270.000         270.000         0.210531E-15
```

```
   MASS(LB/HR   )           17870.2        17870.2       -0.712522E-13
   ENTHALPY(BTU/HR  )      -0.174146E+08   -0.169245E+08   -0.281407E-01


                    *********************
                    ****  INPUT DATA  ****
                    *********************

**** INPUT PARAMETERS ****

 NUMBER OF STAGES                                                10
 ALGORITHM OPTION                                              STANDARD
 ABSORBER OPTION                                               NO
 INITIALIZATION OPTION                                         STANDARD
 HYDRAULIC PARAMETER CALCULATIONS                              NO
 INSIDE LOOP CONVERGENCE METHOD                                BROYDEN
 DESIGN SPECIFICATION METHOD                                   NESTED
 MAXIMUM NO. OF OUTSIDE LOOP ITERATIONS                          25
 MAXIMUM NO. OF INSIDE LOOP ITERATIONS                           10
 MAXIMUM NUMBER OF FLASH ITERATIONS                              50
 FLASH TOLERANCE                                                  0.000100000
 OUTSIDE LOOP CONVERGENCE TOLERANCE                               0.000100000

**** COL-SPECS ****

 MOLAR VAPOR DIST / TOTAL DIST                                    1.00000
 MOLAR REFLUX RATIO                                               4.00000
 MOLAR BOTTOMS RATE            LBMOL/HR                         135.000

****  PROFILES  ****

 P-SPEC          STAGE  1  PRES, PSIA                            90.0000

                     ******************
                     ****  RESULTS  ****
                     ******************


*** COMPONENT SPLIT FRACTIONS ***

                          OUTLET STREAMS
                          --------------
             BUTS           PENTANE
 COMPONENT:
 METHA-01    1.0000       .15177E-09
 ETHAN-01    1.0000       .42673E-06
 PROPA-01    .99990       .95251E-04
 N-BUT-01    .93944       .60562E-01
 ISOBU-01    .98552       .14479E-01
 2-MET-01    .75604E-01   .92440
 N-PEN-01    .27255E-01   .97274
 N-HEX-01    .13926E-03   .99986


***  SUMMARY OF KEY RESULTS  ***
```

```
   TOP STAGE TEMPERATURE          F                        129.701
   BOTTOM STAGE TEMPERATURE       F                        224.291
   TOP STAGE LIQUID FLOW          LBMOL/HR                 540.000
   BOTTOM STAGE LIQUID FLOW       LBMOL/HR                 135.000
   TOP STAGE VAPOR FLOW           LBMOL/HR                 135.000
   BOTTOM STAGE VAPOR FLOW        LBMOL/HR                 523.575
   MOLAR REFLUX RATIO                                        4.00000
   MOLAR BOILUP RATIO                                        3.87834
   CONDENSER DUTY (W/O SUBCOOL)   BTU/HR            -4,490,640.
   REBOILER DUTY                  BTU/HR             4,980,650.

  ****   MAXIMUM FINAL RELATIVE ERRORS   ****

   DEW POINT                        0.72514E-04  STAGE=  6
   BUBBLE POINT                     0.20589E-04  STAGE=  6
   COMPONENT MASS BALANCE           0.27816E-05  STAGE=  4 COMP=ETHAN-01
   ENERGY BALANCE                   0.49973E-04  STAGE=  1


  ****    PROFILES   ****

  **NOTE** REPORTED VALUES FOR STAGE LIQUID AND VAPOR RATES ARE THE FLOWS
           FROM THE STAGE EXCLUDING ANY SIDE PRODUCT.  FOR THE FIRST STAGE,
           THE REPORTED VAPOR FLOW IS THE VAPOR DISTILLATE FLOW.  FOR THE
           LAST STAGE, THE REPORTED LIQUID FLOW IS THE LIQUID BOTTOMS FLOW.

                                              ENTHALPY
 STAGE TEMPERATURE   PRESSURE                BTU/LBMOL           HEAT DUTY
       F             PSIA              LIQUID        VAPOR         BTU/HR

   1   129.70        90.000        -63604.      -54990.       -.44906+07
   2   137.49        91.000        -64224.      -55227.
   3   146.69        92.000        -65165.      -55537.
   4   158.05        93.000        -66366.      -56029.
   5   171.46        94.000        -67588.      -56698.
   6   182.67        95.000        -68294.      -57430.
   9   213.93        98.000        -69833.      -59693.
  10   224.29        99.000        -70377.      -60180.         .49807+07

 STAGE     FLOW RATE                  FEED RATE                PRODUCT RATE
             LBMOL/HR                   LBMOL/HR                  LBMOL/HR
       LIQUID      VAPOR       LIQUID     VAPOR    MIXED       LIQUID    VAPOR
   1  540.0      135.0
135.0000
   2  525.6      675.0
   3  507.0      660.6
   4  485.9      642.0                    86.7366
   5  662.5      534.2      183.2633
   6  660.0      527.5
   9  658.6      526.5
  10  135.0      523.6                                          135.0000

   ****  MASS FLOW PROFILES  ****

 STAGE     FLOW RATE                  FEED RATE                PRODUCT RATE
             LB/HR                      LB/HR                     LB/HR
       LIQUID      VAPOR       LIQUID     VAPOR    MIXED       LIQUID    VAPOR
```

```
   1 0.3190E+05  7820.
7820.2135
   2 0.3175E+05 0.3972E+05
   3 0.3154E+05 0.3957E+05
   4 0.3133E+05 0.3936E+05           5414.6811
   5 0.4435E+05 0.3374E+05  .12456+05
   6 0.4526E+05 0.3430E+05
   9 0.4795E+05 0.3724E+05
  10 0.1005E+05 0.3790E+05                                        .10050+05

                        ****   MOLE-X-PROFILE     ****
   STAGE     METHA-01       ETHAN-01       PROPA-01       N-BUT-01       ISOBU-01
      1     0.10334E-09   0.42944E-04   0.20174E-01   0.50737        0.38459
      2     0.23621E-10   0.13122E-04   0.98441E-02   0.50310        0.31477
      3     0.21483E-10   0.94671E-05   0.62418E-02   0.45190        0.24891
      4     0.21686E-10   0.88058E-05   0.48323E-02   0.36742        0.19071
      5     0.24106E-11   0.27714E-05   0.25666E-02   0.28375        0.13654
      6     0.10750E-12   0.40202E-06   0.83810E-03   0.22197        0.89735E-
01
      9      0.0000       0.99224E-09   0.19260E-04   0.57706E-01   0.14841E-
01
     10      0.0000       0.12408E-09   0.47387E-05   0.29764E-01   0.66397E-
02

                        ****   MOLE-X-PROFILE     ****
   STAGE     2-MET-01       N-PEN-01       N-HEX-01
      1     0.51607E-01   0.36003E-01   0.21006E-03
      2     0.95142E-01   0.75973E-01   0.11614E-02
      3     0.14987       0.13766       0.54005E-02
      4     0.20142       0.21474       0.20856E-01
      5     0.23866       0.28402       0.54468E-01
      6     0.28538       0.34317       0.58915E-01
      9     0.31887       0.48915       0.11941
     10     0.27728       0.48634       0.19997

                        ****   MOLE-Y-PROFILE     ****
   STAGE     METHA-01       ETHAN-01       PROPA-01       N-BUT-01       ISOBU-01
      1     0.27592E-08   0.29077E-03   0.49744E-01   0.46170        0.45193
      2     0.63451E-09   0.92510E-04   0.26088E-01   0.49823        0.39806
      3     0.58270E-09   0.69867E-04   0.17999E-01   0.49464        0.34280
      4     0.59721E-09   0.68624E-04   0.15390E-01   0.45396        0.29160
      5     0.67630E-10   0.22952E-04   0.90955E-02   0.40111        0.23548
      6     0.30275E-11   0.34806E-05   0.32222E-02   0.34875        0.16978
      9      0.0000       0.94778E-08   0.89304E-04   0.11747        0.35379E-
01
     10      0.0000       0.12161E-08   0.23004E-04   0.64910E-01   0.16955E-
01

                        ****   MOLE-Y-PROFILE     ****
   STAGE     2-MET-01       N-PEN-01       N-HEX-01
      1     0.22678E-01   0.13627E-01   0.27853E-04
      2     0.45821E-01   0.31528E-01   0.17361E-03
      3     0.80332E-01   0.63231E-01   0.92976E-03
      4     0.12313       0.11158       0.42707E-02
      5     0.16980       0.17088       0.13608E-01
      6     0.22878       0.23224       0.17232E-01
      9     0.34623       0.44854       0.52286E-01
```

10	0.32959	0.48988	0.98640E-01

**** K-VALUES ****

STAGE	METHA-01	ETHAN-01	PROPA-01	N-BUT-01	ISOBU-01
1	26.694	6.7703	2.4658	0.91002	1.1750
2	26.842	7.0475	2.6500	0.99038	1.2645
3	27.088	7.3751	2.8830	1.0946	1.3770
4	27.497	7.7867	3.1839	1.2355	1.5287
5	28.014	8.2745	3.5426	1.4135	1.7242
6	28.117	8.6491	3.8427	1.5709	1.8915
9	28.191	9.5512	4.6365	2.0356	2.3838
10	28.342	9.8013	4.8546	2.1808	2.5537

**** K-VALUES ****

STAGE	2-MET-01	N-PEN-01	N-HEX-01
1	0.43944	0.37851	0.13261
2	0.48165	0.41508	0.14953
3	0.53608	0.45947	0.17226
4	0.61138	0.51974	0.20490
5	0.71159	0.60176	0.25000
6	0.80177	0.67685	0.29269
9	1.0858	0.91698	0.43787
10	1.1887	1.0073	0.49326

**** MASS-X-PROFILE ****

STAGE	METHA-01	ETHAN-01	PROPA-01	N-BUT-01	ISOBU-01
1	0.28064E-10	0.21859E-04	0.15059E-01	0.49920	0.37841
2	0.62722E-11	0.65308E-05	0.71848E-02	0.48399	0.30281
3	0.55392E-11	0.45752E-05	0.44236E-02	0.42215	0.23252
4	0.53957E-11	0.41066E-05	0.33048E-02	0.33121	0.17192
5	0.57766E-12	0.12448E-05	0.16906E-02	0.24635	0.11854
6	0.25147E-13	0.17627E-06	0.53889E-03	0.18812	0.76052E-01
9	0.0000	0.40980E-09	0.11665E-04	0.46068E-01	0.11848E-01
10	0.0000	0.50119E-10	0.28069E-05	0.23239E-01	0.51840E-02

**** MASS-X-PROFILE ****

STAGE	2-MET-01	N-PEN-01	N-HEX-01
1	0.63030E-01	0.43973E-01	0.30643E-03
2	0.11362	0.90726E-01	0.16566E-02
3	0.17379	0.15963	0.74799E-02
4	0.22539	0.24029	0.27875E-01
5	0.25721	0.30609	0.70114E-01
6	0.30023	0.36103	0.74031E-01
9	0.31599	0.48474	0.14134
10	0.26873	0.47135	0.23149

**** MASS-Y-PROFILE ****

STAGE	METHA-01	ETHAN-01	PROPA-01	N-BUT-01	ISOBU-01
1	0.76415E-09	0.15094E-03	0.37867E-01	0.46326	0.45346
2	0.17299E-09	0.47273E-04	0.19550E-01	0.49213	0.39318
3	0.15604E-09	0.35068E-04	0.13248E-01	0.47990	0.33258
4	0.15625E-09	0.33653E-04	0.11068E-01	0.43031	0.27641
5	0.17179E-10	0.10928E-04	0.63507E-02	0.36915	0.21672
6	0.74689E-12	0.16095E-05	0.21850E-02	0.31172	0.15175

9	0.0000	0.40289E-08	0.55671E-04	0.96527E-01	0.29070E-01
10	0.0000	0.50518E-09	0.14014E-04	0.52121E-01	0.13615E-01

**** MASS-Y-PROFILE ****

STAGE	2-MET-01	N-PEN-01	N-HEX-01
1	0.28246E-01	0.16972E-01	0.41435E-04
2	0.56182E-01	0.38657E-01	0.25426E-03
3	0.96747E-01	0.76151E-01	0.13374E-02
4	0.14488	0.13129	0.60021E-02
5	0.19399	0.19521	0.18568E-01
6	0.25383	0.25768	0.22836E-01
9	0.35315	0.45750	0.63698E-01
10	0.32853	0.48829	0.11743

BLOCK: E-101 MODEL: COMPR

INLET STREAM: FEED
OUTLET STREAM: LPFEED
PROPERTY OPTION SET: PENG-ROB STANDARD PR EQUATION OF STATE

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(LBMOL/HR)	13500.0	13500.0	0.00000
MASS(LB/HR)	253718.	253718.	0.00000
ENTHALPY(BTU/HR)	-0.437132E+09	-0.445822E+09	0.194927E-01

*** INPUT DATA ***

ISENTROPIC TURBINE	
OUTLET PRESSURE PSIA	300.000
ISENTROPIC EFFICIENCY	0.88000
MECHANICAL EFFICIENCY	1.00000

*** RESULTS ***

INDICATED HORSEPOWER REQUIREMENT	HP	-3,415.41
BRAKE HORSEPOWER REQUIREMENT	HP	-3,415.41
NET WORK REQUIRED	HP	-3,415.41
POWER LOSSES	HP	0.0
ISENTROPIC HORSEPOWER REQUIREMENT	HP	-3,881.15
CALCULATED OUTLET TEMP F		-15.8406
ISENTROPIC TEMPERATURE F		-22.7441
EFFICIENCY (POLYTR/ISENTR) USED		0.88000
OUTLET VAPOR FRACTION		0.98444
HEAD DEVELOPED,	FT-LBF/LB	-30,288.2
MECHANICAL EFFICIENCY USED		1.00000
INLET HEAT CAPACITY RATIO		1.50598
INLET VOLUMETRIC FLOW RATE , CUFT/HR		90,988.5
OUTLET VOLUMETRIC FLOW RATE, CUFT/HR		190,657.
INLET COMPRESSIBILITY FACTOR		0.86292
OUTLET COMPRESSIBILITY FACTOR		0.88954
AV. ISENT. VOL. EXPONENT		1.23076
AV. ISENT. TEMP EXPONENT		1.27204
AV. ACTUAL VOL. EXPONENT		1.19283

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  AV. ACTUAL TEMP EXPONENT                                 1.24393

BLOCK:  E-401    MODEL: COMPR
-----------------------------
  INLET STREAM:           HP-N2-C
  OUTLET STREAM:          COLDN2
  PROPERTY OPTION SET:    PENG-ROB  STANDARD PR EQUATION OF STATE

                    ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)           69012.1         69012.1         0.00000
      MASS(LB/HR   )         0.193327E+07    0.193327E+07     0.00000
      ENTHALPY(BTU/HR  )    -0.851661E+08   -0.151940E+09    0.439476

                         ***  INPUT DATA  ***

  ISENTROPIC TURBINE
   OUTLET PRESSURE  PSIA                                 130.000
   ISENTROPIC EFFICIENCY                                   0.88000
   MECHANICAL EFFICIENCY                                   1.00000

                         ***  RESULTS  ***

   INDICATED  HORSEPOWER REQUIREMENT  HP              -26,243.1
   BRAKE      HORSEPOWER REQUIREMENT  HP              -26,243.1
   NET WORK REQUIRED                  HP              -26,243.1
   POWER LOSSES                       HP                    0.0
   ISENTROPIC HORSEPOWER REQUIREMENT  HP              -29,821.7
   CALCULATED OUTLET TEMP  F                             -221.905
   ISENTROPIC TEMPERATURE  F                             -238.525
   EFFICIENCY (POLYTR/ISENTR) USED                         0.88000
   OUTLET VAPOR FRACTION                                   1.00000
   HEAD DEVELOPED,       FT-LBF/LB                    -30,542.6
   MECHANICAL EFFICIENCY USED                              1.00000
   INLET HEAT CAPACITY RATIO                               1.67096
   INLET VOLUMETRIC FLOW RATE , CUFT/HR               279,488.
   OUTLET VOLUMETRIC FLOW RATE, CUFT/HR             1,230,140.
   INLET  COMPRESSIBILITY FACTOR                           0.91658
   OUTLET COMPRESSIBILITY FACTOR                           0.90817
   AV. ISENT. VOL. EXPONENT                                1.47034
   AV. ISENT. TEMP EXPONENT                                1.43458
   AV. ACTUAL VOL. EXPONENT                                1.37335
   AV. ACTUAL TEMP EXPONENT                                1.36487

BLOCK:  E-701    MODEL: COMPR
-----------------------------
  INLET STREAM:           TOCO2EXP
  OUTLET STREAM:          CO2
  PROPERTY OPTION SET:    PENG-ROB  STANDARD PR EQUATION OF STATE

                    ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)           26000.0         26000.0         0.00000
      MASS(LB/HR   )         0.114425E+07    0.114425E+07     0.00000
      ENTHALPY(BTU/HR  )    -0.440540E+10   -0.442685E+10    0.484603E-02
```

```
                         ***  INPUT DATA  ***

  ISENTROPIC TURBINE
   OUTLET PRESSURE  PSIA                                        100.000
   ISENTROPIC EFFICIENCY                                          0.88000
   MECHANICAL EFFICIENCY                                          1.00000

                         ***  RESULTS  ***

   INDICATED  HORSEPOWER REQUIREMENT  HP                  -8,411.12
   BRAKE      HORSEPOWER REQUIREMENT  HP                  -8,411.12
   NET WORK REQUIRED                  HP                  -8,411.12
   POWER LOSSES                       HP                       0.0
   ISENTROPIC HORSEPOWER REQUIREMENT  HP                  -9,558.09
   CALCULATED OUTLET TEMP  F                                -28.7778
   ISENTROPIC TEMPERATURE  F                                -41.4920
   EFFICIENCY (POLYTR/ISENTR) USED                            0.88000
   OUTLET VAPOR FRACTION                                      1.00000
   HEAD DEVELOPED,       FT-LBF/LB                       -16,539.2
   MECHANICAL EFFICIENCY USED                                 1.00000
   INLET HEAT CAPACITY RATIO                                  1.44210
   INLET VOLUMETRIC FLOW RATE , CUFT/HR                  463,178.
   OUTLET VOLUMETRIC FLOW RATE, CUFT/HR                1,111,630.
   INLET  COMPRESSIBILITY FACTOR                              0.89092
   OUTLET COMPRESSIBILITY FACTOR                              0.92462
   AV. ISENT. VOL. EXPONENT                                   1.29127
   AV. ISENT. TEMP EXPONENT                                   1.33822
   AV. ACTUAL VOL. EXPONENT                                   1.23568
   AV. ACTUAL TEMP EXPONENT                                   1.29041

BLOCK:  F-101    MODEL: FLASH2
------------------------------
  INLET STREAM:          REFFEED
  OUTLET VAPOR STREAM:   REFOH
  OUTLET LIQUID STREAM:  REFLUX
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                    ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)          20832.7         20832.7        0.138020E-05
      MASS(LB/HR   )          350113.         350113.        0.158863E-05
      ENTHALPY(BTU/HR  )    -0.722228E+09   -0.722228E+09   -0.777145E-06

                        ***  INPUT DATA  ***
  TWO    PHASE  PQ  FLASH
  PRESSURE DROP         PSI                                       0.0
  SPECIFIED HEAT DUTY   BTU/HR                                    0.0
  MAXIMUM NO. ITERATIONS                                        30
  CONVERGENCE TOLERANCE                                           0.000100000

                         ***  RESULTS  ***
  OUTLET TEMPERATURE    F                                       -160.00
  OUTLET PRESSURE       PSIA                                     290.00
  VAPOR FRACTION                                                  0.59138
```

V-L PHASE EQUILIBRIUM :

COMP	F(I)	X(I)	Y(I)	K(I)
NITRO-01	0.29867E-01	0.96581E-02	0.43831E-01	4.5383
METHA-01	0.94254	0.92820	0.95246	1.0261
ETHAN-01	0.26340E-01	0.59123E-01	0.36875E-02	0.62370E-01
PROPA-01	0.11714E-02	0.28307E-02	0.24947E-04	0.88128E-02
N-BUT-01	0.25891E-04	0.63263E-04	0.69102E-07	0.10923E-02
ISOBU-01	0.48767E-04	0.11898E-03	0.25234E-06	0.21209E-02
2-MET-01	0.17287E-05	0.42294E-05	0.75201E-09	0.17780E-03
N-PEN-01	0.17158E-05	0.41980E-05	0.67306E-09	0.16033E-03
N-HEX-01	0.22156E-07	0.54218E-07	0.16610E-11	0.30635E-04

BLOCK: F-102   MODEL: FLASH2

INLET STREAM: N2-LNG
OUTLET VAPOR STREAM: N2-PURGE
OUTLET LIQUID STREAM: LNG-PROD
PROPERTY OPTION SET: PENG-ROB STANDARD PR EQUATION OF STATE

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(LBMOL/HR)	12320.0	12320.0	0.00000
MASS(LB/HR)	204756.	204756.	0.00000
ENTHALPY(BTU/HR)	-0.446977E+09	-0.446977E+09	0.00000

*** INPUT DATA ***

TWO PHASE PQ FLASH		
SPECIFIED PRESSURE	PSIA	18.0000
SPECIFIED HEAT DUTY	BTU/HR	0.0
MAXIMUM NO. ITERATIONS		30
CONVERGENCE TOLERANCE		0.000100000

*** RESULTS ***

OUTLET TEMPERATURE	F	-259.21
OUTLET PRESSURE	PSIA	18.000
VAPOR FRACTION		0.17872

V-L PHASE EQUILIBRIUM :

COMP	F(I)	X(I)	Y(I)	K(I)
NITRO-01	0.43831E-01	0.10401E-01	0.19746	18.985
METHA-01	0.95246	0.98508	0.80254	0.81469
ETHAN-01	0.36875E-02	0.44885E-02	0.66411E-05	0.14796E-02

```
      PROPA-01         0.24947E-04    0.30375E-04    0.63360E-09    0.20859E-
04
      N-BUT-01         0.69102E-07    0.84139E-07    0.16880E-13    0.20062E-
06
      ISOBU-01         0.25234E-06    0.30725E-06    0.32564E-12    0.10598E-
05
      2-MET-01         0.75201E-09    0.91566E-09    0.33577E-17    0.36670E-
08
      N-PEN-01         0.67306E-09    0.81952E-09    0.25001E-17    0.30506E-
08
      N-HEX-01         0.16610E-11    0.20224E-11    0.21193E-21    0.10479E-
09

 BLOCK:  FN-601   MODEL: RSTOIC
 ------------------------------
   INLET STREAM:          FURNFG
   OUTLET STREAM:         FURNEXH
   PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                      ***  MASS AND ENERGY BALANCE  ***
                              IN           OUT       GENERATION   RELATIVE
DIFF.
   TOTAL BALANCE
   MOLE(LBMOL/HR)         1948.01       1948.01      0.434274E-03  0.116721E-
15
   MASS(LB/HR   )         54828.1       54828.1                    0.398115E-
15
   ENTHALPY(BTU/HR  )   -0.340809E+07 -0.340809E+07               -0.136634E-
15

                          ***  INPUT DATA  ***


   TWO    PHASE  PQ  FLASH
   SPECIFIED PRESSURE    PSIA                              500.000
   SPECIFIED HEAT DUTY   BTU/HR                              0.0
   MAXIMUM NO. ITERATIONS                                   30
   CONVERGENCE TOLERANCE                                     0.000100000
   SIMULTANEOUS REACTIONS
   GENERATE COMBUSTION REACTIONS FOR FEED SPECIES          YES
   COMBUSTION PRODUCT FOR CHEMICALLY BOUND NITROGEN        NO

                           ***  RESULTS  ***
   OUTLET TEMPERATURE    F                                    2345.6
   OUTLET PRESSURE       PSIA                                 500.00
   VAPOR FRACTION                                             1.0000


   COMBUSTION REACTIONS:

      RXN NO    STOICHIOMETRY

      C1        METHA-01 + 2 OXYGE-01 --> CARBO-01 + 2 WATER
      C2        ETHAN-01 + 3.5 OXYGE-01 --> 2 CARBO-01 + 3 WATER
      C3        PROPA-01 + 5 OXYGE-01 --> 3 CARBO-01 + 4 WATER
      C4        N-BUT-01 + 6.5 OXYGE-01 --> 4 CARBO-01 + 5 WATER
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    C5        ISOBU-01 + 6.5 OXYGE-01 --> 4 CARBO-01 + 5 WATER
    C6        2-MET-01 + 8 OXYGE-01 --> 5 CARBO-01 + 6 WATER
    C7        N-PEN-01 + 8 OXYGE-01 --> 5 CARBO-01 + 6 WATER
    C8        N-HEX-01 + 9.5 OXYGE-01 --> 6 CARBO-01 + 7 WATER


  HEAT OF REACTIONS:

    REACTION          REFERENCE          HEAT OF
    NUMBER            COMPONENT          REACTION
                                          BTU/LBMOL
     C1                  METHA-01         -0.34514E+06
     C2                  ETHAN-01         -0.61432E+06
     C3                  PROPA-01         -0.87853E+06
     C4                  N-BUT-01         -0.11426E+07
     C5                  ISOBU-01         -0.11387E+07
     C6                  2-MET-01          0.10000E+36
     C7                  N-PEN-01          0.10000E+36
     C8                  N-HEX-01          0.10000E+36

  REACTION EXTENTS:

     REACTION          REACTION
     NUMBER            EXTENT
                       LBMOL/HR
     C1                 104.94
     C2                0.86838E-03
     C3                0.82849E-07
     C4                0.22084E-11
     C5                0.42581E-10

  V-L PHASE EQUILIBRIUM :

     COMP              F(I)           X(I)           Y(I)           K(I)
     NITRO-01         0.75116        0.75116        0.75116        MISSING
     CARBO-01         0.53871E-01    0.53871E-01    0.53871E-01    MISSING
     OXYGE-01         0.87229E-01    0.87229E-01    0.87229E-01    MISSING
     WATER            0.10774        0.10774        0.10774        MISSING

BLOCK:  HX-102   MODEL: HEATER
------------------------------
  INLET STREAM:          COMP
  OUTLET STREAM:         COMPCOOL
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                     ***  MASS AND ENERGY BALANCE  ***
                                  IN               OUT          RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)           12320.0          12320.0          0.00000
      MASS(LB/HR   )           204756.          204756.          0.00000
      ENTHALPY(BTU/HR  )     -0.366703E+09   -0.380668E+09    0.366843E-01

                         ***  INPUT DATA  ***
  TWO    PHASE  TP  FLASH
  SPECIFIED TEMPERATURE              F                        90.0000
  PRESSURE DROP                      PSI                       5.00000
```

```
 MAXIMUM NO. ITERATIONS                                              30
 CONVERGENCE TOLERANCE                                                0.000100000



                       ***  RESULTS  ***
 OUTLET TEMPERATURE    F                                         90.000
 OUTLET PRESSURE       PSIA                                      720.00
 HEAT DUTY             BTU/HR                                  -0.13965E+08
 OUTLET VAPOR FRACTION                                            1.0000
 PRESSURE-DROP CORRELATION PARAMETER                             1628.9



 V-L PHASE EQUILIBRIUM :

    COMP              F(I)          X(I)           Y(I)           K(I)
    NITRO-01        0.43831E-01    0.35400E-01    0.43831E-01     3.2886
    METHA-01        0.95246        0.95920        0.95246         2.6374
    ETHAN-01        0.36875E-02    0.53552E-02    0.36875E-02     1.8289
    PROPA-01        0.24947E-04    0.46805E-04    0.24947E-04     1.4156
    N-BUT-01        0.69102E-07    0.17093E-06    0.69102E-07     1.0738
    ISOBU-01        0.25234E-06    0.57065E-06    0.25234E-06     1.1745
    2-MET-01        0.75201E-09    0.23799E-08    0.75201E-09    0.83927
    N-PEN-01        0.67306E-09    0.21290E-08    0.67306E-09    0.83969
    N-HEX-01        0.16610E-11    0.64348E-11    0.16610E-11    0.68560

                   ***  ASSOCIATED  UTILITIES  ***

 UTILITY ID FOR WATER                          U-1
 RATE OF CONSUMPTION                     1.4004+06  LB/HR
 COST                                    4.9014+04  $/HR

BLOCK:  HX-201   MODEL: HEATER
------------------------------
 INLET STREAM:          INTERCO2
 OUTLET STREAM:         CO2WARM
 PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                  ***  MASS AND ENERGY BALANCE  ***
                                 IN              OUT        RELATIVE DIFF.
  TOTAL BALANCE
     MOLE(LBMOL/HR)            26000.0         26000.0         0.00000
     MASS(LB/HR   )          0.114425E+07    0.114425E+07     0.00000
     ENTHALPY(BTU/HR  )     -0.442052E+10   -0.440908E+10   -0.258739E-02

                     ***  INPUT DATA  ***
 TWO    PHASE  PQ  FLASH
 PRESSURE DROP                         PSI                         0.0
 SPECIFIED HEAT DUTY                   BTU/HR                      0.114376+08
 MAXIMUM NO. ITERATIONS                                              30
 CONVERGENCE TOLERANCE                                                0.000100000



                       ***  RESULTS  ***
 OUTLET TEMPERATURE    F                                         47.103
```

```
  OUTLET PRESSURE       PSIA                                    100.00
  OUTLET VAPOR FRACTION                                         1.0000
  PRESSURE-DROP CORRELATION PARAMETER                           0.0000



  V-L PHASE EQUILIBRIUM :

     COMP              F(I)          X(I)          Y(I)          K(I)
     CARBO-01          1.0000        1.0000        1.0000        4.4802

BLOCK:  HX-202   MODEL: HEATER
------------------------------
  INLET STREAM:          CO2
  OUTLET STREAM:         INTERCO2
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                     ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)            26000.0         26000.0         0.00000
      MASS(LB/HR   )          0.114425E+07    0.114425E+07     0.00000
      ENTHALPY(BTU/HR  )      -0.442685E+10   -0.442052E+10   -0.143039E-02

                         ***  INPUT DATA  ***
  TWO    PHASE  PQ  FLASH
  PRESSURE DROP                        PSI                          0.0
  SPECIFIED HEAT DUTY                  BTU/HR              6,332,100.
  MAXIMUM NO. ITERATIONS                                           30
  CONVERGENCE TOLERANCE                                             0.000100000



                          ***  RESULTS  ***
  OUTLET TEMPERATURE    F                                      -1.5883
  OUTLET PRESSURE       PSIA                                    100.00
  OUTLET VAPOR FRACTION                                         1.0000
  PRESSURE-DROP CORRELATION PARAMETER                           0.0000



  V-L PHASE EQUILIBRIUM :

     COMP              F(I)          X(I)          Y(I)          K(I)
     CARBO-01          1.0000        1.0000        1.0000        2.5138

BLOCK:  HX-701   MODEL: HEATER
------------------------------
  INLET STREAM:          TOCO2COO
  OUTLET STREAM:         TOCO2EXP
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                     ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)            26000.0         26000.0         0.00000
      MASS(LB/HR   )          0.114425E+07    0.114425E+07     0.00000
```

```
    ENTHALPY(BTU/HR  )      -0.437449E+10   -0.440540E+10    0.701600E-02

                       ***  INPUT DATA  ***
 TWO    PHASE  TP  FLASH
 SPECIFIED TEMPERATURE              F                         90.0000
 PRESSURE DROP                      PSI                        0.0
 MAXIMUM NO. ITERATIONS                                       30
 CONVERGENCE TOLERANCE                                         0.000100000


                        ***  RESULTS  ***
 OUTLET TEMPERATURE    F                                   90.000
 OUTLET PRESSURE       PSIA                                295.00
 HEAT DUTY             BTU/HR                            -0.30908E+08
 OUTLET VAPOR FRACTION                                     1.0000
 PRESSURE-DROP CORRELATION PARAMETER                       0.0000


 V-L PHASE EQUILIBRIUM :

   COMP              F(I)           X(I)           Y(I)           K(I)
   CARBO-01          1.0000         1.0000         1.0000         MISSING

BLOCK:  P-501    MODEL: PUMP
----------------------------
 INLET STREAM:          1
 OUTLET STREAM:         5
 PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                   ***  MASS AND ENERGY BALANCE  ***
                                IN              OUT        RELATIVE DIFF.
  TOTAL BALANCE
    MOLE(LBMOL/HR)          559475.         559475.         0.00000
    MASS(LB/HR   )         0.100791E+08    0.100791E+08     0.00000
    ENTHALPY(BTU/HR  )    -0.693077E+11   -0.693053E+11   -0.351560E-04

                       ***  INPUT DATA  ***
  PRESSURE CHANGE  PSI                                  70.0000
  DRIVER EFFICIENCY                                      1.00000

  FLASH SPECIFICATIONS:
  LIQUID PHASE CALCULATION
  NO FLASH PERFORMED
  MAXIMUM NUMBER OF ITERATIONS                           30
  TOLERANCE                                              0.000100000

                        ***  RESULTS  ***
  VOLUMETRIC FLOW RATE  CUFT/HR                     161,651.
  PRESSURE CHANGE  PSI                                  70.0000
  NPSH AVAILABLE   FT-LBF/LB                            33.3177
  FLUID POWER  HP                                      822.950
  BRAKE POWER  HP                                      957.612
  ELECTRICITY  KW                                      714.091
  PUMP EFFICIENCY USED                                   0.85938
  NET WORK REQUIRED  HP                                957.612
```

```
   HEAD DEVELOPED FT-LBF/LB                              161.665

BLOCK:  P-502    MODEL: PUMP
----------------------------
  INLET STREAM:          2
  OUTLET STREAM:         6
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                    ***  MASS AND ENERGY BALANCE  ***
                                 IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)          559475.         559475.         0.00000
      MASS(LB/HR   )        0.100791E+08    0.100791E+08     0.00000
      ENTHALPY(BTU/HR  )    -0.693077E+11   -0.693053E+11   -0.351560E-04

                        ***  INPUT DATA  ***
   PRESSURE CHANGE  PSI                                   70.0000
   DRIVER EFFICIENCY                                       1.00000

   FLASH SPECIFICATIONS:
   LIQUID PHASE CALCULATION
   NO FLASH PERFORMED
   MAXIMUM NUMBER OF ITERATIONS                            30
   TOLERANCE                                               0.000100000

                         ***  RESULTS  ***
   VOLUMETRIC FLOW RATE  CUFT/HR                      161,651.
   PRESSURE CHANGE  PSI                                   70.0000
   NPSH AVAILABLE   FT-LBF/LB                             33.3177
   FLUID POWER  HP                                       822.950
   BRAKE POWER  HP                                       957.612
   ELECTRICITY  KW                                       714.091
   PUMP EFFICIENCY USED                                    0.85938
   NET WORK REQUIRED  HP                                 957.612
   HEAD DEVELOPED FT-LBF/LB                              161.665

BLOCK:  P-503    MODEL: PUMP
----------------------------
  INLET STREAM:          3
  OUTLET STREAM:         7
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                    ***  MASS AND ENERGY BALANCE  ***
                                 IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)          559475.         559475.         0.00000
      MASS(LB/HR   )        0.100791E+08    0.100791E+08     0.00000
      ENTHALPY(BTU/HR  )    -0.693077E+11   -0.693053E+11   -0.351560E-04

                        ***  INPUT DATA  ***
   PRESSURE CHANGE  PSI                                   70.0000
   DRIVER EFFICIENCY                                       1.00000

   FLASH SPECIFICATIONS:
   LIQUID PHASE CALCULATION
   NO FLASH PERFORMED
   MAXIMUM NUMBER OF ITERATIONS                            30
```

```
   TOLERANCE                                                      0.000100000

                        ***  RESULTS  ***
   VOLUMETRIC FLOW RATE  CUFT/HR                            161,651.
   PRESSURE CHANGE  PSI                                          70.0000
   NPSH AVAILABLE   FT-LBF/LB                                    33.3177
   FLUID POWER  HP                                              822.950
   BRAKE POWER  HP                                              957.612
   ELECTRICITY  KW                                              714.091
   PUMP EFFICIENCY USED                                           0.85938
   NET WORK REQUIRED  HP                                        957.612
   HEAD DEVELOPED FT-LBF/LB                                     161.665

BLOCK:  P-504    MODEL: PUMP
----------------------------
  INLET STREAM:          4
  OUTLET STREAM:         8
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                   ***  MASS AND ENERGY BALANCE  ***
                            IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)          559475.         559475.        0.00000
      MASS(LB/HR   )         0.100791E+08    0.100791E+08    0.00000
      ENTHALPY(BTU/HR  )    -0.693077E+11   -0.693053E+11   -0.351560E-04

                       ***  INPUT DATA  ***
   PRESSURE CHANGE  PSI                                          70.0000
   DRIVER EFFICIENCY                                              1.00000

   FLASH SPECIFICATIONS:
   LIQUID PHASE CALCULATION
   NO FLASH PERFORMED
   MAXIMUM NUMBER OF ITERATIONS                                   30
   TOLERANCE                                                      0.000100000

                        ***  RESULTS  ***
   VOLUMETRIC FLOW RATE  CUFT/HR                            161,651.
   PRESSURE CHANGE  PSI                                          70.0000
   NPSH AVAILABLE   FT-LBF/LB                                    33.3177
   FLUID POWER  HP                                              822.950
   BRAKE POWER  HP                                              957.612
   ELECTRICITY  KW                                              714.091
   PUMP EFFICIENCY USED                                           0.85938
   NET WORK REQUIRED  HP                                        957.612
   HEAD DEVELOPED FT-LBF/LB                                     161.665

BLOCK:  P-601    MODEL: PUMP
----------------------------
  INLET STREAM:          35
  OUTLET STREAM:         36
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                   ***  MASS AND ENERGY BALANCE  ***
                            IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)          1307.17         1307.17        0.00000
```

```
      MASS(LB/HR   )           23549.0        23549.0        0.00000
      ENTHALPY(BTU/HR  )      -0.155934E+09   -0.155931E+09   -0.229092E-04

                      ***  INPUT DATA  ***
   OUTLET PRESSURE  PSIA                                  114.696
   DRIVER EFFICIENCY                                        1.00000

   FLASH SPECIFICATIONS:
   LIQUID PHASE CALCULATION
   NO FLASH PERFORMED
   MAXIMUM NUMBER OF ITERATIONS                             30
   TOLERANCE                                                0.000100000

                       ***  RESULTS  ***
   VOLUMETRIC FLOW RATE  CUFT/HR                          436.338
   PRESSURE CHANGE  PSI                                    20.0000
   NPSH AVAILABLE   FT-LBF/LB                              79.0534
   FLUID POWER  HP                                          0.63467
   BRAKE POWER  HP                                          1.40398
   ELECTRICITY  KW                                          1.04695
   PUMP EFFICIENCY USED                                     0.45205
   NET WORK REQUIRED  HP                                    1.40398
   HEAD DEVELOPED FT-LBF/LB                                53.3633

BLOCK:  P-602    MODEL: PUMP
----------------------------
  INLET STREAM:          33
  OUTLET STREAM:         34
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                   ***  MASS AND ENERGY BALANCE  ***
                                IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)           723.053        723.053        0.00000
      MASS(LB/HR   )           13026.0        13026.0        0.00000
      ENTHALPY(BTU/HR  )      -0.872983E+08   -0.872959E+08   -0.274660E-04

                      ***  INPUT DATA  ***
   OUTLET PRESSURE  PSIA                                   41.6959
   DRIVER EFFICIENCY                                        1.00000

   FLASH SPECIFICATIONS:
   LIQUID PHASE CALCULATION
   NO FLASH PERFORMED
   MAXIMUM NUMBER OF ITERATIONS                             30
   TOLERANCE                                                0.000100000

                       ***  RESULTS  ***
   VOLUMETRIC FLOW RATE  CUFT/HR                          229.675
   PRESSURE CHANGE  PSI                                    20.0000
   NPSH AVAILABLE   FT-LBF/LB                               6.71682
   FLUID POWER  HP                                          0.33407
   BRAKE POWER  HP                                          0.94235
   ELECTRICITY  KW                                          0.70271
   PUMP EFFICIENCY USED                                     0.35451
   NET WORK REQUIRED  HP                                    0.94235
   HEAD DEVELOPED FT-LBF/LB                                50.7804
```

```
BLOCK:  T-301    MODEL: COMPR
-----------------------------
  INLET STREAM:          TOTURB
  OUTLET STREAM:         TURBEXH
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                     ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)            30853.8         30853.8         0.00000
      MASS(LB/HR   )            868400.         868400.         0.00000
      ENTHALPY(BTU/HR  )      -0.539794E+08   -0.478867E+09    0.887277

                          ***  INPUT DATA  ***

  ISENTROPIC TURBINE
   OUTLET PRESSURE  PSIA                                      3.92927
   ISENTROPIC EFFICIENCY                                      0.85000
   MECHANICAL EFFICIENCY                                      1.00000

                          ***  RESULTS  ***

   INDICATED  HORSEPOWER REQUIREMENT  HP               -166,987.
   BRAKE      HORSEPOWER REQUIREMENT  HP               -166,987.
   NET WORK REQUIRED                  HP               -166,987.
   POWER LOSSES                       HP                      0.0
   ISENTROPIC HORSEPOWER REQUIREMENT  HP               -196,455.
   CALCULATED OUTLET TEMP  F                               723.280
   ISENTROPIC TEMPERATURE  F                               401.713
   EFFICIENCY (POLYTR/ISENTR) USED                          0.85000
   OUTLET VAPOR FRACTION                                    1.00000
   HEAD DEVELOPED,       FT-LBF/LB                     -447,929.
   MECHANICAL EFFICIENCY USED                               1.00000
   INLET HEAT CAPACITY RATIO                                1.28205
   INLET VOLUMETRIC FLOW RATE , CUFT/HR               1,869,170.
   OUTLET VOLUMETRIC FLOW RATE, CUFT/HR                     0.996887+08
   INLET  COMPRESSIBILITY FACTOR                            1.00618
   OUTLET COMPRESSIBILITY FACTOR                            1.00006
   AV. ISENT. VOL. EXPONENT                                 1.32435
   AV. ISENT. TEMP EXPONENT                                 1.32212
   AV. ACTUAL VOL. EXPONENT                                 1.21868
   AV. ACTUAL TEMP EXPONENT                                 1.21681

BLOCK:  T-601    MODEL: FLASH2
------------------------------
  INLET STREAM:          HPSTEAMC
  OUTLET VAPOR STREAM:   38
  OUTLET LIQUID STREAM:  37
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                     ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)            1307.17         1307.17         0.00000
      MASS(LB/HR   )            23549.0         23549.0         0.00000
      ENTHALPY(BTU/HR  )      -0.155934E+09   -0.155934E+09    0.177880E-07
```

```
                       ***  INPUT DATA  ***
  TWO    PHASE  PQ  FLASH
  PRESSURE DROP         PSI                                 0.0
  SPECIFIED HEAT DUTY   BTU/HR                              0.0
  MAXIMUM NO. ITERATIONS                                   30
  CONVERGENCE TOLERANCE                                     0.000100000

                        ***  RESULTS  ***
  OUTLET TEMPERATURE    F                                    299.92
  OUTLET PRESSURE       PSIA                                 94.696
  VAPOR FRACTION                                             0.0000


  V-L PHASE EQUILIBRIUM :

     COMP              F(I)          X(I)          Y(I)          K(I)
     WATER             1.0000        1.0000        1.0000        0.69699

BLOCK:  T-602    MODEL: FLASH2
------------------------------
  INLET STREAM:          LPSTEAMC
  OUTLET VAPOR STREAM:   31
  OUTLET LIQUID STREAM:  32
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                   ***  MASS AND ENERGY BALANCE  ***
                                IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)          723.053         723.053        0.00000
      MASS(LB/HR   )          13026.0         13026.0        0.00000
      ENTHALPY(BTU/HR  )    -0.872983E+08   -0.872983E+08    0.123835E-08

                       ***  INPUT DATA  ***
  TWO    PHASE  PQ  FLASH
  PRESSURE DROP         PSI                                 0.0
  SPECIFIED HEAT DUTY   BTU/HR                              0.0
  MAXIMUM NO. ITERATIONS                                   30
  CONVERGENCE TOLERANCE                                     0.000100000

                        ***  RESULTS  ***
  OUTLET TEMPERATURE    F                                    228.55
  OUTLET PRESSURE       PSIA                                 21.696
  VAPOR FRACTION                                             0.0000


  V-L PHASE EQUILIBRIUM :

     COMP              F(I)          X(I)          Y(I)          K(I)
     WATER             1.0000        1.0000        1.0000        0.87948

BLOCK:  V-201    MODEL: VALVE
-----------------------------
  INLET STREAM:          TOFRAC
  OUTLET STREAM:         22
```

```
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                   ***  MASS AND ENERGY BALANCE  ***
                                 IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)           1180.00        1180.00         0.00000
      MASS(LB/HR   )           48961.2        48961.2        0.297213E-15
      ENTHALPY(BTU/HR  )      -0.594620E+08   -0.594620E+08   -0.375900E-15

                       ***  INPUT DATA  ***

   VALVE OUTLET PRESSURE     PSIA                          200.000
   VALVE FLOW COEF CALC.                                   NO

                       FLASH SPECIFICATIONS:
   NPHASE                                                    2
   MAX NUMBER OF ITERATIONS                                 30
   CONVERGENCE TOLERANCE                                     0.000100000

                       ***  RESULTS  ***

   VALVE PRESSURE DROP       PSI                           100.000

BLOCK:  V-202    MODEL: VALVE
-----------------------------
  INLET STREAM:          HEAVY
  OUTLET STREAM:         23
  PROPERTY OPTION SET:   PENG-ROB  STANDARD PR EQUATION OF STATE

                   ***  MASS AND ENERGY BALANCE  ***
                                 IN              OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)           270.000        270.000         0.00000
      MASS(LB/HR   )           17870.2        17870.2       -0.203578E-15
      ENTHALPY(BTU/HR  )      -0.174146E+08   -0.174146E+08    0.213918E-15

                       ***  INPUT DATA  ***

   VALVE OUTLET PRESSURE     PSIA                          100.000
   VALVE FLOW COEF CALC.                                   NO

                       FLASH SPECIFICATIONS:
   NPHASE                                                    2
   MAX NUMBER OF ITERATIONS                                 30
   CONVERGENCE TOLERANCE                                     0.000100000

                       ***  RESULTS  ***

   VALVE PRESSURE DROP       PSI                            99.0000
```

Appendix XI: Problem Statement

12. Natural Gas Liquefaction using a CO_2-Precooled Reverse-Brayton Cycle (recommended by Adam A. Brostow, LNG Process Tech., Air Products and Chemicals)

"Uncommon men require no common trust; give him but the scope and he will set the bounds." -- Friedrich von Schiller

Introduction

Natural gas is a clean-burning fuel with high hydrogen to carbon ratio, a simpler alternative to hydrogen fuel. To transport natural gas from the well to the point of use, it is often liquefied and loaded onto a ship.

LNG (liquid natural gas) is typically prepared by the so-called propane-precooled mixed refrigerant (C3MR) cycle process. Natural gas is pre-cooled by vaporizing propane and liquefied by vaporizing a MR (mixed refrigerant), usually a mixture of hydrocarbons.

An FPSO (Floating Production, Storage and Offloading) is a type of plant that is mounted on a ship or an offshore platform. It can be moved from one location to another. There are two major issues in using the C3MR process in a shipboard application: sensitivity to the vessel motion and the fire hazard associated with the hydrocarbons, especially propane. While propane is volatile, it is sufficiently heavy for a flammable cloud to hover over the area for an extended period causing BLEVEs (Boiling Liquid Expanding Vapor Explosions).

An alternative to C3MR and other MR processes is the reverse-Brayton cycle and the CO_2-precooled reverse-Brayton cycle, typically using gaseous nitrogen as the refrigerant. These cycles are less efficient, but relatively simple, insensitive to motion, and potentially safer. They show promise for smaller plants build on solid ground, with many recent patents issued to various energy companies.

Background Information

In the figures that follow, several reverse-Brayton cycle configurations are shown, beginning with Figure 1, which shows the simplest possible implementation of the reverse-Brayton cycle. These are intended as introduction before the problem statement is presented in the next section.

In Figure 1, gaseous refrigerant (e.g., nitrogen) is compressed in COMP, cooled to about-ambient temperature in an aftercooler, AC, further cooled in the liquefier-heat exchanger, HX, isentropically expanded in the expander (turbine), EXP, and warmed in the HX to provide refrigeration to liquefy natural gas. The heat exchanger, typically a brazed-aluminum core (BAHX) can be simulated using MHEATX in ASPEN PLUS.

Figure 2 shows a *compander* (compressor-expander) and illustrates the power recycle (recovery) concept. Part of the refrigerant compression is done by compressor CMP directly driven by expander EXP.

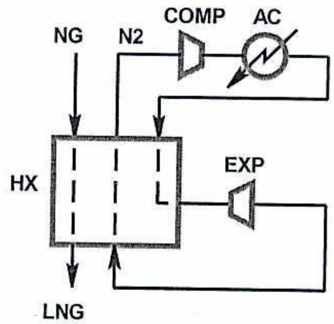

Figure 1 Reverse-Brayton cycle

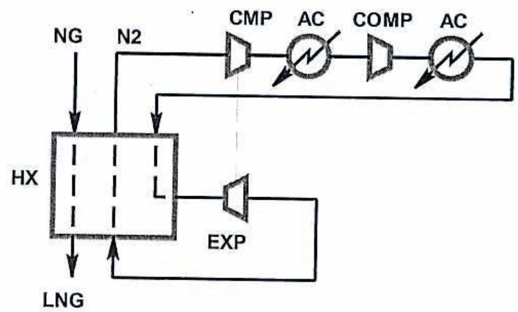

Figure 2. Reverse-Brayton cycle with a compressor-expander

Figure 3 shows a typical 2-expander liquefier. Two expanders, the warmer (EXP1) and colder (EXP2), improve the efficiency of the process. Optionally, they may drive two compressors: CMP1 and CMP2, providing a portion of the compression load.

Figure 4 shows a precooled cycle, where additional refrigerant, such as propane (C3), is condensed, subcooled, throttled, and vaporized. Alternatively, the propane is replaced with CO_2, with multiple temperature levels of CO_2 used. Note that Figure 4 shows just one gaseous refrigerant expander, but two expanders can be used for better efficiency (as shown in Figure 3).

Is it necessary or economical to condense CO_2 prior to throttling? This is an important question to be answered by the design team, the answer to which is not obvious.

Environmentally friendly fluorinated hydrocarbons are alternatives to CO_2. While they don't deplete the ozone layer, they have a greater greenhouse effect than CO_2 and are difficult to generate offshore. Your design team is encouraged to investigate methods of producing CO_2 onsite to initially charge the system and to make up for seal losses.

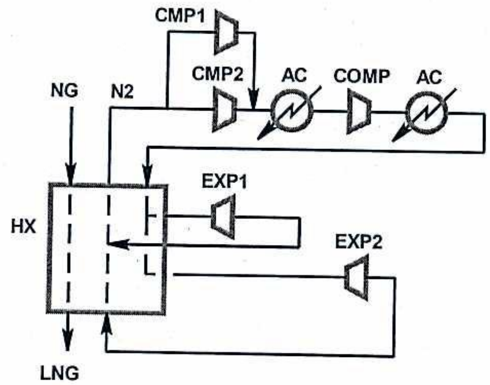

Figure 3. Reverse-Brayton cycle with two expanders

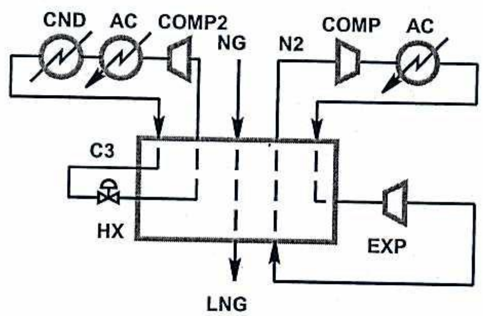

Figure 4. Pre-cooled reverse-Brayton cycle.

Figure 5 shows a liquefier with a scrub column. Here, natural gas is optionally expanded (to improve distillation), cooled in HX, and fed to the scrub column COL. The vapor overhead from the column is optionally recompressed, further cooled in HX, and fed to the phase separator, SEP. The liquid from SEP is used as reflux for COL. Vapor is liquefied to produce the LNG product.

The NGL (natural gas liquid) bottoms product is removed from COL to maintain the LNG heating value, to prevent heavier hydrocarbons from freezing during liquefaction, and to recover valuable products: ethane, LPG (light petroleum gas: propane and butane), and heavier components.

If time permits, the design team is encouraged to model a distillation sequence to recover C2, C3, and C4 in deethanizer, depropanizer, and debutanizer columns (not shown) and calculate the additional revenue from those products.

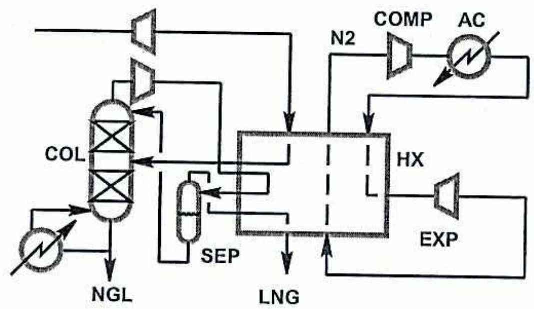

Figure 5. Reverse-Brayton cycle with a scrub column

Figure 6 shows a plant directly driven by a gas turbine (GT). Liquefied natural gas is throttled in a valve and fed to the product separator, PS. The LNG product is recovered from the bottom of the separator. Flash vapor from the separator is warmed in HX, compressed in fuel compressor, FC, and sent to the combustion chamber of the gas turbine, GT.

Air is compressed on the compressor side of the GT. It is then mixed with fuel in the combustion chamber, ignited, and expanded on the expander side to directly drive refrigerant compressor(s), CMP. As an alternative, the GT drives a generator while an electrical motor drives the compressor.

A new emerging technology is "aero derivative" turbines, which are based on jet engines – a more advanced technology.

Problem Statement

13,500 lbmole/hr of natural gas at 68°F and at 725 psia containing 4% N_2, 87% C1 (methane), 5% C2 (ethane), 2% C3 (propane), 0.5% I4 (isobutane), 0.5% C4 (n-butane), 0.3% I5 (isopentane), 0.5% C5 (n-pentane), and 0.2% C6 (hexanes) is being liquefied. This roughly corresponds to 1 MTPA of LNG (1 million metric tons per annum).

The feed is cooled in a liquefier heat exchanger to a certain temperature (to be determined). It is then fed to the scrub column. The column overhead is further cooled in the liquefier heat exchanger. It is then fed to the reflux phase separator. Liquid from the reflux phase separator goes to the top of the column. Vapor from the reflux separator is cooled in the liquefier heat exchanger. The resulting fluid leaves the exchanger at about -230°F. It is throttled to 18 psia in a product valve and fed to the LNG product separator.

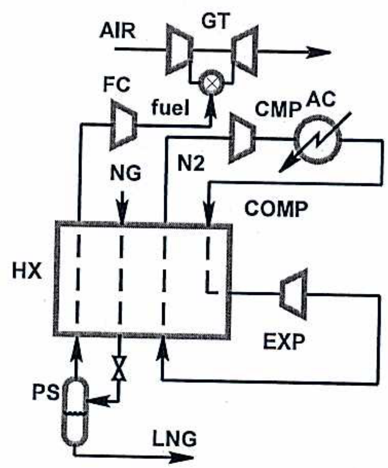

Figure 6. Reverse-Brayton cycle with gas turbine

Liquid from the product separator is recovered as LNG product. Vapor is warmed in the liquefier heat exchanger, compressed to 500 psia, and fed to the combustion chamber of the gas turbine.

Vapor from the reflux phase separator cannot contain more than 0.1% C5+ (I5+C5+C6). Scrub column bottoms product should not contain more than 1% C1.

Column feed can be expanded to improve distillation; in that case the vapor is recompressed with power recovered from the feed expander.

The simplest refrigeration system to be considered is a single-expander nitrogen loop. Optimal process conditions are to be determined. Cooling water for the after-cooler is available at 68°F. Adiabatic (isentropic) efficiency of the compressor is 86% per stage. Adiabatic efficiency of the expander is 88%.

Another compressor driven by the expander is added to handle a portion of the compression load.

Then the second (warm) expander is added.

Finally, the CO_2 precooling loop is added (cooling water for the condenser is at 68°F). The design team is encouraged to find the best strategy to model the process step by step.

Another option is a single-expander system with precooling.

Depending on time available, the design team can model simpler and more complex systems and look at the capital-efficiency (specific power) tradeoff. Even a simple design is valuable as it may be economical for smaller plants. But at least one system should be precooled by at least one stage of CO_2.

The design team is to determine the limitations of using CO_2 as refrigerant and to determine whether it is necessary to condense CO_2 to achieve a working cycle. The team is also encouraged to find a way to generate CO_2 onsite.

Refrigerant compressor(s) are driven by an aero derivative gas turbine. Ambient air at 68°F and 14.7 psia is compressed to 500 psia (adiabatic efficiency of 78%). It is then mixed with fuel. The combustion temperature lies below 2350°F. The flue gas is expanded to about 8 inch Hg (adiabatic efficiency of 85%).

The design team is to model the gas turbine. The power requirement of the compressors determines the size and cost of the turbine. The fuel heating value determines the LNG temperature from the liquefier heat exchanger (initially assumed to be -230°F). In other words, the GT must satisfy both power demand and fuel balance.

If time permits, the design team can design a multiple-stage CO_2 precooling system and/or a ethane, propane, and butane recovery distillation system.

The plant economics data should be scaled to 0.5 MTPA and 2 MTPA to determine the impact of plant's size.

References

U.S. Patent 7,386,966 – describes CO_2-precooled LNG process with a condenser.

U.S. Patent 4,065,278 – describes conventional C3MR process with a scrub column.

Finn, A. J., "Effective LNG Production Offshore" – paper available from W. D. Seider

CO_2 P-H diagram – available from W. D. Seider

Air Products can provide some information about core sizing and costing. Much information is available online. The design team is encouraged to seek additional information and to modify/improve the process.