University of Pennsylvania School of Engineering & Applied Science

Department of Chemical & Biomolecular Engineering

220 South 33rd Street

Philadelphia, PA 19104

Tuesday, April 18, 2023

Dear Professors Warren Seider and Bruce Vrana,

Please find enclosed a report detailing the technical and economic specifications of a process to liquefy 45 metric tons per day (MTD) of vapor feed Hydrogen into liquid Hydrogen (LH2). This is triple the production rate of 15 MTD as proposed by Mr. Adam Brostow. The product LH2 immediately enters the liquid Hydrogen supply chain as saturated liquid at 25 K and 52 psia, at an equilibrium composition of 99.0% p-H2 spin isomer. The process refrigerant will be Neon vapor, purchased from Chinese suppliers at a 10x premium due to the ongoing Russian invasion of Ukraine. Saturated liquid Nitrogen at 77 K and 20 psia is supplied to precool the process.

The vapor feed Hydrogen to the liquefaction process is sourced upstream by electrolytically splitting water into Hydrogen and oxygen. The electricity to split water and to operate the plant comes from a completely renewable power grid. Sensitivity analyses on the fraction of Hydrogen sourced via electrolysis versus Hydrogen sourced by traditional Steam-Methane Reforming (SMR) were performed in order to compare the economic feasibility of the two. The breakeven price of 100% electrolysis was $9.20 per kilogram of LH2 and the breakeven price of 100% SMR was $4.40 per kilogram of LH2, while the cost of liquefaction was $2.71 per kilogram of LH2.

Assuming a cost of capital of 15%, a plant lifetime of 15 years, a sales price of $13 per kilogram of LH2, and 100% of vapor feed Hydrogen sourced via water electrolysis, a plant based on the process design detailed herein has an ROI of 16.57%, an IRR of 18.52%, and an NPV of $44,445,400. $13 per kilogram of LH2 can be restated in terms of its combustible chemical energy as $89 per million BTU.

The plant will be located in St. John's County on the Florida Gulf Coast, allowing the plant to have unhindered access to water, renewable solar and wind power grids, and proximity to consumers in the aerospace industry. The plant will operate continuously: 24 hours a day for 330 days a year, and at 100% production capacity after the third year of operation. Seven days, or 300 metric tons, of safety LH2 inventory are stored onsite during normal operations in one spherical aluminum container to satisfy demand in case of holidays, unexpected plant shutdowns, and random fluctuations in demand.

Sincerely,

Akash Kumashi	Guillermo Ribeiro-Vecino	Evan Bean
Akash R. Kumashi	Guillermo Ribeiro-Vecino	Evan J. Bean

Green Hydrogen Liquefaction by Large- Scale Reverse Brayton Refrigeration

Evan Bean, Akash Kumashi, Guillermo Ribeiro-Vecino

Faculty Advisor: Dr. Warren Seider

Project Author: Mr. Adam Brostow

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School of Engineering & Applied Sciences

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April 18, 2023

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4 ABSTRACT

Over the next decade, global electricity demand is forecast to rise by nearly two-thirds of current demand. Simultaneously, global Carbon Dioxide emissions are projected to increase by up to 9% annually. Green liquid Hydrogen, sourced by splitting water into Hydrogen and oxygen using renewable electricity, and condensed in a deep cryogenic refrigerator at 20 to 25 K, is a promising alternative to traditional fossil fuels. Yet, liquid Hydrogen as a fuel is prohibitively expensive. Between water electrolysis and liquefaction costs, current producers of green liquid Hydrogen must sell their product Hydrogen at a price of at least \$9.20/kg to break even. Breakthroughs in electrolyzer efficiency and electrolyzer capital cost are likely to remedy these unfavorable economics. However, there remain many unknowns in Hydrogen liquefaction process design. We propose a green Hydrogen liquefaction plant that produces 45 metric tons per day (MTD) of liquid Hydrogen. Vapor feed Hydrogen to the liquefaction process will be sourced upstream by electrolytically splitting water into Hydrogen and oxygen. The electricity to split water and to operate the plant will come from a completely renewable power grid. Our plant design has three novel advantages to preexisting green Hydrogen liquefaction plant design. Namely: 1) A successful implementation of Large-scale Reverse Brayton refrigeration cycle, 2) Actualized Heat Exchanger Design, 3) A specific power of 6.24 kilowatt hours per kilogram of liquid Hydrogen, near the state-of-the-art in conceptual liquefiers.

Assuming a cost of capital of 15%, a plant lifetime of 15 years, a sales price of \$13 per kilogram of LH2, and 100% of vapor feed Hydrogen sourced via water electrolysis, a plant based on the process design detailed herein has an ROI of 16.57%, an IRR of 18.52%, and an NPV of \$44,445,500.

5 INTRODUCTION AND OBJECTIVE TIME CHART

5.1 Motivation

On August 27 1859, a prospector named Edwin Drake made a discovery that would change the world. While working for Seneca Oil Company, he became the first American to successfully drill for oil in Titusville, Pennsylvania. Since then, the chemical energy stored in the bonds of hydrocarbon fossil fuels has powered the world economy, producing over 120 terawatt hours of energy per year since 2010 [1].

In 2023, there are many driving forces behind the pursuit of new sources of energy, such as solar and wind energy, and ways to store energy, such as batteries.

First, the accumulation of greenhouse gases, mainly Carbon Dioxide, in the atmosphere due to fossil fuel combustion since Edwin Drake's discovery (figure 5.1) has resulted in a global anxiety about the adverse effects of increased Carbon Dioxide concentrations on the climate (figure 5.2). Activism by climate scientists and environmentalists directed towards reducing Carbon Dioxide emissions has successfully influenced democratically elected policymakers, and as a result we are seeing large-scale public investments for a future 'Carbon zero' economy. Most notably, the Inflation Reduction Act, signed on August 16 2022 by US President Joe Biden, allocates a $369 billion investment over the next ten years to address 'energy security and climate change' [2]. Various European governments have self-imposed restrictions on their energy sources. For example, Germany has planned to phase out all of their coal-powered power plants by 2038, and on April 15, 2023 shut down its three remaining nuclear power plants, declaring nuclear power as 'a distraction from speeding up development of renewable energy' [3].

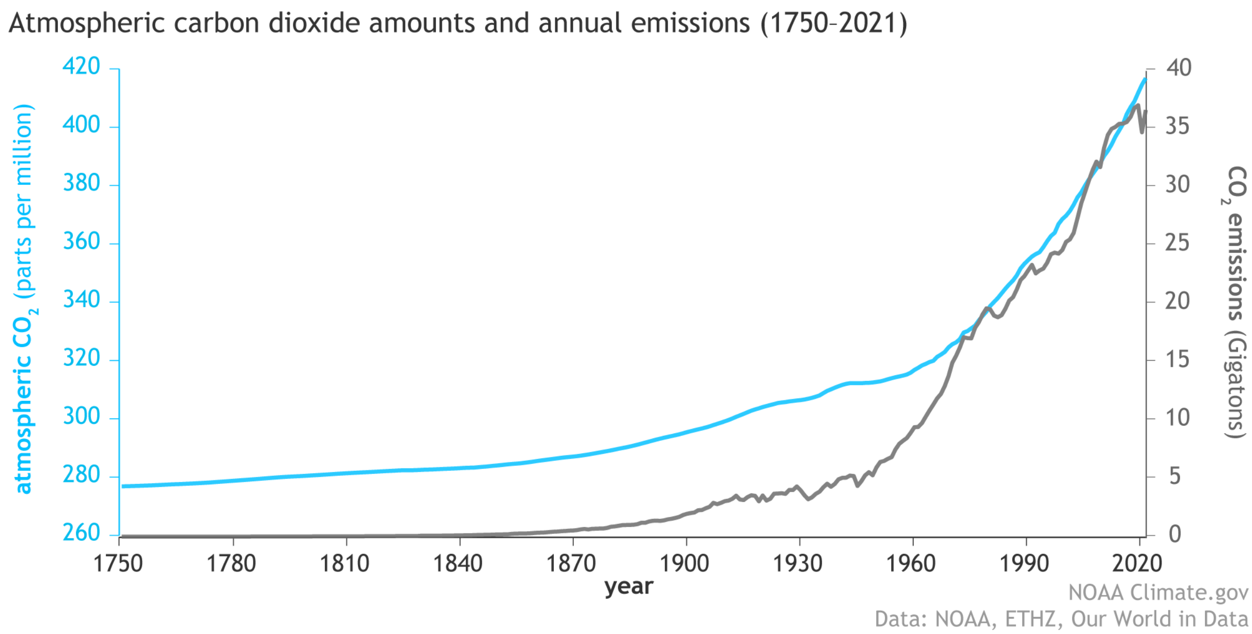

Figure 5.1: source: NOAA Climate.gov [4]. Atmospheric Carbon Dioxide concentrations since 1750 have increased from an equilibrium level of 280 parts per million in 1750 to 412 parts per million in 2020. This increase is mostly due to the release of Carbon Dioxide during the combustion of fossil fuels which humans have been using to power their economy since the 1800s. After being released, the greenhouse gas accumulates in the Earth's atmosphere on a time scale of weeks, while it takes 20-200 years to leave the atmosphere by being absorbed by the ocean's surface, and hundreds of thousands of years via geological processes [5].

Second, geopolitical considerations are driving public investments into renewable energies because countries without access to oil and gas natural resources are currently dependent on other countries' resources. For example, Western Europe's reliance on Russian Natural Gas has become problematic since Russia's recent invasion of Ukraine, because it presents a strategic weakness if Russia ever decides to expand their military aggression on the continent. This dependence has been hypothesized as a principal motive of the mystery saboteurs that destroyed the Nord Stream pipeline that transports Natural Gas from Russia to Germany in September 2022 [6].

Last but not least, technological advances in renewable energies have driven both public and private investments in oil and gas alternatives. Improvements in solar cells, batteries, hydroelectric and wind power, have all made these energy sources economically viable. This is evidenced by the successful transition of many countries to an entirely renewable energy grid such as Iceland, Norway, and Uruguay [7], and the financial profitability of consumer energy businesses like Tesla, Inc. and Enphase Energy. None of these phenomena could happen without the accumulated work of hundreds of thousands of scientists, engineers, and businesspeople.

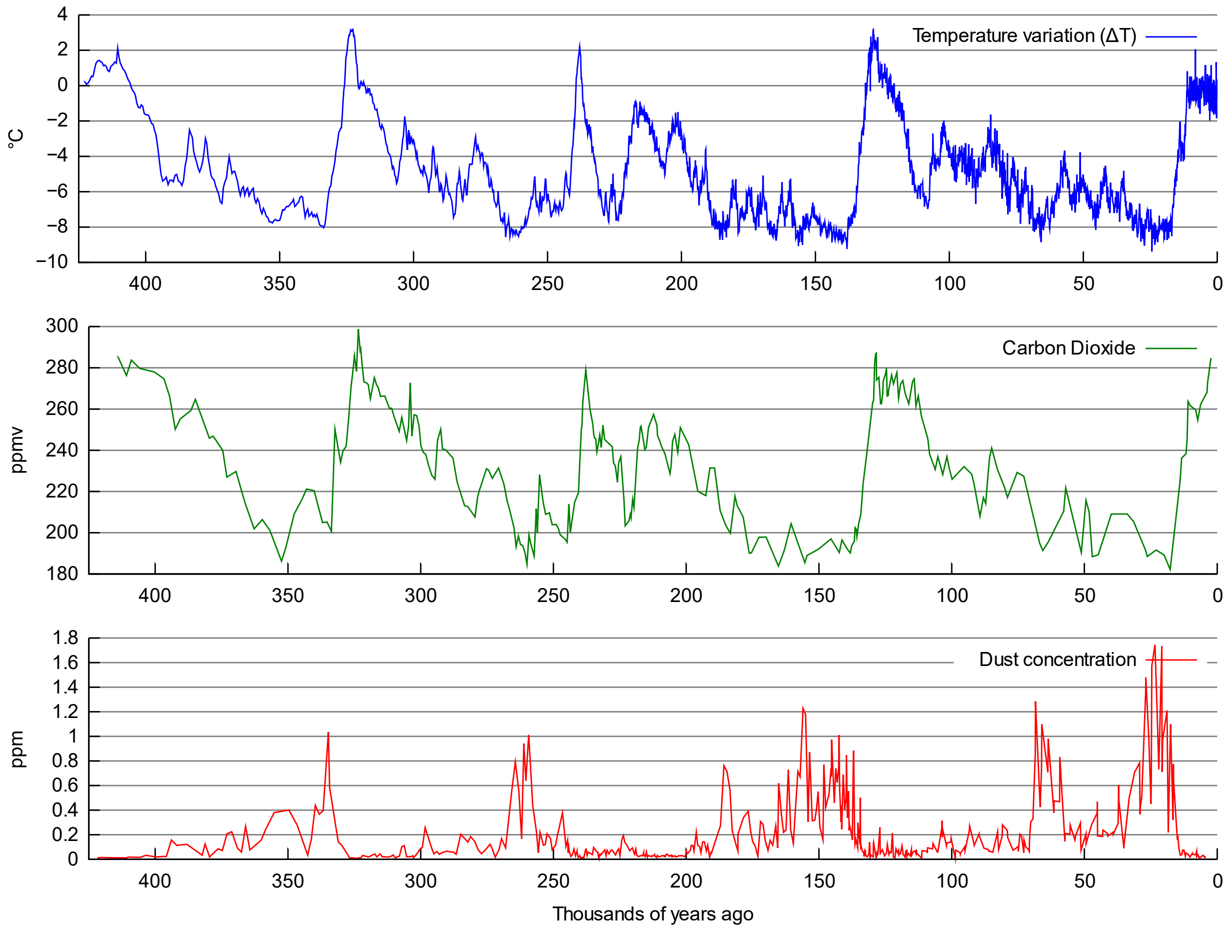

Figure 5.2: Source: NOAA (National Oceanic and Atmospheric Administration) historical data, Vostok ice core. Strong correlation between atmospheric Carbon Dioxide concentrations and average global surface temperatures. The top graph plots the temperature deviation from the 20th century average global surface temperature of 57 °F [8]. By incorporating historical data like these into their atmospheric models, climate scientists estimate that a doubling of CO2 concentration to 560 ppm, which is a likely scenario to reach by 2100, will result in a 4-9 ºF increase in the average global surface temperature [9].

Society is interested in alternative energy sources. These include new, sustainable fuels to power a Carbon neutral economy. Hydrogen is a potential alternative to traditional fossil fuels. The Hydrogen economy and value chain are already being built.

5.2 Background

Hydrogen has emerged as a combustible fuel alternative to traditional fossil fuels. It can be combusted in an internal combustion engine much like fossil fuels, or chemically reacted with oxygen in Hydrogen fuel cells in a more controlled and efficient manner. Unlike hydrocarbons, Hydrogen only releases water vapor (no Carbon Dioxide) when it combusts in the presence of oxygen (Equations 5.1, 5.2).

$$\mathrm{H_2} + \frac{1}{2}\mathrm{O_2} \longrightarrow \mathrm{H_2O} \tag{5.1}$$

$$\Delta\mathrm{H} = -271\,\frac{\mathrm{BTU}}{\mathrm{mol}} \tag{5.2}$$

Like Carbon Dioxide, water vapor is a greenhouse gas. In fact, water vapor is essential for maintaining the temperature conditions that keep the planet livable, accounting for half of Earth's total greenhouse effect [10]. Unlike Carbon Dioxide, however, water vapor is condensable and the concentration of water vapor in the atmosphere is determined by the temperature of the atmosphere. On one hand, this creates a positive feedback loop where more water vapor in the atmosphere increases the temperature, which increases the concentration of water vapor, and it may seem that this would create a runaway effect. On the other hand, there is a negative feedback loop where more water vapor in the atmosphere creates more clouds, which lower the amount of solar energy that reaches the surface of the Earth, but also have an exaggerated greenhouse effect [11]. Evidently the impact of extra water vapor in the atmosphere is not completely understood, and indeed, is an active area of research [12].

In any case, the combustion of Hydrogen is much cleaner than the combustion of hydrocarbons, which can release other greenhouse gases like Carbon monoxide and methane in addition to Carbon Dioxide, and toxic chemicals like nitrogen oxides that contribute to air pollution. It is also more feasible technologically, due to water's relatively high boiling point, to sequester water vapor at the source of combustion as liquid water, than it is to sequester Carbon using the promise of Carbon capture and storage technologies [13]. So while

Hydrogen combustion, despite technically being Carbon neutral, has a healthy amount of skepticism associated with it, Hydrogen's candidacy as an alternative to fossil fuels is a reasonable one.

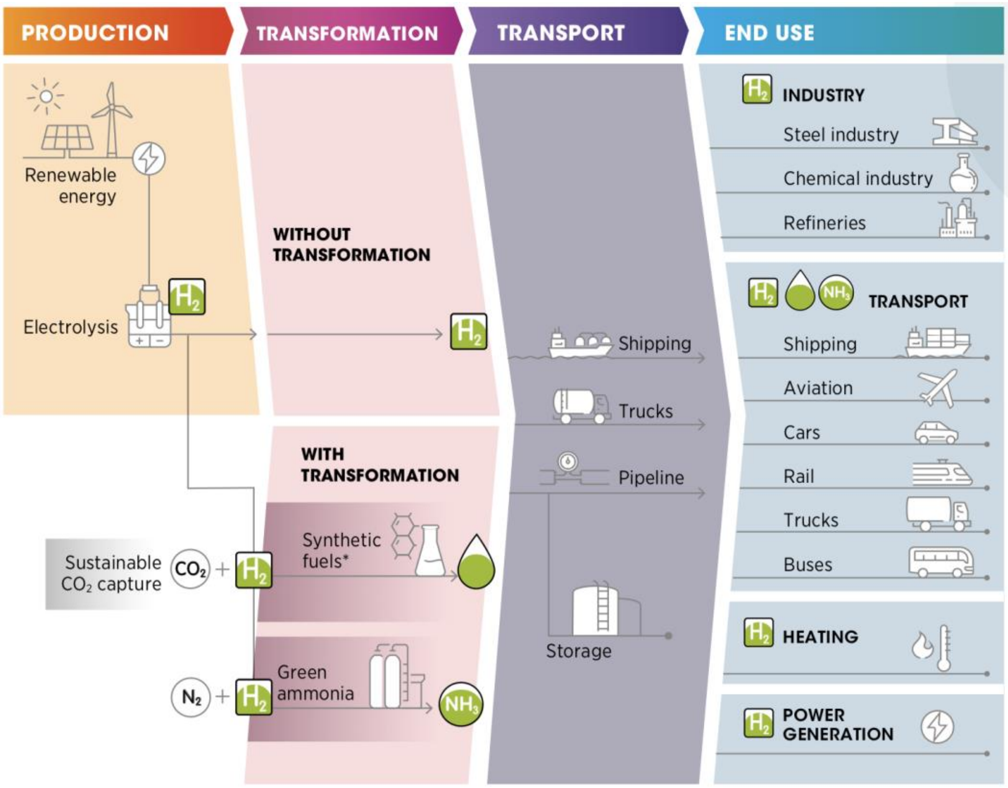

Figure 5.3: Source: IRENA (International Renewable Energy Agency) [14]. The green Hydrogen value chain.

Another drawback to Hydrogen as a fuel is low energy density by volume. Liquid Hydrogen is four times less dense by volume than fossil fuels (figure 5.4).

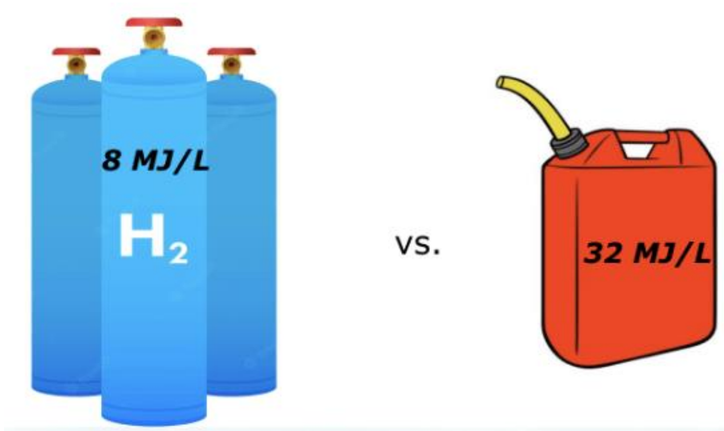

Figure 5.4: Liquid Hydrogen is four times less energy dense than competing hydrocarbons. This presents some issues with the economic feasibility of Hydrogen fuels. In particular, personal vehicles are not easy targets for Hydrogen engines. People are targeting Hydrogen for use in larger vehicles like trucks, ships, planes, and of course, rockets.

There is a possible future in which the world transitions to a Hydrogen economy. In this future, people drive cars with Hydrogen fuel cells and power energy grids with Hydrogen engines (figure 5.3). While most likely such an extreme scenario will not happen, humanity will always need to produce Hydrogen for a myriad of applications. Hydrogen is used as a reactant in the famous Haber-Bosch process to synthesize ammonia for fertilizers, other chemical products, petroleum refining, and as a liquid rocket fuel in the aerospace sector (figure 5.3). As long as Hydrogen is needed, which will be a long time, the production of Hydrogen is a sector that is particularly prepared to be deCarbonized.

Currently, the demand for Hydrogen is being met by Steam-Methane Reforming(SMR), an endothermic conversion of Natural Gas methane to Hydrogen represented by equations 5.3 and 5.4.

$$\mathrm{CH_4 + H_2O \leftrightarrow CO + 3H_2} \tag{5.3}$$

$$\Delta \mathrm{H_{SR}} = 295\ \frac{\mathrm{BTU}}{\mathrm{mol}} \tag{5.4}$$

Currently, 95% or more of global Hydrogen consumed is produced via Steam-Methane Reforming [15]. Hydrogen produced by SMR is not an alternative fuel to fossil fuels because it produces Carbon monoxide, which is a greenhouse gas, as a byproduct. The price of this SMR sourced, ‘grey’ Hydrogen is correlated with Natural Gas prices. This is the cheapest way to produce Hydrogen at scale, but it is not done for any sustainable motives.

A small fraction of SMR plants use Carbon capture to sequester the Carbon monoxide that is released during the process. This is called ‘blue’ Hydrogen.

A promising alternative to SMR is the splitting of water into Hydrogen and oxygen, using electricity and electrolytic cells (figure 5.5). This is exactly the reverse of the exothermic reaction described by equations 5.1 and 5.2. Thus, water electrolysis does not produce any energy; it can only store electrical energy as chemical energy. This is called ‘green’ Hydrogen (figure 5.6). This method of Hydrogen production has the advantage of producing zero Carbon emissions, if the electricity used for the electrolysis reaction is sourced from renewable energy. The price of green Hydrogen is correlated with the price of renewable electricity. Some care must be taken when analyzing the sustainability of a water electrolysis plant because if it is running on an energy grid that is powered by Carbon emitting fossil fuels, then the net emissions of the plant are not zero. That is, a kind of Carbon accounting fraud can be committed.

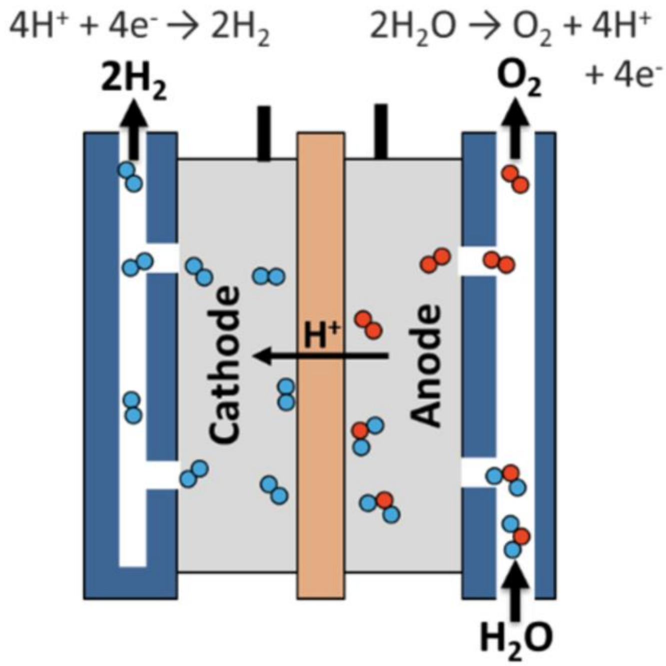

Figure 5.5: Source: DOE [16]. Electrolysis reaction taking place in an electrolytic cell. Water splits into Hydrogen and oxygen in exactly the reverse of the combustion reaction, requiring 271 BTU/mol of electricity in order to run.

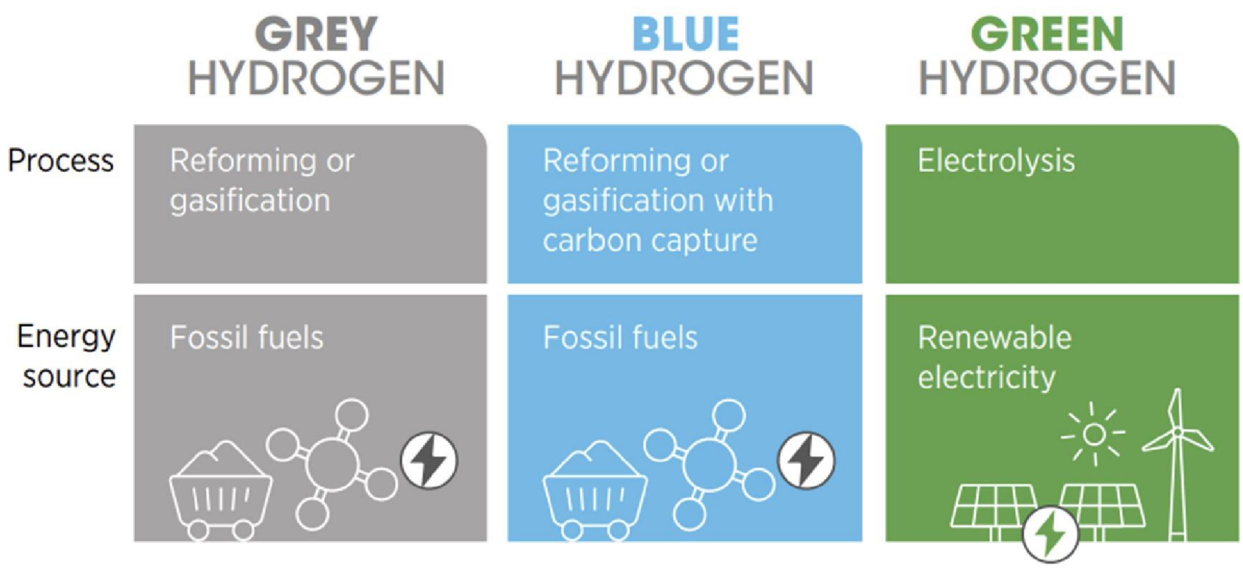

Figure 5.6: The three main types of Hydrogen production. Grey Hydrogen currently dominates the market for economic reasons. Green Hydrogen provides a pathway to truly Carbon neutral Hydrogen production, but is prohibitively expensive. Recent government tax breaks and incentives have begun to make green Hydrogen economically viable (section 6.4). Other promising

methods that may prove lucrative in the future are methane pyrolysis (turquoise Hydrogen) and prospecting of underground Hydrogen reserves.

The economics of green Hydrogen are prohibitive. The process requires a massive amount of electricity and not just any kind of electricity: renewably sourced electricity, which is about 50% more expensive than fossil fuel generated electricity (ten cents per kilowatt hour and seven cents per kilowatt hour, respectively). Add to this the nontrivial capital cost of purchasing industrial scale electrolyzers, an energy efficiency of 70 to 80%, and it becomes clear why only 4% of global Hydrogen is produced via water electrolysis [15].

The current production costs of different Hydrogen production methods are tabulated in table 5.1.

	Carbon Intensity (kg CO_2 per kg H_2)	Production Cost ($ per kg H_2)
Green Hydrogen (Electrolysis)	0	6 to 8
Grey Hydrogen (SMR)	12	2 to 3 (cost of Natural Gas)
Blue Hydrogen (SMR + Carbon Capture)	12	4 to 5
Turquoise Hydrogen (Methane Pyrolysis)	< 0	2 to 3
DOE Hydrogen 'Shot'	0	1

Table 5.1: Source: Solena Group [17]. Methane pyrolysis is a promising technology in its early stages that could prove to be very economical. The DOE has targeted a moonshot goal to reduce the price of green Hydrogen to $1 per 1 kilogram in 1 decade (i.e., by 2030), in a concerted effort to develop Hydrogen as an alternative clean fuel [18]. This would be cheaper than the current cheapest way to produce Hydrogen via Steam-Methane Reformingof Natural Gas.

Other possible Hydrogen sources include methane pyrolysis and naturally occurring underground geological Hydrogen formations.

Methane pyrolysis is a high temperature endothermic reaction that splits methane into solid Carbon and Hydrogen vapor (equations 5.5 and 5.6). By converting methane into Carbon instead of Carbon monoxide or Carbon Dioxide, the Carbon can be easily sequestered and even sold to rubber manufacturers that use pure Carbon in their chemical processes. Two leading private firms are trying to commercialize this process: Solena Group of Washington, D.C., and Monolith Corp of Lincoln, Nebraska [17] [19].

$$\mathrm{CH_4} \longrightarrow \mathrm{C} + \mathrm{H_2} \qquad 5.5$$

$$\Delta H = 71\ \frac{BTU}{mol} \tag{5.6}$$

Commercial scale methane pyrolysis has yet to be tested, but the potential is astronomical. It is cheaper than green Hydrogen, while also being a Carbon-neutral process. A 2022 University of Pennsylvania CBE Senior Design Report reported a 113% return on investment in the third year of a methane pyrolysis plant which produced ten metric tons per day of Hydrogen [20]. In 2021, Monolith Corp. received a 20-year $1 billion loan from the Department of Energy to build a commercial scale methane pyrolysis plant in Hallam, Nebraska [21].

Quite recently, an article was published in Science discussing the possibility of underground Hydrogen stores that could be tapped for Hydrogen [22]. There is a lot of work to be done in order to understand how this Hydrogen forms underground and to develop prospecting methods to predict where to drill for geological Hydrogen. However already there are examples in Africa, Turkey, and the East Coast of the United States of Hydrogen-rich subsurface springs. This Hydrogen could be tapped with minimal Carbon emissions, and would be the cheapest source of Hydrogen available, even cheaper than methane pyrolysis or steam methane reforming. Some preliminary estimates based on geological models estimate that there is enough attainable underground Hydrogen in the world to satisfy humanity's power demands for hundreds of years [23]. If large scale drilling of this natural resource is able to be accomplished, it would be a huge boon for the Hydrogen economy.

Once Hydrogen has been produced by one of the preceding methods, there is the issue of storage. While Hydrogen as vapor is used for its applications in chemical manufacturing, such as ammonia production, Hydrogen for fuel must be stored as liquid in order to have a convenient energy density (figure 5.4). Due to Hydrogen's low boiling point (20.2 K at atmospheric pressure), Hydrogen liquefaction requires the use of a deep cryogenic refrigeration cycle. This liquefier has nontrivial capital and operating costs. The Department of Energy's Hydrogen Delivery Scenario Analysis Model prices a 27 metric ton per day liquefier at $101 million in 2018 [24]. The specific liquefaction cost, which is the operating cost to liquefy

Hydrogen, is estimated in currently operational Hydrogen liquefaction plants to be at $2.75 per kilogram of LH2 [25]. In other words, the cost of liquid grey Hydrogen is about double the price of grey Hydrogen vapor.

Once stored as liquid Hydrogen, a plethora of safety issues arise. Hydrogen is the smallest molecule in the universe, which makes it prone to leaking. As a combustible fuel, this is quite an unsafe phenomenon. This combined with public perception of Hydrogen's lack of safety will be a serious barrier for large scale Hydrogen production to overcome. The public still associates Hydrogen with the Hindenburg Disaster, and the main plot point of a recent Netflix hit blockbuster is the plight of a greedy billionaire to commercialize 'solid Hydrogen' which ends up burning down his idyllic island estate [26]. Given the ability of public sentiment to sway energy policy, perceptions like these can be harmful to develop the technology at scale.

Green Hydrogen is a fascinating topic of great scientific, economic, and political relevance today. Only so much can be written here--the reader is directed towards the sources cited in this section and encouraged to learn more.

Figure 5.7: An infographic prepared by the design group in the early stages of the project, detailing the breakdown of energy consumption, Hydrogen production, and green Hydrogen production in the United States. This was specifically for a plant producing 15 metric tons per day, one third of the production of the final design.

5.3 Goals

The goal of this design project is to contribute to the current wealth of Hydrogen liquefaction plant designs and analyses with our own unique process design and techno-economic model. The design group organized this goal into nine components:

1. Design a process to produce green liquid Hydrogen that improves on the design of US Patent US4765813
2. Complete a working process flow diagram in ASPEN PLUS and optimize for energy efficiency
3. Compare and contrast different refrigerants/working fluids
    a. Neon
    b. Helium
    c. Hydrogen
    d. Mixed Refrigerants
        i. Neon-Helium (Nelium)
        ii. Neon-Hydrogen
        iii. Hydrogen-Helium
4. Accurately model ortho para conversion
5. Design Brazed Aluminum Plate-Fin Heat Exchangers
6. Select a final production amount and working fluid
7. Sensitivity Analysis on the plant ROI vs. the amount of Hydrogen sourced by electrolysis (green) vs. SMR (grey)
8. Calculations of break-even liquid Hydrogen sales price, plant specific power, and specific liquefaction cost
9. Compare different liquefaction process designs and make novel process design

Due to limitations in ASPEN PLUS equation of state modeling (section 9.3), the design group was not able to model mixed refrigerants in ASPEN.

For (6), the design group decided to report results for a 45 metric tons per day liquid Hydrogen plant, using Neon as the process refrigerant.

Due to time limitations, goal (9) was unable to be accomplished.

5.4 Time Chart

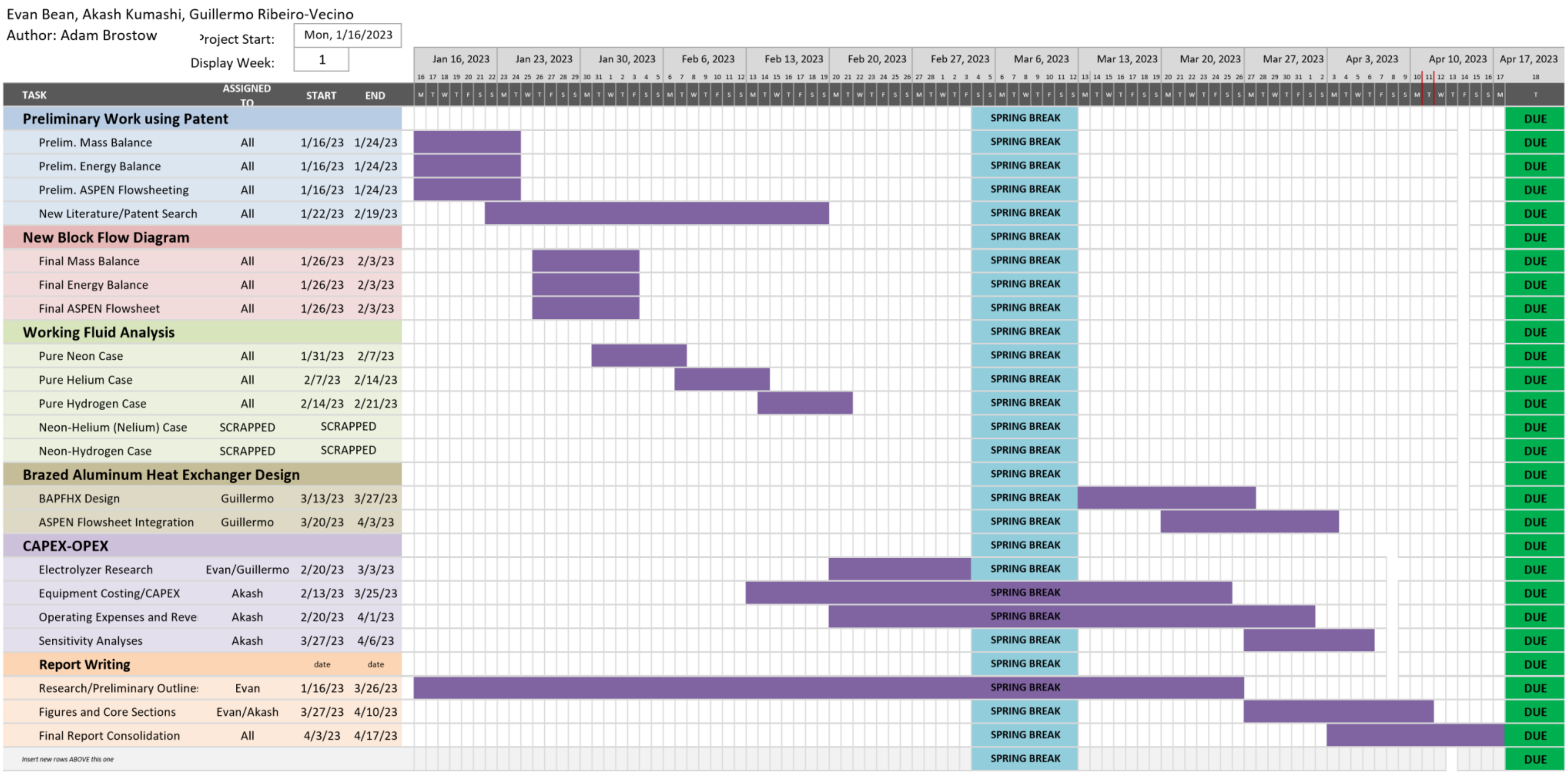

Figure 5.8: Gantt chart of project timeline

6 MARKET AND COMPETITIVE ASSESSMENT

6.1 The Global Green Hydrogen Market

The outlook for Hydrogen brightened over the past decade as governments have collectively mobilized to explore alternatives to greenhouse gas emitting fossil fuels since the Paris Climate Accords in 2016. To this end, the global liquid Hydrogen market is projected to grow at a compound annual growth rate (CAGR of 7.38% to 2030). Currently, the Hydrogen value chain is mostly committed to ammonia production and oil refining (figure 6.1). Future sources of demand for Hydrogen are industries which Hydrogen could help to deCarbonize. This includes the transportation industry (light-duty fuel cells for electric vehicles and large-scale freight vessels such as trailers and ships, and rocket fuel) and the power plant/generation industry.

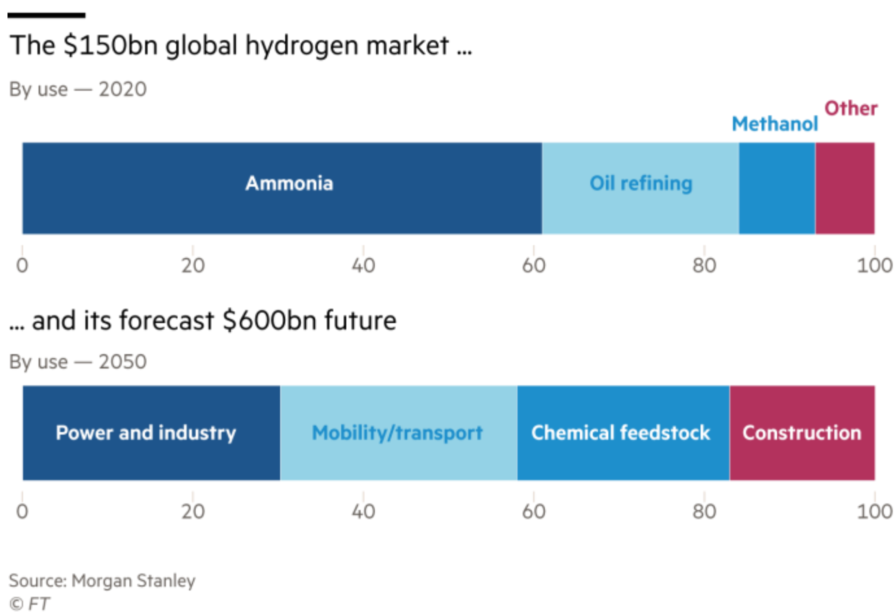

Figure 6.1: Breakdown of 2020 Hydrogen markets [27]. Over sixty percent is used in ammonia production. Experts and policymakers have targeted goals to increase the use of Hydrogen as a clean burning fuel.

The International Renewable Energy Agency (IRENA) published a series of reports outlining a pathway to an optimistic 2050 scenario where 12 percent of global energy demand is met by Hydrogen. The most important development to make green Hydrogen viable at this scale is advances in the economics of renewable energy like solar and wind [28]. In the IRENA scenario, renewable electricity prices drop by 80-90% in order to lower the cost of electrolytic Hydrogen to $1 per kilogram, which is the price necessary to reach 12 percent of global energy demand. Hydrogen liquefaction plants will be ubiquitous and necessary for the Hydrogen supply chain much in the same way Natural Gas liquefaction (LNG) plants exist today, since Hydrogen fueling stations and transport by sea will require the higher energy density of liquid Hydrogen. Natural Gas pipelines can be refitted to transport Hydrogen gas.

National governments have set personal benchmarks for green Hydrogen because they have realized that their goals for Carbon neutrality by 2050 are not possible without clean burning fuels [28]. Renewable electricity will not be enough. By 2030, the International Energy Association (IEA) projects a demand for green Hydrogen of over 60 million tons, about one third of the total Hydrogen demand projection for that same year [29]. In anticipation of the future importance of green Hydrogen in the global energy picture, the industry estimates that $150 billion has already been allocated by governments towards the development of Hydrogen production technologies in the form of loan subsidies and research funding [27]. To get a sense of scale, that is the size of the entire global Hydrogen market in 2020 (figure 6.1).

In the past three years, more water electrolysis plants have been built, or slated to be built, than there were in existence five years ago. For example, an oil refinery in Germany built a 30 MW (4.5 metric tons per day/MTPD) electrolysis plant in 2020, the first step towards Germany's goal of 5000 MW (750 MTPD) of green Hydrogen production by 2030 [30]. A 2021 mega-plant venture between Air Products, ACWA Power, and Neom plans to build a 4000 MW (600 MTPD) renewably powered water electrolysis plant in sunny Kingdom of Saudi Arabia [31]. On October 2021, INEOS billionaire Jim Ratcliffe announced plans to build two billion euros worth of green Hydrogen plants in Europe [32].

6.2 Liquid Hydrogen Market and Competition

Hydrogen liquefaction plants have proliferated along with electrolysis plants, in anticipation of the larger role Hydrogen will occupy in the global economy. Table 6.1 illustrates the recent liquefaction frenzy that is only going to accelerate in the next decade. Air Liquide began operating a 30 MTPD Las Vegas plant in 2022, with Hydrogen sourced from Natural Gas at the moment, but with the ability and expectation to transition towards water electrolysis sourced Hydrogen in the future, anticipating the necessary drops in renewable electricity prices to make this financially profitable [33]. Air Products is spending half a billion dollars to build a 35 MTPD green liquid Hydrogen plant in Massena, New York, that will produce electricity for water splitting from the St. Lawrence River [34].

Location	Operator	Capacity (MTPD)	Year Constructed
Kimitsu, Japan	Nippon Steel Corporation	0.2	2004
Saggonda, India	Andrha sugars	1.2	2004
Osaka, Japan	Iwatani	10	2006
Leuna, Germany	Linde	5	2008
Chiba, Japan	Iwatani	5	2009
Yamaguchi, Japan	Iwatani and Tokuyama	5	2013
Akashi, Japan	Kawasaki Heavy Industries	5	2014
Yamaguchi, Japan	Iwatani and Tokuyama	10	2017
Port of Hastings, Australia	HESC	0.25	2020
Las Vegas, USA	Air Liquide	27.2	2020
Leuna, Germany	Linde	10	2021
La Porte, USA	Air Products	27.2	2021
La Porte, USA	Praxair	27.2	2021
California, USA	Air Products	10 to 30	2021
Ulsan, Korea	Hyosung and Linde	13	2022
2023 CBE Senior Design		45	2026

Table 6.1: Source: [25]. Hydrogen liquefaction plants constructed within the last twenty years. Since 2020 there has been an explosive proliferation of 20+ MTPD Hydrogen liquefaction plants. The last row is for comparison.

In the United States, significant investments have been made towards the production of liquid Hydrogen (LH2) both at a federal and state level. The passage of the Inflation Reduction Act of 2022 (IRA) opened the door to federal grants and tax subsidies for companies producing clean energy alternatives and those that sequestered Carbon from the atmosphere. Table 6.2 in section 6.3 details the tax breaks available for

Hydrogen production. Beyond that, multiple states including California have either through legislation or ad-hoc negotiation incentivized consumer demand for LH2.

The Department of Energy has presented a 2030 moonshot objective of lowering the purchase cost of green LH2 to just $1/kg. Drastic engineering solutions are needed to meet this goal, and for that to be possible there must be a consumer Hydrogen supply chain infrastructure to utilize LH2 at scale: ubiquitous fueling stations, Hydrogen fuel cells in cars, etc. These currently do not meaningfully exist for small vehicles, but the industry agrees that LH2 will more quickly be adopted by large vehicles such as trucks, airplanes and barges. In the context of the current reality of the Hydrogen economy, the design group has chosen to market its liquid Hydrogen product to consumers who are at present are capable of consuming LH2, and consumers who are likely to join the Hydrogen economy in the next decade. Primarily, this means public and private sector rocket launch centers, commercial airports, and seaports.

6.2.1 Cost of Liquefaction

Currently, the specific liquefaction cost, which is the cost of Hydrogen liquefaction, is $2.75 per kilogram of LH2. In order to compare the economic feasibility of its design to this benchmark and other conceptual designs (figure 6.2), the design group calculated the specific liquefaction cost of its plant after a rigorous profitability analysis to be between $2.71 per kg LH2 and $3.40 per kg LH2 (section 20). Targets are to lower the cost to $1 per kg LH2.

In order to understand the long-term viability of liquid Hydrogen as a fuel source, a sensitivity analysis on the relative costs of Steam-Methane Reforming (SMR) versus electrolysis was performed. This is of particular importance because most Hydrogen liquefaction plants use SMR. Figures 6.3 and 6.4 below, generated using our profitability spreadsheet, indicate that for a plant that is 0% green (100% reliant on SMR), the break-even liquid Hydrogen sales price is $4.40 per kilogram of LH2. This is in line with conclusions from preexisting literature. The cost of liquefaction (0% electrolysis) is demonstrated in Section 24.6.2, and the cost of *green* liquefaction is demonstrated in Section 24.6.1.

The cost of *green* liquefaction for our process is \$2.71 per kilogram of LH2. Currently operating Hydrogen liquefaction plants have a specific liquefaction cost (SLC) of \$2.75 per kilogram of LH2, so little to no improvement over these economics was achieved by the design group. Newer theoretical plant designs cite SLCs lower than \$1 per kilogram of LH2, and this has become the fresh target for newly built Hydrogen liquefaction plants.

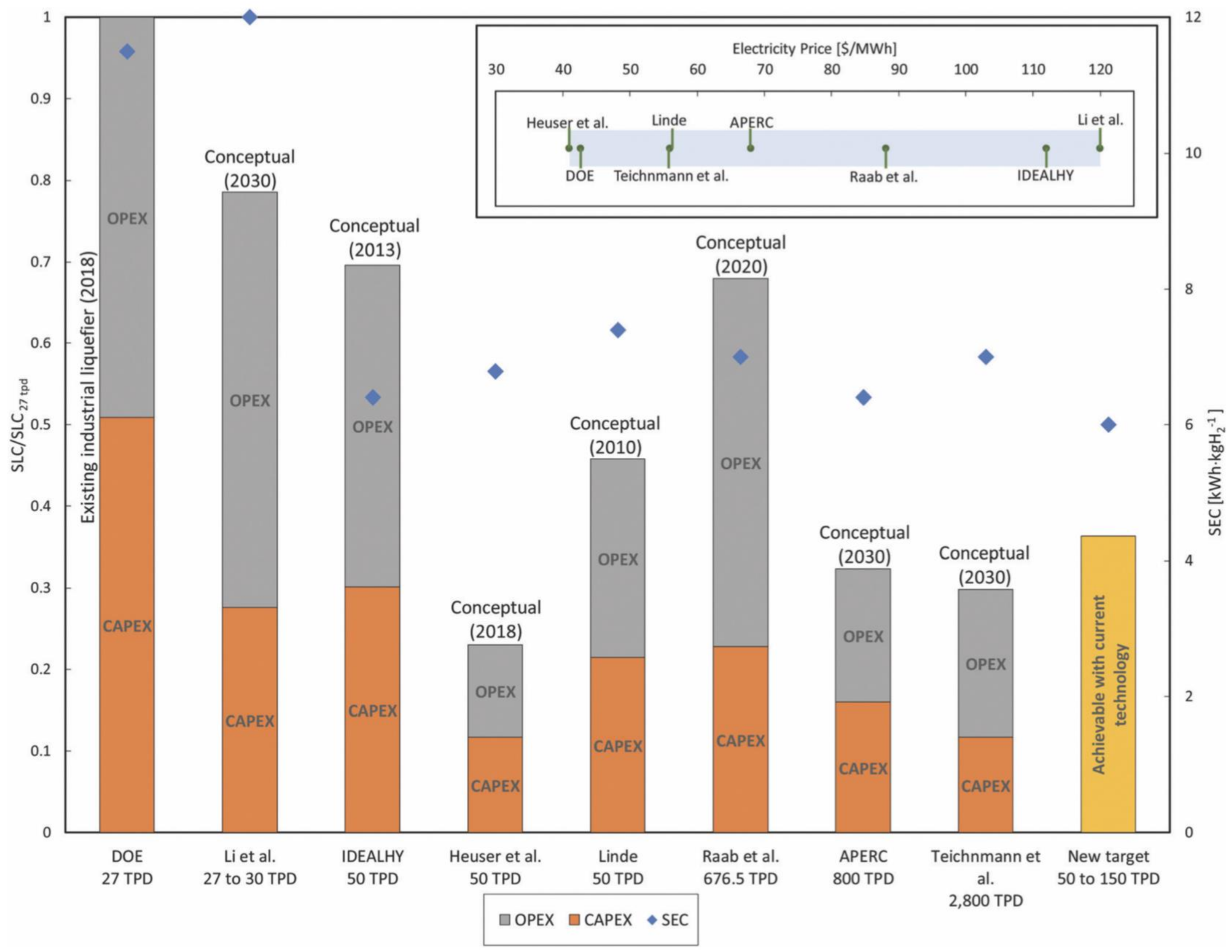

Figure 6.2: [25]. Conceptual liquefier designs rated on their specific liquefaction cost relative to the existing liquefier costs of \$2.75 per kilogram of Hydrogen [24].

The theoretical low limit for the cost of green Hydrogen may be calculated by subtracting the cost of *green* liquefaction from the break-even price of LH2 at 100% electrolysis. This calculation is demonstrated in appendix 24.6 to be \$6.49 per kg, and for a plant operating at 100% SMR, \$1.20 per kg LH2. This is consistent with the economics listed in table 5.1.

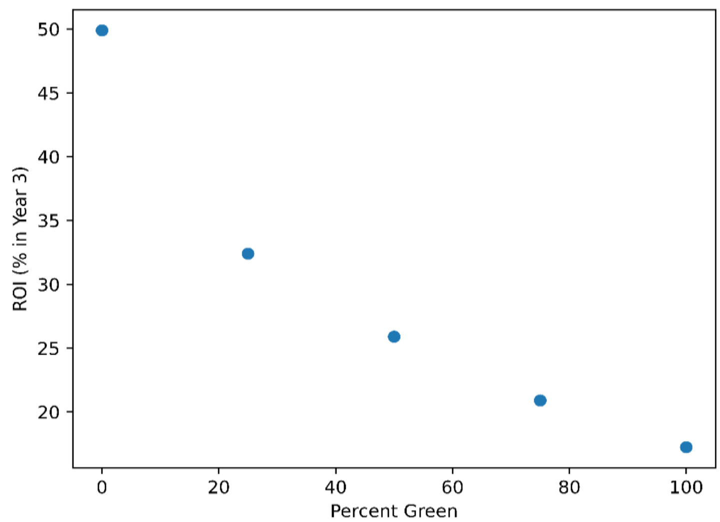

Figure 6.3: Year 3 ROI as a function of 'greenness' of Hydrogen liquefaction. Hydrogen not sourced by electrolysis of water is produced by steam methane reforming

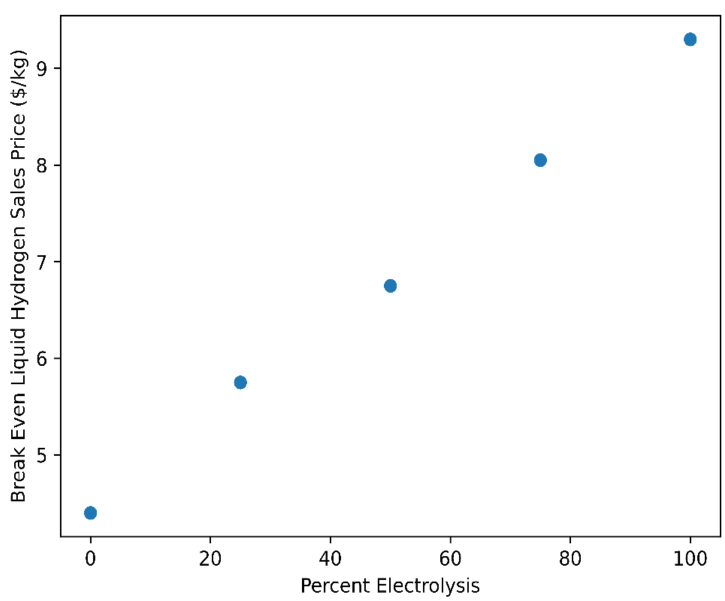

Figure 6.4: breakeven sales price of liquid Hydrogen as a function of 'greenness' of Hydrogen liquefaction. Hydrogen not sourced by electrolysis of water is produced by steam methane reforming.

6.3 Green Premium

Green Hydrogen is significantly more expensive than Hydrogen produced from steam methane reforming: \$6 vs \$1.20.

To incentivize private investments in green Hydrogen, the Inflation Reduction Act of 2022 contains tax subsidies for Hydrogen plants. The subsidy of \$3 per kilogram of Hydrogen for a Carbon neutral plant is relatively huge, taking into account that the production of Hydrogen via SMR is less expensive than that subsidy. These tax breaks were incorporated into the design group's profitability analysis in section 20.

Kg CO2 emitted per kg LH2 produced	***Tax credit (\$/kg LH2)***
2.5 – 4	\$0.60
1.5 – 2.5	\$0.75
0.45 – 1.5	\$1.00
0.00 – 0.45	\$3.00

Table 6.2: Inflation Reduction Act (IRA) tax credits for liquid Hydrogen production, Section 45V [35].

For reference, an SMR plant emits 12 kg Carbon Dioxide per kilogram of Hydrogen.

7 Consumer Requirements and Design Constraints

7.1 Plant Location

Our plant will be located in St. John's County on the Florida Gulf Coast. In lieu of feedstock water requirements, water availability on the Florida Gulf Coast is discussed in appendix section 24.9. California was initially considered because of its preexisting Hydrogen supply chain infrastructure, but was ultimately dismissed due to unfavorable economics.

7.2 Product Specification

Normal Hydrogen has two spin isomers- ortho-Hydrogen (o-H_2) and para-Hydrogen (p-H_2). Hydrogen liquefaction is complicated by o-H_2's spontaneous, exothermic reaction to p-H_2, as liquid Hydrogen is stored at extremely low temperatures and high pressures (Section 24.2). To avoid Hydrogen boil-off and potential explosion hazards from inadvertent vaporization, vendors of LH2 should aim to maximize the fraction of para-Hydrogen present in their product sample.

Our product LH2 will be sold to consumers as 99.0% p-H2 saturated liquid, at 3 bars, -414 F. Legacy producers sell their LH2 product as 99.8+% p-H_2 at 1 bar, -421 F.

7.3 Cryogenic Heat Exchangers

7.3.1 General Constraints

The heat exchangers comprise the central components of the refrigeration cycle. They provide the medium through which Neon and Nitrogen refrigerants can exchange heat with the Hydrogen vapor in order to cool the Hydrogen vapor down near its boiling point.

Additionally, the heat exchangers in this process serve a dual function as reactors packed with ferric oxide catalyst in which conversion of the ortho spin isomer of Hydrogen to the para spin isomer of Hydrogen occurs. This catalyst packing occurs in heat exchangers 26 and 30 in the passages containing the combined feed-recycle Hydrogen stream.

Finally, the sizes of the heat exchangers determine the total volume of material in the heat exchangers and the pipe sizing of the liquefaction plant. Using these sizes, the total one-time purchase cost of the Neon refrigerant is calculated (see section 11.3.1).

For cryogenic systems, brazed aluminum plate fin heat exchangers (BAPFHX) are used due to their superior heat transfer properties compared to the alternative shell and tube heat exchangers. These properties are enumerated by various BAPFHX vendors [36] [37]:

1. Higher heat transfer, 10-20x more UA per volume
2. More compact, 20% size of alternatives and 95% lighter
3. Multiple streams: more flexibility in stream contact design
4. 25-50% lower initial cost
5. Reduced temperature approaches, as low as 2-3 °F (1-2 K) result in downstream reduction in power consumption

Property (3) is particularly valuable, because it means that instead of designing a heat exchanger network of shell and tube heat exchangers to transfer heat between more than two streams, the BAPFHX can take in any number of streams and be optimized to maximize heat transfer efficiency. This is a major source of capital cost reduction between a plant that uses BAPFHX and a plant that uses shell and tube (two-stream) heat exchanger networks.

Property (5) pertains to cryogenic systems involving refrigeration cycles. Reducing approach temperatures in the heat exchanger lowers the pressure drop (expansion) that is needed in the refrigeration cycle to cool the refrigerant, thus reducing the power requirement of the compressor that supplies the mechanical work for the refrigeration cycle.

Deep cryogenic systems like this Hydrogen liquefaction plant require small approach temperatures, as low as 1-2 K or 2-3 °F, in order to minimize the amount of lost work due to entropic processes during heat transfer [38]. BAPFHX are uniquely able to deliver these approach temperatures reliably and thus are a key component to any cryogenic plant.

Despite their advantages, BAPFHX are not used universally for heat transfer applications because of their fragility under high thermal stresses. Accordingly, the local temperature difference at any point along the heat exchanger is limited to $\Delta T \leq 54\,°F\,(30\,K)$, which are typical approach temperatures for systems operating above 77 °F (298 K) [39]. Such gradients weaken the structural integrity of the unit and lead to external material leaks developing over time. Cryogenic systems, which enjoy approach temperatures as small as 2-3 °F (1-2 K), can exploit the advantages afforded by these heat exchangers with no risk of thermal stresses damaging the unit. In this process, particularly, heat exchanger 18 undergoes the largest thermal stresses with a maximum local temperature difference of 36 °F, but this is well within the limits. Heat exchangers 26 and 30, operating at or below -367°F (51.5 K) have tight composite curves in which the maximal temperature differences are just 4°F and 10°F, respectively.

7.3.2 Size Constraints due to Packed Catalyst in Heat Exchanger Passages

Cutting edge Hydrogen liquefaction process design relies on catalyst packed inside heat exchanger passages where hot Hydrogen flows [40]. However, BAHX passages are extremely small and intricately finned. Heat exchanger sizing must account for the necessary volume of packed catalyst within hot Hydrogen passages while simultaneously ensuring that the remaining effective heat transfer area is sufficient to complete required heat exchange. An in-depth discussion on heat exchanger sizing taking into account packed catalyst is covered in Section 24.1.1.

Here, an estimate of the amount of ferric oxide catalyst required for a process configuration producing 45 MTD of LH2 is developed. According to Zhuzhgov et al, sample calculations from a miniature Hydrogen liquefier system with an exiting flowrate of 100 kg/hr LH2 require 80 kg of Fe_2O_3 catalyst [41]. A quick unit conversion may be performed to compare this production capacity to our selected process configuration of 45 MTD (tonnes/day), assuming the catalyst weight requirement is linearly related to the plant production capacity.

$$(100\ kg/hr)\ *\ (24\ hrs/day) *\ (\mathrm{MTD}/1000\ \mathrm{kg})\ =\ 2.4\ \mathrm{tonnes/day}$$

$$(45\ \mathrm{tonnes/day})\ /\ (2.4\ \mathrm{tonnes/day})\ *\ (80\ \mathrm{kg}\ \mathrm{Fe_2O_3})\ =\ 1500\ \mathrm{kg\ of}\ \mathrm{Fe_2O_3}.$$

Heat Exchanger cores in Blocks 26 and 30 must together be able to accommodate this amount of catalyst in order to reach 99.0% p-H2 composition after Block 30. A contingency solution is to pack catalyst that cannot be accommodated in the Heat Exchanger cores in the KO drum. This calculation is discussed in Section 14.7 and Section 24.6.3.

7.4 Refrigerant Availability

Neon, helium, and Hydrogen were evaluated as the three refrigerant candidates for this process. Neon is produced by separating it from air. It comprises 18 ppm of the atmosphere, by molar amount. Typically, Neon is a byproduct of steel production air separation units where process oxygen is purified from air of nitrogen, Carbon Dioxide, and its noble gases. This noble gas distillate contains argon, helium, and krypton as well as Neon, and is distilled further to obtain pure Neon. The major consumer of Neon is the semiconductor industry, which uses Neon as part of its silicon wafer laser etching process (photolithography). Neon use as a cryogenic refrigerant is a secondary market.

50% of Neon production was accounted for by private chemical manufacturing companies in Ukraine, but not anymore since Russia's invasion on February 24, 2022. The major producers were Cryoin, Ingas, and Iceblick, all of which are no longer producing Neon, probably because their factories have either been repurposed for weapons manufacturing, occupied by Russians, or destroyed. The rest was produced by China, which is now responsible for producing the vast majority of the global supply.

Since Neon is a rare gas with relatively low demand (Figure 7.1), it is not available to be publicly traded on the commodity markets. Thus, it does not have a very liquid market and prices are subject to large fluctuations. Moreover, due to the centralization of Neon production in Ukraine and China, the supply chain is sensitive to geopolitical conflicts such as wars or trade disputes. Before the Russian invasion of Ukraine, Neon prices were around \$160 per pound. Due to the supply shortages created by the invasion, prices have soared nearly tenfold, up to \$1300 per pound, according to South Korean news outlets [42].

In any case, the choice of working fluid for Hydrogen refrigeration is dependent on the price of Neon. Helium is cheaper per unit pound than Neon, priced around \$23 per pound. The helium supply chain has been under strain for the past decade but it is mostly sourced from purifying geologically trapped Natural Gas in Texas, Oklahoma, and Kansas instead of air separation. Hydrogen is by far the cheapest refrigerant per unit pound, and the amount of Hydrogen necessary to use as a refrigerant would be sourced at essentially zero cost from the electrolysis units upstream of the liquefaction plant. It is worth noting that Hydrogen and helium are the most common and second most common elements in the universe.

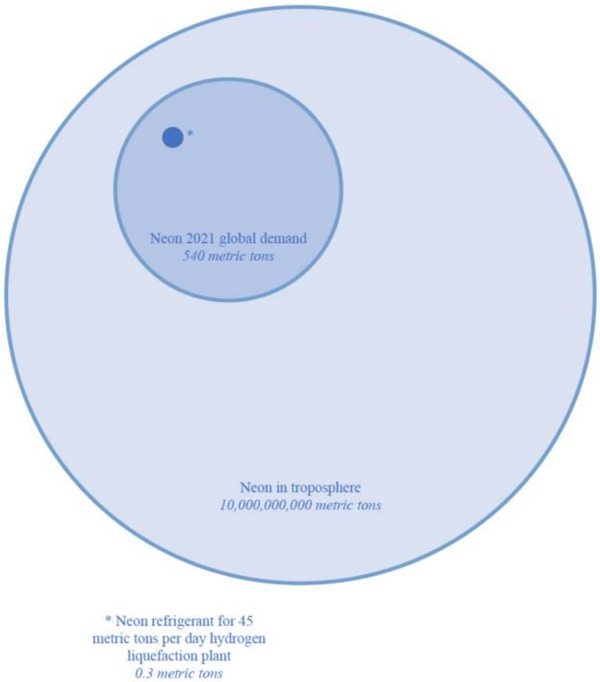

Figure 7.1: Visualization of plant Neon requirement, global Neon demand in 2021 [43], and all the Neon available in the troposphere, which comprises the first ten thousand meters of the atmosphere above sea level. A Hydrogen-based economy should be able to source all the Neon refrigerant it needs for its liquefaction plants.

8 COMPETITIVE (PATENT) ANALYSIS

There are a plethora of Hydrogen liquefaction processes in the literature. Cardella et al. have completed comparative analyses of different plants [44]. The main thermodynamic cycles used in these refrigerators are the Claude cycle, the reverse Brayton cycle, and the Linde cycle. Conventionally, Claude cycles are used for smaller production liquefiers (2 to 15 MTPD), while reverse Brayton cycles scale more appropriately to production rates greater than 15 MTPD. Additionally, different processes use different refrigerants. The most common are Neon, helium, and Hydrogen, but creative processes also used mixed refrigerants using these nonreactive low boiling elements and hydrocarbons. For detailed information on these processes, the reader is directed to [25].

Our process is modeled from USA Patent 4765813 (US4765813), filed in 1988 by Lee Gaumer Jr. and Arthur Winters Jr. of Air Products and Chemicals, entitled "Hydrogen Liquefaction Using a Dense Fluid Expander and Neon as a Precoolant Refrigerant." [45] The patent uses a Reverse Brayton refrigeration cycle (figure 8.1) with Neon as the refrigerant to cool pressurized Hydrogen near its boiling point. The cooled Hydrogen is then expanded adiabatically in a dense fluid expander according to an isentropic efficiency of 70%, thereby liquefying it. A key insight provided by the patent is that the use of a dense fluid expander to liquefy the Hydrogen makes the process more energy efficient than throttling the Hydrogen through a Joule-Thomson valve (an isenthalpic expansion). For needed additional cooling, the process uses liquid Nitrogen at 77 K.

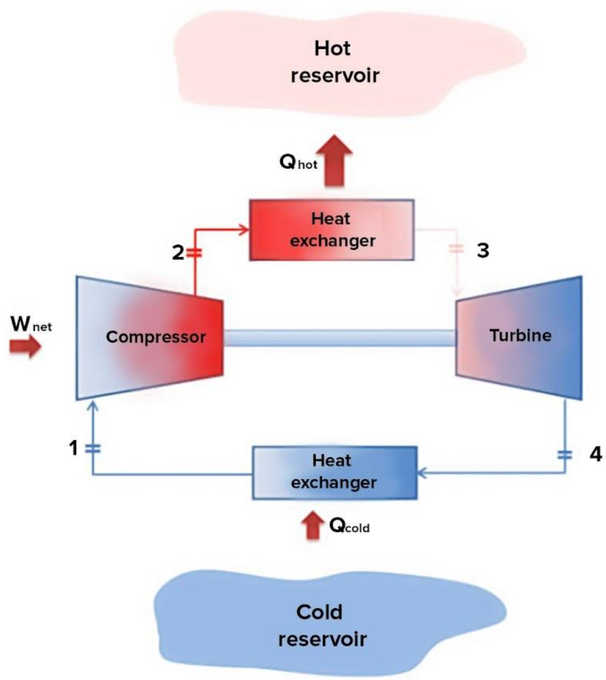

Figure 8.1: An Ideal Reverse Brayton refrigeration cycle. Isentropic compression, isobaric heat exchange of hot fluid to cold fluid, isentropic expansion, isobaric heat exchange of hot fluid to cold fluid. In practice, the pressure changes are never truly isentropic due to the second law of thermodynamics. Instead, the compressors and expanders have associated isentropic efficiencies.

A brief narration of the patent's process is as follows. Pressurized vapor Hydrogen feed at 110 ºF and 650 psia travels through the first two heat exchangers (blocks 18 and 22) and is cooled to cryogenic conditions with a combination of refrigeration delivered from the Neon refrigerant, liquid Nitrogen, and cold recycle p-H2. Next, the vapor Hydrogen at -310 F and 625 psia travels through two dual heat exchanger/reactors (Blocks 26 and 30), in which ferric oxide catalyst is packed within the passages of the heat exchanger where the hot Hydrogen travels. In these blocks, o-H2 is converted into p-H2. Lastly, the liquefied Hydrogen at cryogenic conditions (-404 F, 600 psia) is expanded to both recover power and satisfy customer requirements for liquefied Hydrogen (-414 F, 52 psia). The process flow diagram illustrating this narration is attached on the next page.

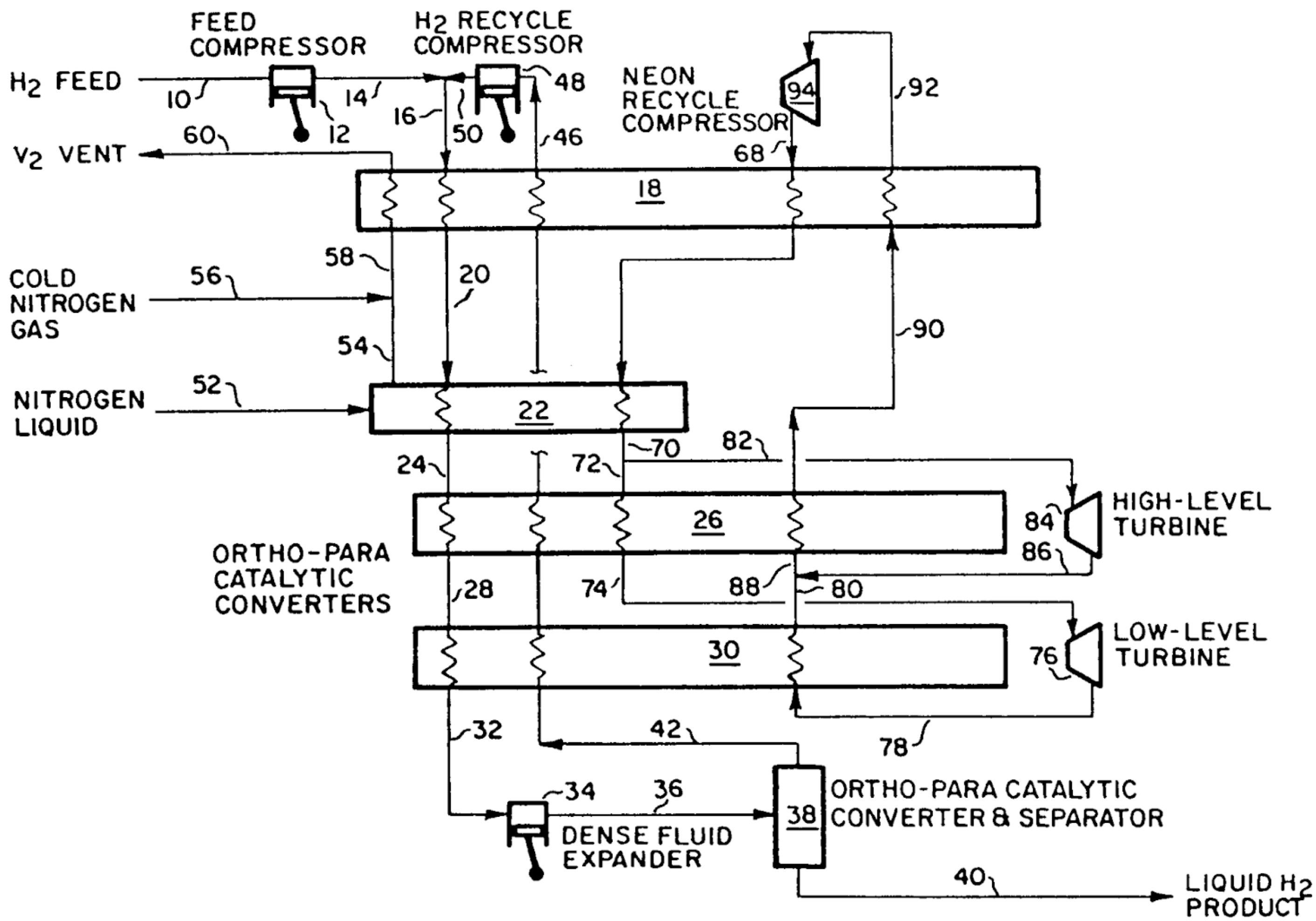

Figure 8.2: US4765813 Process Flow Diagram

A secondary candidate to our selected process was modeled from Patent AU2016344553B2, filed in 2016 by Umberto Cardella, Lutz Decker, and Herald Klein of Linde, entitled "Large-Scale Hydrogen Liquefaction by means of a High Pressure Hydrogen Refrigeration cycle combined to a novel Single Mixed-Refrigerant Precooling." This patent uses a modified Claude refrigeration cycle with Hydrogen as the refrigerant. For needed additional cooling, the process uses a combination of liquid nitrogen and a 5-component mixed refrigerant system at cryogenic temperatures. This process design was dismissed because mixed-refrigerant modeling with Hydrogen liquefaction processes in ASPEN PLUS is prohibitively complex. The process flow diagram depicting Patent AU2016344553B2 is attached on the next page.

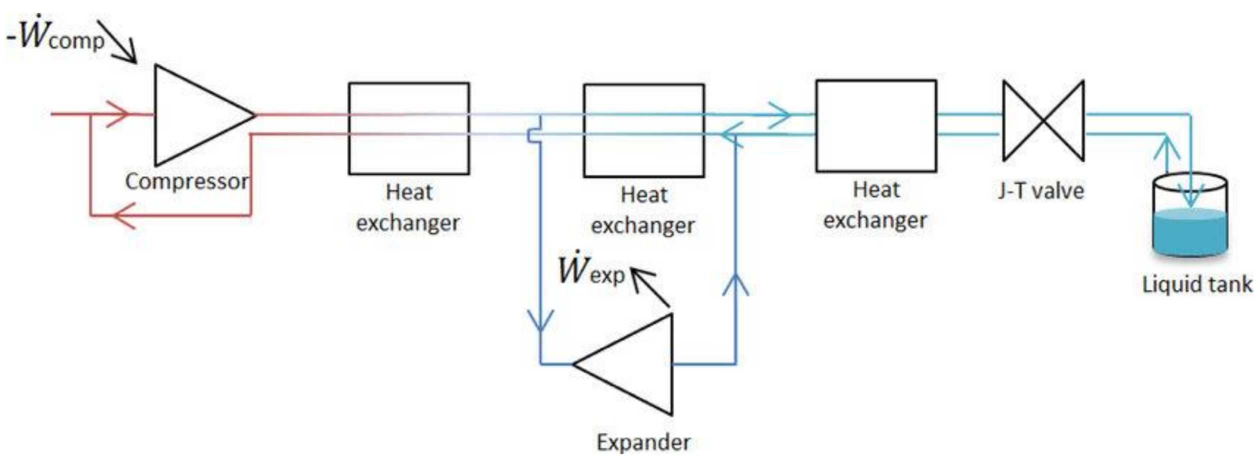

Figure 8.3: Claude refrigeration cycle.

9 PRELIMINARY PROCESS SYNTHESIS

9.1 Refrigeration Cycle Considerations

There are numerous Hydrogen liquefaction plant designs in the literature. In essence, they are all industrial scale refrigerators. There are various ways to design a refrigerator. The main design decisions are the choice of thermodynamic refrigeration cycle, the choice of refrigerant/working fluid, and whether to optimize the design for energy efficiency or ROI.

We worked to reproduce a Hydrogen liquefaction process designed in a 1988 Air Products and Chemicals Inc. patent (Patent US4765813). The patent uses a Reverse Brayton refrigeration cycle with Neon as the refrigerant to cool pressurized Hydrogen near its boiling point. The cooled Hydrogen is then expanded adiabatically in a dense fluid expander according to an isentropic efficiency of 70%, thereby liquefying it. For needed additional cooling, the process uses liquid Nitrogen at 77 K. After reproducing the results of the patent in ASPEN Plus, we repeated the process design using helium and then Hydrogen as the process' refrigerants. We also considered a process that produced its own liquid Nitrogen rather than sourcing it externally.

9.2 Using Reactors to model Ortho-Para conversion

Initial process design used reactors upstream of Heat Exchangers 26 and 30 to convert ortho-Hydrogen to para-Hydrogen. The RSTOICH block in ASPEN PLUS was selected for both reactors. Calculator blocks were used to define the heat of reaction based on the temperature conditions of the stream. However, this injected the total heat of reaction at either the inlet of the heat exchanger or the outlet of the exchanger, depending on the configuration. As such, it was discovered that the RSTOICH blocks were imprecisely calculating the heat of conversion from ortho-Hydrogen to para-Hydrogen and altering the temperature profiles of the system, making an unrealistic approximation of the phenomenon. Moreover, despite certainty in the temperature dependent curve used to calculate heat of reaction, RSTOICH consistently provided unrealistic results in its calculator block. It is unclear if ASPEN's inaccurate calculation of heats of conversion was a downstream consequence of the choice of equation of state or the choice of reactor. For reasons discussed in Section 9.3, reactors were not used in the final process design because the selected EOS for this process does not allow mixed-species streams.

9.3 Thermodynamic Property Modeling/Equation of State Selection

The design group initially chose the Peng-Robinson equation of state (P-R EOS) in order to model the thermodynamic properties of the process due to its broad applicability and its ability to model Natural Gas systems. Three running ASPEN Plus flowsheets were created, one for each working fluid choice, using the P-R EOS. Each of these flowsheets produced 15 metric tons per day (MTPD) of LH2.

While the design group initially did not have intuition for how the system should appear, each of our Peng-Robinson modeled ASPEN Plus flowsheets had large refrigerant compressor power requirements, and the Hydrogen case had the largest. Upon investigating the large compressor power requirements, we realized that the P-R EOS inaccurately modeled the adiabatic expansion (85% isentropic efficiency) of the Hydrogen refrigerant in expanders 76 and 84 (see process flow diagram, figure 11.2). In particular, the C_p to C_v ratio it was calculating for Hydrogen was 50% lower than experimental data provided by NIST. The larger this ratio is for a given material, the more sensitive the temperatures of that material's vapor phase are to isentropic pressure changes. When the pressure drops adiabatically with a certain isentropic efficiency, a material with a larger C_p to C_v ratio will experience a larger decrease in temperature than a material with a smaller one. Peng-Robinson was unable to model the quantum mechanical nature of Hydrogen's spin isomers that dominates in cryogenic temperature ranges (appendix 24.2), and was predicting diatomic ideal gas behavior, i.e., a ratio of four-thirds[1]. As a result, the Hydrogen refrigerant required larger pressure drops to achieve the same temperature drops as Neon and helium, making the Hydrogen case less energy efficient than the Neon and helium cases, although it is well-known that the principal advantage of Hydrogen

[1] Peng-Robinson was also modeling Neon and helium as if they were monatomic ideal gasses, i.e., a ratio of five-thirds. However, Neon and helium do behave as monatomic ideal gasses at essentially any conditions, so the P-R EOS could be used for these two cases.

refrigerant is its superior energy efficiency [44]. This kind of discrepancy was unacceptable because the design group planned to perform a detailed techno-economic comparison of the three working fluids. The pros and cons of each working fluid had to be modeled correctly.

For these reasons, we chose to model thermodynamic properties using the REFPROP model in ASPEN.

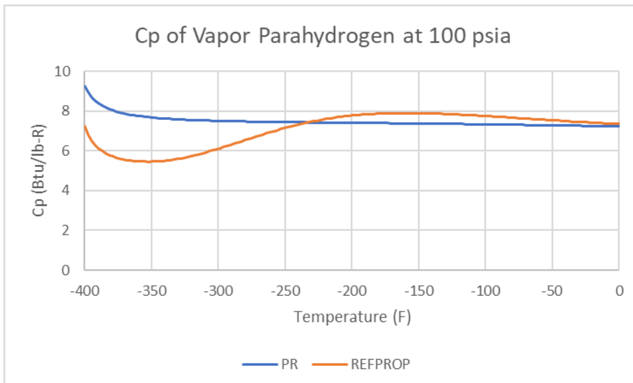

Figure 9.1: significant differences between P-R and REFPROP thermodynamic properties, such as heat capacity of p-H_2, particularly in cryogenic ranges

REFPROP does not calculate any thermodynamic properties using an equation of state, rather it references a databank of experimental properties measured by NIST (National Institute of Standards and Technology). REFROP's values for the C_p to C_v ratio of cryogenic Hydrogen are known to be accurate.

9.3.1 Limitations of REFPROP Databank

Unfortunately, REFPROP data is limited to pure components because it does not estimate properties with an equation of state informed by certain mixing parameters. For this reason, the design group was no longer able to model mixed refrigerant systems such as Neon-helium, Neon-Hydrogen, and Hydrogen-helium reliably via ASPEN.

In addition, selecting REFPROP raised an issue concerning the ortho and para spin isomers of Hydrogen. In ASPEN there is not a material model for o-H_2 (o-H_2), only for p-H_2 (p-H_2) and normal Hydrogen (n-H_2). n-H_2 is Hydrogen at the equilibrium composition of its spin isomers at room temperature: 75% para and 25% ortho (figure 9.2, table 9.3). When modeling the conversion of o-H_2 to p-H_2 in heat exchanger 26, the outlet Hydrogen stream in ASPEN is a mixture of p-H_2 and n-H_2, performing a mass balance so that this corresponds to the actual compositions of o-H_2 and p-H_2. Also, due to the presence of a recycle Hydrogen stream that is nearly pure para-Hydrogen, there is always some combination of o-H_2 and p-H_2 in the combined feed-recycle Hydrogen streams. As a consequence, in ASPEN there is always a mixture of p-H_2 and n-H_2 in the feed-recycle Hydrogen streams, and since REFPROP is the property method, the flowsheet is unable to compute thermophysical quantities for these streams. A workaround is described in the following section.

9.4 Chosen Model for Hydrogen Liquefaction

Since REFPROP cannot model mixtures of ortho-Hydrogen and para-Hydrogen, a combination of tear streams and calculator blocks in ASPEN Plus was used to creatively model the thermodynamic properties of p-H2 and o-H2 mixtures to a sufficiently high accuracy, specified below:

9.4.1 Tear Streams to Ensure Correct Thermodynamics, C_p/C_v, for H_2 vapor

First, it was assumed that the streams in the Hydrogen loop could be reasonably modeled as pure n-H2 if they came before the first ortho-para converter in heat exchanger 26. The reasoning for this was that the nearly pure p-H2 recycle stream was seven times smaller than the pure n-H2 feed, so the actual mole fraction of p-H2 is 34%, which is not too different from 25%. Hydrogen streams after the first ortho-para converter in heat exchanger 26 were modeled as 100% p-H2. This assumption was less reasonable than the previous one because in reality (table 9.2) the stream entering heat exchanger 30 has a mole fraction of 36% o-H_2, not 0%. After heat exchanger 30, however, the approximation is more exact because the actual mole fraction is 5%.

	Feed Hydrogen	***Before HX 26***	***After HX 26***	***After HX 30***	***After DFE***
o-H_2	75 %	75 %	0%	0%	0%
p-H_2	25 %	25 %	100%	100%	100%

Table 9.1: the mole fractions of ortho and para-Hydrogen present in the ASPEN model, because REFPROP cannot model mixtures, only pure components

	Feed Hydrogen	***Before HX 26***	***After HX 26***	***After HX 30***	***After DFE***

o-H_2	75 %	66 %	36%	5%	1%
p-H_2	25 %	34 %	64%	95%	99%

Table 9.2: the actual compositions of o- and p-H_2 present in the liquefaction process. Ideally would be the same as those in the flowsheet if ASPEN had an equation of state method that could model both cryogenic Hydrogen, mixtures, and mixtures of cryogenic Hydrogen spin isomers accurately. As of now, REFPROP is the only method that can accurately model pure components of cryogenic Hydrogen in ASPEN PLUS

	Feed Hydrogen	***Before HX 26***	***After HX 26***	***After HX 30***	***After DFE***
o-H_2	75.0 %	53.7 %	24.3%	3.3%	1.0%
p-H_2	25.0 %	46.3 %	75.7%	96.7%	99.0%

Table 9.3: equilibrium compositions of o- and p-H_2 at the stream temperatures, according to figure 9.2 and the stream temperatures and pressures in section 11

9.4.2 Modeling Ortho-Para Conversion using Calculator Blocks

o-H_2 is gradually converted into p-H_2 in Blocks 26 and 30 of the process. Blocks 26 and 30 (MHEATX 26 and MHEATX 30) serve as "reactors" as well as heat exchangers. In addition to performing heat exchange, the passages which hot Hydrogen passes through are lined with ferric oxide catalyst. Ferric oxide catalyst facilitates the reaction from o-H_2 to p-H_2. As described in Section 7.2, product LH2 must be at 99.0 mol% p-H_2. This composition is achieved through stepwise conversions, as illustrated in Table 9.2. Conversions provided in Table 9.2 come from Patent US4765813 (discussed in Section 8). Alternatively, conversion through HX 26 and HX 30 can be approximated by modeling BAHX passages like a packed bed reactor (PBR). Passage dimension and various stream properties can be used to develop a set of constitutive equations to model ortho/para conversion through the BAHX. This idea is discussed in more detail in section 24.1.4.

As shown in Figure 9.3, the heat of reaction of o-H_2 to p-H_2 as a function of temperature is known. Therefore, it is possible to model the extent of reaction by evolving extra "heat" in Blocks 26 and 30, as o-H_2 reacts to form p-H_2. This is modeled in ASPEN PLUS as a "Heat Leak" in the MHEATX 26 and MHEATX 30 blocks. The specifics of this calculation are provided in Section 24.3.

Using heat leaks as a means of introducing the heat of reaction to the heat exchanger is a suboptimal design choice. Heat leaks are calculated by adding heat to the bulk of the exchanger in no particular isolated area, but evenly distributed on the surface. This is a more reflective means of accounting for the energy released during ortho-para conversion than heat injection before and after the exchanger using RSTOICH, RGIBBS, or HEATX, because the heat of reaction is distributed along the length of the heat exchanger. Still, the heat leak cannot account for the changing heat capacity as the ortho-para mixture composition changes (figure 9.6), so temperature curves are not perfectly representative of the phenomenon. After careful discussion with project mentors, this design of heat integration was agreed to be the most acceptable solution.

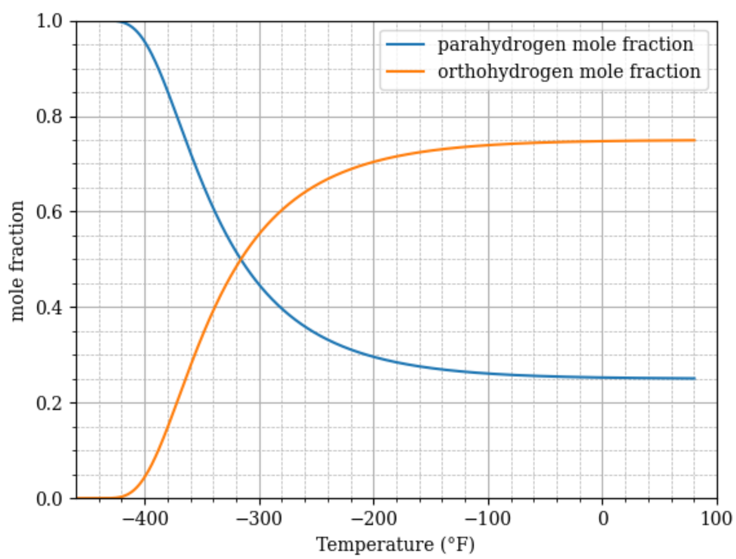

Figure 9.2: equilibrium mole fraction of Hydrogen spin isomers as a function of temperature. methods to calculate these values outlined in appendix 24.4

Based on predicted stream temperatures from Patent US4765813, Fortran code in ASPEN PLUS Calculator blocks was used to calculate the average heat of conversion of o-H2 to p-H2 as a function of temperature. This value was scaled by the expected change in o-H2, according to the last table. Iteration was then performed to fine-tune these guesses. The specifics of these calculations are provided in the appendix 24.2.

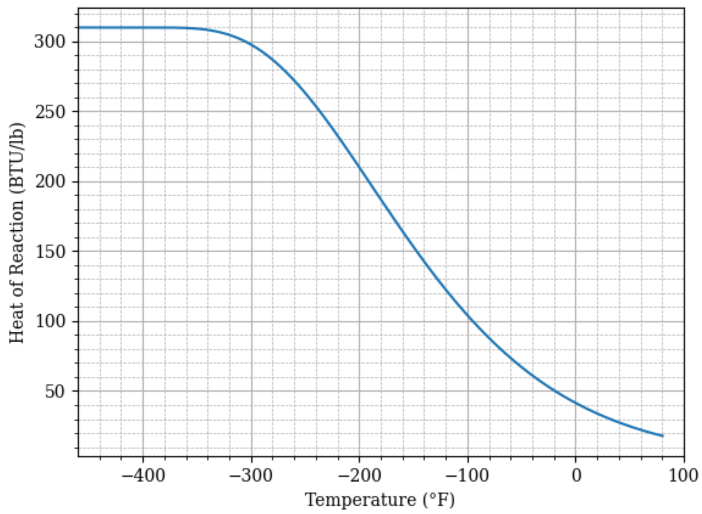

Figure 9.3: heat of reaction of o-H_2 to p-H_2 reaction as a function of temperature. methods to calculate these values outlined in appendix 24.4

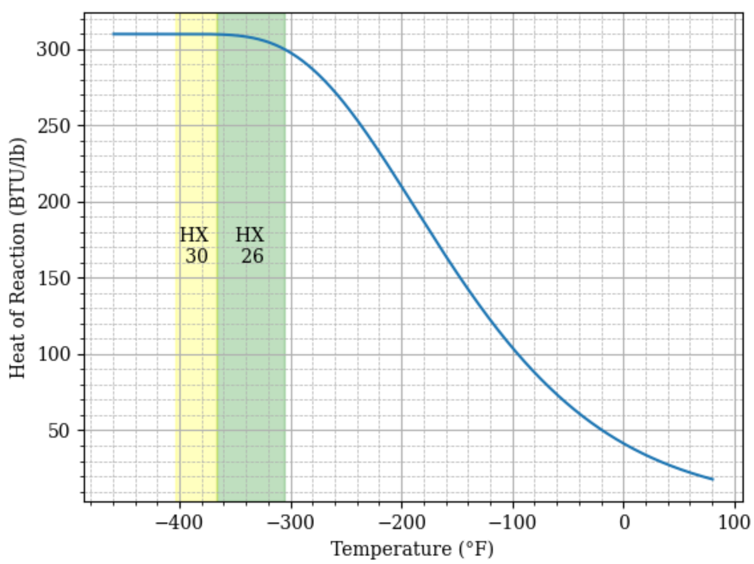

Figure 9.4: temperature ranges of heat exchangers 26 and 30 overlaid on figure 9.3. heat of reaction is approximately constant through heat exchanger 30, and varies slightly through heat exchanger 26

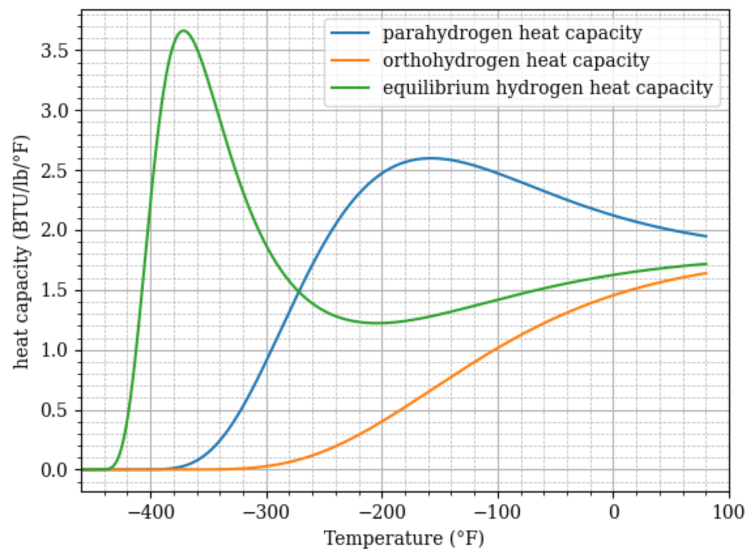

Figure 9.5: rotational heat capacities of pure para-Hydrogen, pure ortho-Hydrogen, and equilibrium Hydrogen, where the equilibrium composition of spin isomers is given in figure 9.2

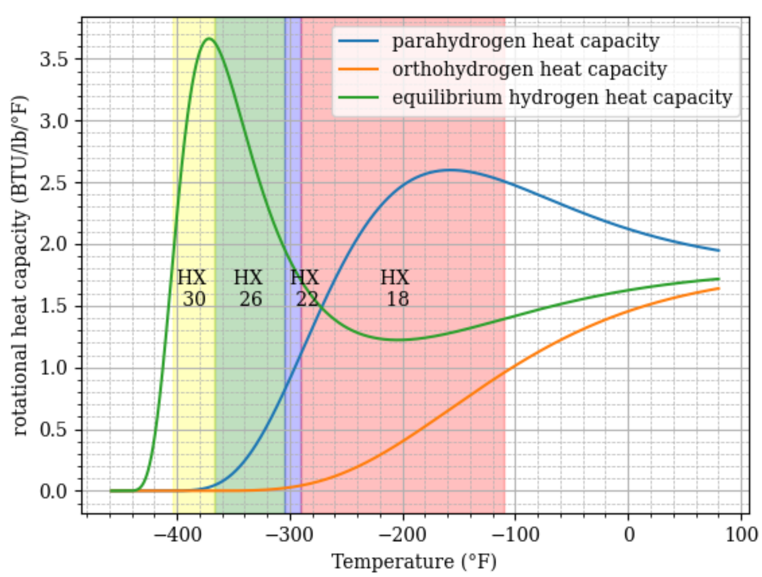

Figure 9.6: temperature ranges of heat exchangers overlaid over figure 9.5. Illustrates the sensitivities of thermophysical properties to the composition of spin isomers in the process.

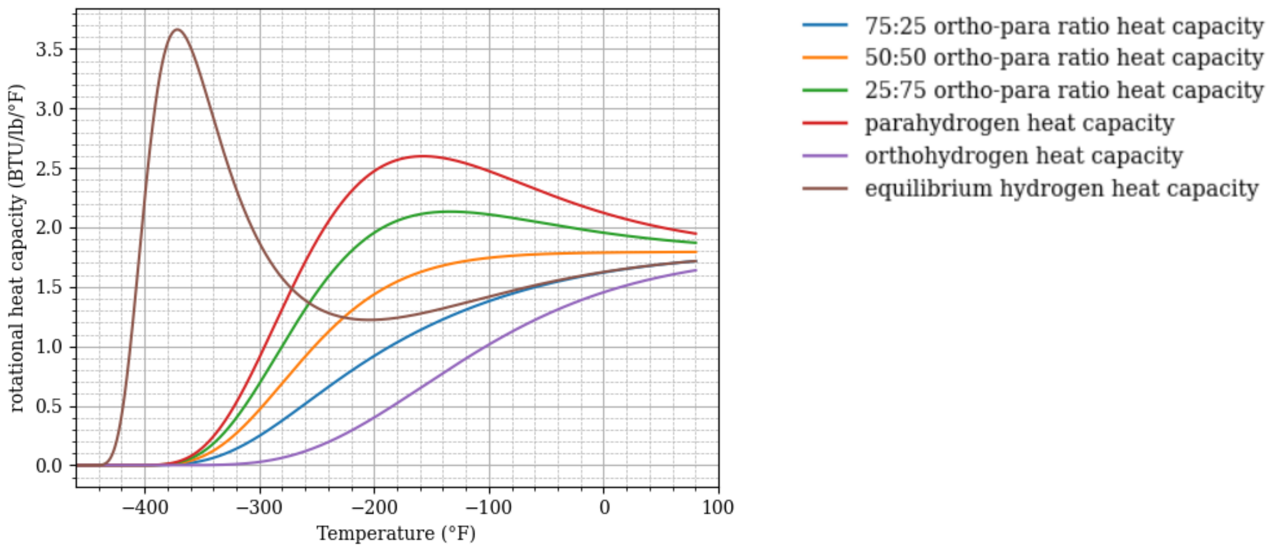

Figure 9.7: rotational heat capacities for varying mixtures of ortho and para spin isomers

9.4.3 Simulating Normal Hydrogen and Para-Hydrogen in Separate Loops

There are many ways to use tear streams and calculator blocks to simulate a Hydrogen liquefaction process in ASPEN PLUS (one method is demonstrated in Sections 9.4.1 and 9.4.2). An alternative model to the chosen process flow model in Sections 9.4.1 and 9.4.2 is developed below.

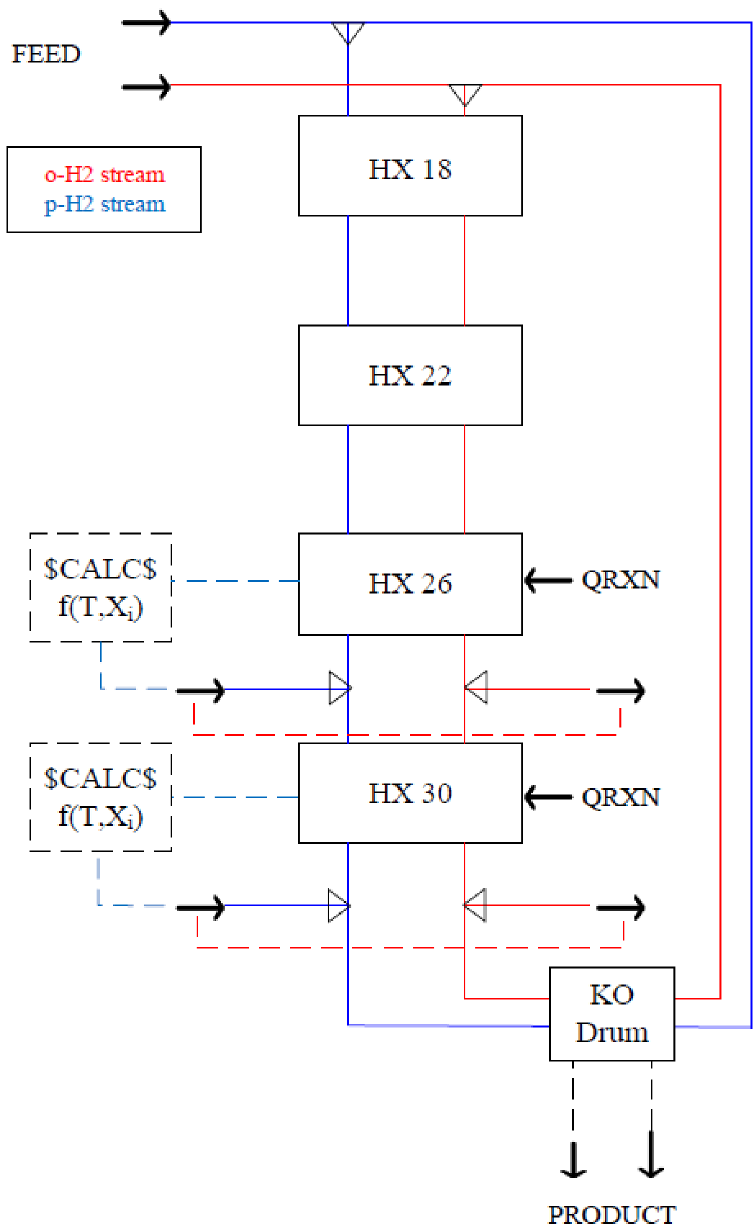

Figure 9.7: Alternative computational configuration of Hydrogen Liquefaction plant, by which exact compositions of o-H_2 and p-H_2 are detailed at every point along the process.

In this model, normal Hydrogen and para-Hydrogen move in separate loops. The advantage to this model compared to the model proposed in Sections 9.4.1 and 9.4.2 is that the compositions of the tear streams would exactly match actual hot Hydrogen compositions (Table 9.2) instead of inaccurately modeled hot Hydrogen compositions (Table 9.1). However, this model would likely complicate heat exchanger sizing and design. This model was ultimately rejected by project author Adam Brostow.

9.5 Design Limitations

It is important to note that the calculations using this ASPEN model are not perfect. A defining characteristic of this design project, which is either a curse or a luxury, depending on perspective, is the ability to create an extremely accurate design due to the relative simplicity of the process. All the involved materials are simple mono or diatomic molecules whose properties have been well documented. There are no complicated, unpredictable, and hard to model separation processes, and plate fin heat exchangers can be designed to extreme detail (section 12.2.1, appendix 24.1). While this gives the project a unique rigor and promising analyses of energy and economic efficiencies, it also requires careful modeling of every aspect of the process.

For example, the final pressures and temperatures of all the streams in the model will differ from those calculated using an equation of state property method that can model the materials used in this process, particularly mixtures. The heat capacity of a mixture of o-H_2 and p-H_2 is different from the heat capacity of pure p-H2, especially at cryogenic temperatures (figure 9.6), and so using tear streams that are pure para-Hydrogen means that the heat exchanger composite curves computed by ASPEN are inaccurate to the extent that the heat capacity is inaccurate. Nonetheless, the values the flowsheet model generates yield accurate enough information to size and design the process equipment.

10 ASSEMBLY OF DATABASE

10.1 Raw materials and products

Raw Material	***Unit Cost (USD/1000 gal)***
Process water	$0.80

Table 10.1.1: Process feedstock components and their unit prices

Green Hydrogen liquefaction plants obtain feedstock Hydrogen through electrolysis of water. Thus, process water is the only raw material required.

Refrigerant	***Unit Cost (USD/lb)***
Neon	$1600
Helium	$23
Hydrogen	~$5
Nitrogen	$0.023

Table 10.1.2: Process refrigerant candidates and their unit prices (section 7.3 for more detail)

Hydrogen liquefaction process design utilizes refrigeration. Table 10.1.2 contains the refrigerants considered for all discussed process designs and their unit prices. To account for supply chain shocks due to Russia's invasion of Ukraine (section 7.3), the pre-Russian invasion price of Neon is adjusted by an order of magnitude ($160/lb to $1600/lb). The price of *green* vapor Hydrogen is calculated by subtracting the cost of liquefaction from the breakeven sales price of liquid Hydrogen.

Additionally, data was gathered from NIST's Material Database on material qualities ranging from Aluminum strength and deformation properties to heat conductivity metrics. [46]

10.2 Safety and Toxicity of Major Chemicals

For the selected process configuration, Neon and Hydrogen are the only two chemical species present in the process. Process water and oxygen are found upstream of the process. MSDS sheets for all four chemical species present can be found in Sections 24.12.1 to 24.12.4. Hydrogen is also an explosion hazard, and appropriate precautions are discussed in Section 19.7. Most importantly, Hydrogen liquefaction is performed at cryogenic temperatures. Exposure to process streams or unit operations below 32 F may result in frostbite.

10.3 Yield Data

Catalysts are used in Hydrogen liquefaction processes to convert ortho-Hydrogen to Hydrogen's more stable para-Hydrogen spin isomer. Catalyst selection for the selected process configuration was based on catalyst cost, yield, and industry standards. The two candidates considered are listed in table 10.3.

Polyukhov et al. (*Efficient MOF-Catalyzed Ortho-Para Hydrogen Conversion for Practical Liquefaction and Energy Storage)* claim that Ni-MOF-74 catalyst is estimated to have an effective rate constant that is two orders of magnitude greater than that of ferric oxide. However, Ni-MOF-74 is new and untested at an industrial scale. Upon recommendation from project author Adam Brostow, ferric oxide catalyst (IONEX) was selected. Ferric oxide catalyst is the current industry standard for ortho para conversion in Hydrogen liquefaction processes.

Catalyst	***Unit Cost (USD/lb)***
IONEX (Fe2O3)	$9
Ni-MOF-74	$2960

Table 10.3: Catalyst candidates for ortho-para conversion in a Hydrogen liquefaction process. Unit prices are taken from Section 24.8.3 and Das et al. (*An Efficient Synthesis Strategy for Metal-Organic Frameworks: Dry-Gel Synthesis of MOF-74 Framework with High Yield and Improved Performance)* [47]

11 PFD AND MATERIAL BALANCE

11.1 Upstream Electrolyzers

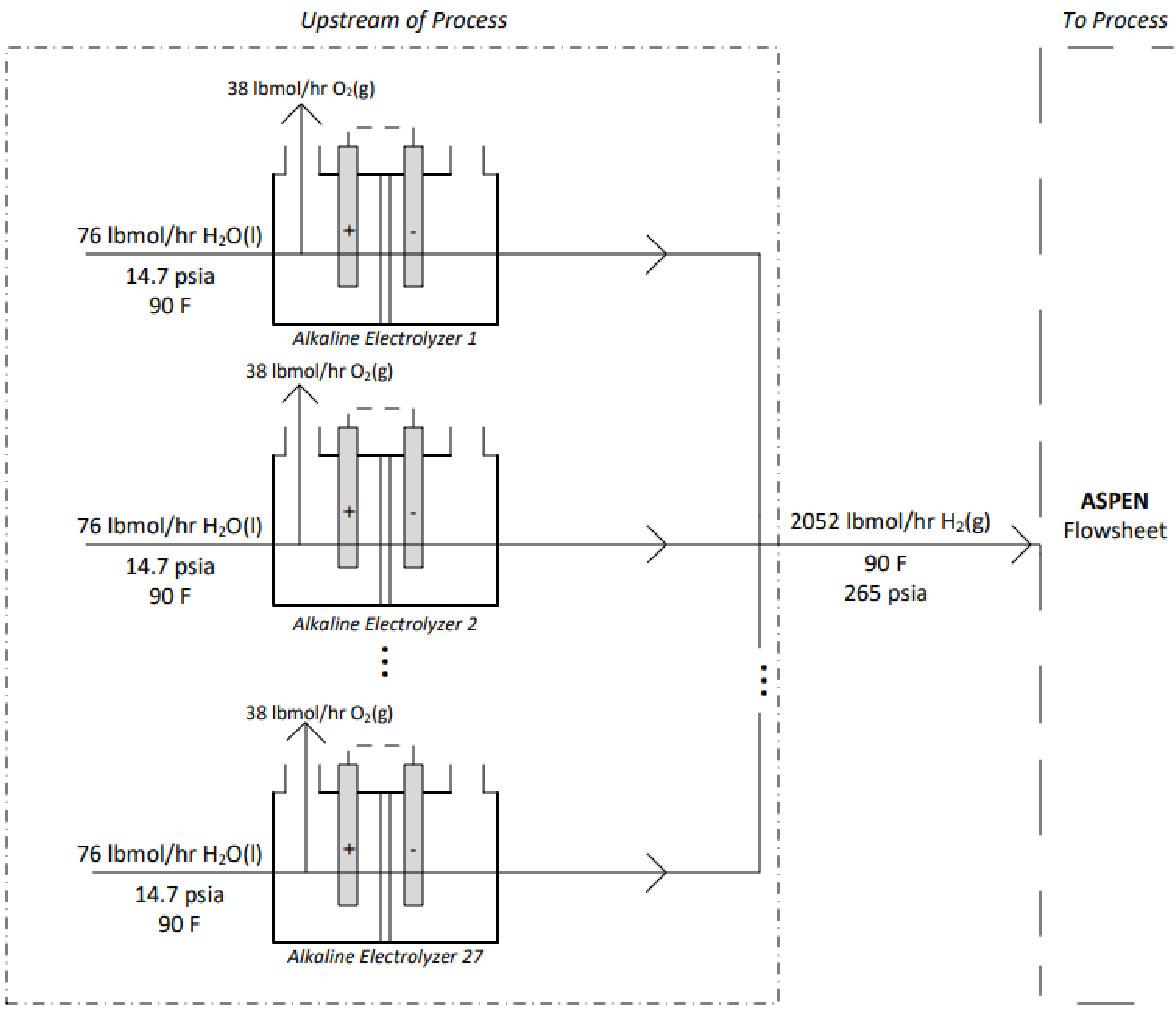

Figure 11.1: water electrolysis takes place upstream from the liquefaction plant

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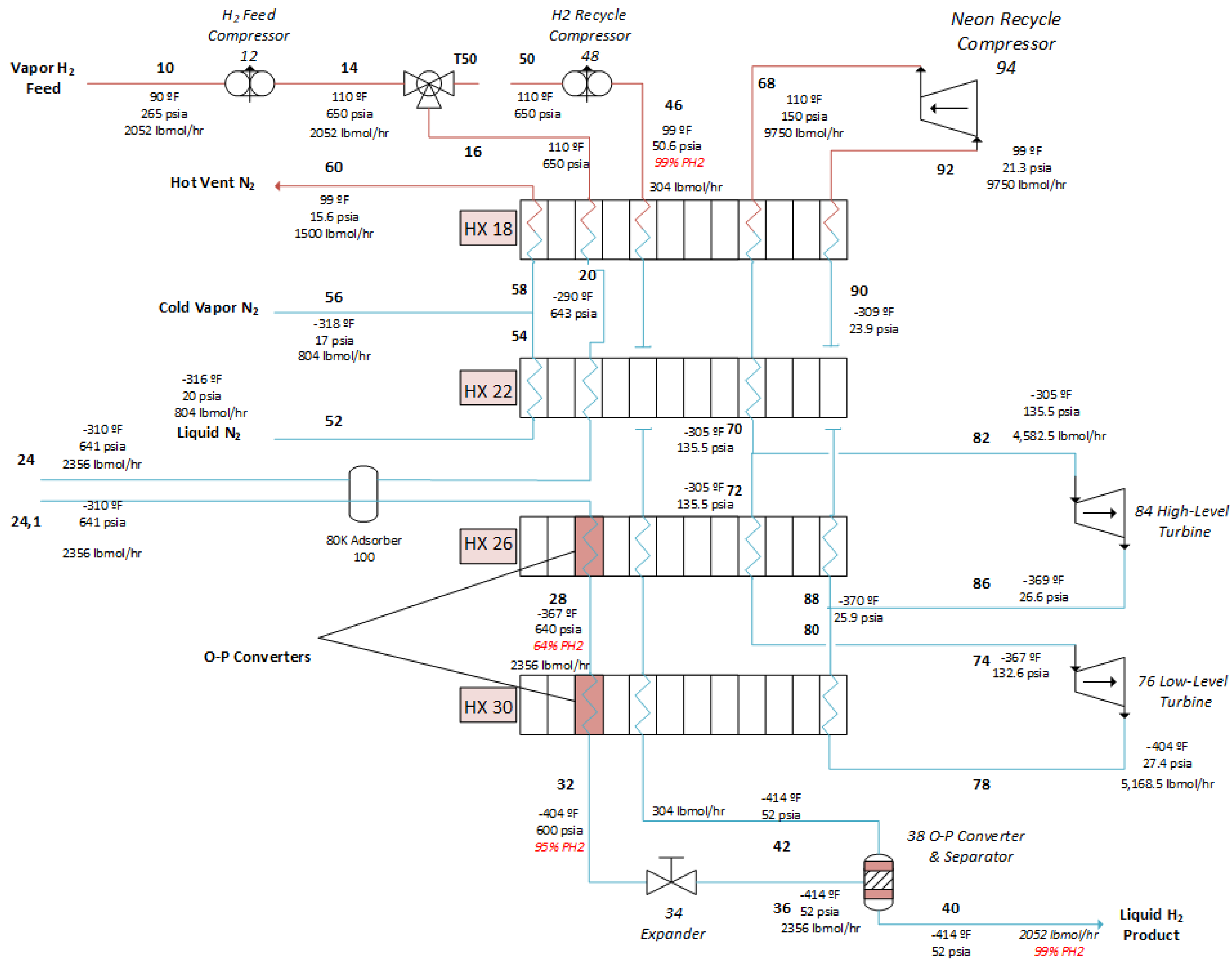
11.2 Process Flow Diagram



11.3 Mass Balances

Stream	***Pressure***	***Temperature***	***Flowrate (lbmol/hr)***	***Vapor***	***Mol Frac***	***Mol Frac.***	***Mol Frac.***	***Mol Frac.***	***Mol Frac.***
10	265	90	2052	1	1	0	0	0	0
14	650	110	2052	1	1	0	0	0	0
16	650	110	2356	1	1 (0)	0 (0.66)	0 (0.34)	0	0
20	643	-290	2356	1	1 (0)	0 (0.66)	0 (0.34)	0	0
24	-640.7	-305	2356	1	1 (0)	0 (0.66)	0 (0.34)	0	0
24,1	*-640.7*	*-305*	*2356*	*1*	*0* (0)	*0 (0.66)*	*1 (0.34)*	*0*	*0*
28	640.1	-367	2356	1	0 (0)	0 (0.36)	1 (0.64)	0	0
32	640.0	-404	2356	1	0 (0)	0 (0.05)	1 (0.95)	0	0
36	52	-414	2356	0.13	0 (0)	0 (0.05)	1 (0.95)	0	0
40	52	-414	2052	0	0 (0)	0 (0.01)	1 (0.99)	0	0
42	52	-414	304	1	0 (0)	0 (0.01)	1 (0.99)	0	0
46	50.6	99	304	1	0 (0)	*0* (0.01)	1 (0.99)	0	0
50	650	110	304	1	0 (0)	*0* (0.01)	1 (0.99)	0	0
T50	*650*	*110*	*304*	*1*	*1* (0)	*0* (0.01)	0 (0.99)	*0*	*0*
52	20	-316	696	0	0	0	0	0	1
56	19	-316	804	1	0	0	0	0	1
58	17	-318	1500	0.81	0	0	0	0	1
60	15.6	99	1500	1	0	0	0	0	1
68	150	110	9750	1	0	0	0	1	0
70	135.5	-305	9750	1	0	0	0	1	0
72	135.5	-305	5168.5	1	0	0	0	1	0
74	132.6	-367	5168.5	1	0	0	0	1	0
78	27.4	-404	5168.5	1	0	0	0	1	0
80	25.9	-370	5168.5	1	0	0	0	1	0
82	135.5	-305	4582.5	1	0	0	0	1	0
86	26.6	-369	4582.5	1	0	0	0	1	0
88	25.9	-370	9750	1	0	0	0	1	0
90	23.9	-309	9750	1	0	0	0	1	0
92	21.3	99	9750	1	0	0	0	1	0

Table 11.1: Pressures, temperatures, flowrates and compositions of all streams. Values in red represent the inputted compositions of the respective streams in ASPEN. Values in parentheses represent the actual compositions of the respective streams and are used for subsequent modeling and calculations.

11.3.1 Material Volume, Pipe Sizing, & Neon Purchase Amount

The decision about which refrigerant to use is made for economic reasons, so it is necessary to know how much refrigerant is in the working fluid loop of the refrigeration cycle. Because the working fluid loop has no inlet or outlet streams, the mass contained in it is constant and will be determined by the total volume of the pipes and heat exchanger cores that comprise the physical loop, and the density of the working fluid in the loop. For this reason, the BAPFHX sizing informs the total one-time purchase cost of the Neon refrigerant. The calculation of BAPFHX core volume, pipe sizes, and amount of Neon in the process is found in appendix 24.1.3.

Additionally, due to safety considerations concerning the risk of Hydrogen fueled explosions (section 19.2), this analysis is also applied to the Hydrogen streams to calculate how much Hydrogen is in the plant at a given time.

The results are tabulated below:

BAPFHX	18	22	26	30	Total
Neon Core Volume (ft^3)	30	2	20	1	63
Hydrogen Core Volume (ft^3)	0.4	0.1	0.2	0.2	0.9

Table 11.2: core volumes of Neon and Hydrogen passages of the heat exchangers

	Pipe Diameter (ft)	Total Length of Pipes (ft)	Average Density (lb/ft^3)	Total material amount inside process (lb)
Neon	1	200	1.0	220
Hydrogen	0.2	225	0.75	7

Table 11.3: pipe sizing and total material amounts in the process, for Neon and Hydrogen

12 PROCESS DESCRIPTIONS

12.1 High-Level Process Overview

Upstream of the process, 1026 lbmol/hr water is electrolyzed across 27 parallel alkaline electrolyzers. Hydrogen gas is evolved at the cathode and oxygen gas is evolved at the anode of each electrolyzer. Each gas is sent to a separate stream where oxygen gas is collected and sold as is, and Hydrogen gas is compressed by an oil-injected screw compressor operating at 226.25 hp, 90-110 ºF, and 85% isentropic efficiency (Block 12). The compressed Hydrogen is then mixed with a recycle Hydrogen stream and enters a refrigeration loop at a starting temperature of 110 ºF and starting pressure of 650 psia.

The loop consists of two non-catalytic (Blocks 18 & 22) and two catalytic (Blocks 26 & 30) heat exchangers in series. The hot Hydrogen stream is cooled in heat exchangers 18 and 22 to a temperature of -310 ºF and pressure of 625 psia. In heat exchangers 26 and 30, this Hydrogen is further cooled, but also reacted to preferentially form the p-H2 spin isomer. The resulting hot Hydrogen outlet is at -404 F, 600 psia, and is fully liquid. Lastly, power is recovered from the high energy supercritical p-H2 fluid to reach -414 F, 52 psia, at which the liquid p-H2 is packaged and sold, and the vapor p-H2 is sent back through a recycle to deliver additional cooling.

In this chosen process, Neon is selected as the refrigerant. A Reverse Brayton refrigeration cycle is used to deliver cooling to the hot Hydrogen. 9750 lbmol Neon/hr moves through the refrigeration loop. Seven production days' worth of product LH2 is stored in a spherical containment unit which then dispenses product to mobile containment units such as trailers on a continuous basis. A diagram of this process is shown in section 11.2.

12.2 Process Optimization

12.2.1 Minimum Approach Temperature

Hydrogen liquefaction plants can be optimized for energy efficiency or profitability. Theoretical plant designs usually optimize energy efficiency, while actualized plant designs (plants that have been built and are operating today) usually optimize profitability. Consequently, actualized plants tend to have lower capital costs but higher operating costs and lower efficiencies (See Section 20.4 for more information). In order to decide whether to optimize for energy efficiency or profitability, a sensitivity analysis was performed on one fundamental parameter in the refrigerant loop: heat exchanger minimum approach temperature, ΔT_{min}. In general, for cryogenic processes, the energy efficiency increases as approach temperatures decrease (tighten). This comes at the expense of an increased capital cost, because heat exchanger sizes increase as approach temperatures tighten. The tradeoff between these two effects will be different for every process design. To determine the optimal approach temperature with the ideal tradeoff, a sophisticated optimization algorithm must be performed that maximizes the return on investment (ROI) of the plant.

The ΔT_{min} for heat exchangers 26 and 30 determine the energy efficiency of the process. The smaller (tighter) that the approach temperature for heat exchanger 30 is, the warmer that the cold Neon stream going into heat exchanger 30, stream 78, can be. This means that the outlet pressure of turboexpander 76 can be higher and as a result, the Neon recycle compressor, 94, has a smaller power requirement when the Neon stream reaches it to be compressed.

Likewise, the tighter that the approach temperature for heat exchanger 26 is, the warmer that the cold Neon stream going into heat exchanger 26, a combination of stream 86 and 80, can be. This means that the outlet pressure of turboexpander 84 can be higher and as a result, the Neon recycle compressor, 94, has a smaller power requirement when the Neon stream reaches it to be compressed.

In the process design, there are two parameters that affect the minimum approach temperatures of the heat exchangers, the flow rate of the Neon refrigerant, and the outlet pressure of the low level turboexpander,

76. The outlet pressure of the high level turboexpander, 84, also affects the minimum approach temperatures, but is constrained by the outlet pressure of 76 because the two outlet streams of each turboexpander combine in stream 88. Raising the flow rate of the Neon refrigerant or the outlet pressures of the turboexpanders tightens the approach temperatures, while lowering either does the opposite.

These two parameters are responsible for deciding the bulk of in-process costs (Heat Exchangers, Compressors, Turboexpanders). They determine the minimum approach temperature on all heat exchangers, thus informing heat exchanger sizing, and the power requirement of the Neon recycle compressor - the largest, most expensive unit operation in the process. Since approach temperatures cannot be changed directly, these two parameters are varied in order to vary the approach temperatures and observe the effect of approach temperature on energy efficiency and ROI, indirectly. Tables 12.2.1 through 12.2.6 outline the results of this sensitivity analysis.

	Molar Flowrate of Neon Refrigerant (lbmol/hr)								
Low Level Turboexpander, Power Recovered (psia)	---	9000	9250	9500	9750	10237.5	10725	11212.5	11700
	32	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	29	N/A	N/A	N/A	16.63%	16.51%	16.26%	16.19%	15.69%
	27.4	N/A	16.57%	16.67%	16.59%	16.34%	16.06%	15.75%	15.45%
	26	16.65%	16.68%	16.58%	16.45%	16.16%	15.85%	15.54%	15.22%
	23	16.50%	16.35%	16.22%	16.02%	15.68%	15.35%	15.01%	14.67%
	20	16.04%	15.85%	15.66%	15.47%	15.09%	14.72%	14.35%	13.98%
	17	15.37%	15.16%	14.95%	14.74%	14.33%	13.93%	13.53%	13.14%
	14	14.47%	14.23%	14.00%	13.77%	13.33%	12.89%	12.46%	12.04%

Table 12.2.1: Return on Investment (ROI) in the third year at various refrigerant flowrates and expander power specifications. ROI increases as the approach temperatures tighten (see tables 12.2.3 through 12.2.6). The yellow cell marks the chosen process configuration of 45 MTPD production with Neon refrigerant. The red cells mark flowsheets that did not converge because of temperature crossovers in the MHEATX blocks of the ASPEN PLUS flowsheet. The orange cells mark flowsheets that either had the same refrigerant flowrate or the same low-level turboexpander specification as the selected process configuration.

	Molar Flowrate of Neon Refrigerant (lbmol/hr)								
Low Level Turboexpander, Power Recovered (psia)	---	9000	9250	9500	9750	10237.5	10725	11212.5	11700
	32	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	29	N/A	N/A	N/A	46.10%	44.20%	42.40%	42.60%	39.30%
	27.4	N/A	46.40%	45.40%	44.36%	42.90%	41.20%	39.60%	38.20%
	26	45.90%	44.90%	43.80%	42.90%	41.10%	39.40%	37.90%	36.50%
	23	42.42%	41.42%	40.41%	39.56%	37.89%	36.37%	34.96%	33.65%
	20	38.78%	37.86%	36.97%	36.13%	34.60%	33.19%	31.90%	30.70%
	17	34.92%	34.08%	33.28%	32.52%	31.13%	29.85%	28.67%	27.58%
	14	30.71%	29.96%	29.25%	28.57%	27.33%	26.20%	25.20%	24.20%

Table 12.2.2: Liquefaction plant Energy Efficiencies at various refrigerant flowrates and expander power specifications. Efficiency increases as approach temperatures tighten. The yellow cell marks the chosen process configuration of 45 MTPD production with Neon refrigerant. The red cells mark flowsheets that did not converge because of temperature crossovers in the MHEATX blocks of the ASPEN PLUS flowsheet. The orange cells mark flowsheets that either had the same refrigerant flowrate or the same low-level turboexpander specification as the selected process configuration.

	Molar Flowrate of Neon Refrigerant (lbmol/hr)								
Low Level Turboexpander, Power Recovered (psia)	---	9000	9250	9500	9750	10237.5	10725	11212.5	11700
	32	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	29	N/A	N/A	N/A	10.3 ºF	11.6 ºF	12.9 ºF	14.0 ºF	15.1 ºF
	27.4	N/A	9.8 ºF	10.5 ºF	11.3 ºF	12.6 ºF	13.9 ºF	15.0 ºF	16.1 ºF
	26	9.9 ºF	10.7 ºF	11.4 ºF	12.2 ºF	13.5 ºF	14.8 ºF	15.9 ºF	17.0 ºF
	23	11.8 ºF	12.6 ºF	13.4 ºF	14.2 ºF	15.5 ºF	16.8 ºF	18.0 ºF	19.1 ºF
	20	14.0 ºF	14.8 ºF	15.6 ºF	16.4 ºF	17.8 ºF	19.1 ºF	20.3 ºF	21.4 ºF
	17	16.4 ºF	17.2 ºF	18.1 ºF	18.8 ºF	20.2 ºF	21.5 ºF	22.8 ºF	23.9 ºF
	14	19.1 ºF	20.0 ºF	20.8 ºF	21.6 ºF	23.0 ºF	24.4 ºF	25.6 ºF	26.8 ºF

Table 12.2.3: Minimum Approach Temperature of Heat Exchanger 1 (Block 18) at various refrigerant flowrates and expander power specifications. The yellow cell marks the chosen process configuration of 45 MTPD production with Neon refrigerant. The red cells mark flowsheets that did not converge because of temperature crossovers in the MHEATX blocks of the ASPEN PLUS flowsheet. The orange cells mark flowsheets that either had the same refrigerant flowrate or the same low-level turboexpander specification as the selected process configuration.

Low Level Turboexpander, Power Recovered (psia)	Molar Flowrate of Neon Refrigerant (lbmol/hr)								
	---	9000	9250	9500	9750	10237.5	10725	11212.5	11700
	32	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	29	N/A	N/A	N/A	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF
	27.4	N/A	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF
	26	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF
	23	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF
	20	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF
	17	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF
	14	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF	10.6 ºF

Table 12.2.4: Minimum Approach Temperature of Heat Exchanger 2 (Block 22) at various refrigerant flowrates and expander power specifications. The yellow cell marks the chosen process configuration of 45 MTPD production with Neon refrigerant. The red cells mark flowsheets that did not converge because of temperature crossovers in the MHEATX blocks of the ASPEN PLUS flowsheet. The orange cells mark flowsheets that either had the same refrigerant flowrate or the same low-level turboexpander specification as the selected process configuration.

Low Level Turboexpander, Power Recovered (psia)	Molar Flowrate of Neon Refrigerant (lbmol/hr)								
	---	9000	9250	9500	9750	10237.5	10725	11212.5	11700
	32	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	29	N/A	N/A	N/A	0.6 ºF	1.8 ºF	2.9 ºF	3.8 ºF	4.6 ºF
	27.4	N/A	0.3 ºF	1.1 ºF	1.8 ºF	3.1 ºF	4.1 ºF	5.0 ºF	5.8 ºF
	26	0.5 ºF	1.4 ºF	2.2 ºF	3.0 ºF	4.2 ºF	5.3 ºF	6.2 ºF	7.0 ºF
	23	3.1 ºF	4.0 ºF	4.8 ºF	5.5 ºF	6.8 ºF	7.9 ºF	8.8 ºF	9.6 ºF
	20	5.8 ºF	6.8 ºF	7.6 ºF	8.3 ºF	9.6 ºF	10.7 ºF	11.6 ºF	12.4 ºF
	17	8.9 ºF	9.8 ºF	10.7 ºF	11.4 ºF	12.7 ºF	13.8 ºF	14.8 ºF	15.6 ºF
	14	12.4 ºF	13.4 ºF	14.2 ºF	15.0 ºF	16.3 ºF	17.4 ºF	18.3 ºF	19.1 ºF

Table 12.2.5: Minimum Approach Temperature of Heat Exchanger 3 (Block 26) at various refrigerant flowrates and expander power specifications. The yellow cell marks the chosen process configuration of 45 MTPD production with Neon refrigerant. The red cells mark flowsheets that did not converge because of temperature crossovers in the MHEATX blocks of the ASPEN PLUS flowsheet. The orange cells mark flowsheets that either had the same refrigerant flowrate or the same low-level turboexpander specification as the selected process configuration.

	Molar Flowrate of Neon Refrigerant (lbmol/hr)								
Low Level Turboexpander, Power Recovered (psia)	---	9000	9250	9500	9750	10237.5	10725	11212.5	11700
	32	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	29	N/A	N/A	N/A	1.0 ºF	1.0 ºF	1.0 ºF	1.0 ºF	1.0 ºF
	27.4	N/A	2.0 ºF	2.0 ºF	2.0 ºF	2.0 ºF	2.0 ºF	2.0 ºF	2.0 ºF
	26	1.9 ºF	2.8 ºF	2.8 ºF	2.8 ºF	2.8 ºF	2.8 ºF	2.8 ºF	2.8 ºF
	23	3.9 ºF	4.7 ºF	4.7 ºF	4.7 ºF	4.7 ºF	4.7 ºF	4.7 ºF	4.7 ºF
	20	5.5 ºF	5.5 ºF	5.5 ºF	5.5 ºF	5.5 ºF	5.5 ºF	5.5 ºF	5.5 ºF
	17	6.4 ºF	6.4 ºF	6.4 ºF	6.4 ºF	6.4 ºF	6.4 ºF	6.4 ºF	6.4 ºF
	14	7.5 ºF	7.5 ºF	7.5 ºF	7.5 ºF	7.5 ºF	7.5 ºF	7.5 ºF	7.5 ºF

Table 12.2.6: Minimum Approach Temperature of Heat Exchanger 4 (Block 30) at various refrigerant flowrates and expander power specifications. The yellow cell marks the chosen process configuration of 45 MTPD production with Neon refrigerant. The red cells mark flowsheets that did not converge because of temperature crossovers in the MHEATX blocks of the ASPEN PLUS flowsheet. The orange cells mark flowsheets that either had the same refrigerant flowrate or the same low-level turboexpander specification as the selected process configuration.

Something interesting and perhaps unexpected occurs here. As the molar flowrate of Neon refrigerant decreases and the power specification for the low level turboexpander increases, *both* energy efficiency (table 12.2.2) and return on investment in the third year (ROI) (table 12.2.1) increase. In other words, the optimal configuration has the best energy efficiency and the best ROI. In effect, for this process design, optimizing energy efficiency results in the same design as optimizing for ROI. This is a result specific to this process, and is likely due to the relatively large Neon recycle compressor that provides almost all of the needed energy for the process to run. The design group hypothesized that this result may be limited to Reverse-Brayton Hydrogen liquefaction processes with large production capacities, that have such a centralized power requirement.

12.2.2 Heat Exchanger Design Informs the Entire Process Design

The BAPFHX design (appendix 24.1) is dependent on certain parameters obtained from the ASPEN model. Namely, in order to design the heat exchangers of the process, one needs the composite curves of the MHEATX blocks in ASPEN, and the properties of the streams flowing in and out of the MHEATX blocks. That is, flow rates, temperatures, pressures, phases, and compositions. The pressure drops through the MHEATX blocks are specified as user inputs prior to running the flowsheet.

Along with these parameters, the heat exchangers are then designed rigorously by varying their heat transfer fins (geometry, height, thickness, frequency, effective heat transfer area), the number of passages per material stream, and the heat exchanger width and length.

Once a design has been obtained in this way that satisfies the constraints in appendix 24.1.2, the total design process of the BAPFHX is not actually complete. This is because in the design of the heat exchangers, one creates a design with pressure drops through them that are not necessarily the same as those in the ASPEN model that was used to inform the BAPFHX design in the first place. Due to this discrepancy, the ASPEN model must be adjusted by changing the pressure drops through the MHEATX blocks from what they were at the beginning of the design to the new values determined by the BAPFHX design process.

Running the flowsheet with these new values results in a new ASPEN model with different composite curves and stream properties. This necessitates a redesign of the heat exchangers. This process is repeated until the pressure drops determined by the BAPFHX design are equal to those which were specified in the inputs to the ASPEN model that informed the BAPFHX design, as illustrated in figure 12.1. In practice, this process took three iterations before the designed pressure drops converged to the ASPEN user input pressure drops.

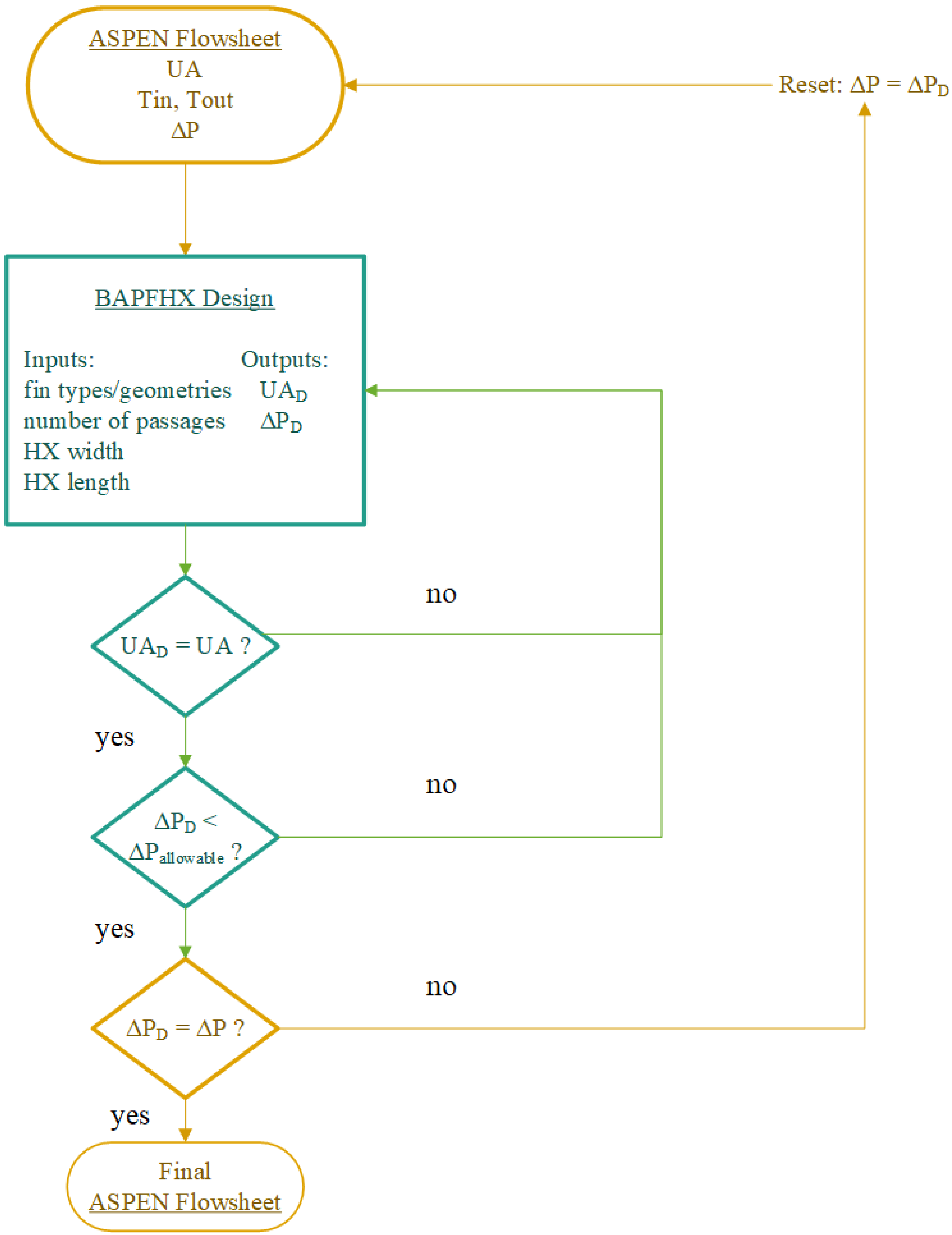


Figure 12.1: Algorithm for simultaneous ASPEN Flowsheet and BAPFHX design. Illustrates the intricate connection between BAPFHX design and the ASPEN model

13 ENERGY BALANCE AND UTILITY REQUIREMENTS

13.1 Upstream Energy Requirement

Upstream	*Annual Energy requirement (kW-hr/year)*
Electrolyzers (27 total)	731,109,019

Table 13.1: Upstream Annual Energy requirement

13.2 In-Process Energy Requirement

Process	*Annual Energy requirement (kW-hr/year)*
Hydrogen feed compressor	3,251,650
Hydrogen recycle compressor	7,043,575
Neon recycle compressor	71,636,760

Table 13.2: In-Process Annual Energy requirement

Process	*Annual Energy recovered (kW-hr/year)*
High-level turboexpander	3,286,139
Low-level turboexpander	1,985,248
Dense-fluid expander	208,720

Table 13.3: Power recovered from Turboexpanders in process

Our ASPEN PLUS flowsheet converges with no errors, therefore the selected process design is in mass and energy balance. Since a number of tear streams are used in the selected process model, overall energy balances are not shown. Instead, a nominal energy balance is demonstrated around "cryogenic" Blocks 26 and 30. Stream IDs in black text represent Stream IDs from ASPEN PLUS flowsheet. Stream IDs in red text represent streams on the PFD, in Section 11.2.

	Inlet enthalpy H_{in} (Btu/hr)	*Outlet enthalpy H_{out} (Btu/hr)*
S-72 or 72	-9889855	---
S-74 or 74	---	-11615384
S-(24,1) or 24,1	-6762200	---
S-(28,1) or 28		-7831605
S-(46,-2)	-962003	---
S-(46,+R)	---	-859221
S-88 or 88	-21710507	---
S-90 or 90	---	-18703821

Table 13.2.1: Stream enthalpies for Heat Exchanger 3 (Block 26)

$$H_{in,26} = H_{out,26}$$

$$\sum(Inlet\ enthalpies) = -9889855\frac{Btu}{hr} - 6762200\frac{Btu}{hr} - 962004\frac{Btu}{hr} - 21710507\frac{Btu}{hr}$$

$$= -3.93 * 10^7 Btu/hr$$

$$\sum(Outlet\ enthalpies) = --11615384\frac{Btu}{hr} - 7831605\frac{Btu}{hr} - 859221\frac{Btu}{hr} - 18703821\frac{Btu}{hr}$$

$$= -3.90 * 10^{7} Btu/hr$$

	Inlet enthalpy H_{in} (Btu/hr)	***Outlet enthalpy H_{out} (Btu/hr)***
S-(28,1) or 28	-7831606	---
S-32 or 32	---	-8662309
S-78 or 78	-12482416	---
S-80 or 80		-11534612
S-42 or 42	-1037327	---
S-(46,-2)	---	-962004

Table 13.2.1: Stream enthalpies for Heat Exchanger 4 (Block 30)

$$H_{in,30} = H_{out,30}$$

$$\sum(Inlet\ enthalpies) = -7831606\frac{Btu}{hr} - 12482416\frac{Btu}{hr} - 1037327\frac{Btu}{hr}$$

$$= -2.14 * 10^{7} Btu/hr$$

$$\sum(Outlet\ enthalpies) = -8662309\frac{Btu}{hr} - 11534612\frac{Btu}{hr} - 962004\frac{Btu}{hr}$$

$$= -2.12 * 10^{7} Btu/hr$$

14 EQUIPMENT LIST AND UNIT DESCRIPTIONS

14.1 Electrolyzer

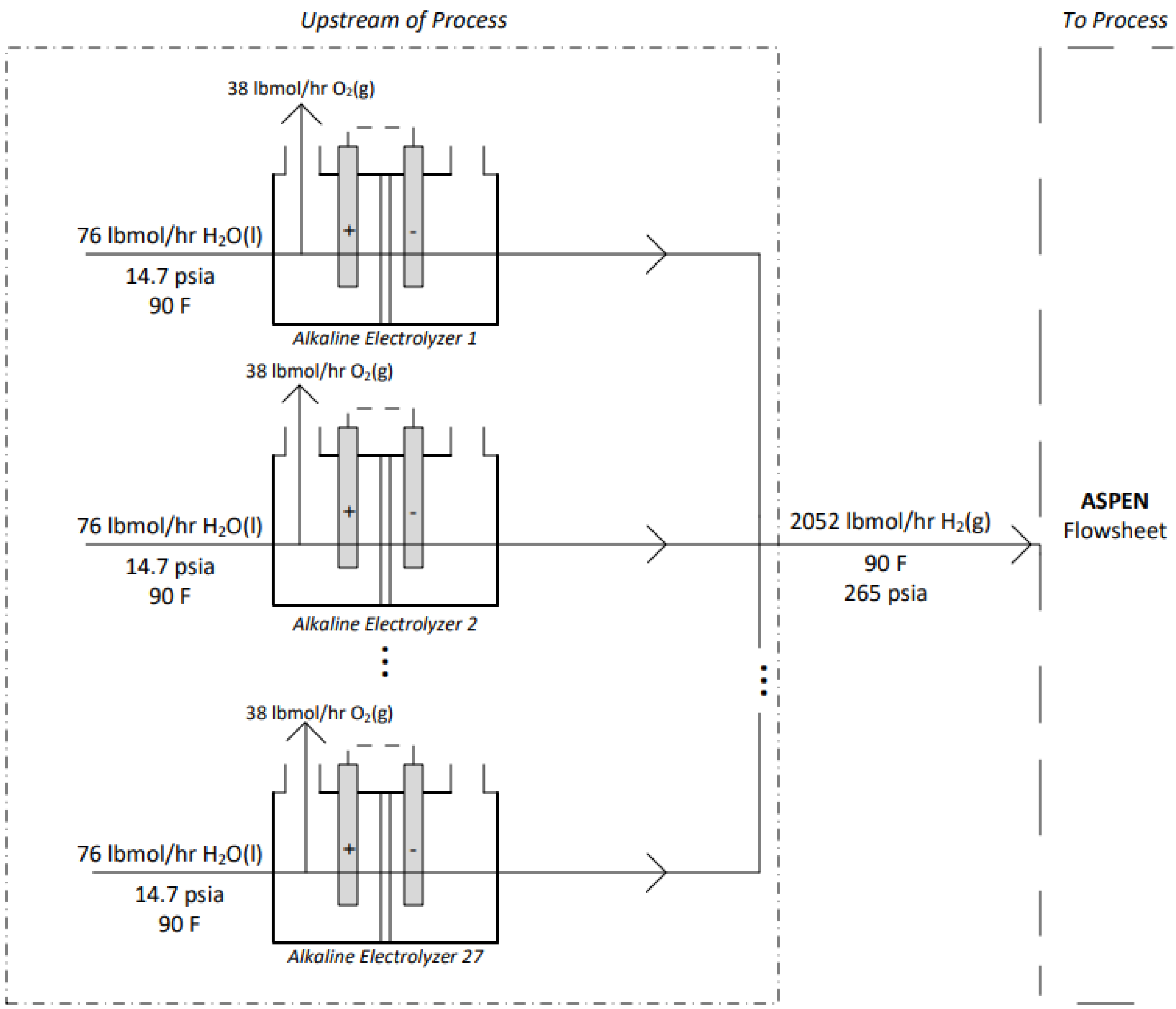


Figure 14.1: Electrolyzer mass balance, feeding 2052 lbmol/hr H2(g) to feed pre-compressor. Process water is sourced from local water providers and distributed to 27 McPhy McLyzer 800 Nm^3/hr (Normal cubic meters per hour) electrolyzers. Hydrogen gas is released at 90F and approximately 265 psia.

Figure 14.2: three electrolyzers in parallel

14.2 Oil-Injected Screw Compressors: Blocks 12 and 48

Figure 14.3: Image of oil-injected screw compressor [48]

After vaporizing and separating from Oxygen at the anode, Hydrogen is delivered to a feed compressor where it is compressed to 650 psia, resulting in a temperature increase to 110F. Compression is performed as the first step of the refrigeration cycle. The second oil-injected compressor, compressor 48, returns recycled Hydrogen to mix with the outlet of compressor 12.

14.3 Non-Catalytic BAHX: Blocks 18 and 22

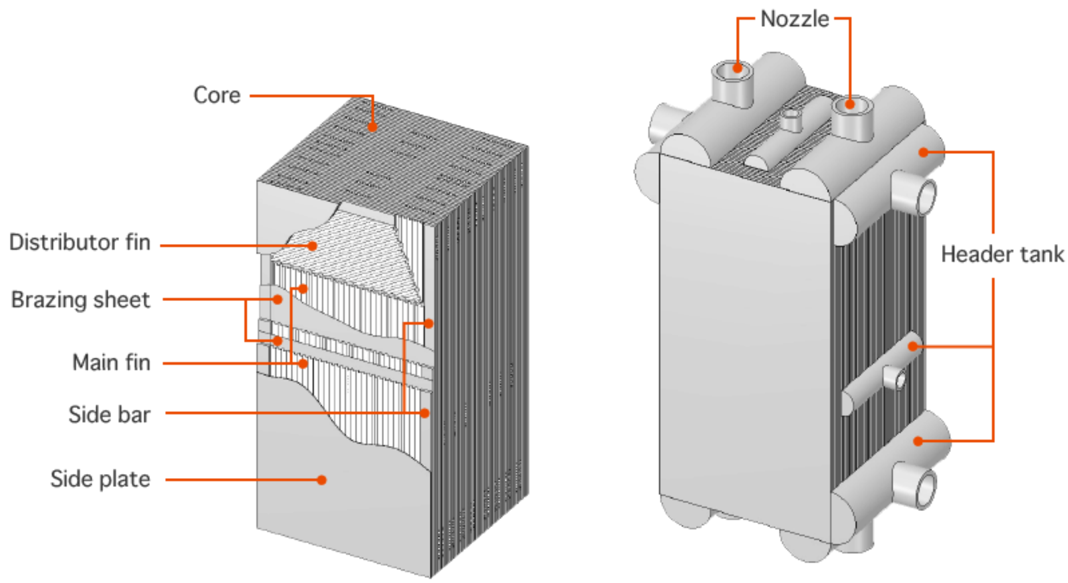


Figure 14.4: Diagram of a typical Brazed Aluminum Plate-Fin Heat Exchanger (BAHX), fitted with nozzles, header tanks, and distributor fins. [49]

The mixed stream of the outlets of compressors 12 and 48 are delivered to a 5-stream counter-current brazed aluminum heat exchanger, Heat Exchanger 18. Hydrogen enters and is distributed to a singular perforated passage, according to Figures 14.6 through 14.9. Elaborate heat exchanger design is discussed in appendix 24.1. Results are tabulated and illustrated on the following pages.

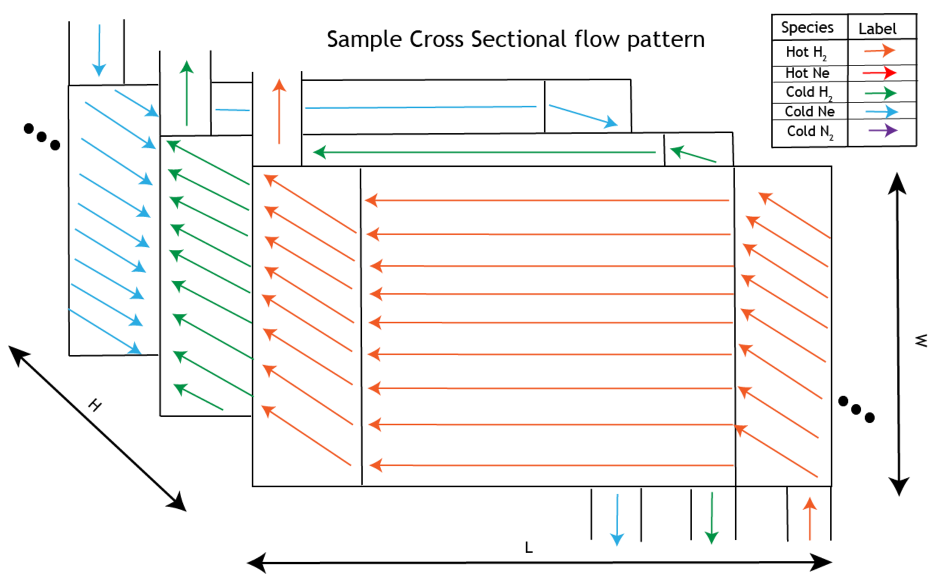


Table 14.1: cross section of heat exchanger core

Dimensions	
Height (ft)	2.877
Width (ft)	2.5
Length (ft)	8
No. of Passages	107
Total heat transfer area (ft^2)	21,502
Total effective heat transfer area (ft^2)	16,631
Overall heat transfer coefficient U (BTU/(hr-ft^2-R))	78

Table 14.2: HXer 18 dimensions

Performance Specifications							
Stream	-	A	B		D	E	F
Fluid	-	Hydrogen, Hot Side	Neon, Hot Side		Hydrogen, Cold Side	Neon, Cold Side	Nitrogen, Cold Side
Flow Rate	lb/hr	4,750	196,752		613	196,752	42,020
Inlet Temperature	°F	110	110		-309	-318	-309
Outlet Temperature	°F	-290	-290		99	99	99
Inlet Pressure	psia	650	150		51	24	17
Total Surface Area	ft^2	73	2,203		73	15,322	3,830
Fin Type	-	Perforated	Perforated		Perforated	Lanced	Lanced
Fin Height	in	0.200	0.200		0.200	0.375	0.375
Fin Thickness	in	0.025	0.025		0.025	0.008	0.008
Fin Frequency	1/in	6	6		6	15	15
Allowable Pressure Drop	psia	15	10		2	2	2
Designed Pressure Drop	psia	6.99	6.19		0.47	2.65	1.43
No. of passages	-	1	30		1	60	15
Passage Pattern	-	EEEFBEEBEFBEEEFBEEBEBEEEFBEEBEFBEEEFBEEBEEEBEFBEEBEFB EEADEFBEEEBEEBEFBEEBEFBEEBEFBEEEBEEEFBEEEFBEEBEFBEEEB					

Table 14.3: HXer 18 passage details

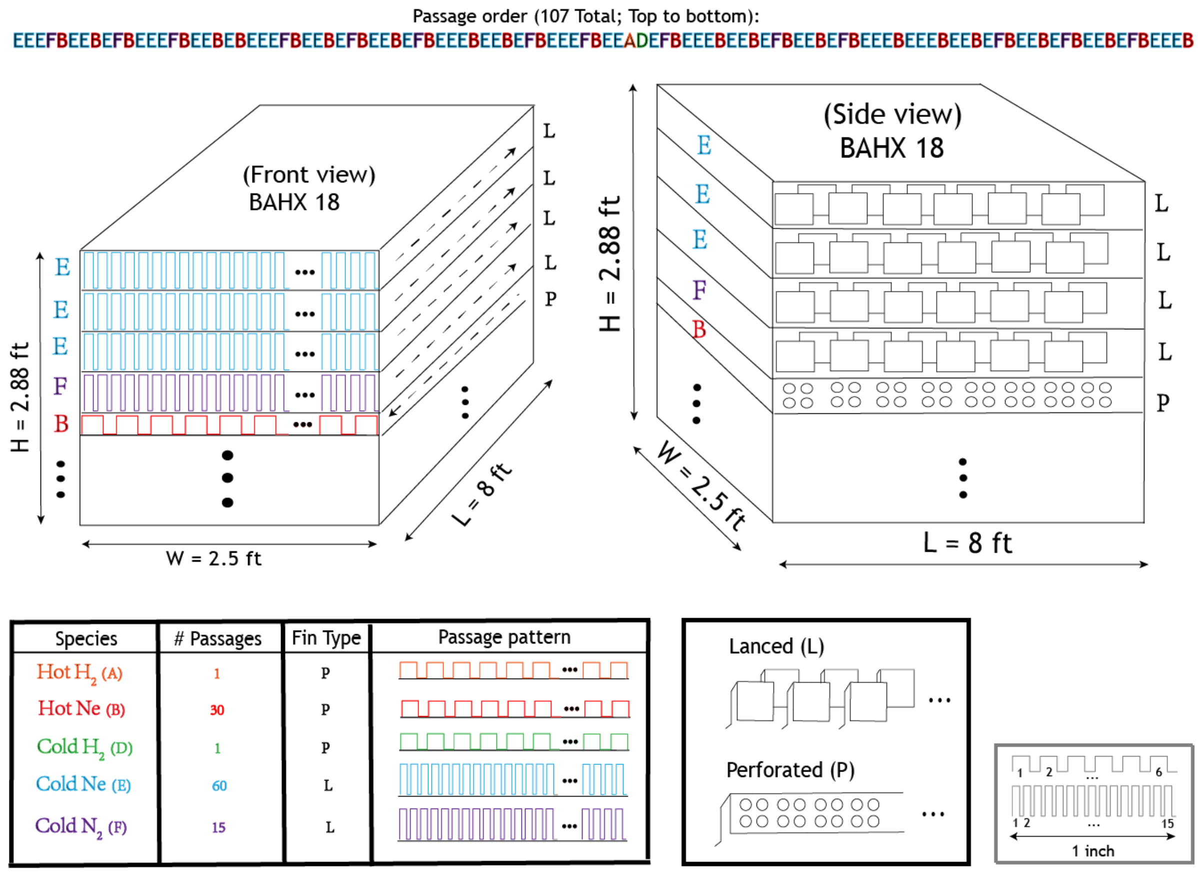


Figure 14.5: Exchanger 18, a five-stream exchanger with displayed passage numbers, passage pattern, fin type, and fin density. HX dimensions are labeled. Design details are provided in the appendix

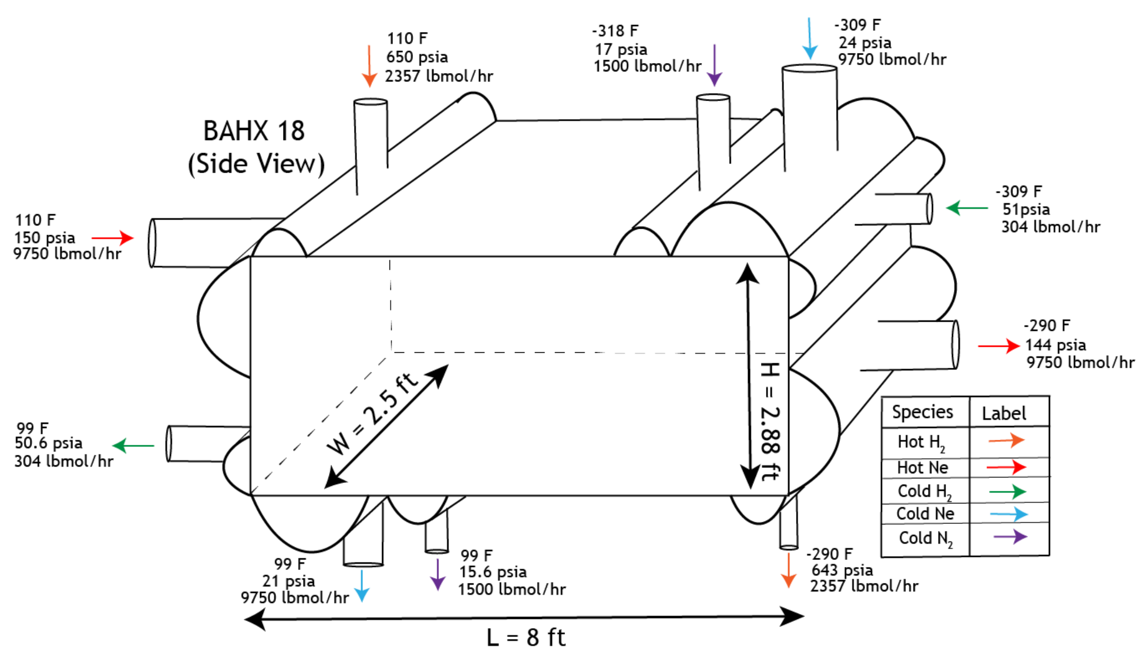


Figure 14.6: Exchanger 18 with displayed headers, nozzles, and stream data

Dimensions	
Height (ft)	0.3
Width (ft)	2
Length (ft)	3
No. of Passages	18
Total heat transfer area (ft^2)	375
Total effective heat transfer area (ft^2)	329
Overall heat transfer coefficient U (BTU/(hr-ft^{2}-R))	179

Table 14.4: HXer 22 dimensions

Performance Specifications					
Stream	-	A	B		F
Fluid	-	Hydrogen, Hot Side	Neon, Hot Side		Nitrogen, Cold Side
Flow Rate	lb/hr	4,750	196,752		19,497
Inlet Temperature	°F	-290	-290		-316
Outlet Temperature	°F	-305	-305		-316
Inlet Pressure	psia	643	144		20
Total Surface Area	ft^2	22	330		44
Fin Type	-	Perforated	Perforated		Perforated
Fin Height	in	0.2	0.2		0.2
Fin Thickness	in	0.025	0.025		0.025
Fin Frequency	1/in	6	6		6
Allowable Pressure Drop	psia	10	10		1
Designed Pressure Drop	psia	2.31	8.33		1.13
No. of passages	-	1	15		2
Passage Pattern	-	BBBBBFBBBABBFBBBBB			

Table 14.5: HXer 22 passage details

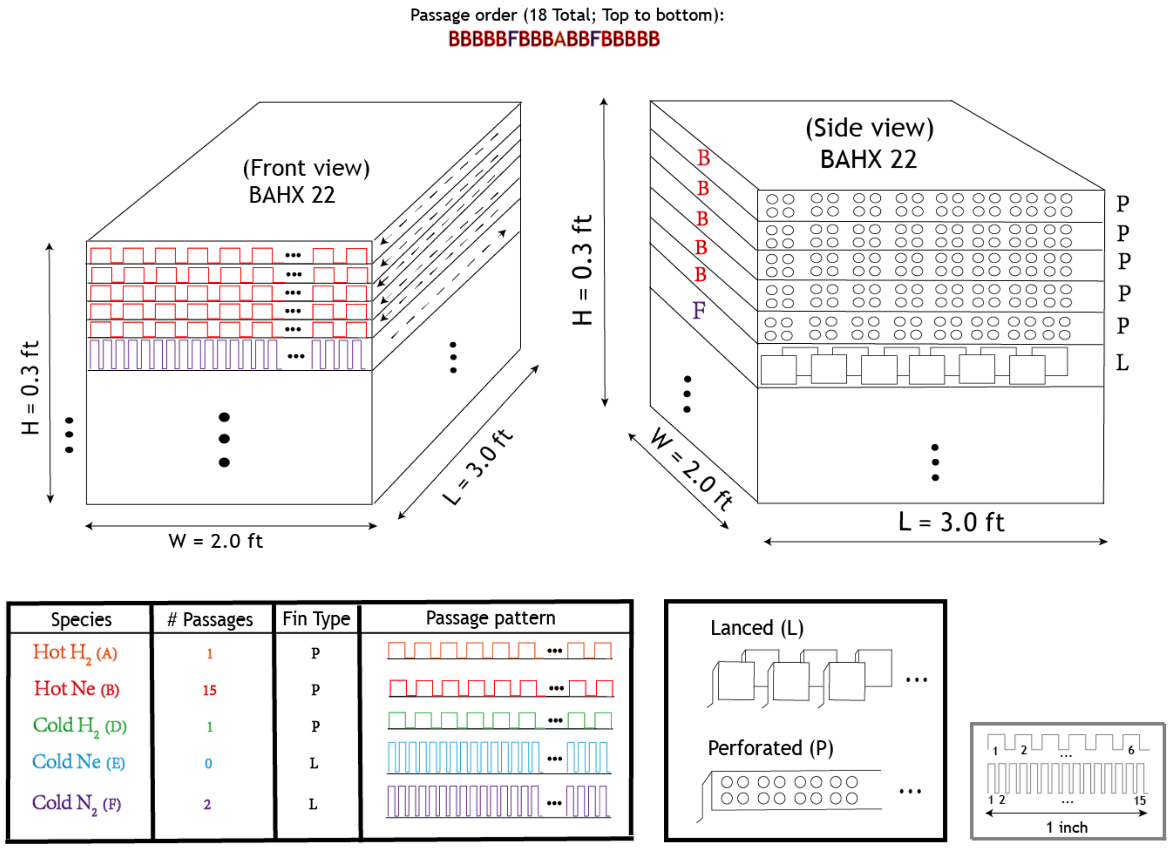


Figure 14.7: Exchanger 22, a three-stream exchanger with displayed passage number, passage pattern, fin type, and fin density. HX dimension are labeled. Design details are provided in the appendix.

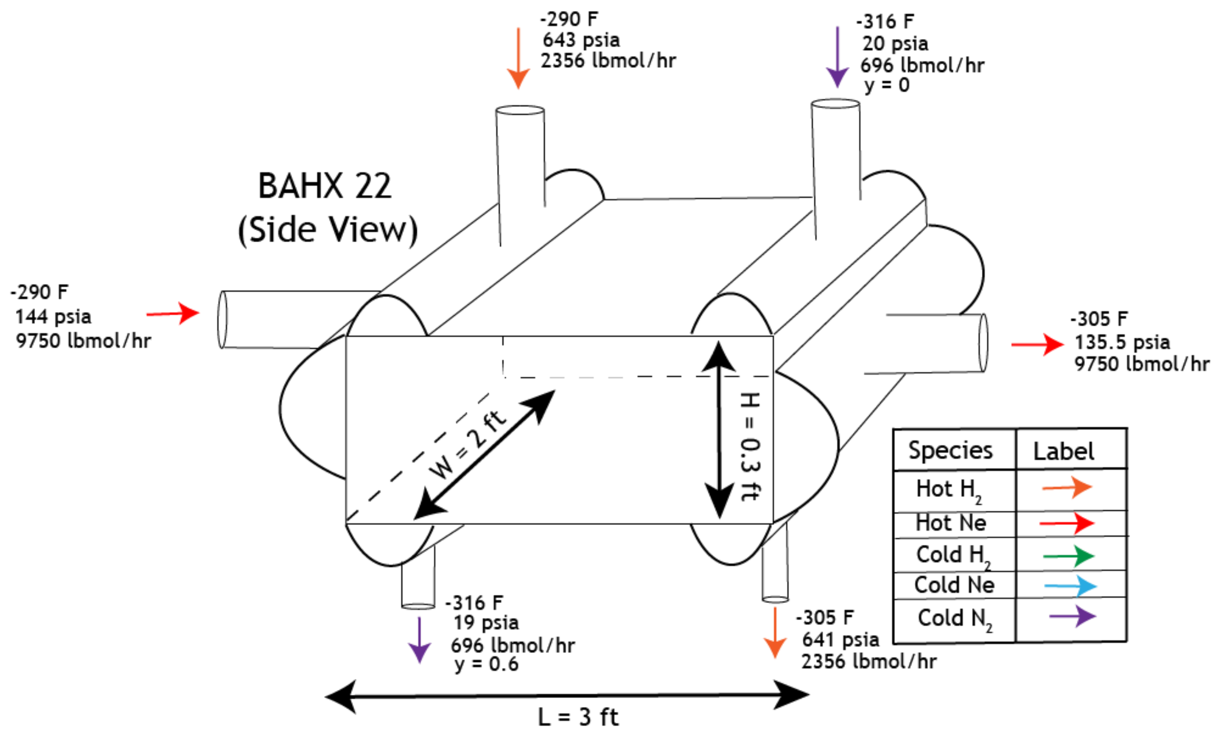


Figure 14.8: Exchanger 22 with displayed headers, nozzles, and stream data

14.4 Hydrogen Gas Purifier – 80K Adsorber

Figure 14.9: Hydrogen impurity temperature-swing adsorption column offered by Ability Engineering [50]

After leaving Heat Exchanger 22, the hot Hydrogen stream enters a Hydrogen purification system known as an 80K adsorber. Using activated Carbon with temperature-swing adsorption, the adsorber entraps residual impurities present in the feed stream. The apparatus requires no utilities, offers a filter shelf life that does not invoke additional plant shutdowns beyond those already planned throughout a given year, and boasts an outlet purity of 99.999%+. No replacement filters are required, as the filter is self-regenerative through a drying process which completes in under 6 hours. Pressure drop is assumed to be negligible. Given a Hydrogen source of process water, impurities are considered to be trace amounts of oxygen and other air gases which are water soluble.

14.5 Catalytic BAHX: Blocks 26 and 30

In heat exchangers 26 and 30, hot Hydrogen undergoes simultaneous cooling and reaction to form the p-H2 spin isomer. Both hot Hydrogen passages are packed with *Ionex*, a commercially available Iron Oxide catalyst that is packed into the exchanger during fabrication. Outlet compositions are estimated at 64% and 95% p-H2 for exchangers 26 and 30, respectively. Elaborate heat exchanger design that accounts for the mass of packed catalyst within is discussed in the appendix 24.1. Results are tabulated and illustrated on the following pages.

Dimensions	
Height (ft)	1.550
Width (ft)	2.5
Length (ft)	8
No. of Passages	58
Total heat transfer area (ft^2)	11,536
Total effective heat transfer area (ft^2)	9,295
Overall heat transfer coefficient U (BTU/(hr-ft^2-R))	127

Table 14.6: HXer 26 dimensions.

Performance Specifications						
Stream	-	A	B		D	E
Fluid	-	Hydrogen, Hot Side	Neon, Hot Side		Hydrogen, Cold Side	Neon, Cold Side
Flow Rate	lb/hr	4,750	104,279		613	196,752
Inlet Temperature	°F	-305	-305		-371	-370
Outlet Temperature	°F	-367	-367		-309	-309
Inlet Pressure	psia	641	136		52	26
Total Surface Area	ft^2	147	1,102		73	10,214
Fin Type	-	Perforated	Perforated		Perforated	Lanced
Fin Height	in	0.2	0.2		0.2	0.375
Fin Thickness	in	0.025	0.025		0.025	0.008
Fin Frequency	1/in	6	6		6	15
Allowable Pressure Drop	psia	10	10		5	1
Designed Pressure Drop	psia	0.60	2.94		0.58	1.94
No. of passages	-	2	15		1	40
Passage Pattern	-	EEBEEBEEBEEEEBEEBEEBEEADEEBEE BEEBEEAEEBEEBEEBEEEEBEEBEEBEE				

Table 14.7: HXer 26 passage details

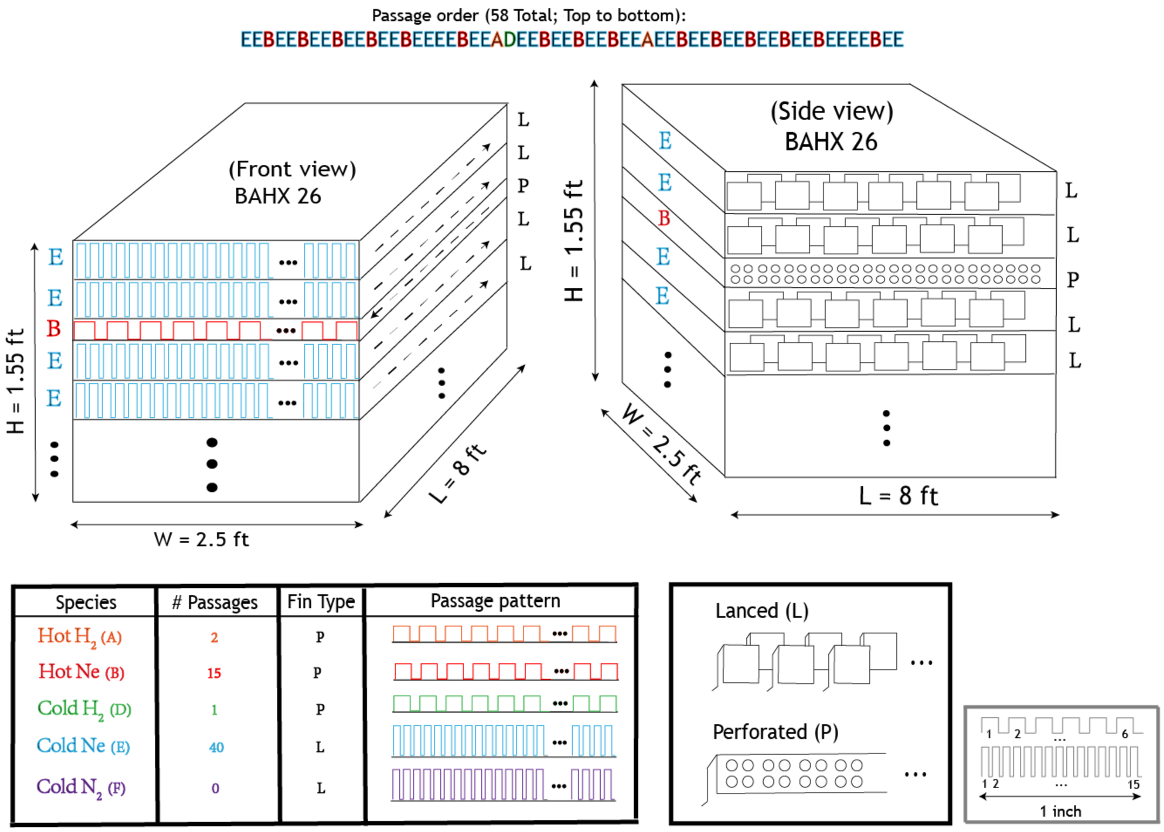


Figure 14.10: Exchanger 26, a four-stream exchanger with displayed passage number, passage pattern, fin type, and fin density. HX dimensions are labeled. Design details provided in the appendix

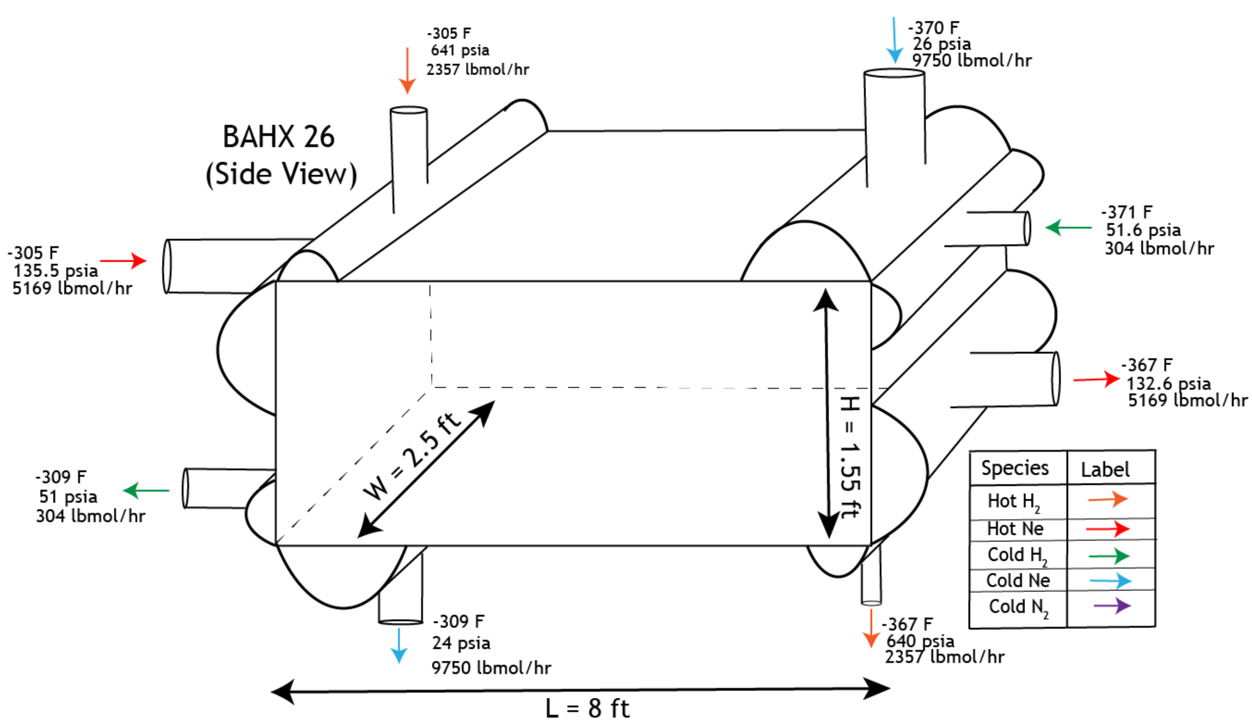


Figure 14.11: Exchanger 26 with displayed headers, nozzles, and stream data

Dimensions	
Height (ft)	0.758
Width (ft)	1
Length (ft)	1.5
No. of Passages	28
Total heat transfer area (ft^2)	427
Total effective heat transfer area (ft^2)	326
Overall heat transfer coefficient U (BTU/(hr-ft^2-R))	589

Table 14.8: HXer 30 dimensions

Performance Specifications					
Stream	-	A		D	E
Fluid	-	Hydrogen, Hot Side		Hydrogen, Cold Side	Neon, Cold Side
Flow Rate	lb/hr	4,750		613	104,279
Inlet Temperature	°F	-367		-413.9	-405
Outlet Temperature	°F	-404		-371	-371
Inlet Pressure	psia	640		52	27
Total Surface Area	ft^2	39		6	383
Fin Type	-	Perforated		Perforated	Lanced
Fin Height	in	0.2		0.2	0.375
Fin Thickness	in	0.025		0.025	0.008
Fin Frequency	1/in	6		6	15
Allowable Pressure Drop	psia	10		5	1
Designed Pressure Drop	psia	0.21		0.36	1.53
No. of passages	-	7		1	20
Passage Pattern	-	EEEEAEEAEEAEEADEEAEEAEEAEEEE			

Table 14.9: HXer 30 passage details

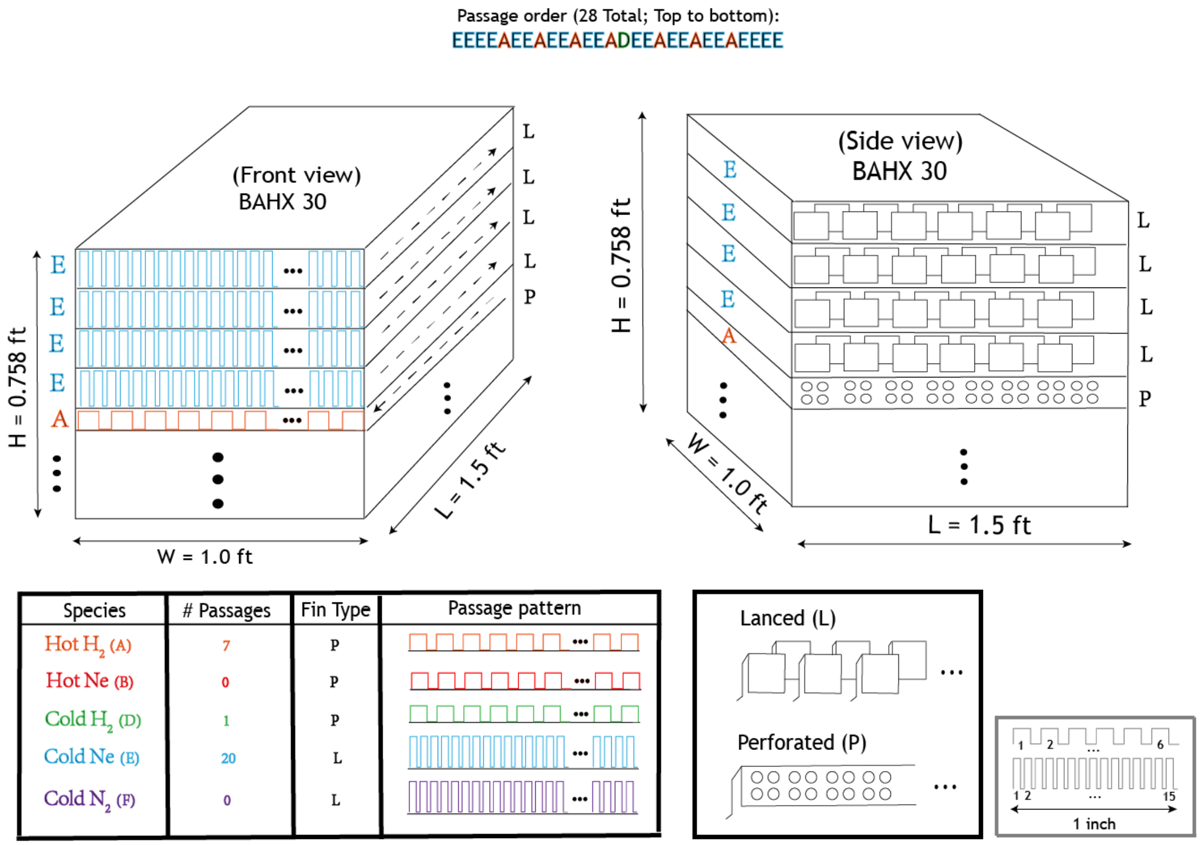


Figure 14.12: Exchanger 30, a three-stream exchanger with displayed passage number, passage pattern, fin type, and fin density. HX dimensions are labeled. Design details are provided in the appendix.

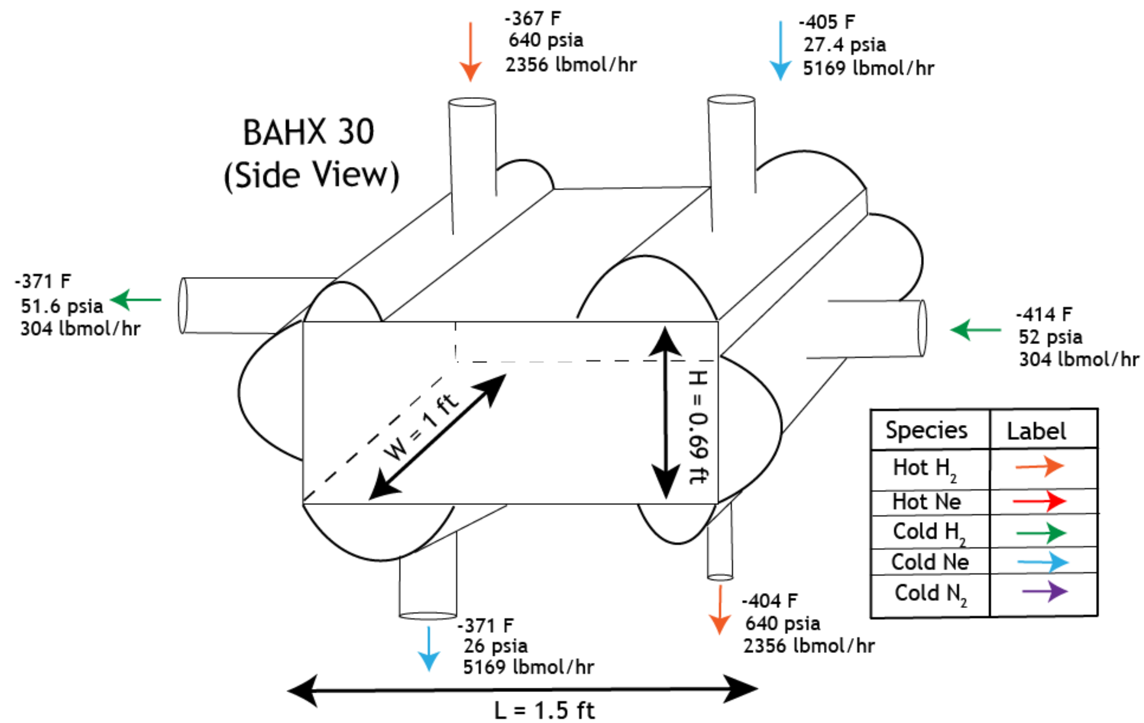


Figure 14.13: Exchanger 30 with displayed headers, nozzles, and stream data

14.6 Dense Fluid Expander

Figure 14.14: Dense fluid expander

The dense fluid expander is an expansion unit designed for supercooled cryogenic gases and vapor-liquid mixtures. When entering the expander, Hydrogen is fed as a pure gas. After expanding from 600 psi to 52 psi and generating power for the grid, Hydrogen leaves as a vapor-liquid equilibrium to be separated at the catalytic phase separator.

14.7 Knockout Drum (O-P Converter and Separator)

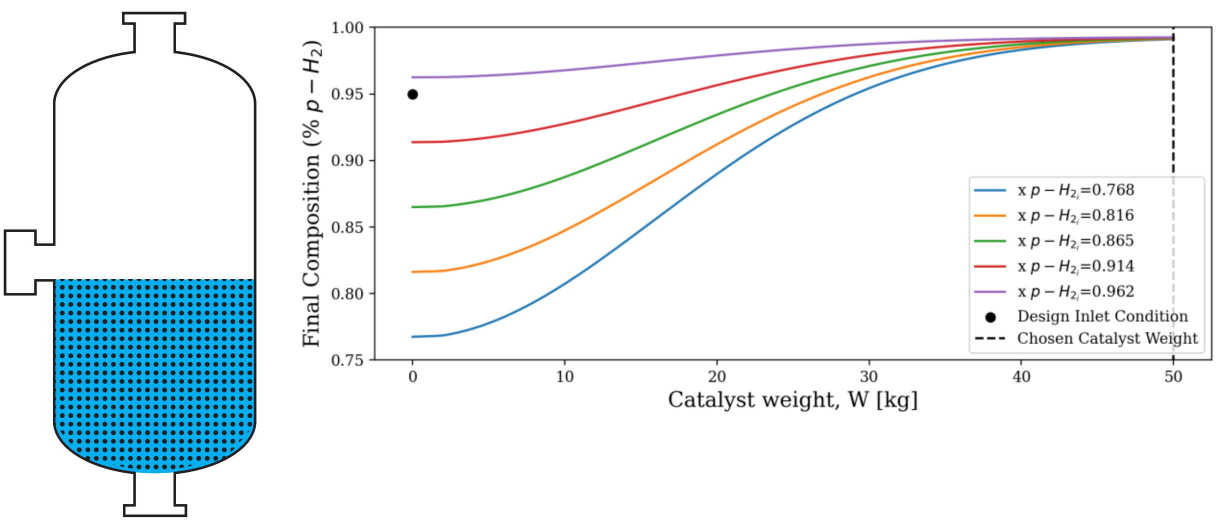


Figure 14.15: Left: Rudimentary diagram of a knockout drum filled with solid particle catalyst in its bottom half, by which liquid is reacted to further extent and gas is returned to recycle stream. Right: conversion graph accounting for specification sheet dimensions of vessel, kinetic data, and equilibrium constraints using PBR model.

Unlike a traditional knockout drum, the phase separator is packed to the calculated liquid fill volume with Ionex® ferric oxide catalyst to react residual o-H_2 close to the equilibrium p-H2 concentration of 98.7%. The exothermic heat released is assumed to be absorbed by the heat of vaporization of the liquid phase, as this condition would create a higher specific energy requirement and is used to offer a conservative efficiency metric. The gas phase is returned through Heat Exchangers 30, 26, 22, and 18 as a coolant until repressurized and mixed the feed Hydrogen. The liquid phase is transported to a Hydrogen storage tank 100 feet away (Section 19.8). The previously stated assumption, however, neglects possible temperature changes that would thermodynamically constrain conversion. To determine which of the two models offers higher accuracy would require a temperature gradient of the drum to be generated, which at such a low conversion is considered unnecessary.

The phase separator may be approximated as a PBR of length equal to the steady-state liquid balance in the tank, and residence time proportional to said length and the liquid flow rate, 45 MTD. Using first-order reaction kinetics and constants provided by Xia Teng et. al [51], a series of composition-weight curves were generated based on temperature and pressure conditions specified in the proposed flowsheet. With a drum volume of 5.78 ft^3, and cross section 1.431 ft^2, a liquid linear velocity of 0.08 ft/s required, leads to a catalyst contact residence time of 16.39 seconds – a realistic constraint for liquid passing through a packed bed. Due to the relatively high density of liquid Hydrogen, significantly faster reaction rates are possible than those possible for gaseous Hydrogen. The total occupied volume of the liquid Hydrogen and the solid catalyst occupies approximately 4.25 ft^3, leaving sufficient void volume for 85% flooding and tolerable linear gas velocity. Due to the small cost of the unit operation, upscaling can be performed with insignificant economic effect. Under these specifications, process inefficiencies in product conversion caused by plant aging could be offset at this final operation, at the cost of higher evaporation rates and lower maximum conversion.

14.8 High and Low Turboexpanders

To provide Neon working fluid at two different temperatures for the inlets to heat exchanger 30 and 26, the Neon hot stream is split between heat exchanger 22 and 26, with one stream continuing cooling in 26. Both streams are expanded through turboexpanders for refrigerative cooling and utilized as the primary cold fluid for the next exchanger after the exchanger that each had last been cooled in. The streams rejoin prior to entering heat exchanger 26.

14.9 Centrifugal Compressor: Neon Recycle Compressor

Figure 14.15 [52]

The Neon recycle compressor is the single most expensive unit operation aside from electrolysis. By compressing the working fluid, it prepared the hot stream to be cooled by Nitrogen and ultimately turboexpanded to refrigeratively cool.

15 SPECIFICATION SHEETS

The location of each unit operation is written on the upper left-hand side of each specification sheet. If it is part of the process itself and not upstream, the ASPEN PLUS block number is provided.

15.1 Hydrogen Recycle Compressor

Hydrogen Recycle Compressor					
Identification:	**Item**	*Oil-injected Screw Compressor*			
Block 48	Item No.	*(ASPEN)*			
	No. Required	1			
Function:	Compress 100% parahydrogen recycle to match feed conditions				
Operation:	Continuous				
Materials Handled:	S-46	---	---	---	---
Mass flow (lb/hr):	613				
Volumetric Flow (cuf/hr):	36083				
Mass Frac. (Liq phase):	0				
Mass Frac. (Vap phase):	1				
Temperature (F):	99				
Composition:	---				
H20	(trace)				
O2	(trace)				
o-H2	0				
p-H2	1				
Ne	0				
N2	0				
Design Data:	Discharge Pressure:	*650 psia*			
	Type:	*Isentropic*			
	Isentropic efficiency:	*0.7*			
	Material of Construction:	*Stainless steel*			
	Power	*543 hp*			
	---	---			
Utilities:	Green Process Electricity				
Comments and drawings:	See PFD and Appendix 24.6.1				

Please note the listed concentration of p-H_2 and o-H_2. These are written as approximations used in the ASPEN flowsheet preparation. In practice, there is residual o-H_2 in this unit.

15.2 Hydrogen Feed Compressor

		Feed Precompressor				
Identification:	**Item**	*Oil-injected Screw Compressor*				
Block 12	Item No.	*(ASPEN)*				
	No. Required	1				
Function:	Compress incoming hydrogen feed					
Operation:	Continuous					
Materials Handled:	Feed (S-10)	---	---	---	---	
Mass flow (lb/hr):	4137					
Volumetric Flow (cuf/hr):	46161					
Mass Frac. (Liq phase):	0					
Mass Frac. (Vap phase):	1					
Temperature (F):	90					
Composition:	---					
H20	(trace)					
O2	(trace)					
o-H2	0.75					
p-H2	0.25					
Ne	0					
N2	0					
Design Data:	Discharge Pressure:	*650 psia*				
	Type:	*Isentropic*				
	Isentropic efficiency:	*0.7*				
	Material of Construction:	*Stainless steel*				
	Power	*1193 hp*				
	---	---				
Utilities:	Green Process Electricity					
Comments and drawings:	See PFD and Appendix 24.6.1					

15.3 Neon Recycle Compressor

Neon Recycle Compressor						
Identification:	**Item**	*Centrifugal Compressor*				
Block 94	Item No.	*(ASPEN)*				
	No. Required	1				
Function:	Compress Ne refrigerant to return through Heat Exchanger blocks as the hot fluid					
Operation:	Continuous					
Materials Handled:	S-92	---	---	---	---	
Mass flow (lb/hr):	196752					
Volumetric Flow (cuf/hr):	2747456.8					
Mass Frac. (Liq phase):	0					
Mass Frac. (Vap phase):	1					
Temperature (F):	99					
Composition:	---					
H20	0					
O2	0					
o-H2	0					
p-H2	0					
Ne	1					
N2	0					
Design Data:	Discharge Pressure:	*150 psia*				
	Type:	*Isentropic*				
	Isentropic efficiency:	*0.7*				
	Material of Construction:	*Carbon steel*				
	Power	*13368 hp*				
	---	---				
Utilities:	Green Process Electricity					
Comments and drawings:	See PFD and Appendix 24.6.1					

15.4 High-level Turboexpander

High-level Turboexpander						
Identification:	**Item**	*Turboexpander*				
Block 84, 1	Item No.	*(ASPEN)*				
	No. Required	1				
Function:	Recover energy from high pressure refrigerant (Ne)					
Operation:	Continuous					
Materials Handled:	S-82	---	---	---	---	
Mass flow (lb/hr):	93422					
Volumetric Flow (cuf/hr):	59054					
Mass Frac. (Liq phase):	0					
Mass Frac. (Vap phase):	1					
Temperature (F):	-305					
Composition:	---					
H20	0					
O2	0					
o-H2	0					
p-H2	0					
Ne	1					
N2	0					
Design Data:	Discharge Pressure:	*25 psia*				
	Type:	*Isentropic*				
	Isentropic efficiency:	*0.85*				
	Material of Construction:	*Stainless steel*				
	Power recovered	*561 hp*				
	---	---				
Utilities:	None					
Comments and drawings:	See PFD and Appendix 24.6.2					

15.5 Low-level Turboexpander

Low-level Turboexpander						
Identification:	**Item**	*Turboexpander*				
Block 74, 1	Item No.	*(ASPEN)*				
	No. Required	1				
Function:	Recover energy from high pressure refrigerant (Ne)					
Operation:	Continuous					
Materials Handled:	S-74	---	---	---	---	
Mass flow (lb/hr):	105348					
Volumetric Flow (cuf/hr):	39306					
Mass Frac. (Liq phase):	0					
Mass Frac. (Vap phase):	1					
Temperature (F):	-367					
Composition:	---					
H20	0					
O2	0					
o-H2	0					
p-H2	0					
Ne	1					
N2	0					
Design Data:	Discharge Pressure:	*26.5 psia*				
	Type:	*Isentropic*				
	Isentropic efficiency:	*0.85*				
	Material of Construction:	*Stainless steel*				
	Power recovered	*339 hp*				
	---	---				
Utilities:	None					
Comments and drawings:	See PFD and Appendix 24.6.2					

15.6 Dense Fluid Expander

Dense Fluid Expander						
Identification:	**Item**	*Turboexpander*				
Block 34, B	Item No.	*(ASPEN)*				
	No. Required	1				
Function:	Recover energy from supercritical hydrogen product pre-sale					
Operation:	Continuous					
Materials Handled:	S-36, B1	---	---	---	---	
Mass flow (lb/hr):	4750					
Volumetric Flow (cuf/hr):	1184					
Mass Frac. (Liq phase):	1					
Mass Frac. (Vap phase):	0					
Temperature (F):	-404					
Composition:	---					
H20	0					
O2	0					
o-H2	0					
p-H2	1					
Ne	0					
N2	0					
Design Data:	Discharge Pressure:	*52 psia*				
	Type:	*Isentropic*				
	Isentropic efficiency:	*0.7*				
	Material of Construction:	*Stainless steel*				
	Power recovered	*35 hp*				
	---	---				
Utilities:	None					
Comments and drawings:	See PFD and Appendix 24.6.2					

15.7 Heat Exchanger 1

		Multi-stream Heat Exchanger			
Identification:	**Item**	*Brazed Aluminum Heat Exchanger*			
Block 18	Item No.	*BAHX 18*			
	No. Required	1			
Function:	Cool the entering feed hydrogen				
Operation:					
Materials Handled	[Hot] S-16	[Hot] S-68	[Cold] S-90	[Cold] S-58	[Cold] S-46, +R
Quantity (lb/hr):	4750	198770	198770	42020	613
Composition:	---	---	---	---	---
H20	(trace)	0	0	0	0
O2	(trace)	0	0	0	0
o-H2	0.65	0	0	0	0
p-H2	0.35	0	0	0	1
Ne	0	1	1	0	0
N2	0	0	0	1	0
Inlet Temperature (F):	110	110	-309	-318	-309
Inlet Pressure (psia):	650	150	24	17	50
Inlet vapor fraction:	1	1	1	0.813	1
Outlet Temperature (F):	-290	-290	98.7	98.7	98.5
Outlet Pressure (psia):	643	143.8	21.3	15.5	49
Outlet vapor fraction:	1	1	1	1	1
Pressure drop (psia):	-7	-6.7	-2.7	-1.5	-1
Design Data:	Length:	*8 ft*			
	Width:	*2.5 ft*			
	Height:	*2.71 ft*			
	Construction Material:	*Aluminum*			
Utilities:	None				
Comments and drawings:	See PFD and Appendix 24.1, 24.6.4				

15.8 Heat Exchanger 2

		Multi-stream Heat Exchanger				
Identification:	**Item**	*Brazed Aluminum Heat Exchanger*				
Block 22	Item No.	*BAHX 22*				
	No. Required		1			
Function:	Cool the entering feed hydrogen					
Operation:						
Materials Handled	[Hot] S-20, L	[Hot] S-68, 1	[Cold] S-T52	---	---	
Quantity (lb/hr):	4750	198770	19497	---	---	
Composition:	---	---	---	---	---	
H20	(trace)	0	0	---	---	
O2	(trace)	0	0	---	---	
o-H2	0.65	0	0	---	---	
p-H2	0.35	0	0	---	---	
Ne	0	1	0	---	---	
N2	0	0	1	---	---	
Inlet Temperature (F):	-290	-290	-316	---	---	
Inlet Pressure (psia):	643	143.8	20	---	---	
Inlet vapor fraction:	1	1	0	---	---	
Outlet Temperature (F):	-305	-305	-316.5	---	---	
Outlet Pressure (psia):	640.7	135.5	18.87	---	---	
Outlet vapor fraction:	1	1	0.59	---	---	
Pressure drop (psia):	-2.3	-8.3	-1.13	---	---	
Design Data:	Length:	*3 ft*				
	Width:	*2 ft*				
	Height:	*0.3 ft*				
	Construction Material:	*Aluminum*				
Utilities:	None					
Comments and drawings:	See PFD and Appendix 24.1, 24.6.4					

15.9 Heat Exchanger 3

	Dual Multi-stream Heat Exchanger and Reactor				
Identification:	**Item**	*Brazed Aluminum Heat Exchanger*			
Block 26	Item No.	*BAHX 26*			
	No. Required	1			
Function:	Cool and react the entering feed hydrogen from orthohydrogen to parahydrogen				
Operation:					
Materials Handled	[Hot] S-72	[Hot] S-24, 1	[Cold] S-46, -2	[Cold] S-88	---
Quantity (lb/hr):	105348	4750	613	198770	---
Composition:	---	---	---	---	---
H20	0	0	0	0	---
O2	0	0	0	0	---
o-H2	0	0	0	0	---
p-H2	0	1	1	0	---
Ne	1	0	0	1	---
N2	0	0	0	0	---
Inlet Temperature (F):	-305	-305	-370.8	-369.8	---
Inlet Pressure (psia):	135.5	640.7	51.6	25.9	---
Inlet vapor fraction:	1	1	1	1	---
Outlet Temperature (F):	-367	-367	-308.8	-308.8	---
Outlet Pressure (psia):	132.6	640.1	51	23.9	---
Outlet vapor fraction:	1	1	1	1	---
Pressure drop (psia):	-0.6	-2.94	-1.94	-0.58	---
Design Data:	Length:	*8 ft*			
	Width:	*2.5 ft*			
	Height:	*1.467 ft*			
	Construction Material:	*Aluminum*			
	Packed Catalyst:	*Ferric Oxide*			
Utilities:	None				
Comments and drawings:	See PFD and Appendix 24.1, 24.6.4				

15.10 Heat Exchanger 4

		Dual Multi-stream Heat Exchanger and Reactor				
Identification:	**Item**	*Brazed Aluminum Heat Exchanger*				
Block 30	Item No.	*BAHX 30*				
	No. Required		1			
Function:	Cool and react the entering feed hydrogen from orthohydrogen to parahydrogen					
Operation:						
Materials Handled	[Hot] S-28, 1	[Cold] S-78	[Cold] S-42	---	---	
Quantity (lb/hr):	4750	105348	613	---	---	
Composition:	---	---	---	---	---	
H20	0	0	0	---	---	
O2	0	0	0	---	---	
o-H2	0	0	0	---	---	
p-H2	1	0	1	---	---	
Ne	0	1	0	---	---	
N2	0	0	0	---	---	
Inlet Temperature (F):	-367	-405	-414	---	---	
Inlet Pressure (psia):	640.1	27.4	52	---	---	
Inlet vapor fraction:	1	1	1	---	---	
Outlet Temperature (F):	-404	-370.9	-370.8	---	---	
Outlet Pressure (psia):	640	25.9	51.6	---	---	
Outlet vapor fraction:	0	1	1	---	---	
Pressure drop (psia):	-0.11	-1.53	-0.36	---	---	
Design Data:	Length:	*1.5 ft*				
	Width:	*1 ft*				
	Height:	*0.692 ft*				
	Construction Material:	*Aluminum*				
	Packed Catalyst:	*Ferric Oxide*				
Utilities:	None					
Comments and drawings:	See PFD and Appendix 24.1, 24.6.4					

15.11 Knockout Drum (O-P Converter and Separator)

Knockout Drum						
Identification:	**Item**	*Knockout Drum*				
Block 38	Item No.	*(ASPEN)*				
	No. Required	1				
Function:	Separate supercritical liquid hydrogen from its vapor					
Operation:	Continuous					
Materials Handled:	S-36, B2	---	---	---	---	
Mass flow (lb/hr):	4750					
Volumetric Flow (cuf/hr):	1345					
Mass Frac. (Liq phase):	1					
Mass Frac. (Vap phase):	0					
Temperature (F):	-414					
Composition:	---					
H20	0					
O2	0					
o-H2	0					
p-H2	1					
Ne	0					
N2	0					
Design Data:	Design Pressure:	*51 psia*				
	Allowable vapor volumetric flowrate:	*0.0178 cuf/s*				
	Vessel height:	*4.83 ft*				
	Vessel internal diameter:	*1.61 ft*				
	Vessel thickness	*0.3125 in*				
	Material of construction:	*Stainless steel*				
Utilities:	None					
Comments and drawings:	See PFD and Appendix 24.6.3					

Please note the listed concentration of p-H_2 and o-H_2. These are written as approximations used in the ASPEN flowsheet preparation. In practice, there is residual o-H_2 in this unit.

15.12 Cold Box

		Cold Box			
Identification:	**Item**	*Vertical Pressure Vessel*			
(External to flowsheet)	Item No.	*(ASPEN)*			
	No. Required	1			
Function:	Insulate unit operations at cryogenic conditions				
Operation:	---				
Unit Operations contained:	MHeatX 1 (B-18)	MHeatX 2 (B-22)	MHeatX 3 (B-26)	MHeatX 4 (B-30)	KO Drum (B-30)
Volume (cuf):	44	2	29	1	6
Temperature (F):	---	---	---	---	-414
Design Data:	Design Pressure:	*50 psia*			
	Vessel height:	*45 ft*			
	Vessel internal diameter:	*7.5 ft*			
	Vessel thickness:	*1 inch*			
	Material of Construction:	*Stainless steel*			
	---	---			
Utilities:	None				
Comments and drawings:	See PFD and Appendix 24.6.5				

Dimensions for this Cold Box were generated based on heuristics provided by Prof. Vrana regarding the minimum distance between each unit operation and the outer wall, to then be filled with perlite.

15.13 Storage Vessels

		Storage Vessel				
Identification:	**Item**	*Vertical Pressure Vessel*				
(External to flowsheet)	Item No.	*(ASPEN)*				
	No. Required	1				
Function:	Store liquefied hydrogen offsite					
Operation:	---					
Volume held (cuf):	46161					
Temperature (F):	-414					
Composition:	---					
H20	0					
O2	0					
o-H2	0					
p-H2	1					
Ne	0					
N2	0					
Design Data:	Design Pressure:	*50 psia*				
	Vessel height:	*45 ft*				
	Vessel internal diameter:	*7.5 ft*				
	Vessel thickness:	*2.5 inch*				
	Material of Construction:	*Stainless steel*				
	---	---				
Utilities:	None					
Comments and drawings:	See Appendix 24.6.5					

Please note the listed concentration of p-H_2 and o-H_2. These are written as approximations used in the ASPEN flowsheet preparation. In practice, there is residual o-H_2 in this unit.

15.14 80K Adsorber

		80K Adsorber				
Identification:	**Item**	*80 K Adsorber*				
Block 100	Item No.	*(ASPEN)*				
	No. Required	1				
Function:	Remove trace impurities from hot hydrogen stream using an activated carbon filter					
Operation:	Continuous					
Materials Handled:	S-(24,1)	---	---	---	---	
Mass flow (lb/hr):	4750					
Volumetric Flow (cuf/hr):	5981					
Mass Frac. (Liq phase):	0					
Mass Frac. (Vap phase):	1					
Temperature (F):	-305					
Composition:	---					
H20	(trace)					
O2	(trace)					
o-H2	0.75					
p-H2	0.25					
Ne	0					
N2	0					
Design Data:	---	---				
	---	---				
	---	---				
	---	---				
	---	---				
	---	---				
Utilities:	None					
Comments and drawings:	See PFD and Appendix 24.6.7					

Insufficient technical information is publicly available on this unit operation to provide a detailed specification sheet. Discussion on unit operation will be presented throughout.

15.15 Electrolyzer

		Electrolyzer				
Identification:	**Item**	*Electrolyzer*				
(External to flowsheet)	Item No.	*(ASPEN)*				
	No. Required	27				
Function:	React process water to form vapor hydrogen feed					
Operation:	Continuous					
Materials Handled:	Pre-Feed	---	---	---	---	
Mass flow (gal/yr):	35057511					
Mass Frac. (Liq phase):	1					
Mass Frac. (Vap phase):	0					
Temperature (F):	90					
Composition:	---					
H20	1					
O2	0					
o-H2	0					
p-H2	0					
Ne	0					
N2	0					
Design Data:	Material of construction:	*Stainless steel*				
	Rated H2 Output	*800 cum/hr*				
	Conditions	*1 atm, 25 C*				
	---	---				
Utilities:	Green Process Electricity					
Comments and drawings:	See PFD and Appendix 24.6.6					

Rated H2 output is measured in cubic meters per hour at 1 atm, 25 ºC.

16 Equipment Cost Summary

The total bare module cost for all equipment is $146.6 MM, as shown in Table 17.2. Table 16.1 shows the equipment cost for the process. Table 16.2 shows the costs for fabrication equipment. Table 16.3 shows the costs for processing equipment. Compressors and expanders contribute the most to the cost, and the minimization of their power requirement is a primary objective for this process. Stainless steel is used for all streams in contact with Hydrogen to prevent Hydrogen embrittlement.

Equipment	***Purchase Cost ($, 2023)***	***Source***
H_2 recycle compressor	$1,417,155	Seider et al.
H_2 feed compressor	$2,507,376	Seider et al.
Neon recycle compressor	$5,778,481	Seider et al.
Dense fluid expander	$141,790	Professional consultant
High-level turboexpander	$2,209,762	Professional consultant
Low-level turboexpander	$1,363,027	Professional consultant
Knockout Drum	$23,703	Seider et al.
HX 18	$1,269,020	Seider et al.
HX 22	$163,559	Seider et al.
HX 26	$680,460	Seider et al.
HX 30	$165,936	Seider et al.
Spherical Storage Vessels	$4,514,860	Seider et al.
Electrolyzers	$59,400,078	Internet (McPhy)
Cold Box	$660,674	Seider et al.
90 K Adsorber	$100,000	Internet

Table 16.1: List of Purchase Cost of All Equipment (In-Process and Upstream)

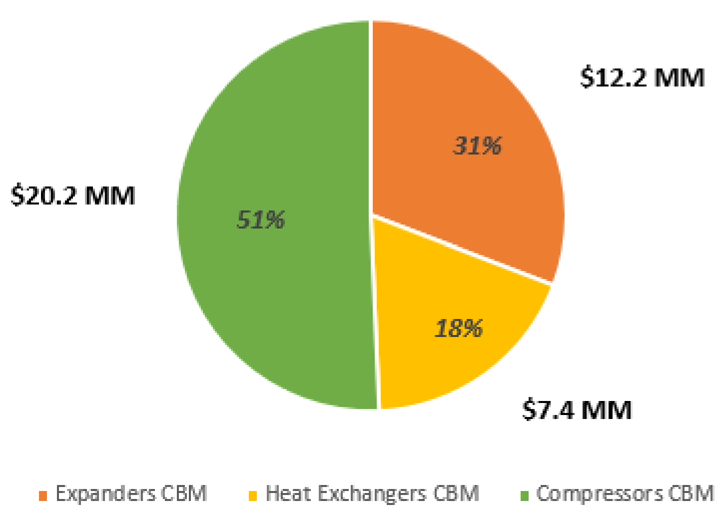


Figure 16.1: Pie chart visualizing the breakdown of equipment costs by type. Knockout Drum and 80K Adsorber are excluded from this pie chart because they are less than one percent of equipment bare module cost

ID	Equipment Name	Purchase Cost	Bare Module Factor	Bare Module Cost
B-48	H_2 recycle compressor	$1,417,155	2.15	$3,046,833
B-12	H_2 feed compressor	$2,507,376	2.15	$5,390,859
B-94	Neon recycle compressor	$5,778,481	2.15	$12,423,733
B-34B	Dense fluid expander	$141,790	3.3	$467,904
B-86,1	High-level turboexpander	$2,209,762	3.3	$7,292,213
B-76,1	Low-level turboexpander	$1,363,027	3.3	$4,497,988
B-38	Knockout Drum	$23,703	4.16	$98,606
B-18	Heat Exchanger 1	$1,269,020	3.2	$4,060,863
B-22	Heat Exchanger 2	$163,559	3.2	$523,389
B-26	Heat Exchanger 3	$680,460	3.2	$2,177,472
B-30	Heat Exchanger 4	$165,936	3.2	$530,995
-----	Cold Box	$660,674	4.16	$2,748,400

Table 16.2: Purchase cost, Bare Module Factor, and Bare Module Cost of all Fabricated equipment. Purchase costs and Bare Module All costs are in 2023 dollar values.

ID	Name	Qty.	C_p/Unit	Bare Module Factor	Bare Module Cost
---	Electrolyzer	27	$2,200,000	*1.5*	$89,100,000
B-100	80K Adsorber	1	$100,000	1.5	$150,000

Table 16.3: Equipment name, quantity, purchase cost per unit, Bare Module Factor, and Bare Module Cost of all processing equipment. All costs are in 2023 dollars. Electrolyzer bare module factor is a subject of a later sensitivity analysis.

ID	Equipment Name	Quantity	Purchase	Bare Module Factor	Bare Module Cost
---	Spherical Storage	1	$4,514,854	3.05	$13,770,306

Table 16.4: Purchase cost, Bare Module Factor, and Bare Module Cost of Storage equipment. All costs are in 2023 dollars.

Spare equipment was not accounted for in the upcoming profitability analysis for this process design. It is envisioned that 3 electrolyzers will be needed for contingencies to supplement the 27 that will be used upstream of the process.

ID	Equipment Name	Quantity (lb)	Purchase Cost ($, 2023)
---	Ferric Oxide	1500	$13,608
---	Neon	220	$350,000

Table 16.5: One time purchase cost of ferric oxide catalyst and Neon refrigerant, expressed in 2023 dollars. The details of one time purchase cost calculations for the catalyst and working fluid amount required are shown in 24.7.1 and 24.7.3.

Section 24.6 in the Appendix details the costing calculations of the equipment. Two spreadsheets were used in assessing the economic viability of this process. The first spreadsheet was put together by Akash Kumashi, and the second was put together by Professor Russell Dunn at Vanderbilt University.

Costing correlations used include equations from Chapter 16 of Seider et al and heuristics from project author Adam Brostow.

17 FIXED CAPITAL INVESTMENT SUMMARY

Fixed Capital Investment Summary:

Using the methods presented in Ch. 16 of Seider et al *Product and Process Design Principles* and Professor Dunn's Equipment Costing Spreadsheet, the plant's total permanent investment was calculated to be $231.5 MM. Table 17.1 below shows the assumptions made for the calculation of total permanent investment.

Site Factor	1.00 (Gulf Coast)
Year of Total Permanent Investment	100% in 2024
Cost of Site Preparation	10.00% of Total Bare Module Costs
Cost of Service Facilities	10.00% of Total Bare Module Costs
Allocated Cost for Utility Plants	$0
Cost of Contingencies and Contractor's Fees	18% of Direct Permanent Investment
Cost of Land	2% of Total Depreciable Capital
Cost of Royalties	$0
Cost of Plant Startup	10% of Total Depreciable Capital

Table 17.1: Heuristics used in the calculation of Total Permanent Investment (TPI). All conventions taken from Seider et al. Cost of Site preparations and Cost of Service facilities heuristics given by Professor Bruce Vrana.

The direct permanent investment (DPI) calculation is demonstrated below.

Total Bare Module Costs (C_{TBM})

Fabricated Equipment	$43,259,320
Process Equipment	$89,250,000
Spares	---
Storage	$13,770,306
Other	---
Catalyst	$13,608
Refrigerant	$350,000
Computer/Software	--
Total Bare Module Costs	$146,643,234

Table 17.2: Components of Total Bare Module Cost for the selected process configuration. All costs in the right hand column are summed to give the Total Bare Module Cost of the selected process configuration.

Direct Permanent Investment (C_{DPI})

Site Preparation Cost, C_{site}	$14,664,323
Service Facilities Cost, $C_{service}$	$14,664,323
Allocated Cost for Utilities Plants/Related Facilities	$0
Total Bare Module Costs	$146,643,234

Direct Permanent Investment	$175,971,880

Table 17.3: Components of Direct Permanent Investment for the selected process configuration. All costs in the right hand column are summed to give the Total Direct Permanent Investment of the selected process configuration.

Total Depreciable Capital (C_{TDC})

Contingencies Cost, C_{site}	$31,674,938
Direct Permanent Investment, C_{DPI}	$175,971,880
Total Depreciable Capital	$207,646,818

Table 17.4: Components of Total Depreciable Capital for the selected process configuration. All costs in the right column are summed to give the Total Depreciable Capital.

Total Permanent Investment (C_{TPI})

Cost of Land, C_{land}	$4,152,936
Cost of Royalties, $C_{royalties}$	$0
Cost of Startup, $C_{startup}$	$20,764,682
Total Depreciable Capital, C_{TDC}	$207,646,818
Total Permanent Investment	$232,564,436

Table 17.5: Components of the Total Permanent Investment for the selected process configuration. A site factor of 1 is taken, as the plant will be located in the Gulf Coast (Florida). All costs in the right column are summed to give the Total Depreciable Capital.

18 OPERATING COST – COST OF MANUFACTURE

18.1 Raw Materials

The singular feedstock used for this process is process water. For a later sensitivity analysis done on SMR versus electrolysis, the cost of Natural Gas is also included. Both unit costs are in the table below.

Raw Material	***Cost (USD/1000 gal.)***	***Annual Cost (USD)***
Process Water	$0.80	$28,046

Table 18.1: Unit cost and total annual purchase cost of feedstock water.

Raw Material	***Cost (USD/1000 SCF)***	***Annual Cost (USD)***
Natural Gas	$7.40	---

Table 18.2: Unit cost of Natural Gas.

18.2 Product Revenues

Product	***Sales Price ($/kg)***	***Annual Revenue (USD)***
Liquid H_2	$12	$178,200,000
Pure O_2	*$0.044*	*$5,196,604*

Table 18.3: Oxygen is italicized because it can be sold if the plant is located near a short oxygen pipeline. Since our team did not have enough information on the locations of short oxygen pipelines in the Florida Gulf Coast, potential revenue generated from pure oxygen was not factored into the profitability analysis.

Electrolysis of feedstock process water creates byproduct oxygen, in addition to Hydrogen that is further liquefied. The ability to sell this oxygen is dependent on proximity to short oxygen pipelines, which are prevalent in Gulf Coast. Information on the location of short oxygen pipelines in the Florida Gulf Coast is not known, so potential revenue generated from the sale of pure O_2 was not factored into the profitability calculation.

18.3 Utility Costs

Utility prices were given in Chapter 17 of Seider et al. (*Product and Process Design)*. These costs are outlined below. Literature suggests that the cost of green electricity might even be underestimated. The cost of regular process electricity is provided for reference.

Utility	*Unit Price ($/kW-hr)*	*Upstream Annual Cost (USD)*	*In-process Annual Cost (USD)*
"Green" electricity	$0.10	$73,110,902	$7,645,188
Standard process	$0.07	---	---

Table 18.4: Unit prices of "Green" grid electricity and standard grid electricity, Upstream annual electricity cost, and In-process annual electricity cost.

18.4 Other Variable Costs

Heuristics from *Product and Process Design Principles (*Seider et al.) were used to develop costs for the following general expenses. These costs are taken at 100% plant capacity.

Selling/Transfer Expenses	3.00% of Sales	$5,791,500
Direct Research	4.80% of Sales	$9,266,400
Allocated Research	0.50% of Sales	$965,250
Administrative Expense	2.00% of Sales	$3,861,000
Management Incentive Compensation	1.25% of Sales	$2,413,125
Total General Expenses	---	$22,297,275

Table 18.5: Components of Total General Expenses. All costs in the right column are summed to give Total General Expenses.

	Price per unit ($/kg)	**Total Annual Cost (USD)**
Raw Materials	$0.00188	$28,046
Byproducts	---	---
Utilities	$5.837	$86,685,592
Total General Expenses	----	$22,297,275
Total Variable Cost	----	$109,010,913

Table 18.6: Components of Total Variable Cost. All costs in the right column are summed to give Total Variable Cost.

18.5 Fixed Cost

Similar to the methods above, fixed costs were mostly kept as default values in the spreadsheet. The process will require 1 remote operator per shift, earning $40/hr, for 5 shifts. There will also be 0.3 (a fraction of) a technical assistance engineer and 0.3 (a fraction of) a control laboratory engineer, each earning a salary of $200k per year, including benefits. This is a convention in cryogenics plants, where oftentimes, 1 operator and 1 engineer remotely monitor multiple plants. Table 18.5 shows the fixed costs of the plant.

18.6 Working Capital

The table shows the working capital required in the first three years of plant operations.

	2024($)	***2025 ($)***	***2026 ($)***
Accounts Receivable	$7,933,562	$3,966,781	$3,996,781
Cash Reserves	$4,535,458	$2,267,729	$2,267,729
Accounts Payable	$ (3,563,574)	$(1,781,787)	$(1,781,787)
Liquid Hydrogen Inventory	$1,851,164	$925,582	$925,582
Raw Materials	$77	$38	$38
Total	$10,756,687	$5,378,343	$5,378,343
Present Value at 15%	$9,353,641	$4,066,800	$3,536,348

Table 18.7: Working Capital available in 2024(Construction), 2025 (Production, Year 1, 50% capacity), and 2026 (Production, Year 2, 75% capacity).

18.7 Total Capital Investment

Refer to tables 17.2, 17.3, 17.4, and 17.5 for the calculation of Total Bare Module Cost, Direct Permanent Investment, Total Depreciable Capital, and Total Permanent Investment. Total Capital Investment is calculated by adding calculated Total Permanent Investment to the Present Value of Working Capital at 15% in Years 2024, 2025, and 2026. The Present Value of Working Capital at 15% in Years 2024-2026 is provided in Table 18.7.

$$\text{Total Capital Investment} = \$232{,}564{,}436 + \$9{,}353{,}641 + \$4{,}066{,}800 + \$3{,}536{,}348$$

$$\text{Total Capital Investment} = \$249{,}521{,}225$$

19 OTHER IMPORTANT CONSIDERATIONS

19.1 Environmental Factors

Environmental considerations are limited for this process, as it relies primarily on electricity from the grid that is purchased at a “green” premium. Furthermore, green Hydrogen liquefaction does not produce any harmful byproducts. However, there are a selection of relevant environmental factors to consider:

1) **Energy and materials consumption** – Cold box construction requires perlite, a product that only forms from volcanic eruptions.
2) **Production Emissions** –Hydrogen, Oxygen, and Neon are likely to escape the process in small quantities. An in-production assessment must be performed on this plant to evaluate the leakage risks. Beyond that, Nitrogen is vented continuously.
3) **Water Availability** – Previously discussed.

19.2 High Level Startup Procedure

There are several factors that make plant start up and shutdown challenging. First, the rate of temperature change in each unit operation should be limited as the process decreases in temperature from ambient to cryogenic, or vice-versa in cool down. This is to prevent unit operations from cracking. BAHXs, due to their large surface area, thin fins, and multi-stream components, are immediately and unevenly affected by the sudden flow of an extremely high or low energy stream. Based on the guidance provided by industry experts in cryogenics, this team was advised to cool in-process BAHXs at a maximum of 5ºF per hour. At this rate of cooling, the coldest of the unit operations in this process (90ºF to -423ºF) will require 4.5 days for startup. Assuming an initial temperature of 90oF (302K), a coolant must be prepared and progressively cooled to facilitate the controlled temperature reduction of the heat exchangers. An attractive coolant option for initial precooling is creating a mixture of previously captured exhaust Nitrogen (vented at 100oF) and feed saturated Nitrogen gas at -321F (77K). The mixture may be altered throughout the duration of the startup process to provide cooling from the full range of temperatures between the two gases. However, it will only be used until the exit temperature of the first heat exchanger is reached, as cryogenic cooling to the entire process would be inefficient. Instead, once the first heat exchanger is thermally primed, the saturated nitrogen feed stream will be initiated, as well as the flow of the Neon refrigeration cycle. Simultaneously, the nitrogen fed through the hot Hydrogen stream will be purged with Hydrogen. Importantly, the 80K adsorber should be bypassed until all Nitrogen has been purged in order to not saturate the activated Carbon filter and cause a shutdown.

Once the Neon refrigeration loop begins, a challenge arises: the second heat exchanger must be cooled but liquid Nitrogen injection could result in harsh thermal gradients that the operation is unable to endure. The passages are designed and fitted for a liquid stream, so it is unclear if its contents could feasibly be replaced for startup procedures. Considering this, the cooling previously discussed in the hot Hydrogen stream must be sufficient to cool the second heat exchanger, causing an overshoot in the first heat exchanger. Once the first two heat exchangers are primed, a graduated increase in the flowrate of the Neon recycle loop offers the required cooling to prepare heat exchangers 3 and 4.

Importantly, the hot Hydrogen stream must be run above the freezing point of Nitrogen until there is confidence of near-complete Nitrogen evacuation. This will prevent accumulation and increased pressure drops in the passages of the lower heat exchangers.

Once all heat exchangers are primed, all other unit operations previously not running at production pace will ramp up within the safety limits of the individual unit operation. In the case of compressors, this may involve limiting the acceleration of motors. A comparison of similar plant procedures has yielded an expected plant startup time between 1 and 2 days.

This process is guided heavily by similar patented proposals offered in US20100281915A1. [53]

19.3 Public and Employee Safety Health & Welfare

This plant will have minimal employees as later discussed. For those operators staffing the plant on a given day, it is important to provide sufficient safety training and protective equipment to ensure operational safety. Members of the public are another concern that must be evaluated when choosing a plant location, as this report later discusses.

19.4 Global Cultural Social & Ethical Factors

As discussed in the motivation section of this report, there is an ethical imperative in this generation to mediate the acceleration of greenhouse-gas-induced climate change across the globe. Developing means of harnessing high-energy-density fuel sources from renewable and non-invasive means has been presented as an ultimatum.

Relevant Regulations & Standards

All relevant regulations and standards for a plant producing LH2 are listed below in short form. For those regulations that are particularly unique or impactful to this process, there will be a discussion in "Safety Considerations"

19.4.1 Occupational Safety and Health Administration (OSHA) regulations:

Human impact, industrial and worker standards, technical regulation.

19.4.2 Environmental Protection Agency (EPA) regulations:

Emissions, waste disposal, water utilization, environmental impact, etc.

19.4.3 National Fire Protection Association (NFPA) codes:

Hydrogen handling, storage, and production regulations

19.4.4 National Electrical Code (NEC) regulations:

All electrical regulations, pertinent to all unit operations but primarily electrolyzers.

19.4.5 Department of Transportation (DOT) regulations:

Shipping and transmitting of Liquid Hydrogen.

19.4.6 Process Safety Management (PSM) regulations:

May be a part of multiple different regulatory bodies, but relevant to the functioning of process on macroscopic scale.

19.4.7 Permitting and licensing requirements:

Local, regional, and national authorities that will vary on a site-by-site basis.

19.4.8 Startup/Shutdown

There are several factors that make starting up and shutting down this process challenging. The principal challenge is limiting the rate of temperature change in each unit operation as the process decreases in temperature from ambient to cryogenic, or vice-versa in cool down. As will be discussed in section 19.6, lowering the temperature necessarily creates temperature gradients on each unit operation employed. BAHX particularly, due to their large surface area, thin fins, and multi-stream components, are immediately and unevenly affected by the sudden flow of a given stream. Based on the guidance provided by industry experts in cryogenics, this team was advised to cool the process using the BAHX as limiting features, by which they can cool at a maximum of 5^{o}F per minute. Assuming this rate is reached, that necessitates a start-up at an absolute minimum of 8.5 hours to traverse 90^{o}F to -423^{o}F for the coldest of the unit operations in this process. There are multiple set variables and multiple degrees of freedom that may be modulated to provide the necessary cooling. Set variables include the output power of each compressor, the total volume of working fluid, and the temperature of inlet nitrogen. Degrees of freedom include the flowrate of feed Hydrogen, modulated by the current applied in the electrolyzer system and the number of electrolyzers used, the decision to keep compressors on or off, and the flow rate of nitrogen employed.

Assuming an initial temperature of 90^{o}F (302K), a coolant must be prepared and progressively cooled to facilitate the controlled temperature reduction of the heat exchangers. An attractive coolant option for initial precooling is creating a mixture of previously captured exhaust Nitrogen (vented at 100^{o}F) and feed saturated Nitrogen gas

19.5 Contingency Storage:

Many industrial chemical processes require some amount of contingency storage for their reactants, working fluids, and other transient inventory. However, with high and consistent availability to our sole reactant, water, there is no need to build infrastructure for long-term contingency inventory. With working fluid leakage rates ranging between 1-5% per year depending on the selected fluid, a preferred design choice to schedule periodic top-ups has been favored over on-hand storage to minimize capital and operational costs.

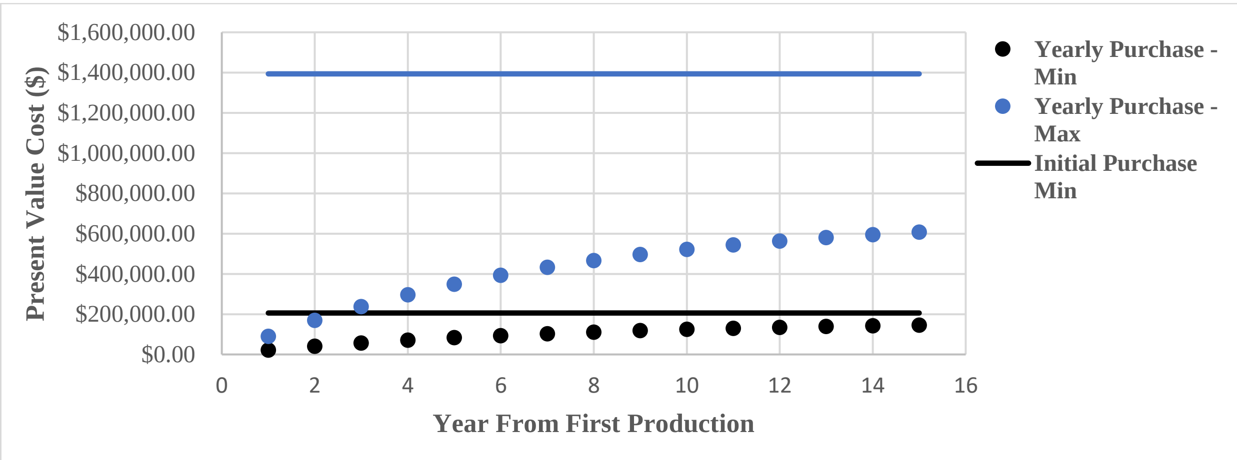


Figure 19.1: present value of cash for yearly purchase of replenishing working fluid vs. initial purchase, including escape and containment considerations

As seen in the graph above, for the anticipated 15-year lifespan of this plant, a scheduled working fluid replenishment calendar reduces the total appreciated costs, particularly as leakage rates increase. However, one will note the asymptotic behavior between one-time purchases and yearly replenishments for the lowest leakage cases. With the instability in Neon markets over the past two decades, the margins between options may be sufficiently small to warrant investment into a one-time purchase at a stable price to prevent future unexpected costs or operational comprises due to unavailability.

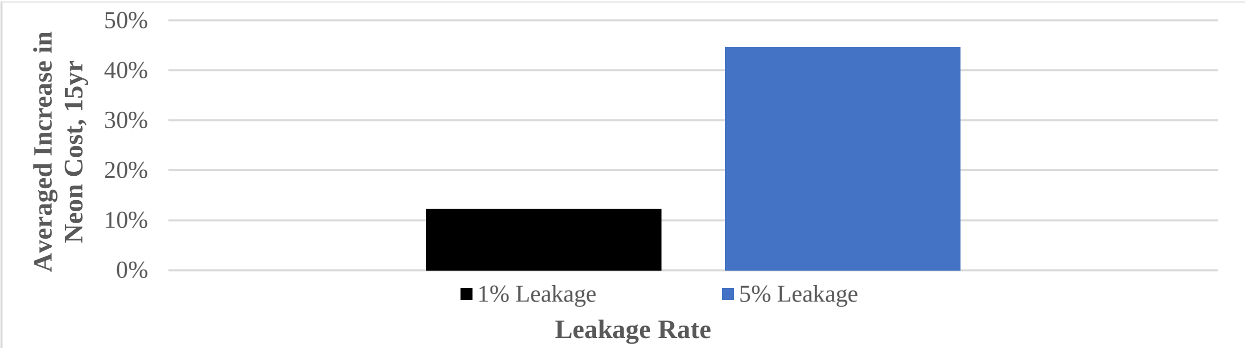


Figure 19.2: minimum average increase in Neon bulk price to justify purchasing one time investment of on-site Neon storage

Despite uncertainty of the in-conflict current bulk price of Neon as global demand exceeds global supply, realistic crisis-point cost evaluations on Neon can be extrapolated from the previous supply shortage caused by the Crimean War in 2014, at which time the price of Neon increased by 600% and lasted for over 1 calendar year. Accounting for inflation, such a temporary increase would result in a fifteen-year average bulk price increase of ~33%. Neon prices fluctuate pseudo-linearly with steel prices as the noble gas is a byproduct of steel refining. Any long-term conflict that threatened permanent access to steel and Neon would result in new or increased steel production in unafflicted regions. As such, a permanent step increase in the market price of Neon is not expected, but bulk cost restoration could take upwards of 5 years to account for the design and installation of new production plants.

The last contingency reservoir to be required by this process is of consumer-ready LH2. During national holidays, equipment malfunctions and repairs, and other production-threatening events, there must be a reservoir of sellable product to maintain supply to consumers. With projected primary clients of government-sector, low-error-tolerance entities such as NASA and the US Air Force, as well as potential commercial aviation companies with needs for consistently priced and available fuel, shipment lapses could result in contract terminations. With such niched clients and fragile contract threats, significant lapses are considered to be the paramount threat to long-term profitability and must be eliminated.

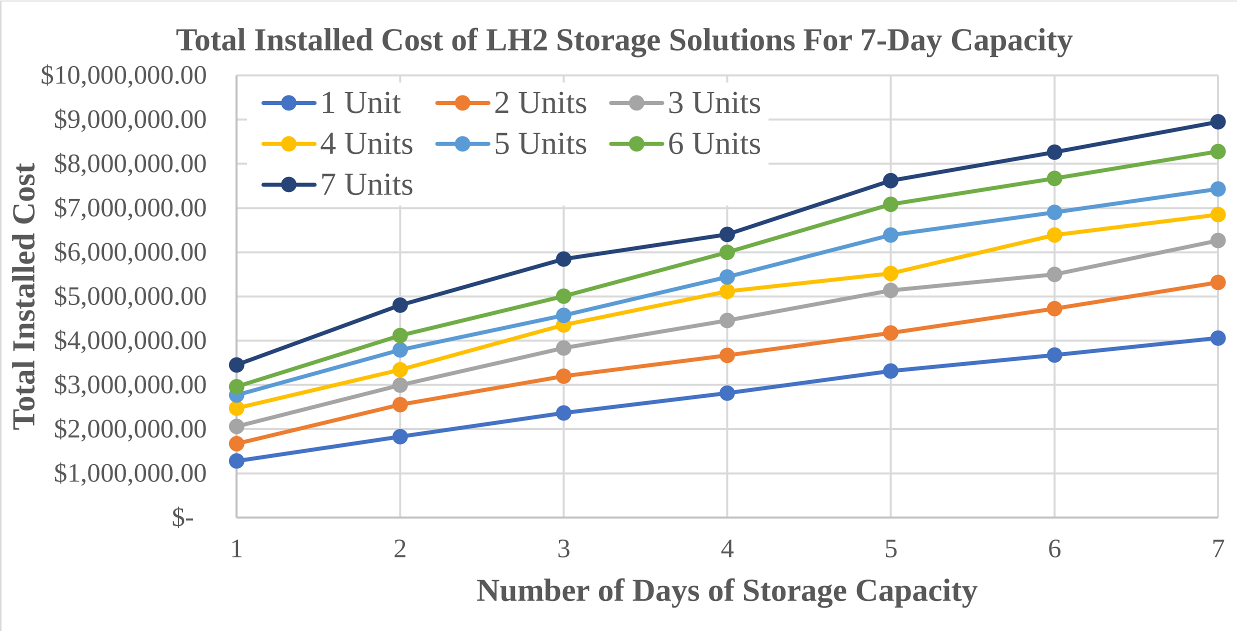


Figure 19.3: cost comparison of total installed cost of onsite LH2 storage systems, varying days of production capacity and number of storage tanks

While there is a need for capacity of up to 7 days of production for holiday events, normal operations throughout a given calendar year require significantly smaller actual storage – approximately 3 days has been suggested to this group by industrial consultants. With the margin between maximum and average operational storage capacity, certain advantages may exist in investing in several smaller units as opposed to one large storage unit. Because heat transfer is directly correlated with heat transfer area – and therefore the inner diameter a storage unit – a single larger unit at maximum capacity would evaporate stores slower than multiple smaller units. However, at average operational storage capacity, having multiple units allows engineers the flexibility to utilize only a portion of their total units, as illustrated in the figure below.

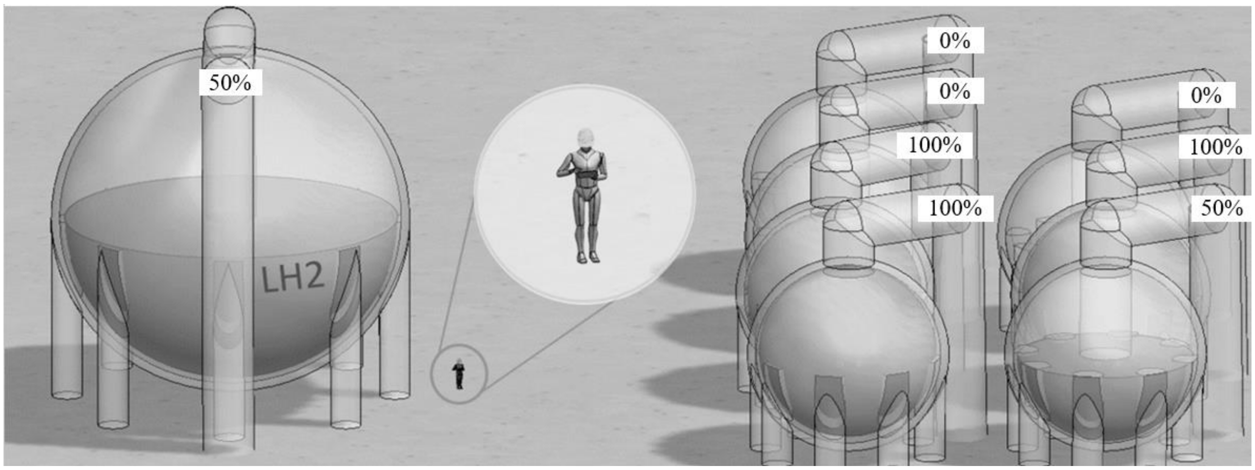


Figure 19.4: comparison of one large storage tank at half maximum capacity (left) against seven smaller storage tanks of equal volume, with three units at maximum capacity, one unit at half maximum capacity, and three units at zero capacity

Such an allocation to a portion of smaller units would reduce the total heat transfer area and therefore the amount of product evaporated. It should be noted, however, that despite the above diagram representing the configuration of smaller storage tanks that minimizes evaporative losses, any storage unit left empty will gradually equilibrate to atmospheric temperature and will require system cooling before subsequent utilization. Such a precooling operation would require LN2 off-gas cooling, following by cryogenic Hydrogen gas cooling and finally LH2 loading, a non-trivial process that could be most easily completed during a whole plant startup, but otherwise would be completed by an ad hoc procedure.

Built into the installed costs of this plant is pipework to return any evaporated material to the feed stream. With evaporative leak rates ranging between 35-45 kg/day depending on the number of units used, the contribution does not make a significant thermal sink to the 45 MTD stream and is thus neglected in the ASPEN model of this design.

The cost of evaporative losses is thus valued as the product of mass flow rate of evaporation, the SEC of the process, and the local cost of electricity. Cost projections of evaporative losses may be seen below for all of the investigated configurations.

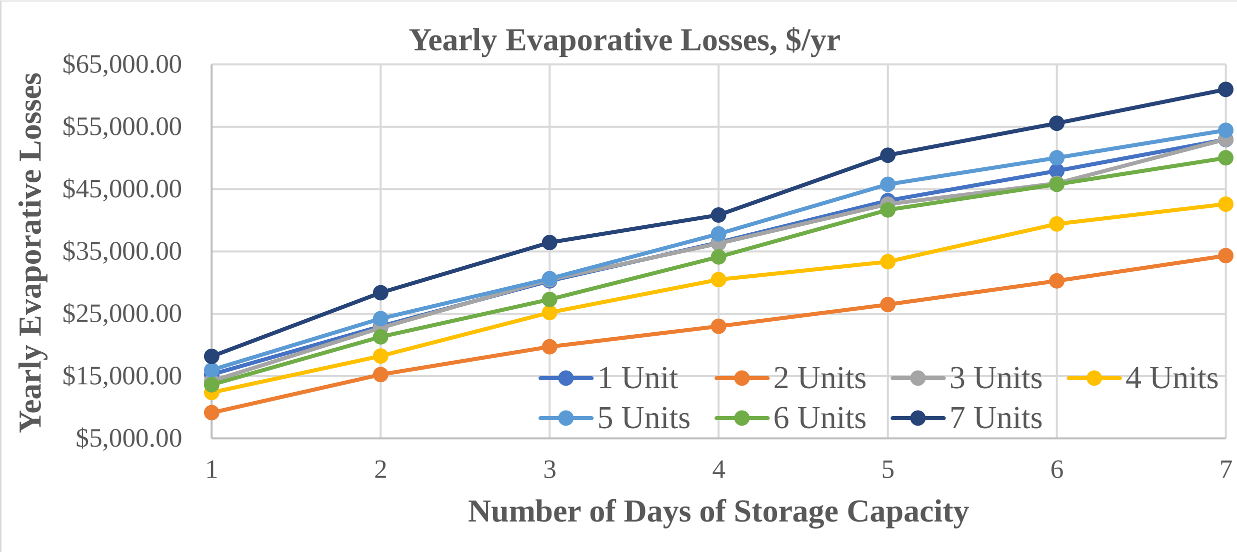


Figure 19.5: yearly evaporative losses evaluated for multiple onsite LH2 storage systems, varying days of production capacity and number of storage tanks under 50% relative maximum capacity

As one can observe, evaporative losses may be minimized by installing a pair of two storage tanks rather than just one unit. However, to properly account for the total cost considerations of operational and capital expenditures for a storage system over a 15-year lifespan, all future operational costs must be accumulated and discounted relative to their current present value.

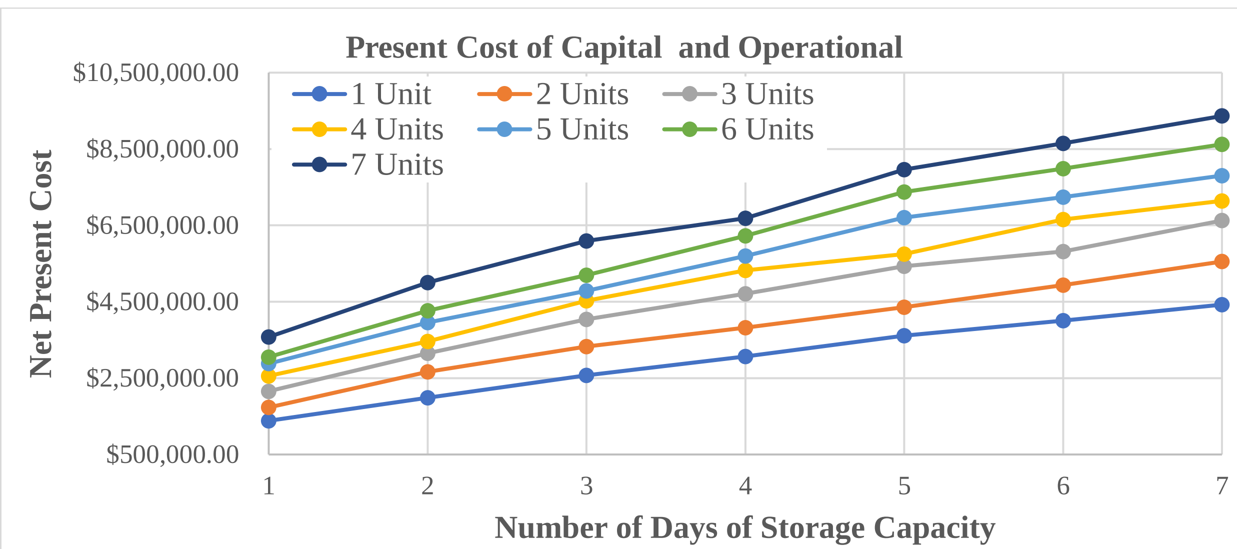


Figure 19.6: total present cost of permanent storage systems

With the costs projected, installing one unit offers approximately $1M in accumulated savings over the next most competitive design of two installed units, acknowledging that above the observed deficit, systems of smaller units would either incur higher evaporative losses or would require multiple costly precooling operations each year.

To approximate the precooling cost of a unit, an "empty" unit is defined as one at atmospheric temperature and unit design pressure filled with gaseous H_2. This assumption would not hold for the first loading of a tank in which gas composition and initial pressure would be dictated by the state of the vessel delivered by the vendor. However, for subsequent pre-cooling and loading, the described state acts as a boundary case. To precool the system is to have the unit cool from atmospheric temperature to the exit process temperature of our unit. The remainder of this subsection will offer a procedure, in great detail, on how to cool the selected cryogenic tank unit.

The cooling of a storage unit has two rate limiting considerations: gas flow rate and maximum allowable thermal stress. The temperature difference between the bulk gas and the outer surface of the aluminum shell is the driving force of thermal stress, and is limited by shell temperature, as throughout the cooling process, bulk gas is cycled to maintained to progressively reduce temperature and prevent thermal equilibrium until the shell reaches the temperature of the inlet gas flow. Flow rate is controlled by mechanical limitations such as gas flow rate through the tank's volume and gas availability on site. Thermal stress limitations of a given storage unit, on the other hand, require a materials analysis to elucidate.

First, a weak-point analysis must be performed on the aluminum shell to determine the stress limits on surface. Assuming a well-supported storage tank that distributes load away from the bottom surface of the shell and to multiple supports that are unaffected by temperature shifts. Under this description, the cross-section that bears the largest stress is at a point directly below the equator, with internal pressure, shell weight, thermal stress, and atmospheric pressure simultaneously inducing stress. Due to the curvature of the sphere, weight at such a point is projected outwards, tangent to the sphere surface. By accumulating all of the stresses, one can determine the maximum allowable induced thermal stress.

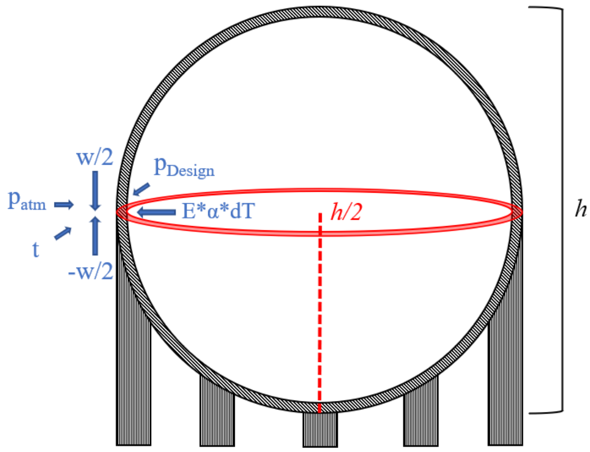


Figure 19.7: Force balance around equator of Hydrogen storage tank used for stress calculations

Using the thermal stress equation, stress curves were generated from NIST Young's Modulus and Linear Expansion correlation constants for the chosen core alloy: Aluminum 6061. By sampling temperature differences ranging from 20K-120K, a comprehensive thermal stress chart was created, displaying possible routes to shell cooling. These curves were limited by a reduction of the NIST-provided alloy yield strength, discounting material strength to account for weld-seam weaknesses, load, and a design margin of 75% to prevent deformation in severe operation changes.

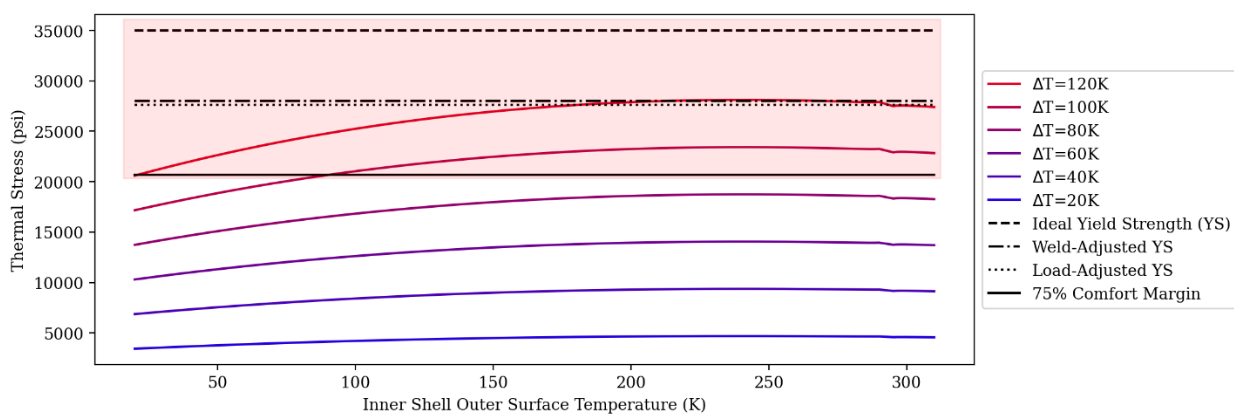


Figure 19.7: display of thermal stresses induced by various surface to surface temperature differences, assuming no axial temperature gradients. Ideal yield strength, here defined as the minimum yield strength at temperature T, and weld based strength reductions are provided by Fergusson Perforating. Weight contributions are calculated by pressure induced at the surface tangent vector immediately below sphere equator

General heuristics can be generated from the above chart. For any of the tank sizes proposed, the difference in temperature between the bulk temperature of the gas and the outer surface of the aluminum shell may not exceed 80K from shell temperatures 320K to 75K and may not exceed 100K from shell temperatures 75K to 20K. Using this limitation, a dynamic heat balance was performed to generate cooling timelines at different process settings for each tank size.

Newton's Law of Cooling and the First Law of Thermodynamics are used together to characterize the cooling rate. Both convective and conductive transfer are considered. Due to the low heat transfer coefficients of perlite and vacuum insulation, external heat transfer from the atmosphere onto the outer shell of the heat exchanger are not considered at this time scale.

To gradually reduce temperature, saturated gaseous Nitrogen is first pumped into the vessel at design pressure. Local temperature gradients induced by directed gas flow towards the surface of the shell must be avoided to prevent unsuitable stress. To prevent such gradients, cooling gas is injected at the volumetric center of the tank and processed through a static gas distributor such that coolant must first mix with ambient gas under assumed ideal mixing in every radial direction before contacting aluminum. As feed is injected, bulk gas is removed in equal molar volume for simplicity of calculation. This procedure could be changed to reduce variability in internal pressure reduction as cooling condenses bulk gas. Applying the same molar flowrate to each of 7 tank volume settings, the following concentration profiles were generated.

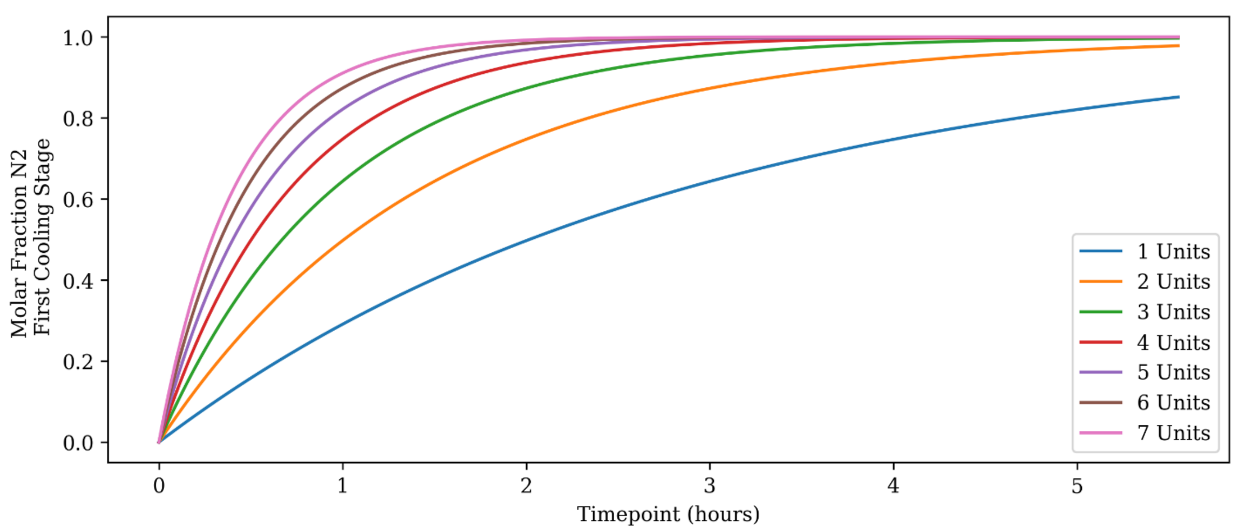


Figure 19.8: Molar fraction of Nitrogen during initial precooling of Hydrogen storage tank, sampled across tanks of 7 different volumes, corresponding to the previous 7 options in tank use. Feed is assumed to replace contents of tank equally and is held constant at an optional limit of 50 mol/sec.

With such flow rates and concentration shifts, the following temperature curves are generated. Importantly, temperature differences between gas and aluminum do not exceed the calculated maximum. In these configurations, the first stage of cooldown to approach 130K is not reached until after between 2 and 5 for the smallest and largest vessels, respectively. A curation of generated data shows the relationship between tank volume and time required to cool to the first stage.

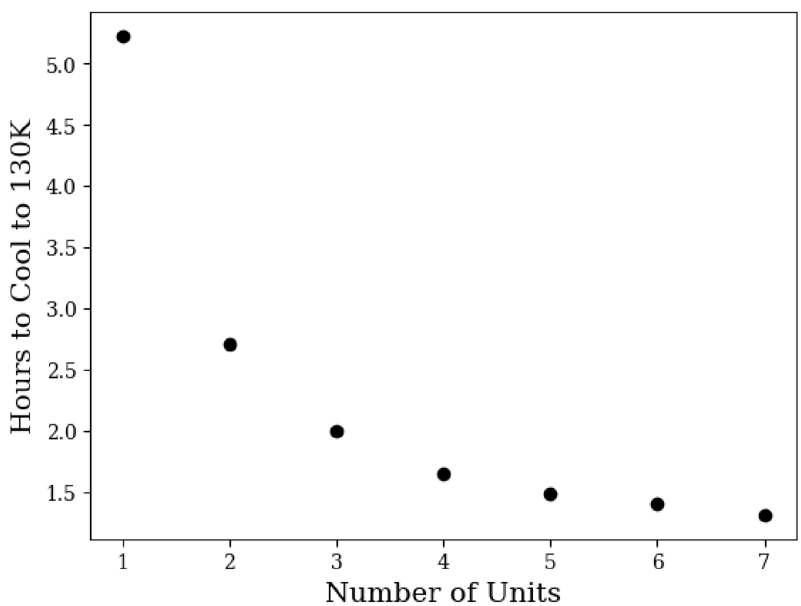


Figure 19.9: Time to cool to 130K using previously described cooling method, across 7 configurations of tank sizes.

Once the first temperature stage of 130K is reached, cooling is continued by injecting gaseous Hydrogen at 80K. Hydrogen is assumed to be 100% p-H_2 if it is the product of the liquefaction process. An important consideration in the continued cryogenic cooling of the storage unit is the molar concentration of Nitrogen left in the tank once liquid Hydrogen is added. Any residual Nitrogen present in the tank once the temperature falls below the melting point of Nitrogen (-346°F, or 63K) at operating pressure will solidify, leading to potentially dangerous impurities in storage that could have adverse effects such as abrasions on end-use processes for liquid Hydrogen, such as fuel cells. With liquid Hydrogen and solid nitrogen densities of 4.432 lb/ft^3 and 53.064 lb/ft^3, respectively, Nitrogen would either adhere to the tank walls or sink to the bottom of the tank where product is ejected. It is therefore critical to account for Nitrogen's near-complete removal.

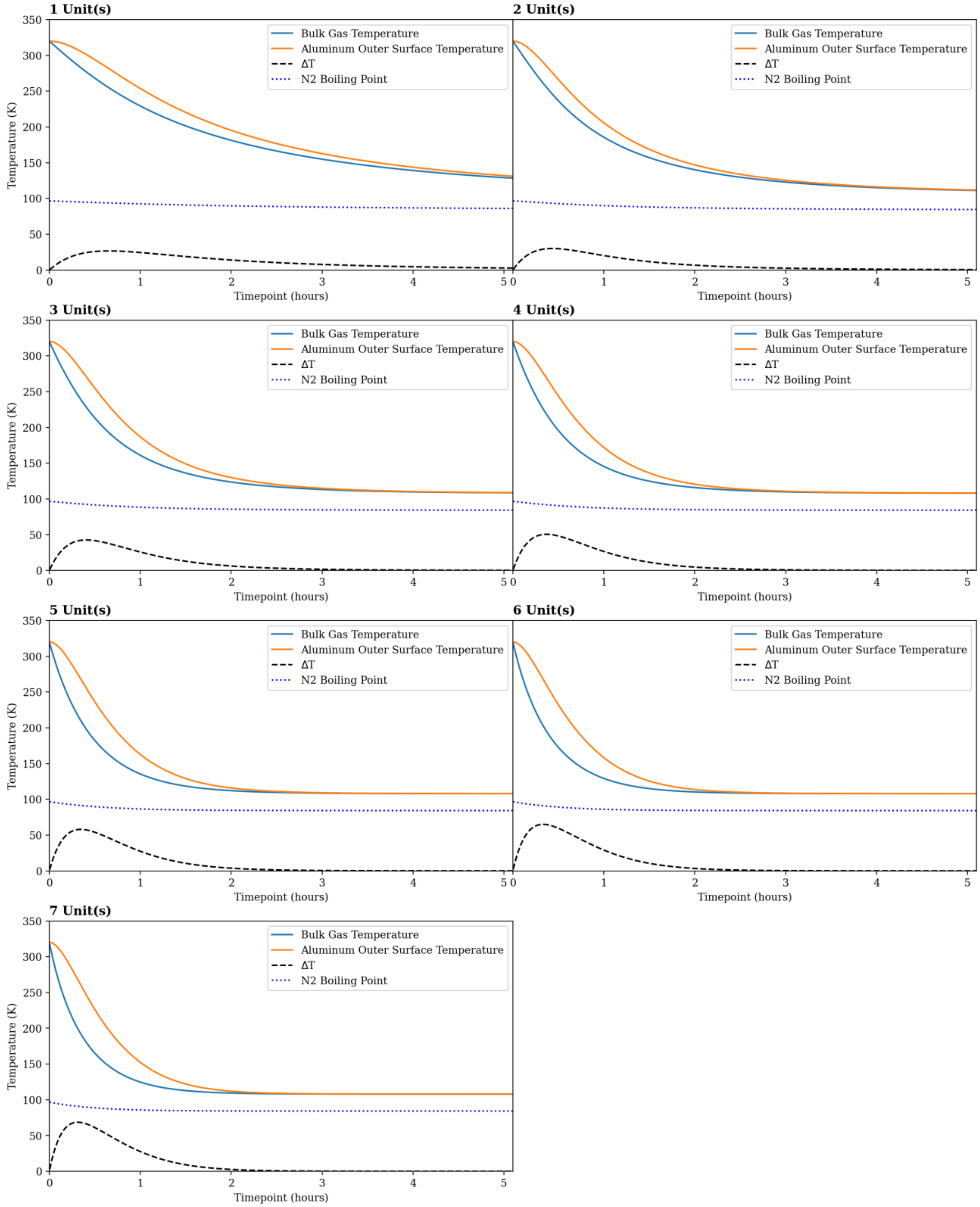


Figure 19.10: Comparison of bulk gas temperature to aluminum shell temperature, considering N_2 boiling point at operational pressure and maximum temperature difference limits.

Certain assumptions are made in the mass balance of Nitrogen replacement. Gas composition is assumed to be perfectly mixed at all times. To account for the Nitrogen cooling step in the second stage of Nitrogen purging and Hydrogen cooling, initial composition is considered to be the composition of the tank gas when shell temperature reaches 130K. In shorter words, the concentration of N_2 reached to reach 130K is the concentration at which feed will switch to H_2 to purge out N_2 impurities. It is assumed that once Nitrogen's boiling point at operating pressure is reached, no further removal of Nitrogen may be accomplished through gas purging and will remain as an impurity to be gradually purged through product release.

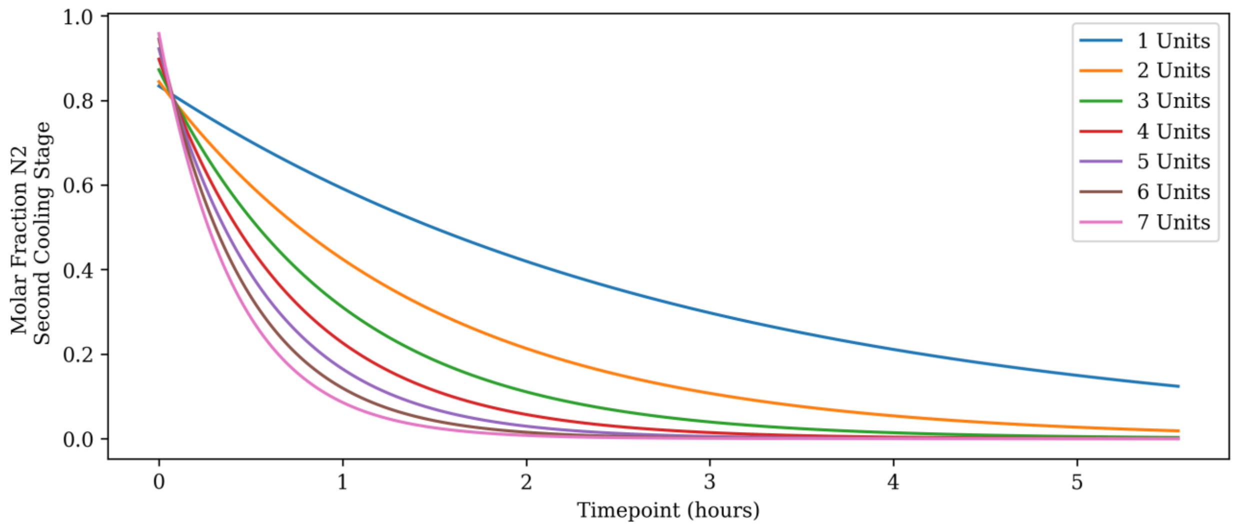


Figure 19.11: Concentration profiles of N_2 as it is purged from the tank

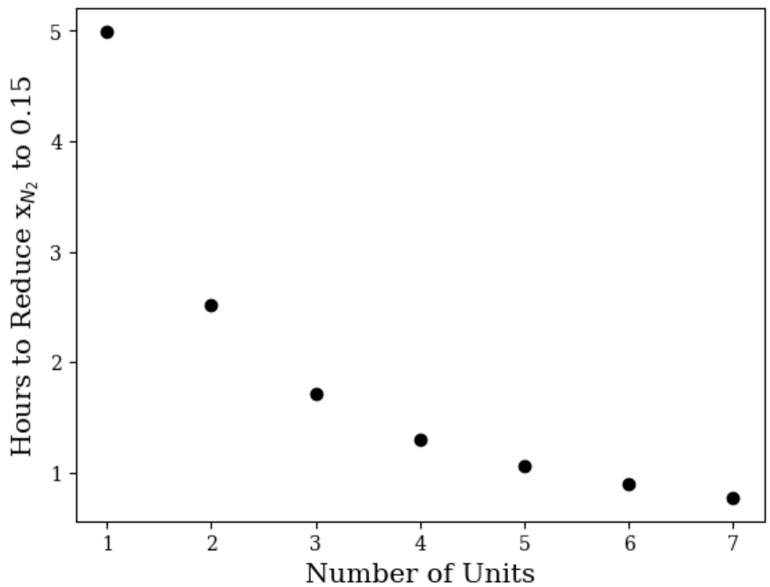


Figure 19.12: Time to reduce N_2 concentration considerably to 0.15, a benchmark chosen as an operational limit that allows for further temperature reduction in-tank.

The limiting factor in gas removal becomes the comfortability with Nitrogen impurities in the liquid product. The design engineer may choose a different tolerable Nitrogen concentration to allow for faster cooling, but the objective here was to minimize Nitrogen. Cooling rate could be further slowed and Nitrogen concentration lowered by using a warmer Hydrogen gas in this stage and then further cooling it later.

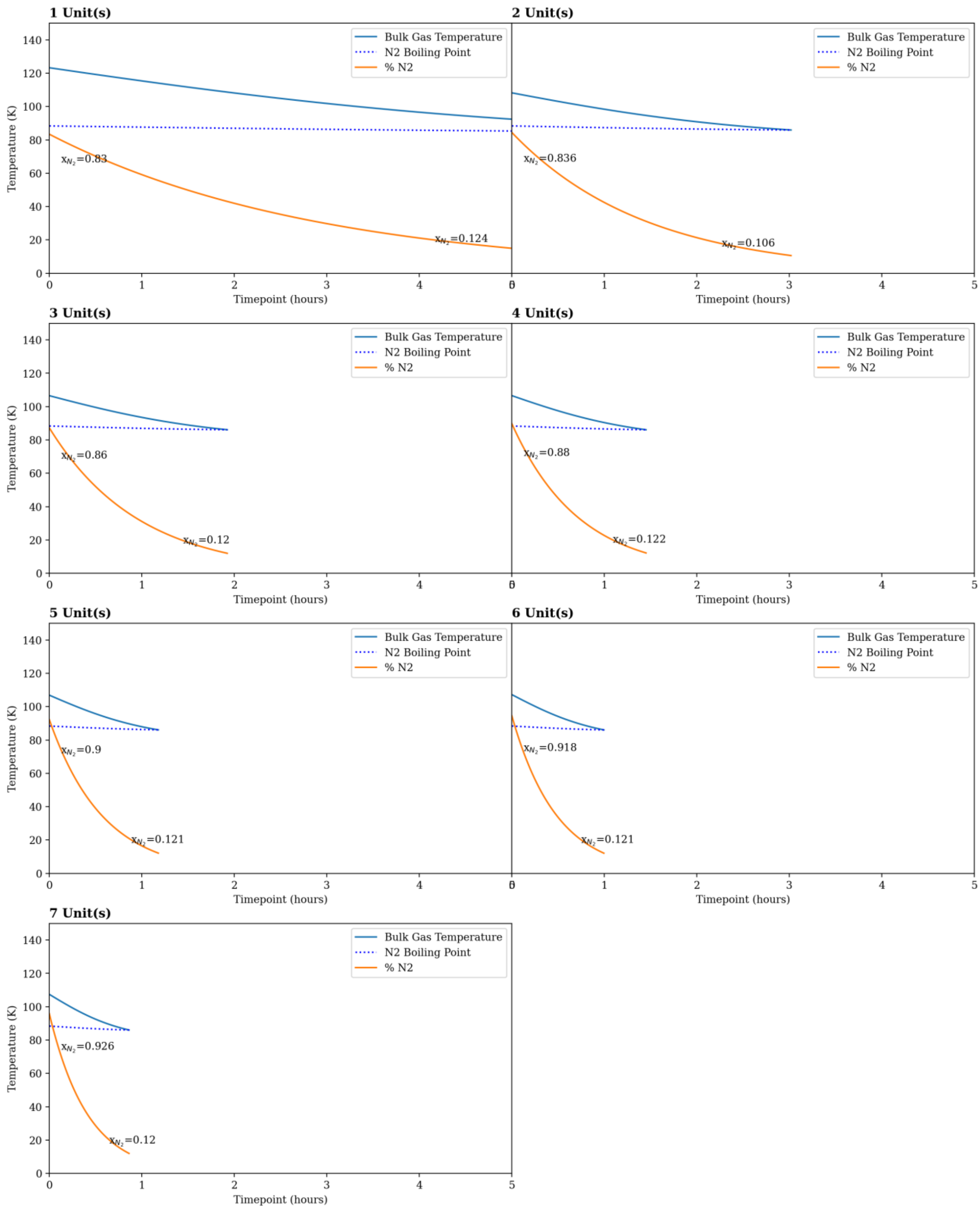
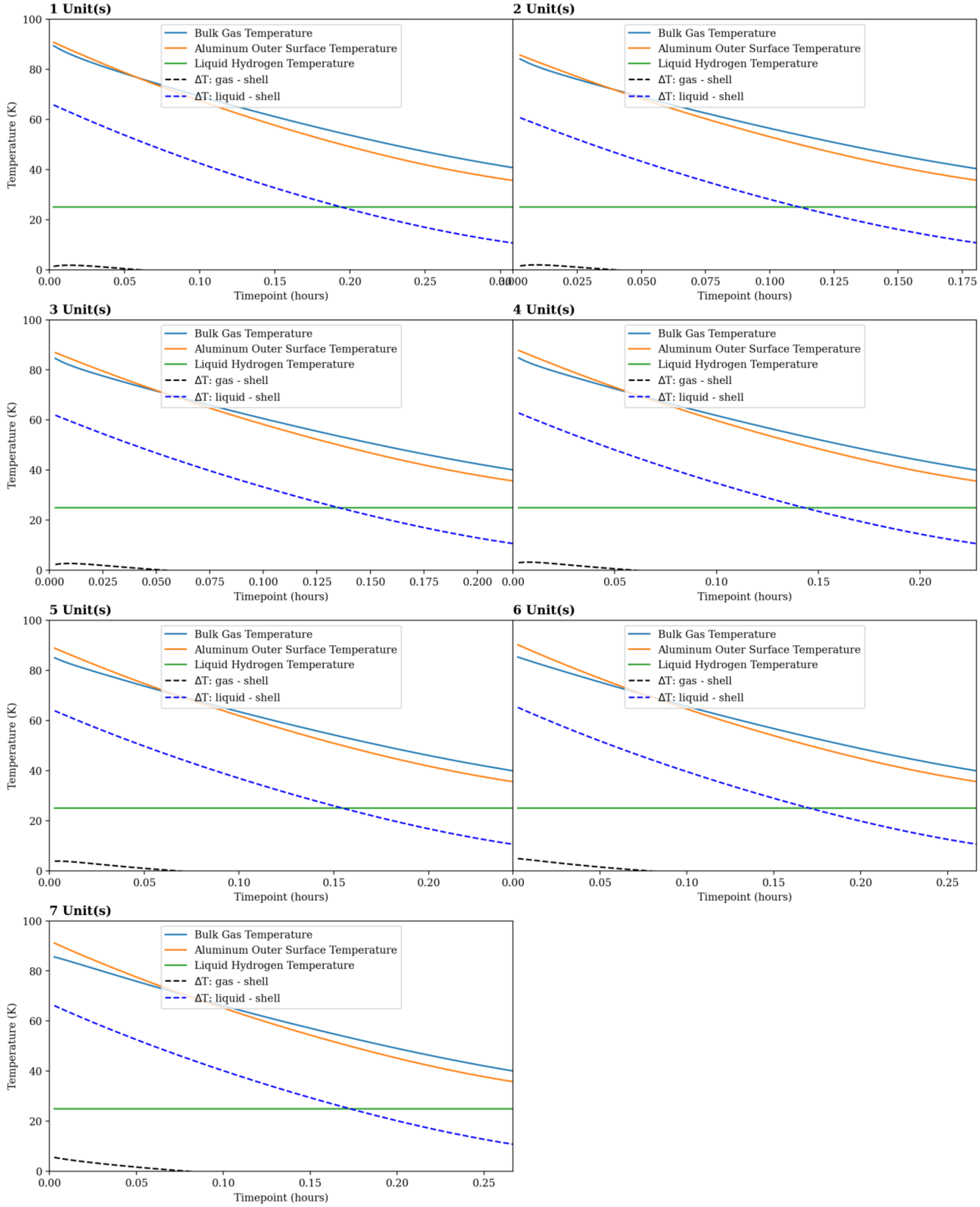


Figure 19.13:Residual gas N_2 concentration as the bulk temperature of gas approaches N_2's boiling point in the defined operational conditions.

The reader may observe that final compositions at Nitrogen's boiling point all approximate 12%. This percentage may either be increased or decreased by lowering or raising the temperature of inlet Hydrogen, depending on the purity requirements of the plant consumer.

The final step of storage pre-cooling involves loading with liquid Hydrogen. With final second stage temperatures less than 90 degrees, loading with liquid Hydrogen may occur at any point within the vessel without concern for stress-induced vessel deformation per the results detailed in figure. Liquid evaporation is considered in heat transfer and composition shifts of the bulk gas in the vessel. Hydrogen is loaded at a rate equal to the production rate of the plant, 45 MTD.



[54] Figure 19.14: Cooling curves of tank when liquid Hydrogen is added at defined flow rate.

On the whole, setting all operational costs to be equal by normalizing flow rates in each case, the total time to cool an individual tank under the previously defined constraints follows the curve below:

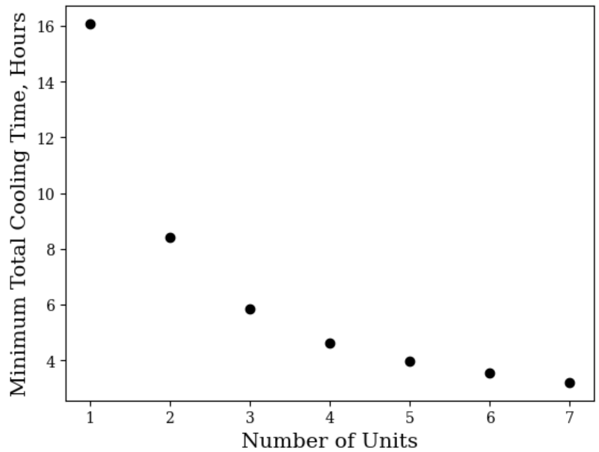


Figure 19.15: Total time to cool based on offered procedure, varying chosen tank schematic

The tank is then loaded over multiple days. For first use conditions, the tank is filled to 90% capacity rather than the steady state 50%. This is to allow initial flexibility in the production schedule to account for any variability caused by startup and/or tank preparation.

Under the described conditions, Nitrogen is considered a minor impurity. However, a simple molar breakdown is insufficient.

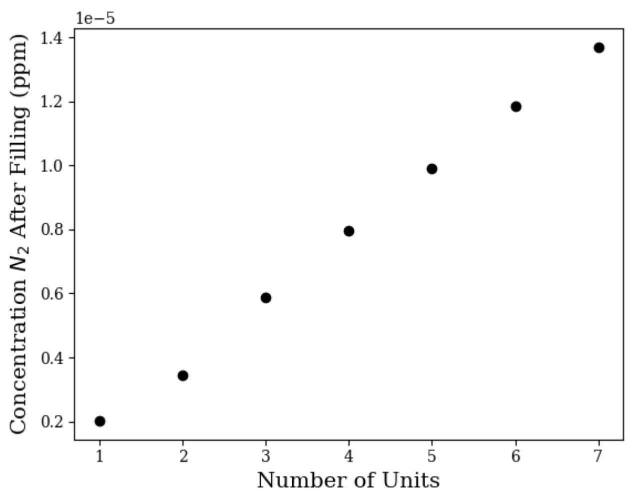


Figure 19.16: Nitrogen concentration in liquid Hydrogen tank once loaded to 90% capacity.

Instead, a spherical diffusion model is used considering an evenly caked Nitrogen solid across the surface of the shell. There is not a wealth of information on Nitrogen solubility in Liquid Nitrogen. However, there is data available for pressure ranges as low as double the operating pressure of this tank. Solubility in this plant's tank conditions is assumed to be in the area of $5x10^{-7}$. Diffusion constants are unknown, and thus a dynamic model cannot be properly generated. However, the solubility falls significantly below the purity requirements of Hydrogen consumers. As such, in-tank Nitrogen is not considered a product risk.

[55]

Spontaneous In-Unit Conversion

As long residence times in containment systems may lead to residual conversion towards p-H2 and accelerated evaporation, additional losses and final composition metrics must be tabulated for operational decision-making and consumer reports, respectively. For consistent availability of orders of various volumes, orders will be dispatched from the tank rather than the main production line. The storage tank will be fitted to load into transportation vectors, either 67,000L trucks or 110,000L cryogenic rail cars such as those offered by Chart Industries. Under steady state business operations, this will lead to an average residence time equivalent to the number of days of production stored in the system, approximately 3 days. Tank conditions are defined as the following:

1) Minimum inlet product purity of 98.5% para-Hydrogen – based on a kinetic sensitivity analysis
2) No mixing beyond thermal stratification and surface disturbance from added material, which are neglected to provide a worst-case conversion.

Literature reviews yielded little and inconsistent values for the spontaneous rate constant and activation energy. [56] However, it was found that that the process of n-H_2 approaching its equilibrium composition at 20K may take several months. Using 6 months as a basis, such a reaction is assumed as a first-order reaction in a batch reactor, using the equation:

$$k = -\frac{\ln\left(\frac{[o-H_2]_t}{[o-H_2]_0}\right)}{t} = -\frac{\ln\left(\frac{0.001}{0.75}\right)}{183\ days * 24\ hours * 3600\ seconds} \approx 5E-07\,\mathrm{sec}^{-1}$$

Applying this order of magnitude approximation to the spherical tank, we find:

$$[o-H_2]_{delivered} = [o-H_2]_{inintial} * e^{-kt} = 0.015 * e^{-k(3\ days*24\ hours*3600\ seconds)} = 0.013467$$

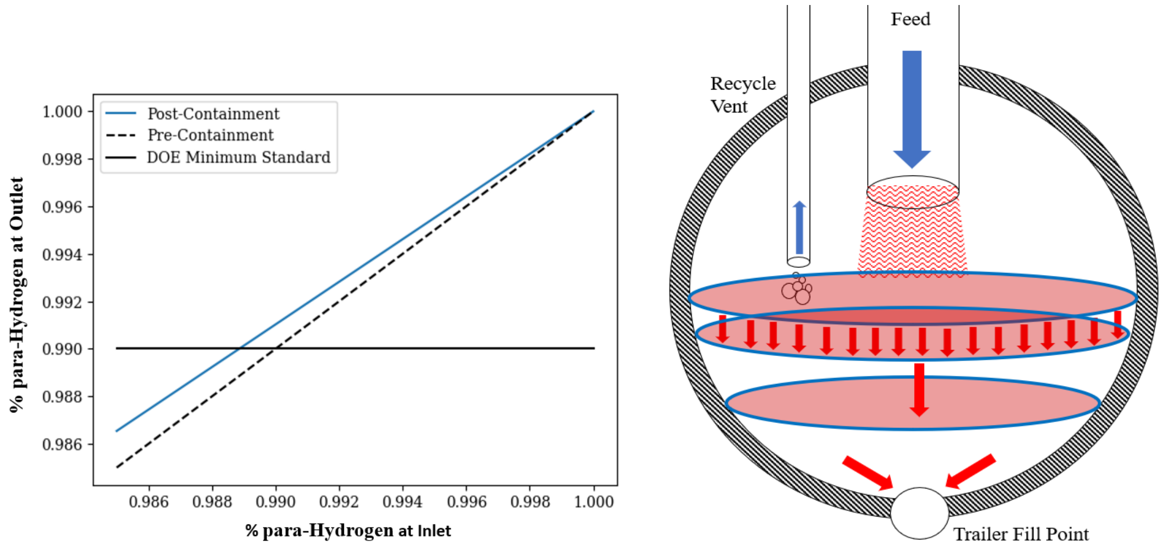


Figure 19.17: Exit composition of tank assuming spontaneous reaction over the course of 3 day residence time.

The result is a graph that relays the average change in composition of product due to spontaneous conversion within the tank. The evaporative losses of such an in-unit conversion have been detailed more explicitly in the earlier portion of this discussion. As one can observe, the minimum tank inlet composition should be ~98.5% p-H_2 to satisfy the DOE benchmark upon loading into cryogenic tankers for delivery. However, it must also be acknowledged that some non-trivial conversion will occur in transportation, leading to marginally higher purity upon delivery. However, it is best practices to report the minimum conversion value that one can validate in-house rather than projecting theoretical conversion accomplished off-site.

19.6 Transportation and Supply Chain Considerations

On top of the base production price of liquid Hydrogen is the cost of transportation. Although plant location has been carefully chosen based on its proximity to target consumers that would ideally significantly reduce supply chain costs, a sensitivity analysis is required nonetheless to bolster the need for careful plant placement.

There are 3 primary methods of transporting liquid Hydrogen domestically: by rail, by trailer, and by pipeline. The last option is primarily for very short distances and is not considered in this project. Rail and trailer transportation, however, are much more common. Rail transportation will vary greatly depending on the rail line chosen and the time of year, but trailer transportation is much more regular, outside of monthly fluctuations in fuel cost [57]. Assuming that rail transportation is cheaper than trailer transportation for large volumes of product, an analysis of supply chain costs has been produced for trailer shipped product, assuming the use of the Chart 67,000 L cryogenic tanker. For operational ease in the transportation of large volumes of liquid Hydrogen on a continual basis, a constraint on trailer use has been defined as always having 10% of tank volume filled with liquid Hydrogen. This is to ensure that the trailer never warms and is required to undergo the expensive and time-intensive pre-cooling process.

 [58]

Figure 19.18: Chart Industries 67,000 L cryogenic storage tanker

An analysis similar to that performed for the heat leak of the Hydrogen storage tank was performed for the capped cylindrical shell of the Chart trailer with specifications provided from their website, with work inserted into the Appendix. The results, when accounting for cost of labor, transportation fuel, and evaporative losses, generate the following curve:

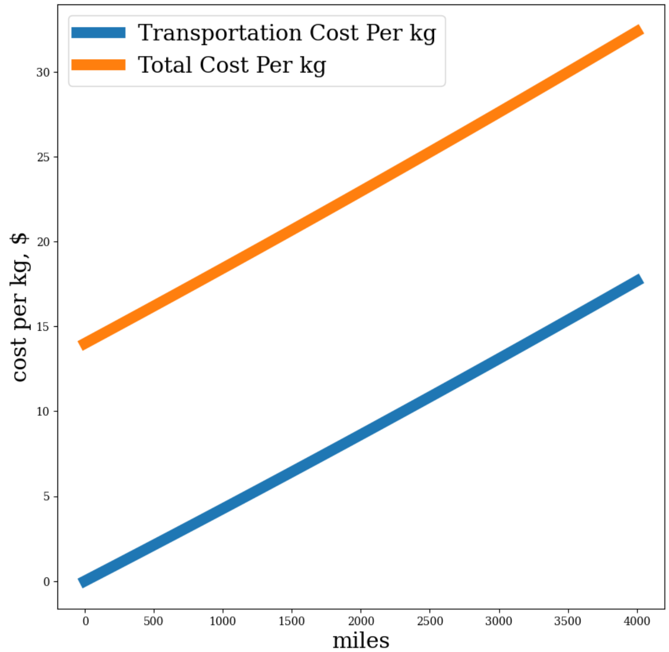


Figure 19.19: cost of transporting Hydrogen

One can observe the significant cost associated with transporting a kilogram of liquid Hydrogen. These costs figures are driven primarily by the cost of fuel to travel the long distances and the wages of the driver. Evaporative losses contribute approximately 0.5% per day in this containment configuration, which causes the slight curvature in the graph, compounding the cost of lost material onto the cost of delivered material. Sensibly, it would be preferred to ship product under 100 miles to minimize costs. Thus, our plant is designed near potential consumers.

19.7 Safety Considerations

Safety considerations are informed primarily from the Occupational Health and Safety Agency (OSHA) Standard 1910.103, Hazardous Materials, for Hydrogen. Beyond these, standard OSHA best practices are implemented but not discussed for brevity.

19.7.1 **Containers**

All storage units carrying liquid or vapor Hydrogen will be clearly labeled with “Hydrogen” per 1910.103(b)(1)(i)(c).

19.7.2 **Stationary Hydrogen Storage Unit**

Must be constructed with substantial noncombustible supports on firm noncombustible foundations per OSHA 1910.103(b)(1)(i)(b).

Will be fitted with safety pressure release valves at the top of the unit, facing upwards with unobstructed airflow per 1910.103(b)(1)(ii)(b). Pressure release valves and vents must be designed to avoid moisture freezing and buildup on the vent. Mitigation methods to prevent moisture freezing include:

(1) Proper insulation placement around vents and valves

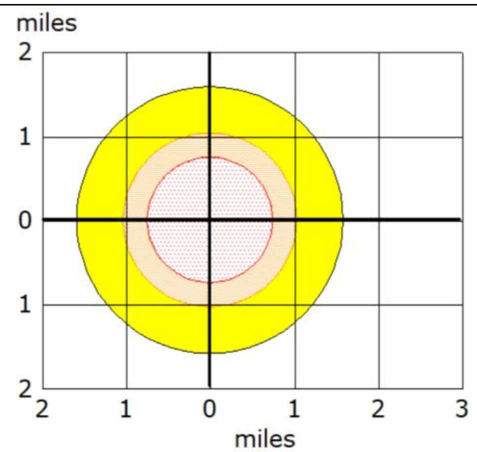
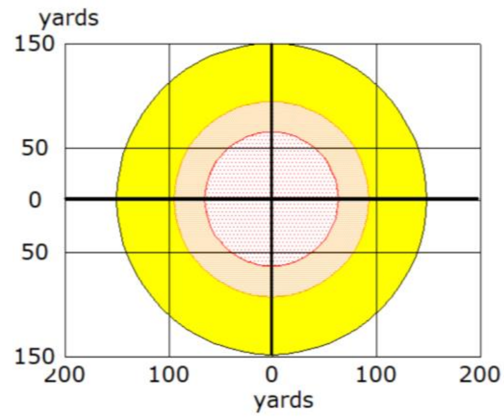
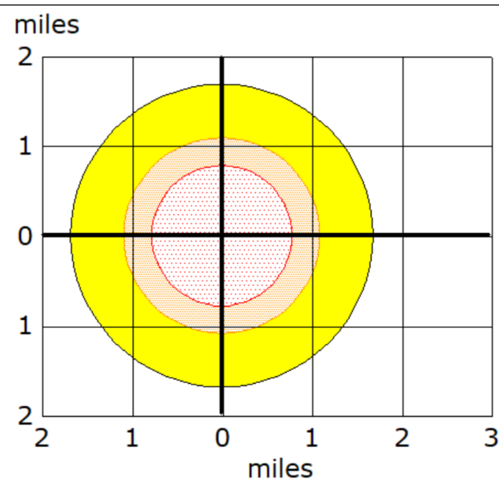
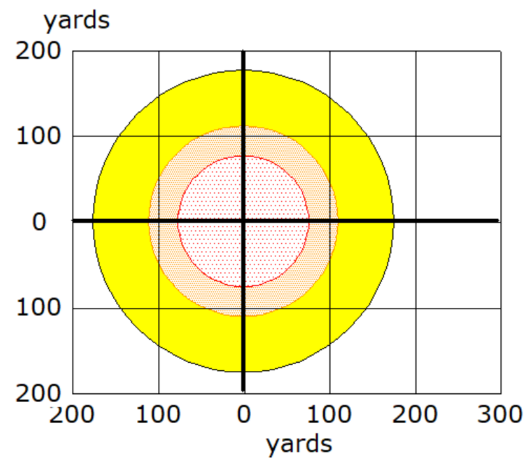
(2) A vent heater, which preheats vented gas above cryogenic temperatures

(3) An air dryer present at the vent-atmosphere interface

(4) Regular maintenance

Current NFPA 2 guidelines do not offer options for Hydrogen storage systems through generic certifications above a volume of 283,906 liters – nearly a third of a percent of the proposed storage system for this plant. However, as previously discussed, tanks of similar volume to that proposed in this project exist in small number currently and are likely to increase in appearance as Hydrogen liquefaction increases in popularity. The current procedure required by NFPA to construct a cryogenic storage vessel with a volume of our proposed dimensions is to seek Performance-Based Option certification. However, a common critique of NFPA 2's Performance-Based Option is that its language is not explicit, and leaves the step towards large storage vessel approval vague. For that reason, we propose a storage system that we see as most efficient but offer additional size proposals and cost comparisons in the event the initial volume proposal is denied. It is not expected that our smallest tank proposal would be denied approval, as it is a small fraction of the size of the NASA unit. Moreover, in 2021, the Department of Energy created a taskforce led by Shell to demonstrate the viability of producing a LH_2 storage system on the order of magnitude of 20,000 to 100,000 cubic meters, 10x-50x the volume of this proposed container. As such, while present innovation may have limited examples of storage on par with the proposed unit, there is interest and expected viability for units of significantly greater complexity.

Preliminary safety models have been generated to consider the two highest-impact risk events that could occur in the storage unit – BLEVE and dispensing valve close failure – on the largest and smallest tank models proposed. As Hydrogen is significantly less dense than air and is non-toxic, the primary concerns for plant safety are explosive and fire risks. Using ALOHA, a free EPA-provided chemical risk software, risk radii were generated for plant design decisions.

<table>
<tr><td colspan="2">Inner Circle: >10 kW/m^2 (lethal in 60 sec)
Middle Circle: >5 kW/m^2 (second degree burns in 60 sec)
Outer Circle: >2 kW/m^2 (pain in 60 sec)</td></tr>
<tr><td>BLEVE</td><td>Valve Leak</td></tr>
<tr><td colspan="2">Half-Full “Operational Volume” Tank, Largest Volume Under Normal Conditions</td></tr>
<tr><td>
</td><td>
</td></tr>
<tr><td colspan="2">Maximum Capacity Tank, Largest Volume Under Normal Conditions</td></tr>
<tr><td>
</td><td>
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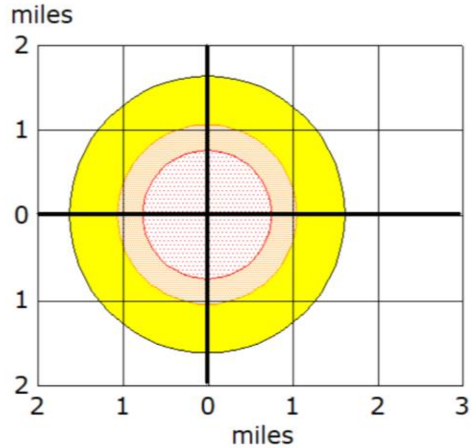
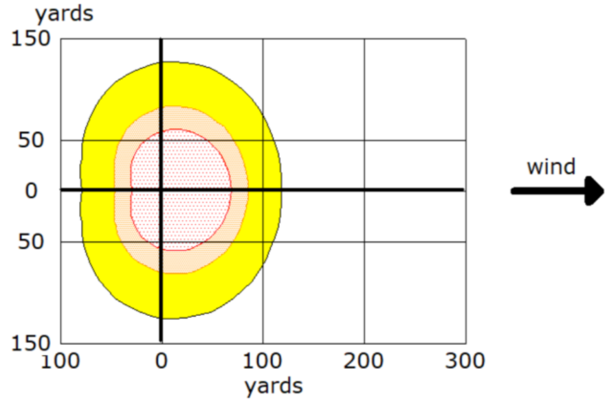
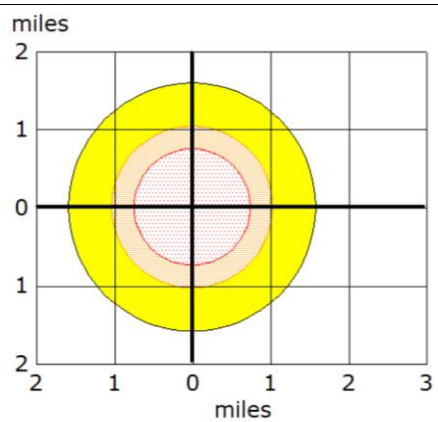
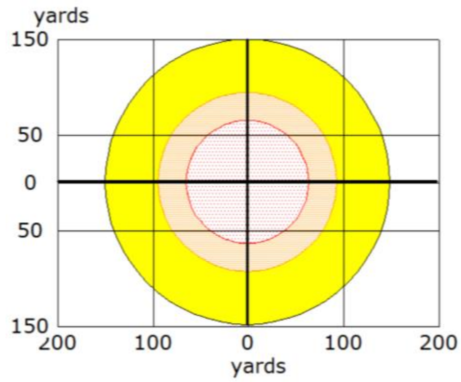
Maximum Capacity Tank, Largest Volume Under Hurricane Conditions



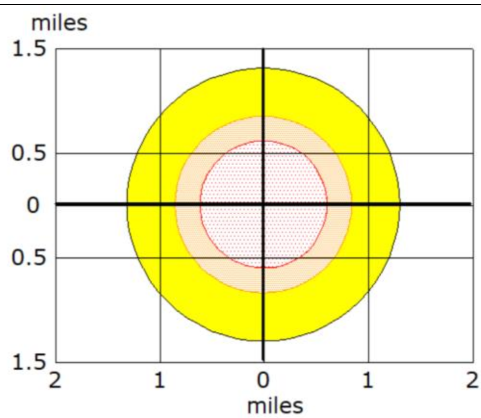


Maximum Capacity Tank, Smallest Volume Under Normal Conditions





Maximum Capacity Tank, Smallest Volume Under Hurricane Conditions



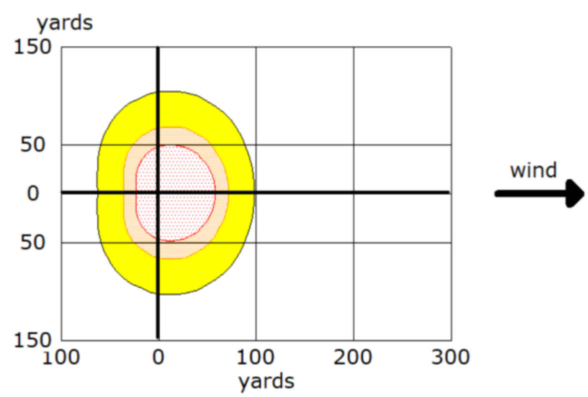


Figure 19.20 – Risk zones of Hydrogen storage leaks

Here, “Hurricane Conditions” are defined as simulation settings with westward wind speeds exceeding 150 mph, high relative humidity, and full cloud cover. “Normal Conditions” experience no wind, no cloud cover, 10% relative humidity, and 90°F.

Acknowledging the large volume of Hydrogen stored in each configuration, the plant must be designed in such a way to be able to withstand BLEVE-induced heat. As the reader will observe in the diagrams produced above, a BLEVE of any size in any condition would cause over a mile radius of 2 kW/m^2 heat induction. The main process will have to be designed to withstand prolonged exposure to heat induction of more than 10 kW/m^2. Best practices for liquid Hydrogen fires are to allow for a controlled burn until the source of Hydrogen is depleted.

To accommodate local and OSHA safety regulations, the plant will not be installed near residential or school zones and will have a separation distance of at least 1 mile from densely populated areas. This is a limitation that plant architects must consider.

With regard to separating distances from production areas, a review of currently installed plants have suggested that a combination of low risk in leakage and the use of durable design equipment (i.e. hurricane-proof Cold Box with significant perlite barrier) has led to the decision to position the storage tank immediately adjacent to the plant, though both are fairly separated from the significant electricity users – the electrolyzers.

Figure 19.21: Visual aid to demonstrate proximity of cold box and storage tank currently in use at Air Liquide.

[59]

19.7.3 **Delivery Vectors: Trailers and Rail Cars**

Must be purchased/leased from a vendor who designed, constructed, tested, and maintained equipment in accordance with U.S. Department of Transportation Specifications and Regulations per OSHA 1910.103(b)(1)(i)(a)(2).

19.7.4 **Pipes, Tubing, and Fitting**

Piping and tubing must follow rules outlined in OSHA's "Industrial Gas and Air Piping" - Code for Pressure Piping, ANSI B31.1-1967 with addenda B31.1-1969.

Pipe and tube joints must be made in the following preference order, depending on individual unit viability and design specifications: welded, brazed, flanged, threaded, socketed, compression fitted. Gaskets and thread sealants, at a minimum, must be used where applicable per OSHA 1910.103(b)(1)(iii)(c).

All tubing must have a minimum MDMT (Minimum Design Metal Temperature) of -425F and may only be used after impact testing has been performed and passed per NFPA 11.2.3.2.2.

19.7.5 **Equipment Assembly**

System installation will be supervised by engineers and personnel experienced with proper Hydrogen system construction and use practices per OSHA 1910.103(b)(1)(iv)(b)

All components of plant shall be protected against physical damage and against tampering, including hurricane proofing for the target regions of Florida and California coast per OSHA 1910.103(b)(1)(iv)(c).

Housing units, including the process cold box and other containment for rotary equipment shall be adequately ventilated per OSHA 1910.103(b)(1)(iv)(d).

19.7.6 Testing.

Between construction and first operation, all piping, tubing, rotary equipment, and welds must be proved as Hydrogen gas tight at maximum operating pressure per OSHA 1910.103(b)(1)(vi).

19.7.7 Location and Facilities

System will be above ground, underneath electrical power lines, and significantly displaced by any piping carrying flammable fluids, including Natural Gas pipelines per OSHA 1910.103(b)(2)(i)(b-d) and 1910.103(b)(2)(i)(c).

Due to the required proximity between the production site and the storage unit, certain safety design choices must be made. Any process equipment near LH2 storage must either be located on ground higher than the storage tank, unless one or many of the following plant design choices mitigate risk: diversion curbs, grading, or a barrier wall to prevent LH2 accumulation underneath the plant per 1910.103(b)(2)(i)(e). Diking may only be used for the redirection of spill flows. Due to the low temperatures involved, through placement away from storage and design of insulation, plant must be able to withstand heat released from plant deflagration or prolonged BLEVE.

All piping will remain insulated, especially those close to the temperature of LH2, to prevent the accumulation of liquid air. Asphalt and other combustible materials will not be used as construction materials near the site.

The plant and storage system will be grounded and bonded.

19.7.8 Process Controls

Pressure monitoring will occur for each stream and unit operation in the plant. Should compressor discharge pressures exceed design pressures at any point, the pump is required to shut down by NFPA standards.

However, to turn off just one compressor would lead to unsteady heating of BAHXs in the process, since each compressor is a member of a refrigeration loop that cycles through a BAHX. Unsteady heating and temperature elevation of any BAHX could lead to stress-induced fracturing, leading to gas leakage and fire risks. As such, any discharge pressure exceeding the design pressure of its compressor will lead to the complete controlled shutdown of the plant, by which all feed is paused and gas flow rates halt. There will be an emergency shutoff at the bulk source of the feed in three forms: at the water inlet, the electrolyzer mains, and at a stop valve between the electrolyzer and the feed pre-compressor. Should an adverse process event occur at any of the storage tank, platform, or electrolyzer grid, the plant will be guided to a safe shutdown.

19.8 Plant Diagram

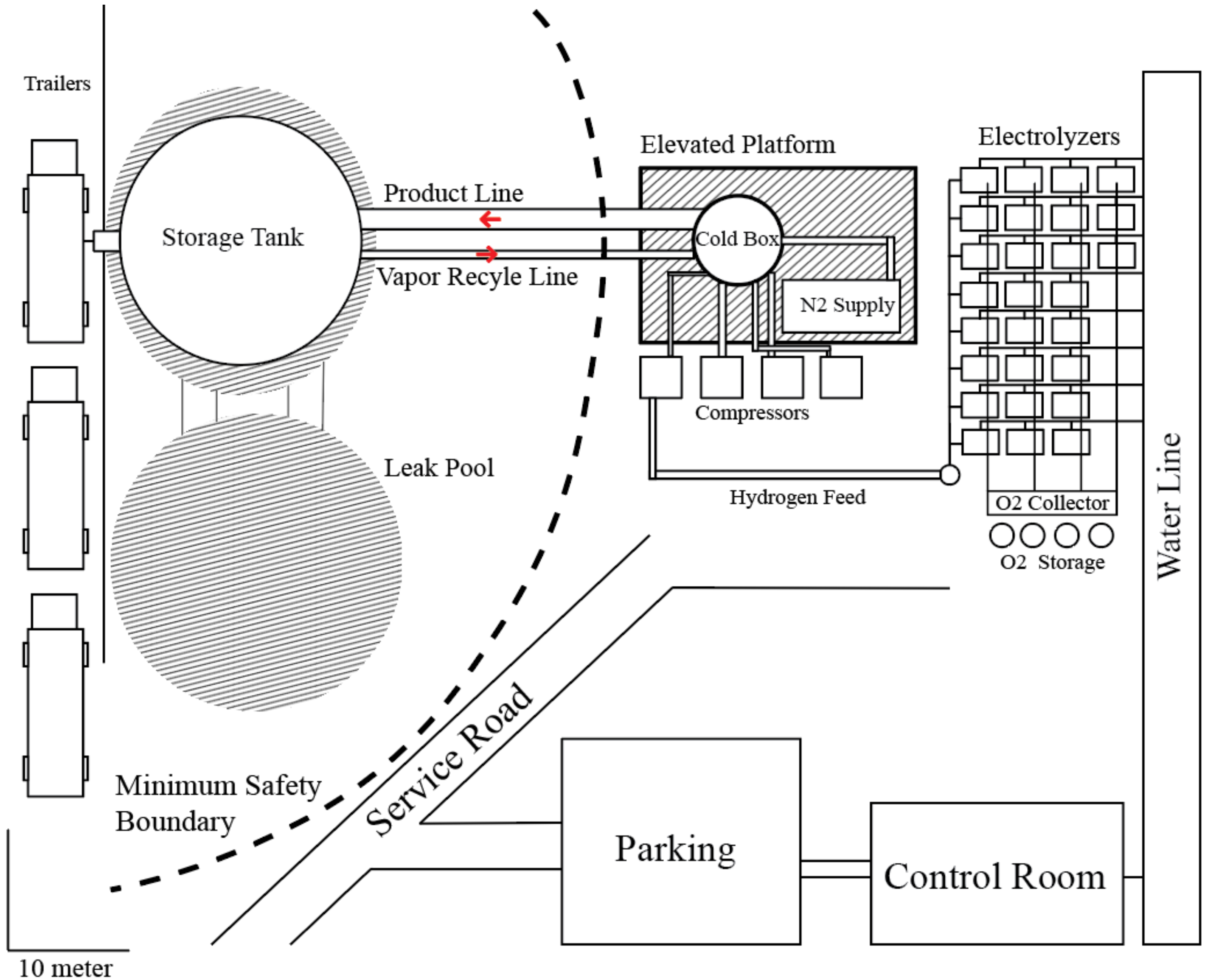


Figure 19.21: plant layout. Storage tank, platform, and electrolyzers to scale. Trucks parking and control room not necessarily to scale

The proposed plant schematic accomplishes multiple objectives set forth for the operation of the proposed process. The storage tank is positioned in a welled feature that is capable of collecting and redirecting liquid Hydrogen towards a depressed leak pool. Designing a leak pool sufficiently deep to contain the entire volume of the storage unit is considered out of scope. Instead, an ideal plant location would offer natural drainage alternatives towards areas with no impacted individuals downstream of the leak. In the event of a leakage event, the surface area of the leak pool should be sufficiently large to promote sufficient evaporation to prevent overflow. With a lower explosive limit (LEL) of approximately 5%, boiloff from the leak pool is a serious threat that must be considered. A proposed mitigation strategy is to install outdoor industrial strength dilution air blowers to blow upwards at a rate sufficient to dilute Hydrogen below its LEL.

In the event of overflow, the nearest unit operations have been placed on a raised surface. Service access roads give emergency access to each unit operation, as well as the control areas and water line. Tankers are able to drive next to storage tank for loading purposes. A future change to this plant that could increase safety would be to instead have a supply line leave from the storage tank and travel a distance to the tanker. This however, would function similarly to a dead leg when not in loading and would require either constant pressure and increased footprint. If vacated between loads, it would require precooling between each load. Therefore, loading at the edge of the storage tank is the chosen loading method. Control room and parking may be and may need to be removed significantly from current placement upon OSHA review process.

I

20 PROFITABILITY ANALYSIS – BUSINESS CASE

20.1 Approximate Profitability Analysis Guides choice of process configuration

To generate the most economically viable configuration of our Hydrogen liquefaction plant, a sensitivity analysis was performed on the choice of working fluid and the amount of liquid Hydrogen produced. Return-on-investment (ROI) calculations on the third year of plant operations were performed for all cases. For each ROI calculation, straight line depreciation was used, ASPEN PLUS UA results were used to estimate heat exchanger sizing, and ΔT_{min} approaches in all HXs were bounded between 2 ºF and 5 ºF. Heat Transfer coefficients were found from by-hand Heat Exchanger Design (Consult Sections 14.3, 14.3, and 24.1) and are provided in Table 20.1

Heat Exchanger	*Heat Transfer Coefficient, U [Btu/(lb-ft^2-R)]*
HX 1 (Block 18)	78
HX 2 (Block 22)	175
HX 3 (Block 26)	127
HX 4 (Block 30)	589

Table 20.1: Heat Transfer Coefficients used in Approximate Profitability Analysis from Heat Exchanger Design by Hand (Consult Section 14.3 and Section 14.5)

The specific profitability calculations are outlined in the Appendix section.

	Neon	*Helium*	*Hydrogen*
15 MTPD	12.36%	11.20%	8.22%
30 MTPD	14.88%	12.50%	8.10%
45 MTPD	16.59%	13.28%	8.17%
60 MTPD	16.72%	13.38%	8.89%
75 MTPD	17.43%	14.07%	9.42%
90 MTPD	17.61%	13.95%	9.19%

Table 20.1: Sensitivity Analysis on the Hydrogen liquefaction plant production capacity and Refrigerant choice. Return on Investment values calculated for the third year.

Third year ROIs for each working fluid case were approximately monotonically increasing with the plant production rate. Higher plant outputs yielded higher ROIs, reflecting an economy of scale in the equipment purchase costs. Approximate profitability analysis shows that Neon is the most economical choice of working fluid, and that higher ROIs can be generated by producing more liquid Hydrogen. Since third year ROIs plateau from 45 MTD to 90 MTD and the largest Hydrogen liquefaction plant in the world produces 45 MTD of liquid Hydrogen, we selected 45 MTD of Liquid Hydrogen produced with Neon as the choice of working fluid as the most viable process configuration. Cardella et al. demonstrate increased profitability from producing upwards of 100 MTD of liquid Hydrogen, although for a Hydrogen liquefaction process using a modified Claude refrigeration cycle.

20.2 Rigorous Profitability Analysis on Selected Process Configuration

Rigorous profitability analyses of the plant were conducted with the Profitability Analysis spreadsheet. The plant will commence production in 2025 at 50% capacity. Production will be ramped up to 75% in 2026, and reach 100% capacity in 2027.

Table 20.2 shows the profitability measures of the *Green* Hydrogen liquefaction plant, with a cost of capital of 15%. At current utilities, feedstock and product prices, capital investment, and fixed costs, the plant will have an NPV of \$44,445,400, an ROI of 16.57% in the 3rd year, and an IRR of 18.52%.

ROI Analysis (Third Production Year)

Annual Sales	193,050,000
Annual Costs	(132,688,130)
Depreciation	(18,272,920)
Income Tax	-
Net Earnings	42,088,950
Total Capital Investment	254,077,809
ROI	16.57%

Table 20.2:

Cash Flow Summary

Year	Percentage of Design Capacity	Product Unit Price	Sales	Capital Costs	Working Capital	Var Costs	Fixed Costs	Depreciation	Depletion Allowance	Taxible Income	Taxes	Net Earnings	Cash Flow	Cumulative Net Present Value at 15%
2023	0%		-	-	-	-	-	-	-	-	-	-	-	-
2024	0%		-	(232,564,400)	(10,756,700)	-	-	-	-	-	-	-	(243,321,100)	(211,583,600)
2025	50%	$13.00	96,525,000	-	(5,378,300)	(54,505,500)	(23,677,200)	(10,382,300)	-	7,960,000	-	7,960,000	12,964,000	(201,781,000)
2026	75%	$13.00	144,787,500	-	(5,378,300)	(81,758,200)	(23,677,200)	(19,726,400)	-	19,625,700	-	19,625,700	33,973,800	(179,442,700)
2027	100%	$13.00	193,050,000	-	-	(109,010,900)	(23,677,200)	(17,753,800)	-	42,608,100	-	42,608,100	60,361,900	(144,930,600)
2028	100%	$13.00	193,050,000	-	-	(109,010,900)	(23,677,200)	(15,988,800)	-	44,373,100	-	44,373,100	60,361,900	(114,920,000)
2029	100%	$13.00	193,050,000	-	-	(109,010,900)	(23,677,200)	(14,389,900)	-	45,971,900	-	45,971,900	60,361,900	(88,823,900)
2030	100%	$13.00	193,050,000	-	-	(109,010,900)	(23,677,200)	(12,936,400)	-	47,425,500	-	47,425,500	60,361,900	(66,131,700)
2031	100%	$13.00	193,050,000	-	-	(109,010,900)	(23,677,200)	(12,251,200)	-	48,110,700	-	48,110,700	60,361,900	(46,399,300)
2032	100%	$13.00	193,050,000	-	-	(109,010,900)	(23,677,200)	(12,251,200)	-	48,110,700	-	48,110,700	60,361,900	(29,240,700)
2033	100%	$13.00	193,050,000	-	-	(109,010,900)	(23,677,200)	(12,271,900)	-	48,089,900	-	48,089,900	60,361,900	(14,320,100)
2034	100%	$13.00	193,050,000	-	-	(109,010,900)	(23,677,200)	(12,251,200)	-	48,110,700	-	48,110,700	60,361,900	(1,345,800)
2035	100%	$13.00	193,050,000	-	-	(109,010,900)	(23,677,200)	(12,271,900)	-	48,089,900	-	48,089,900	60,361,900	9,936,300
2036	100%	$13.00	193,050,000	-	-	(109,010,900)	(23,677,200)	(12,251,200)	-	48,110,700	-	48,110,700	60,361,900	19,746,800
2037	100%	$13.00	193,050,000	-	-	(109,010,900)	(23,677,200)	(12,271,900)	-	48,089,900	-	48,089,900	60,361,900	28,277,700
2038	100%	$13.00	193,050,000	-	-	(109,010,900)	(23,677,200)	(12,251,200)	-	48,110,700	-	48,110,700	60,361,900	35,695,800
2039	100%	$13.00	193,050,000	-	21,513,400	(109,010,900)	(23,677,200)	(12,271,900)	-	48,089,900	-	48,089,900	81,875,200	44,445,400

Table 20.3: cash flows for 15 years of plant operations and initial design and construction years

Product Price	Total Permanent Investment										
	$116,282,218	$139,538,662	$162,795,105	$186,051,549	$209,307,993	***$232,564,436***	$255,820,880	$279,077,323	$302,333,767	$325,590,211	$348,846,654
$6.50	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR
$7.80	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR	Negative IRR
$9.10	5.21%	1.01%	-2.69%	-6.16%	-9.58%	-13.16%	-17.19%	-22.14%	-28.98%	Negative IRR	Negative IRR
$10.40	19.68%	14.91%	11.06%	7.81%	4.98%	2.43%	0.08%	-2.13%	-4.26%	-6.35%	-8.44%
$11.70	31.11%	25.35%	20.85%	17.17%	14.07%	11.38%	9.00%	6.85%	4.88%	3.05%	1.33%
$13.00	41.31%	34.49%	29.23%	25.01%	21.51%	18.52%	15.92%	13.62%	11.56%	9.68%	7.94%
$14.30	50.79%	42.91%	36.89%	32.09%	28.14%	24.80%	21.94%	19.42%	17.19%	15.18%	13.35%
$15.60	59.79%	50.87%	44.07%	38.68%	34.28%	30.58%	27.42%	24.66%	22.23%	20.06%	18.11%
$16.90	68.41%	58.47%	50.92%	44.95%	40.08%	36.01%	32.54%	29.54%	26.91%	24.56%	22.45%
$18.20	76.72%	65.80%	57.51%	50.96%	45.63%	41.19%	37.42%	34.17%	31.31%	28.79%	26.52%
$19.50	84.78%	72.90%	63.88%	56.77%	50.99%	46.18%	42.11%	38.59%	35.52%	32.81%	30.39%

Table 20.4: sensitivity analysis of liquid Hydrogen sales price

20.3 Additional Sensitivity Analyses

Green Hydrogen liquefaction is sensitive to electrolyzer price, efficiency, and heat exchanger minimum approach temperature.

20.3.1 Electrolyzer Cost

Electrolysis is required to create Hydrogen feedstock without emitting Carbon Dioxide. Unfortunately, our team was unable to get detailed information on a purchase price for our electrolyzers. Our selected vendor, McPhy, did not respond to any of our email requests for information. To complete capital cost calculations, we used a number from a previous CBE senior design report (2019) and scaled up the price using the 2023 CE index to give a purchase price of ~$2.2M per electrolyzer.

We anticipate that McPhy's 800 Nm^3/hr electrolyzer uses stainless steel as the material of construction and can be used off the rack, with minimal alterations. For this reason, we used a bare module factor of 1.5 for our profitability analysis. However, the choice of bare module factor is likely more complex and is dependent on whether McPhy's unit is skid-mounted, whether it includes rectifiers, local piping, and/or controls. For this reason, we performed a sensitivity analysis on the electrolyzer bare module factor. As the figures below indicate, a mere doubling of the bare module factor could take a green Hydrogen liquefaction process from having a reasonable ROI to being unprofitable. Subsequent green Hydrogen liquefaction process designs that use electrolyzers should confirm exactly what features are included in intended electrolyzers.

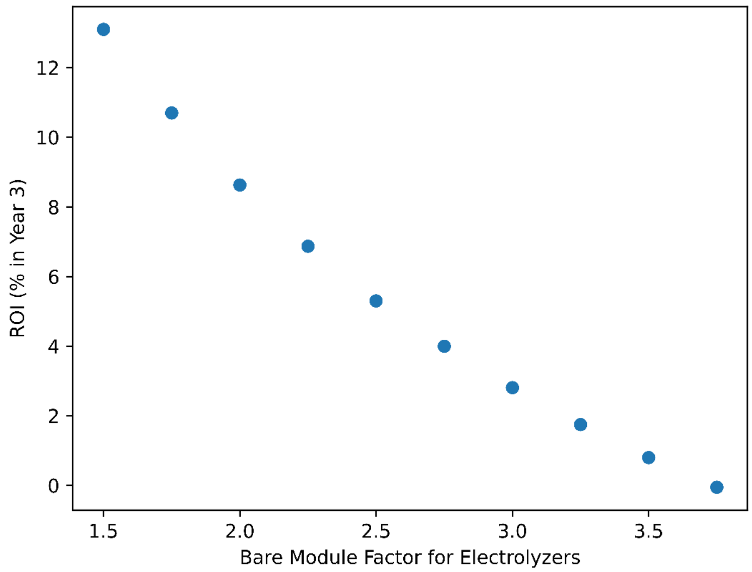


Figure 20.1: Year 3 ROI as a function of electrolyzer bare module factor.

20.3.2 Electrolyzer Efficiency

Electrolyzers are already quite efficient. However, due to the magnitude of green process electricity required to produce our Hydrogen feedstock, incremental improvements to electrolyzer efficiency have significant consequences for the profitability of our process. Our process assumes a conservative efficiency of 80%. We anticipate that as the next generation of low cost, high efficiency electrolyzers is available for purchase, profitability of green Hydrogen liquefaction processes will skyrocket. Alternatively, green Hydrogen liquefaction plants can outsource Hydrogen production to large scale electrolyzer plants. Currently, large PEM electrolyzers with efficiencies close to 95% that could handle the capacity that our plant is producing exist. Economies of scale allow large scale electrolyzer plants to produce Hydrogen at a cheaper unit price than our plant will be able to.

20.3.3 Minimum Approach Temperature

In cryogenics, BAHX minimum approach temperature is generally constrained to under 5 ºF.

Consequently, cryogenic BAHXs have enormous effective heat transfer areas. There is a tradeoff between energy efficiency and capital cost of heat exchangers, related to exchanger sizing, that is worth exploring in order to understand profitability sensitivities.

Minimum approach temperatures and resulting BAHX areas from our process configuration are listed below, in Table 20.1.3.1. As shown in Figure 16.1, Heat Exchangers comprise 18% of the total bare module cost of In-Process equipment. As the minimum approach temperature is relaxed, the relative fraction of BAHX capital costs should decrease. Conversely, as the minimum approach temperature is tightened, the relative fraction of BAHX capital costs should increase.

	Minimum Approach Temperature (ºF)	***Area (ft^2)***
HX 1 (Block 18)	11.3	15989
HX 2 (Block 22)	10.6	317
HX 3 (Block 26)	1.8	9443
HX 4 (Block 30)	2.0	328

Table 20.1.3.1: Minimum approach temperatures and areas of all process BAHXs

20.4 Comparison of Energy Efficiencies of Working Fluids

The minimum work required to liquefy Hydrogen is the difference in specific exergy between the liquid Hydrogen product and the inlet feed. For the inlet feed and product specifications of this process, this comes out to 2.8 kWh per kg LH2. The exergy efficiency is defined as the minimum work required divided by the specific power, or the actual work required to liquefy the Hydrogen, which is calculated by summing the power requirements of the compressor in the process and subtracting the electricity generated by the turboexpanders [44].

$$\eta_{ex} = \frac{(h_{product} - h_{feed}) - T_0(s_{product} - s_{feed})}{w_{real}} \qquad 20.1$$

$$w_{real} = w_{94} + w_{12} + w_{48} - w_{84} - w_{76} - w_{34} \qquad 20.2$$

Current operating liquefiers operate at a lower exergy efficiency than is possible because they are optimized for financial profitability instead of energy sustainability. The upper limit of efficiency of theoretical process designs is 4.0 kWh per kg LH2 (figure 20.2).

In practice, today's Hydrogen liquefaction plants operate at around 10 to 14 kWh per kg of LH2. This is because these plants are optimized for economic output (return on investment) and so these plants have lower capital costs but higher operating costs and lower efficiencies. It is important to keep in mind that higher efficiencies are more sustainable because they consume less energy for every kilogram of liquid Hydrogen that is produced. Thus, for a technology that is meant to be sustainable, philosophically the energy optimized plant is more appropriately aligned with the green Hydrogen liquefaction process than the ROI optimized plant.

Ultimately, the design group optimized the selected process configuration (45 MTPD, Neon working fluid) for energy efficiency instead of economic output. This was partly for philosophical reasons and partly because the ROI objective function has more parameters to worry about than specific power. However, after conducting some sensitivity analyses (see section 12.2.1), the design group discovered a strong one-to-one correlation between the energy efficiency and the ROI of the plant, which suggested that optimizing the energy efficiency produces the same results as optimizing ROI, in the case of this particular Hydrogen liquefaction process design. The process attained a specific power of 6.24 kWh per kg LH2, corresponding to an exergy efficiency of 44.41%, in line with state-of-the-art conceptual designs.

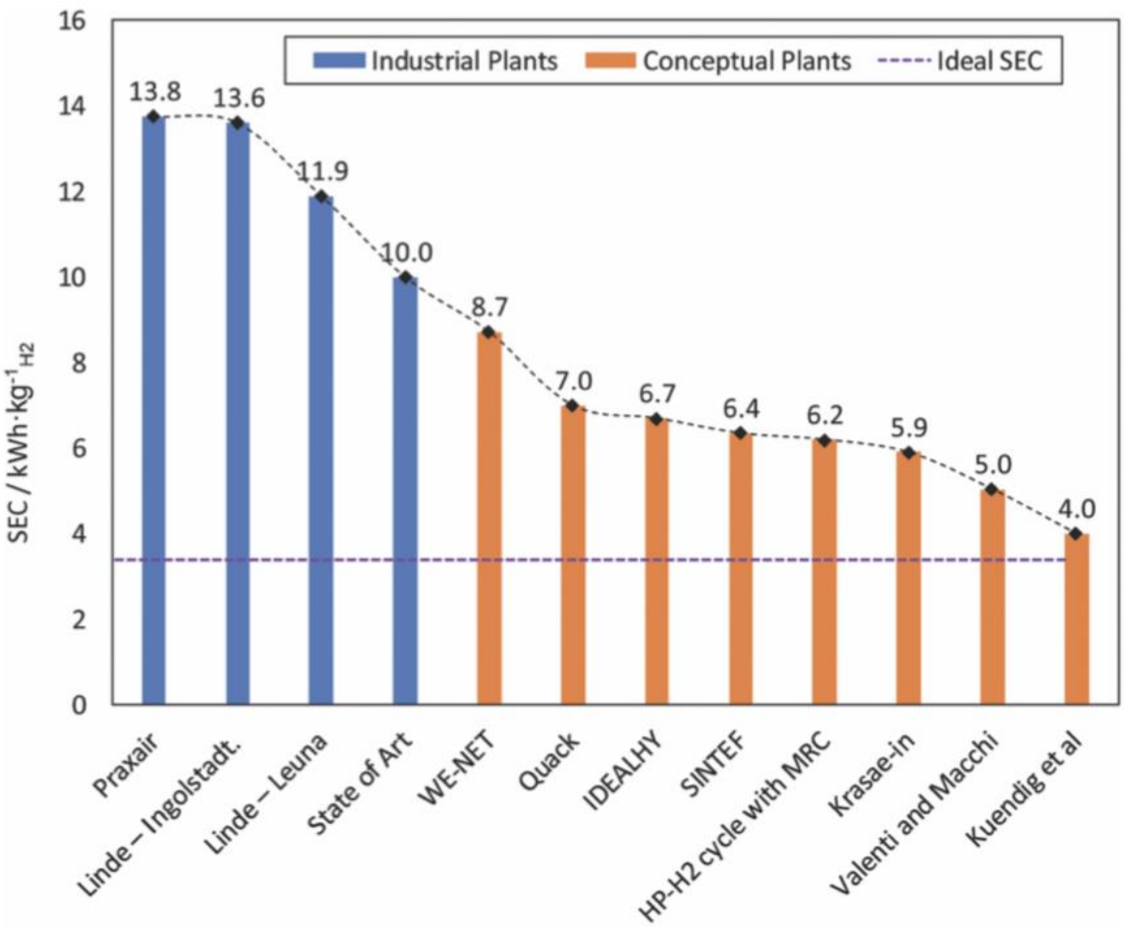


Figure 20.2: Source: Cardella paper [25]. List of currently operating and conceptual Hydrogen liquefaction processes. Conceptual plants have lower specific powers than operating plants because the operating plants are designed to maximize ROI while the conceptual plants are designed to minimize specific power, and contain assumptions that cannot yet be realized with today's technology.

We also performed a sensitivity analysis for the specific power/exergy efficiency of all three working fluids over a production range of 15 to 90 metric tons per day. The results are visualized in figures 20.3 through 20.5.

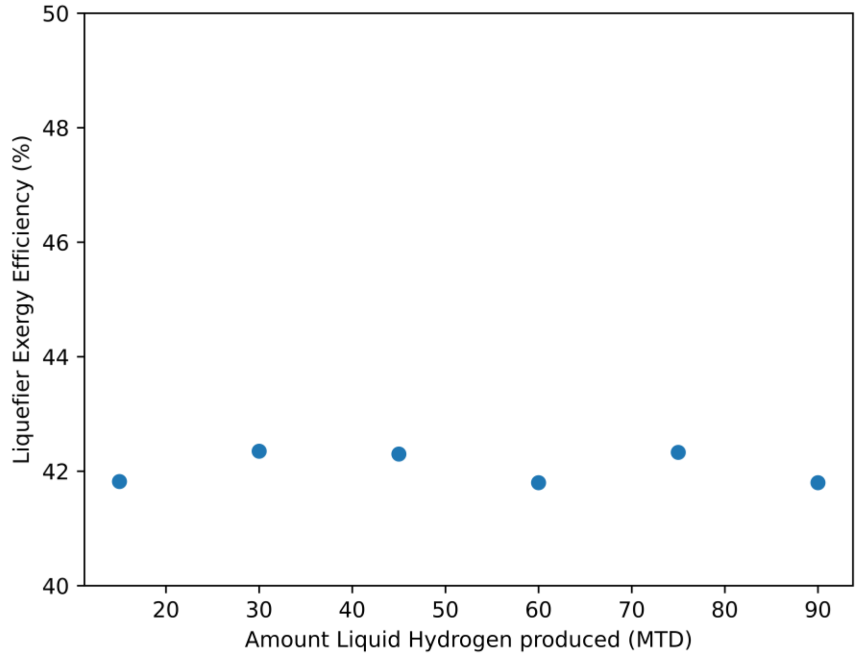


Figure 20.3: Plot of Liquefier Exergy Efficiency versus Amount of Liquid Hydrogen produced (MTD) for 15, 30, 45, 60, 75, and 90 MTD of Hydrogen produced, with Helium as the refrigerant.

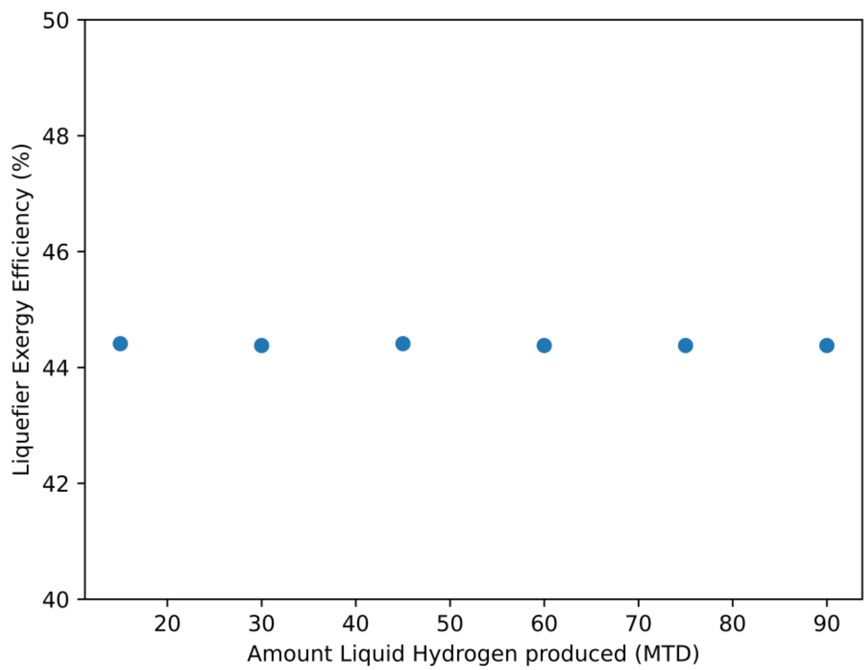


Figure 20.4: Plot of Liquefier Exergy Efficiency versus Amount of Liquid Hydrogen produced (MTD) for 15, 30, 45, 60, 75, and 90 MTD of Hydrogen produced, with Neon as the refrigerant.

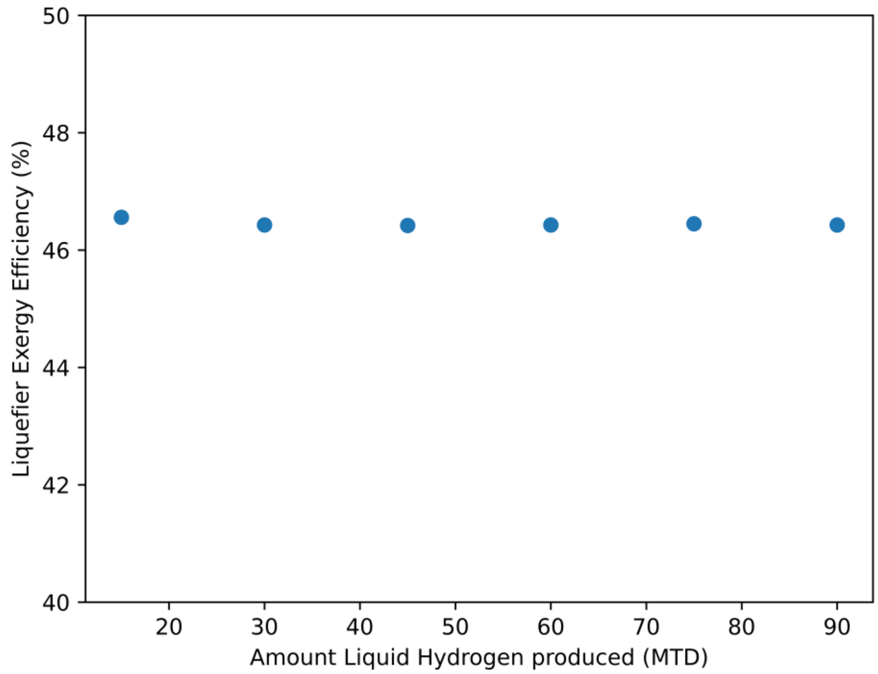


Figure 20.5:Plot of Liquefier Exergy Efficiency versus Amount of Liquid Hydrogen produced (MTD) for 15, 30, 45, 60, 75, and 90 MTD of Hydrogen produced, with Hydrogen as the refrigerant.

The conclusions are: Hydrogen refrigerant provides the most energy efficient plant, at the cost of paying for a stainless steel recycle compressor. Thus, it provides the most sustainable process design, which is important because the reasons to build this plant are related to sustainability. Also, energy efficiency is not sensitive to production rate. For some reason, helium provides the most energy inefficient plant. This influences the decision to design the process using Neon refrigerant instead of helium when the ROIs between the two are similar (table 20.1).

21 CONCLUSIONS AND RECOMMENDATIONS

At an NPV of $44,445,400, our Hydrogen liquefaction process is profitable. However, this number is deceiving, because we are able to charge a *green* premium for our product, selling Hydrogen at $13 per kilogram. This is approximately twice the market price for liquid Hydrogen.

As described in Section 20.3.3, the cost of liquefaction is $2.71 per kilogram of LH2, and the break-even price of liquid Hydrogen created using 100% electrolysis is $9.20 per kilogram. This means that the theoretical low limit for the cost of *green* Hydrogen is $6.49/kg, which is still about 33% more expensive than Natural Gas ($4.50/kg). More work must be done to get the cost of liquefaction down to the set target of $1 per kilogram.

Currently, there are no good equations of state that can accurately model Hydrogen at cryogenic conditions in ASPEN PLUS. This is why the REFPROP Databank was used. In order for ASPEN to reliably model Hydrogen liquefaction processes in the future, we recommend that equations of state that can properly model H_2's behavior in cryogenic ranges are implemented soon in ASPEN. One may hope that since H_2 is the simplest molecule, that calculations of Hydrogen's properties from first principles be incorporated into process designs. The design group was able to calculate Hydrogen's heat capacity and ortho para heat of conversion using statistical and quantum mechanics methods. There seems to be an overreliance on traditional equations of state in the literature for modeling cryogenic Hydrogen, where it may be more appropriate to model cryogenic Hydrogen using computational chemistry methods.

We were unable to perform any mixed refrigerant analysis due to incompatibility with our REPPROP database, but that is the logical next step for the design group.

For large scale processes like ours, we should also investigate the downstream consequences to profitability of creating an in-house Nitrogen loop. While the capital costs of additional heat exchangers, compressors, and expanders would increase the total capital investment, $7M in Nitrogen utilities expenditure would be saved annually.

The design group analyzed this process design rigorously and took great care to create an accurate process design. This effort can be directed into exploring other process design, comparing their economics, and selecting the most favorable one. This was an initial goal of the group that had to be scrapped due to time limitations.

Going forward, the tremendous amount of public and private interest in Hydrogen will only continue to increase. Given the financial commitments of numerous parties towards the development of green Hydrogen infrastructure in recent years, the design group predicts that you will know someone in twenty years who drives a Hydrogen car.

22 ACKNOWLEDGEMENTS

We would like to extend our most heartfelt thanks to the team of advisors who worked tirelessly to make our project possible. We thank Professor Seider for his unyielding patience, his detailed preparation for each of our weekly meetings, and his frank feedback. We thank Professor Vrana for his involved help with our profitability analysis, instant and detailed email replies, and accommodation of our senior year schedules to organize weekly meetings. We would like to thank Adam Brostow, our project author, for his know how in cryogenics. Without Adam's tireless work in both responding to emails and agreeing to take a look at our flowsheet during the weekends, breaks, and late hours n of the night, we would not have made the progress we did. Lastly, we would like to thank Professor Fabiano for his help in BAHX design.

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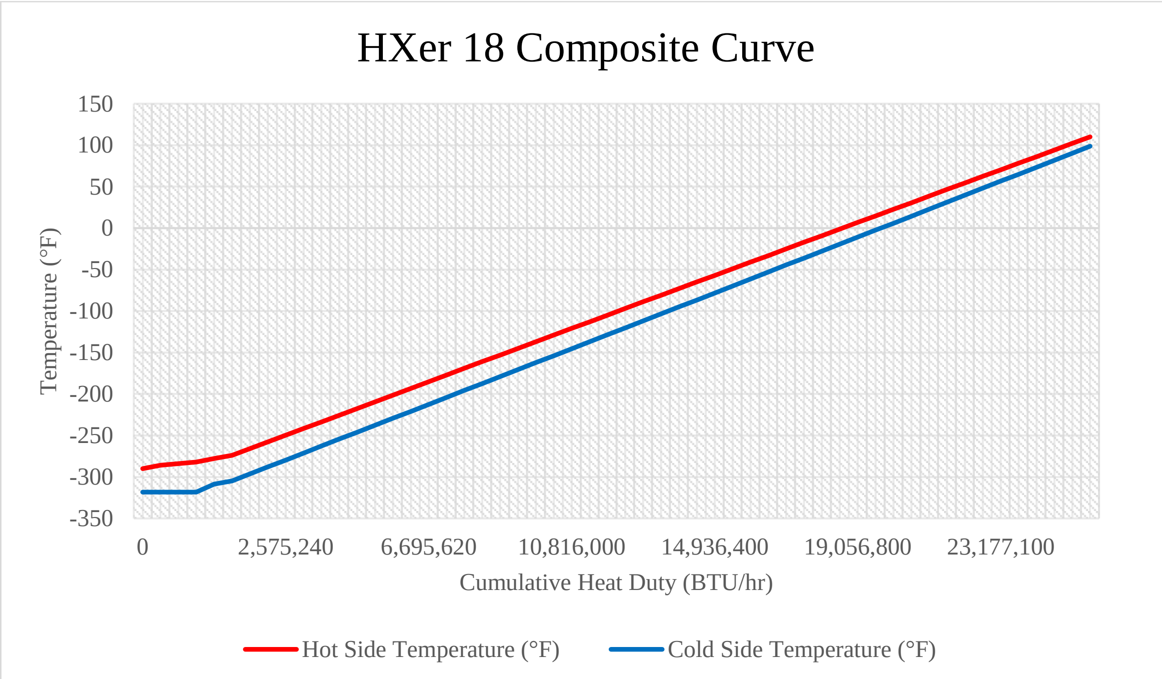
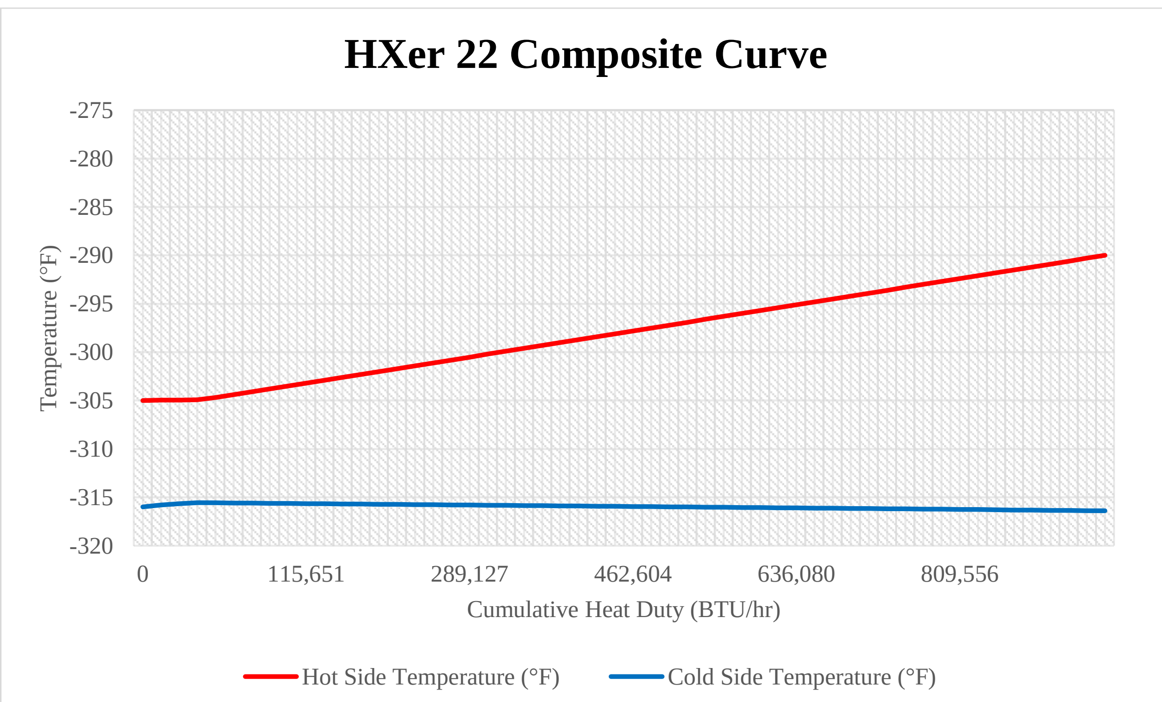
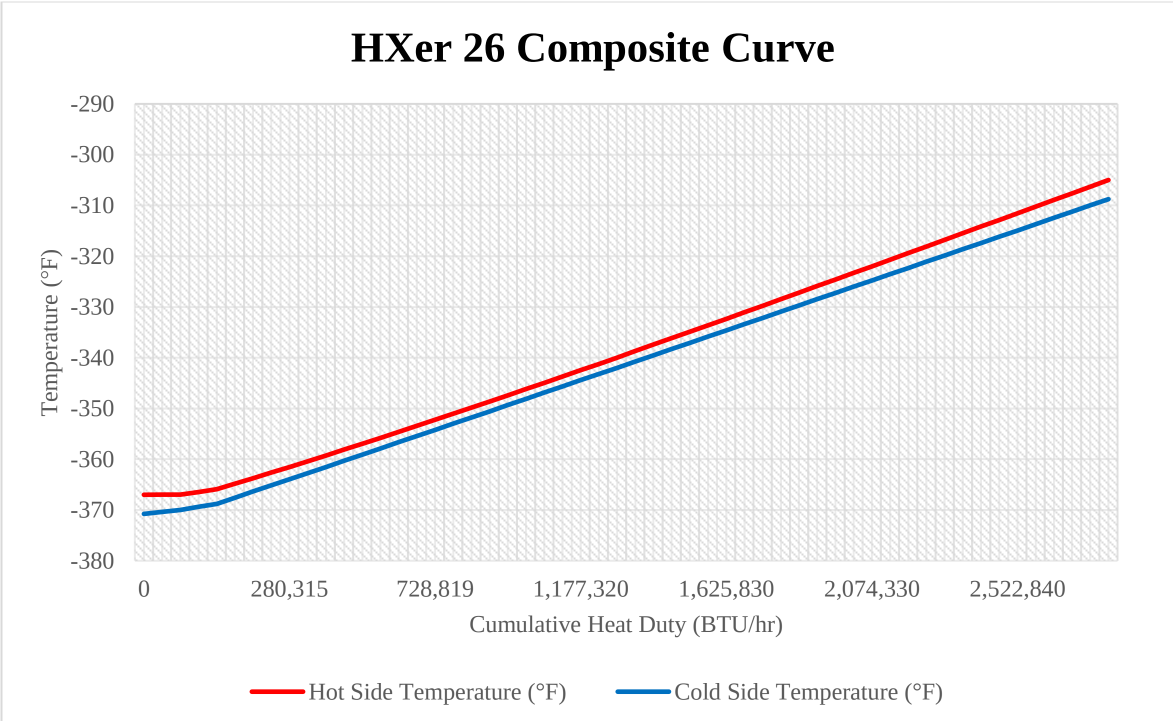
24 APPENDIX

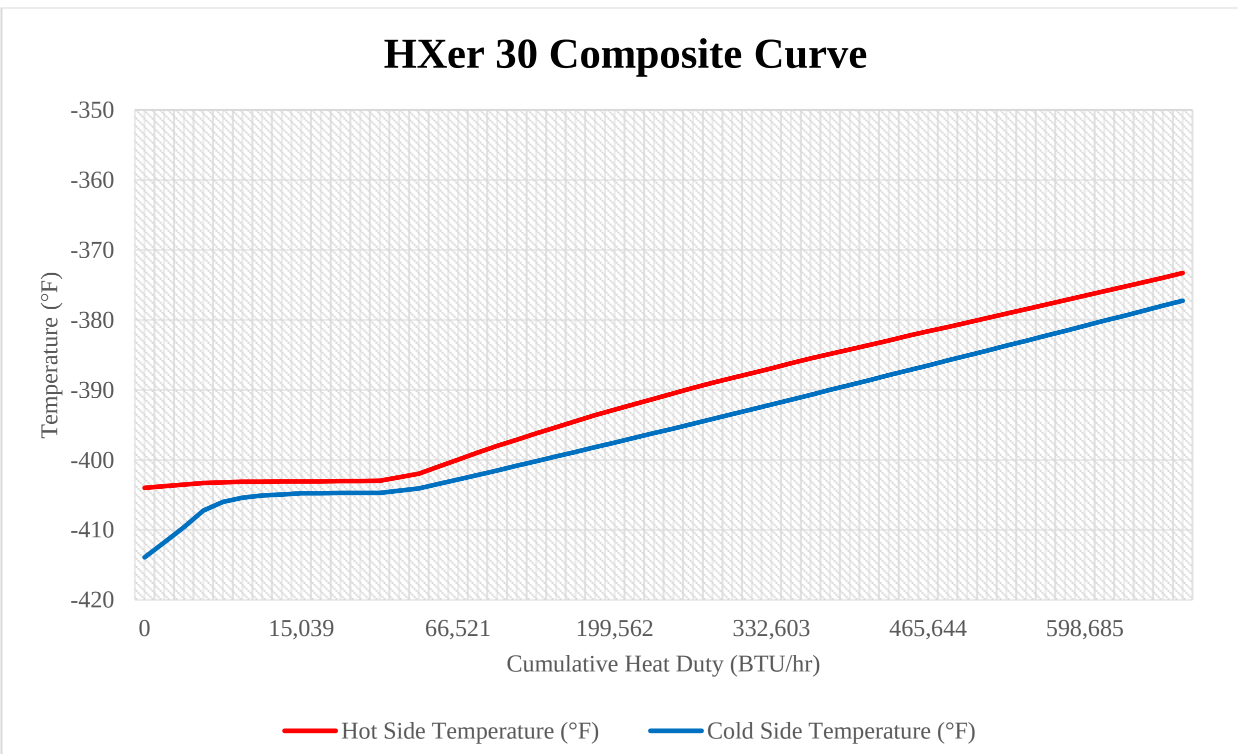
24.1 BAPFHX Design

24.1.1 Heat Exchanger Composite Curves






HXer 26 Composite Curve
-290
-300
-310
-320
-330
-340
-350
-360
-370
-380
Temperature (°F)
0
280,315
728,819
1,177,320
1,625,830
2,074,330
2,522,840
Cumulative Heat Duty (BTU/hr)
Hot Side Temperature (°F)
Cold Side Temperature (°F)


HXer 30 Composite Curve
-350
-360
-370
-380
-390
-400
-410
-420
Temperature (°F)
0
15,039
66,521
199,562
332,603
465,644
598,685
Cumulative Heat Duty (BTU/hr)
Hot Side Temperature (°F)
Cold Side Temperature (°F)

24.1.2 Design Procedure

The design procedure of the process BAPFHX was accomplished according to methods outlined in a Stewart Warner brochure from the 1960s, provided generously by Professor Leonard Fabiano from his early career at Air Products [36]. The design procedure for such systems has not changed significantly since then. One can also design these units using the recent published standards of ALPEMA, the Brazed Aluminum Plate-Fin Heat Exchanger Manufacturers' Association, since the design methodology for heat exchangers has hardly changed in fifty years [60]. Nevertheless, the fin types used are from the Stewart Warner brochure and it may be possible to develop better designs using modern fin types.

The design begins by calculating the overall thermal resistance, $\frac{1}{UA}$, of the heat exchanger from its composite curve, obtained from MHEATX Zone Analysis in Aspen Plus. The composite curve is divided into discrete temperature and pressure intervals and heat transfer in each interval is calculated according to heat transfer properties at that temperature, pressure, and composition. This yields composite curves that are nonlinear, as opposed to some of the simpler textbook examples students may be accustomed to, in which curves are piecewise linear [38]. These heat transfer properties are slowly varying, meaning that to a first order approximation the curves are approximately piecewise linear. Due to the small temperature approaches in this process, however, this approximation is not appropriate for a sophisticated design of the heat exchangers.

UA is calculated by summing all the resistances in each temperature interval of the zone analysis:

$$\mathrm{UA_i} = \frac{\mathrm{Q_i}}{\mathrm{LMTD_i}} \tag{24.1}$$

$$\mathrm{LMTD_i} = \frac{\Delta \mathrm{T_{i+1}} - \Delta \mathrm{T_i}}{\ln\left(\frac{\Delta \mathrm{T_{i+1}}}{\Delta \mathrm{T_i}}\right)} \tag{24.2}$$

$$\mathrm{UA} = \sum_{\mathrm{i}}^{\mathrm{intervals}} \mathrm{UA_i} = \sum_{\mathrm{i}}^{\mathrm{intervals}} \frac{\mathrm{Q_i}}{\mathrm{LMTD_i}} \tag{24.3}$$

Q_i	: heat duty in temperature interval i	$(\frac{BTU}{hr})$
ΔT_i	: temperature differences between hot and cold streams at each end of section i	(°F)
$LMTD_i$	: log-mean temperature difference, a ubiquitous quantity in chemical engineering that represents the driving force of heat transfer in heat exchanger flows	(°F)

Hot and cold streams are arranged in counterflow in the exchanger.

The designed thermal resistance, $\frac{1}{UA_D}$, is calculated by summing the thermal resistances of each layer passage in the heat exchanger. Technically this sum should also include the thermal resistances of the aluminum parting sheets dividing each layer, but this effect is neglected due to the high thermal conductivity of aluminum and the thinness of the sheets.

$$\frac{1}{\mathrm{UA_D}} = \left(\sum_{\mathrm{hi}}^{\mathrm{hotstreams}} \mathrm{h_{hi}}\,\mathrm{A_{hi}}\right)^{-1} + \left(\sum_{\mathrm{ci}}^{\mathrm{coldstreams}} \mathrm{h_{ci}}\,\mathrm{A_{ci}}\right)^{-1} \qquad 24.4$$

hi, ci	: hot or cold stream i	
h	: convective heat transfer coefficient of a stream	$(\frac{BTU}{hr\cdot ft^2\cdot°F})$
A	: effective heat transfer area of a passage	(ft^2)

The convective heat transfer coefficient, h, of a stream is dependent on its material properties and the geometry of the passage it flows through. It can be estimated using the Chilton-Colburn j-factor correlation with experimental data relating the j-factor to the Reynolds number of the stream.

Material properties used to compute h are the average of the properties at the inlet and outlet of the stream passage. A more accurate design can be done by computing the designed thermal resistance for each temperature interval in the zone analysis, using the material properties corresponding to the mean temperature, pressure, and composition in the interval, and then summing their inverses to get an ultimate UA_D, as in equation 24.3. The design group decided that this was not worth the effort.

Since the passages of plate-fin heat exchangers have an atypical geometry involving fins of varying characteristics, it is not appropriate to use many of the correlations in the literature to relate the two dimensionless quantities, Reynolds number and j-factor. For a given fin geometry, experimental data should be used to construct a custom correlation[2]. The Stewart Warner brochure only provided data for one out of the eleven fin geometries it provides specifications for (fin #10 in table 24.1).

A sixth-order polynomial fit to the data was made using Excel. An assumption was made that the j-factor asymptotically approaches a value of 0.005 as Reynolds number approaches infinity to account for design cases where the Reynolds number of streams in the heat exchanger exceeds 10,000. This fit was applied to all fin types for lack of information (figure 24.1). A better design would have a different correlation for each fin type.

$$\lim_{\mathrm{Re}\to\infty} \mathrm{j} = 0.005 \tag{24.5}$$

[2] Alternatively, researchers have suggested a BAPFHX-modified Dittus-Boelter Nusselt number correlation for turbulent flows:

$$\mathrm{Nu_D} = .000702\mathrm{Re}^{1.243}\mathrm{Pr}^{\frac{1}{3}}$$

$$\mathrm{h} = \frac{\mathrm{Nu_D}\kappa}{4\mathrm{R_h}}$$

This correlation was not used in the design but provides values for h within 90% of the values estimated by the Chilton-Colburn correlation for fin #10 [69]. [citation: https://dl.acm.org/doi/pdf/10.1145/3386762.3386780]

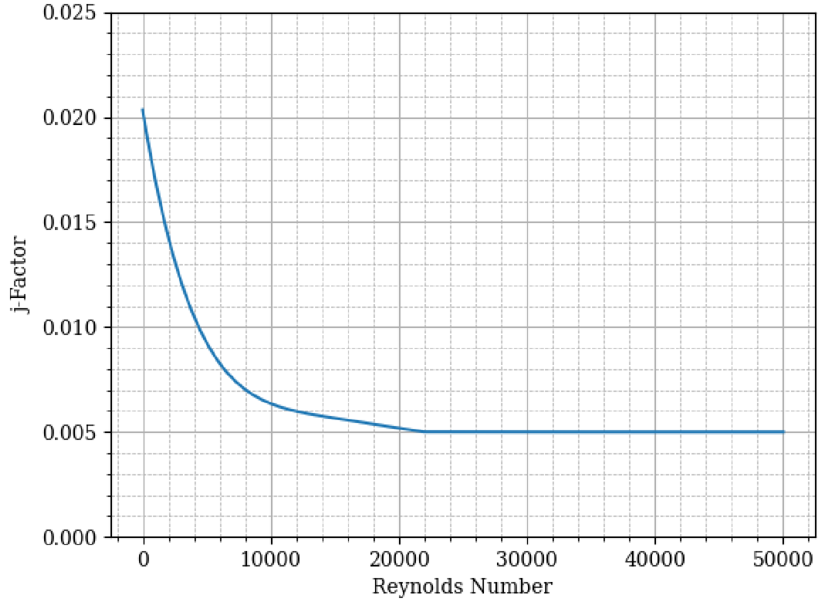


Figure 24.1: Chilton-Colburn j-factor correlation for lanced fin (.375'' height, 15 fins per inch—.008'' thick). This correlation was utilized for every fin type in table 24.1 due to lack of data.

$$Re = \frac{4R_h G}{\mu} \tag{24.6}$$

$$R_h = \frac{D_h}{4} = \frac{A_c}{P} \tag{24.7}$$

$$G = \frac{\dot{m}}{A_c} \tag{24.8}$$

$$Pr = \frac{\mu C_p}{\kappa} \tag{24.9}$$

$$h = \frac{jGC_p}{Pr^{\frac{2}{3}}} \tag{24.10}$$

R_h	: hydraulic radius	(ft)
A_c	: cross-sectional area of passage	(ft^2)
P	: wetted perimeter of passage	(ft)
$\dot{m}$	: mass flow rate of stream through passage	($\frac{lb}{hr}$)
$\mathrm{C_p}$	: constant pressure heat capacity of material	($\frac{BTU}{lb\cdot°F}$)
G	: Mass velocity of stream through passage	($\frac{kg}{m^2\cdot s}$)
μ	: dynamic viscosity of stream material	($\frac{lb}{ft\cdot hr}$)
κ	: thermal conductivity of stream material	($\frac{BTU}{ft\cdot hr\cdot°F}$)
Pr	: Prandtl number	(-)

Fin #	Fin Height, b (inches)	Type of Surface	Fin Spacing (Fins per inch)	Fin Thickness δ (inches)	A'_c (square feet per inch)	A''_{ht} (feet per inch)	R_h (ft)	$\frac{A_f}{A_{ht}}$
1	0.200	Perforated	6	0.025	0.001032	0.306	0.003262	0.537
2	0.200	Perforated	10	0.025	0.000911	0.399	0.002187	0.687
3	0.200	Perforated	14	0.008	0.001183	0.57	0.001986	0.74
4	0.250	Perforated	15	0.02	0.001118	0.658	0.001616	0.822
5	0.375	Perforated	15	0.012	0.002067	0.939	0.001979	0.854
6	0.375	Perforated	20	0.008	0.00214	1.292	0.00157	0.891
7	0.200	Lanced	18	0.01	0.001081	0.706	0.001531	0.806
8	0.250	Lanced	15	0.012	0.001355	0.731	0.001852	0.813
9	0.250	Lanced	14	0.02	0.00115	0.656	0.001751	0.817
10	0.375	Lanced	15	0.008	0.002242	1.064	0.002107	0.862
11	0.375	Lanced	18	0.008	0.002181	1.243	0.001754	0.865

Table 24.1: Geometrical factors of commonly used Stewart Warner fin surfaces. These factors are from the 1960s, and superior fin designs are likely to exist nowadays which may boost the heat transfer efficiency of BAPFHX.

Cross-sectional area, A_c, and hydraulic radius, R_h, of each passage were obtained from table 24.1, for the appropriate fin type, where:

$$\mathrm{A_c = A'_c \times W_p} \qquad 24.11$$

$A_c{}'$	: cross-sectional area per width of passage	$(\frac{ft^2}{inch})$
W_p	: passage width	(ft)

The effective heat transfer area of each passage, A, is calculated using

$$\mathrm{A} = \mathrm{A_{ht}} \times \eta_\mathrm{e} \qquad 24.12$$

$$\mathrm{A_{ht}} = \mathrm{A''_{ht}} \times \mathrm{W_p} \times \mathrm{L} \times \mathrm{N} \qquad 24.13$$

$$\eta_\mathrm{e} = 1 - \frac{\mathrm{A_f}}{\mathrm{A_{ht}}}(1 - \eta_\mathrm{f}) \qquad 24.14$$

Calculating the effective heat transfer area of each passage, A, is somewhat not straightforward due to the aluminum fins that enhance heat transfer between passages via conduction. Temperature gradients dictated by Fourier's law of conduction form along the fins, yielding the following fin efficiency factor:

$$\eta_\mathrm{f} = \frac{\mathrm{tanh(ml)}}{\mathrm{ml}} \qquad 24.15$$

$$ml = \frac{b}{2}\sqrt{\frac{2h}{\kappa\delta}} \qquad 24.16$$

This expression is approximate and assumes that the fin is (i) rectangular, and (ii) has an active tip, meaning convective heat transfer between the fin and the material stream occurs at the borders of the fins [61].

A_{ht}	: heat transfer area	(ft^2)
A''_{ht}	: heat transfer area factor	$(\frac{ft}{in})$
W_p	: passage width	(ft)
L	: heat exchanger length	(ft)
N	: number of passages for a liquid stream	$(-)$
η_e	: weighted surface effectiveness factor	$(-)$
η_f	: fin efficiency factor	$(-)$
$\frac{A_f}{A_{ht}}$	: ratio of fin area to total area	$(-)$
ml	: fin geometry and material factor	$(-)$
b	: fin height	(ft)
h	: convective heat transfer coefficient of stream material	$(\frac{BTU}{ft^2 \cdot hr \cdot {}^\circ F})$
κ	: thermal conductivity of stream material	$(\frac{BTU}{ft \cdot hr \cdot {}^\circ F})$
δ	: fin thickness	(ft)

This procedure obtains the quantities needed to compute UA_D from equation (24.4).

One important design note is that for the hot Hydrogen streams in heat exchangers 26 and 30, which are packed with ferric oxide catalyst, an adjustment must be made to the convective heat transfer coefficient. Broadly speaking, two possible phenomena occur which affect it. The first is the thermal resistance generated by conductive heat transfer through the catalyst material, which will decrease the apparent convective heat transfer coefficient. The second is the competing effect of increased turbulence and Reynolds number which will increase the convective heat transfer coefficient.

According to experimental data in [62], the convective heat transfer coefficient of a passage packed with ferric oxide increases by approximately a factor of two compared to a vacant passage. This suggests that the turbulence increase effect dominates the conductive resistance effect of the catalyst packing. To account for this, the convective heat transfer coefficients for the hot Hydrogen streams in heat exchangers 26 and

30 were multiplied by a factor of two. This resulted in a 22% and 36% increase in the overall heat transfer coefficients of the exchangers, respectively.

Passage pattern is determined by heuristics derived from studies of rigorous BAPFHX optimization, such as [63]. These optimizations show that the passage orderings that produce the most uniform temperature gradients in BAPFHX, and thus which minimize entropic heat losses, are those which alternate between hot and cold streams. This is an intuitive result. When designing passage patterns, therefore, the ordering that was selected was the one that maximized the number of hot streams that were adjacent to cold streams, and vice versa.

The pressure drop of a stream through the heat exchanger can be roughly broken up into seven components:

1) expansion loss from nozzle into header
2) contraction loss from header into HX core entry
3) Drop across inlet distributor
4) Drop across HX core
5) Drop across outlet distributor
6) expansion loss from HX core exit into header
7) contraction loss from header into nozzle

The most significant contributor to the total pressure drop is the pressure drop across the HX core, (4).

Drop across heat exchanger core:

$$\Delta P = f\frac{G^2 L}{2\rho R_h} \qquad 24.17$$

Fanning friction factor for turbulent flows:

$$f = \frac{1}{4}\ 1.35\ (1.82\ \log_{10}\mathrm{Re} - 1.64)^{-2} \qquad 24.18$$

Expansion and contraction losses:

$$\Delta P = K\frac{G^2}{2\rho} \qquad 24.19$$

Detailed information about expansion and contraction loss coefficients, K, can be found at [64]. Pressure drops across inlet and outlet distributors were neglected since L in equation (24.1.16) accounts for the total heat exchanger length which includes the distributors.

f	: Fanning friction factor	$(-)$
G	: mass velocity of stream through passage	$(\frac{kg}{m^2 \cdot s})$
L	: heat exchanger length	(ft)
ρ	: density of material stream	$(\frac{lb}{ft^3})$
R_h	: hydraulic radius	(ft)
K	: expansion/contraction loss coefficient	$(-)$

Pressure drop in the catalyst-packed Hydrogen streams of exchangers 26 and 30 could be modeled by the Ergun equation, but this model has been found to overestimate the actual pressure drop in BAPFHX by about five times [63]. Due to the unreliability of the Ergun equation for this geometry and the small effect of the catalyst packing on the pressure drop, the effect was neglected and equation (24.17) was used to calculate pressure drops for these streams as well.

24.1.3 Design Constraints

The design of the heat exchanger is subject to two constraints that must be met.

The first is that the designed thermal resistance, $\frac{1}{UA_D}$, must equal the actual thermal resistance, $\frac{1}{UA}$:

$$\mathrm{UA_D} = \mathrm{UA} \qquad 24.20$$

UA is determined from the composite curve of the heat exchanger by equation (24.3).

UA_D is determined by the various specifications of the heat exchanger, namely its length, width, number of passages, fin types, and by the material properties of the streams flowing through it. It is calculated via Chilton-Colburn correlations provided in the Stewart Warner brochure and the procedure outlined in the previous section. The specifications are varied until the constraint (24.20) is met.

The second constraint relates to the pressure drops along the heat exchanger. Due to larger scale considerations of the energy efficiency of the Hydrogen liquefaction plant, there are certain limits to how large of a pressure drop there can be for a particular stream when it passes through a heat exchanger. For example, pressure drops for the Neon refrigerant should not exceed 2-3 psi in the streams where it is responsible for cooling down other streams (i.e., is a cold stream). The reason for this is to minimize the power requirement of the Neon recycle compressor, 94, when it recompresses the working fluid in stream 92 to 150 psia in stream 68.

Other pressure drop constraints are listed in the BAPFHX performance specification sheets in section 14.

$$\Delta\mathrm{P} \leq \Delta\mathrm{P_{allowable}} \qquad 24.21$$

This constraint is particularly troublesome because maintaining a low pressure drop requires maintaining a small mass velocity through the passages of the BAPFHX, which reduces the heat transfer through those passages, lowering the efficiency of the heat exchanger. Thus, the unconstrained overall heat transfer coefficient would be larger than the constrained overall heat transfer coefficient that we are forced to use.

This results in a larger heat exchanger area, A_D, thereby increasing purchase cost of these heat exchangers. Evidently, there is a tradeoff between energy efficiency/operating cost and capital cost. This is one of the reasons why there is a difference between optimizing the energy efficiency of the plant and optimizing the ROI of the plant.

24.1.4 Calculation of Material Volume, Pipe Sizing, and Purchase Amount

The core volume for a stream of a BAPFHX is calculated as

$$V_{stream} = A_c \times N \times W_p$$

The pipe diameter of the pipe that connects the inlet and outlet of the stream to the BAPFHX is estimated by assuming that the cross-sectional area of the pipe equals the total cross-sectional area of the stream through the heat exchanger:

$$A_{pipe} = A_c \times N$$

I.e., the pipe sizing is chosen to minimize expansion and contraction pressure losses. The pipe is cylindrical so

$$D_{pipe} = \sqrt{\frac{4A_{pipe}}{\pi}}$$

The total heat exchanger core volume for the Neon refrigerant streams was 63 cubic feet, the optimal pipe diameter was about 1 foot, and the average density of the Neon through the loop is 1 pound per cubic foot. Assuming the total length of the pipe system to be 200 feet, the total mass of Neon in the plant was estimated at 220 pounds, about 0.1 metric tons. At a mass flowrate of 196,572 pounds per hour, the average Neon molecule takes four seconds to go through the loop. This corresponds to a total one-time purchase cost of $35,000 at pre-Russian invasion of Ukraine prices. With the foreseeable uncertainty in Neon price and the shortage in supply, a realistic purchase cost would be ten times this, or $350,000.

Additionally, due to safety considerations concerning the risk of Hydrogen fueled explosions, this analysis is applied to the Hydrogen streams to calculate how much Hydrogen is in the plant at a given time. The total heat exchanger core volume for the Hydrogen streams was 0.9 cubic feet, the optimal pipe diameter was about 0.2 feet, and the average density of the Hydrogen through the loop is 0.75 pounds per cubic foot.

Assuming the total length of the pipe system to be 225 feet, the total mass of Hydrogen in the plant was estimated at 7 pounds.

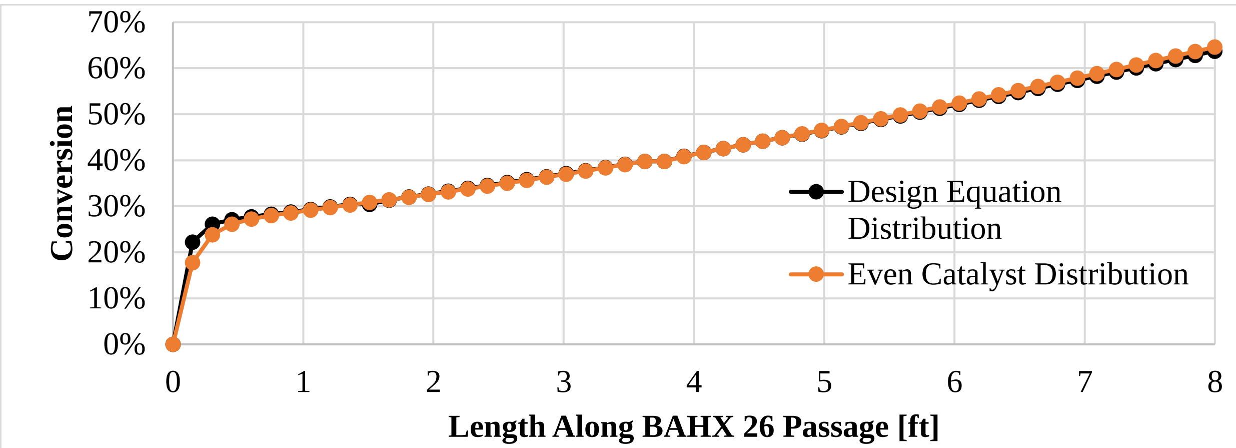
24.1.5 Packed Bed Reactor Model

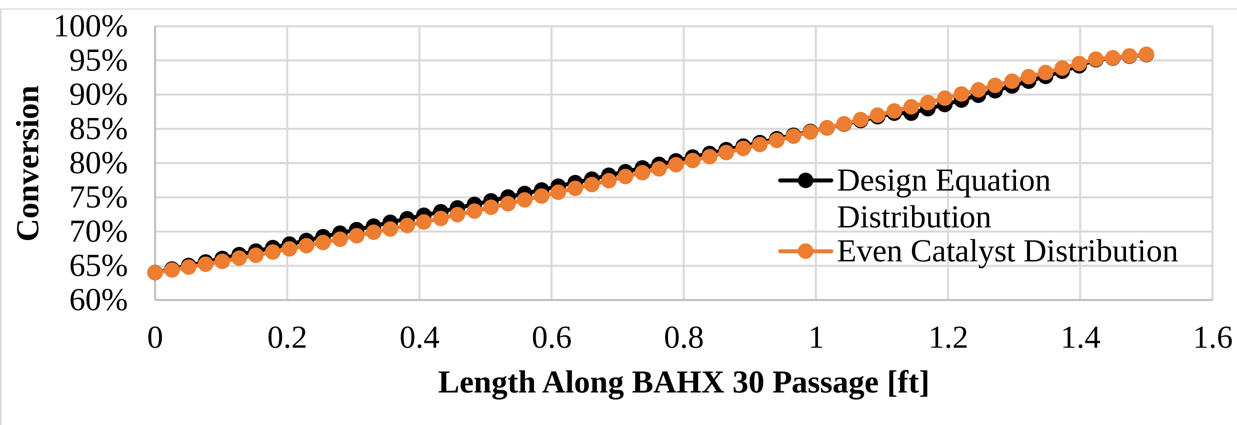
The Packed Bed reactor model has been used to approximate the reactions occurring in Heat Exchangers 26 and 30, as well as in the catalytic phase separator. In the separator, as discussed in Section 14.7, isothermal assumptions are made due to the unique nature of the reaction occurring on the liquid phase of a vapor-liquid equilibrium. In this case, all calculations were performed using a Python script that simply solved multiple differential PBR design equations with technical details provided such as the particle diameter of the catalyst. In the case of the knock-out drum, the correlation to a PBR was very straight-forward: a cylindrical vessel with no axial mixing can be easily modeled by fluid traveling down a bed of catalyst. However, the BAPFHX is a complex system with several non-idealities that diverge from the PBR assumption. In this section, we detail our considerations in moving forward with the PBR model and display our results.

With a BAPFHX, streams are partitioned into non-tubular passages that are interrupted by fins that aid in the temperature distribution of the unit. The geometry is explicitly non-cylindrical, and turbulent reactant flows (76,000-157,000 Re for HX 30, and 100,500-107,000 Re for HX 26) in such a geometry can lead to unpredictable flow patterns especially around bends other internal BAHX design features. From a fluid flow perspective, there are many reasons to be suspicious about the results produced through the PBR model. However, given the unknown internalities of the exchanger, there are few model options that offer reasonable improvements with high confidence. The PBR is useful in a number of ways: its differential equations allow for discrete integration, through which Python-generated data, such as equilibrium conversion, may be paired with ASPEN-generated data such as viscosities and densities of fluids at discrete points in the heat exchanger.

To make variable assignments, A_c is defined as the total passage space, excluding the surface area fin thickness take, z is defined as the length of the exchanger, which here is considered the passage length, and mass velocities are detailed by the aforementioned A_c.

Determining the required mass of catalyst in each heat exchanger through the PBR model also provided insights into the reaction distribution across each heat exchanger.





Earlier discussion hinted at the concern over modeling the heat of reaction for conversion within the length of the BAHX. One particular concern was the distribution of heat across the reactant stream, since it was unknown how the reaction would unfold in the system and how specific areas would elevate in temperature. Ideally, the bulk of the heat would travel with the gas through and out of the exchanger, absorbed by the heat sinks along the way. However, concern over thermal gradients in the aluminum still existed. Using the

PBR model, a spatial distribution of conversion was created, showing a linear increase in conversion in HX 30 across the length, but a region of high conversion in HX 26. This simulated phenomenon can be attributed to introducing paramagnetic catalysts sharply rather than prior the third heat exchanger. It appears as though prior reaction in the colder region of HX 22 or in an interstitial operation would reduce the uneven distribution of conversion in HX 26 should it pose a threat or point of improvement to this process.

Using discrete integration also allowed for catalyst distribution manipulation, by which the calculated total required catalyst was distributed equally at each node of the heat exchanger passage(s). In both cases of HX 26 and HX 30, conversion curves were nearly undistinguishable between the optimally distributed catalyst map calculated from discrete integration, and the evenly distributed counterpart. This is a production success, as an evenly distributed catalyst is undoubtably a simpler configuration to produce.

Moreover, by accumulating each "dW" across the length of the heat exchanger, the required mass of catalyst installed is understood to be approximately 305 kg for HX 26 and approximately 25 kg for HX 30. The reason for the vastly different mass requirements are twofold: HX 30 has both a lower temperature range with faster reaction kinetics and a colder material that is more condensed, causing increased surface interactions between the gas and the catalyst.

24.2 Ortho-Para conversion physical chemistry

A Hydrogen molecule consists of two Hydrogen atoms, each of which has one proton and one electron. The H_2 molecule thus consists of two protons and two electrons. The two electrons form a symmetric covalent bond. The two protons each have a spin 1/2 degree of freedom, which gives rise to two spin isomers of Hydrogen: the triplet o-H2 state and the singlet p-H2 state. This is a quantum mechanical phenomenon which was predicted in 1927 by Werner Heisenberg and Friedrich Hund by the new quantum theory developed by Heisenberg, Schrodinger, and other legendary scientists. This prediction was one of the early triumphs of the quantum theory and was cited as the most noteworthy application of the theory when Heisenberg was awarded the 1932 Nobel Prize in Physics.

The explanation is as follows. Each proton has two spin degrees of freedom: each proton spin can be either up (+1/2), or down (-1/2). Since there are two protons, the combined nuclear system has an overall spin of $\frac{1}{2}+\frac{1}{2}=1$, and has $2 \cdot 2 = 4$ possible spin states, which are listed in their natural basis:

$$|\uparrow\uparrow\rangle;\ |\uparrow\downarrow\rangle;\ |\downarrow\uparrow\rangle;\ |\downarrow\downarrow\rangle \qquad 24.22$$

Since the two protons each have a spin of 1/2, which is a fraction, it turns out for quantum mechanical reasons that the overall state of each proton must be antisymmetric, meaning that if you exchanged the two protons, the state of each proton would be the same as it was before except for a minus sign. Only the first and fourth states in equation (24.2.1) are symmetric and none of them are antisymmetric. Systems with integer spins, for example, the nucleus of the Hydrogen atom has total spin 1, must have a symmetric overall state. To have symmetric and antisymmetric spin states, the natural basis is reorganized into a symmetric/antisymmetric basis, consisting of three 'triplet' symmetric states, and one 'singlet' antisymmetric state [65]:

$$\text{triplet};\ S = 1: \begin{cases} |\uparrow\uparrow\rangle \\ 2^{-1/2}(\,|\uparrow\downarrow\rangle + |\downarrow\uparrow\rangle\,) \\ |\downarrow\downarrow\rangle \end{cases}$$

$$\text{singlet; } S = 0: \{\, 2^{-1/2}(\,|\uparrow\downarrow\rangle - |\downarrow\uparrow\rangle\,)\tag{24.23}$$

This basis is the origin of the ortho and para spin isomers of Hydrogen. Ortho-Hydrogen are the molecules of Hydrogen with triplet spin states, and para-Hydrogen are the molecules with singlet spin states. In addition to spin states, protons have rotational states described by a quantum number J (J = 0, 1, 2, 3, …), which can be symmetric (J is even) or antisymmetric (J is odd). For the overall state of a proton to be anti-symmetric, there are two possibilities. The first is a symmetric triplet spin state composed with an odd J, antisymmetric rotational state. All the spin states have the same energy eigenvalues, so this is said to be triply degenerate, $g = 3$, since there are three symmetric spin states. The second is an antisymmetric single spin state composed with an even J, symmetric rotational state. This is singly degenerate, $g = 1$.

The energy eigenvalue of the proton states comes from the rotational state and can be calculated by applying the rigid rotor approximation:

$$E_J = \frac{J(J+1)\hbar^2}{2I} \tag{24.24}$$

Rotational states have corresponding degeneracy

$$g_J = 2J + 1 \tag{24.25}$$

A higher order expression can be obtained [25] but equation (24.24) is more than sufficient to calculate the thermodynamic properties of ortho and para-Hydrogen.

The moment of inertia is known from experimental measurement:

$$\frac{\hbar^2}{kI} = 174.98\text{K} \tag{24.26}$$

$\hbar$	: Reduced Planck's constant	$(J \cdot s)$
J	: rotational quantum number (0, 1, 2, …)	$(-)$
I	: moment of inertia	$(kg\ m^2)$
k	: Boltzmann constant	$(\frac{J}{molK})$

With this information, one can use statistical methods to determine thermodynamic properties of Hydrogen at a given temperature [66]. The total partition function for the entire two atom system is a product of the nuclear, rotational, positional, vibrational, and electronic partition functions of the molecule (equation 24.27). However, only the nuclear and rotational partition functions are different between ortho and para spin isomers, and of these two, only the rotational partition functions have different energy eigenvalues. Therefore, only the rotational partition function needs to be calculated.

$$\mathrm{q_{tot}} = \mathrm{q_{elec} q_{vib} q_{rot} q_{trans} q_{nuc}} \qquad 24.27$$

Rotational partition functions were constructed in order to calculate the equilibrium compositions of o- and p-H_2 at a certain temperature, as well as the enthalpy heat of reaction of the ortho-para conversion reaction at a certain temperature. Ortho-Hydrogen's partition functions are constructed by summing only over odd J rotational states, and para-Hydrogen by summing only over even J rotational states, including J = 0, the lowest energy state. That para-Hydrogen occupies the lowest energy state while ortho-Hydrogen does not, is the origin of why para-Hydrogen is more stable than ortho-Hydrogen, and why there is an exothermic conversion that occurs in the first place. As the temperature goes to infinity, the enthalpy heat of reaction goes to nearly zero as all the higher energy J states of both spin isomers are filled and the difference between the two decreases. As the temperature goes to zero, the heat of reaction increases until the temperature is low enough where the only states that are occupied are the lowest energy states: J = 0 for para-Hydrogen and J = 1 for ortho-Hydrogen. The difference between the two energy levels for this state is 310 BTU per pound of Hydrogen, which is the heat of reaction observed near the boiling point of Hydrogen. Figures in section 13 were produced using these quantum and statistical mechanics methods.

$$Z_{ortho} = 3\sum_{oddJ}^{\infty}(2J+1)e^{-\frac{E_j}{kT}} \quad 24.28$$

$$Z_{para} = \sum_{evenJ}^{\infty}(2J+1)e^{-\frac{E_j}{kT}} \quad 24.29$$

$$y_{para} = \frac{Z_{para}}{Z_{para}+Z_{ortho}} \quad 24.30$$

$$y_{ortho} = 1 - y_{para} \quad 24.31$$

$$\langle E_{ortho}\rangle = \frac{3\sum_{oddJ}^{\infty} E_J(2J+1)e^{-\frac{E_J}{kT}}}{Z_{ortho}} \quad 24.32$$

$$\langle E_{para}\rangle = \frac{\sum_{evenJ}^{\infty} E_J(2J+1)e^{-\frac{E_J}{kT}}}{Z_{para}} \quad 24.33$$

$$\langle E_{eqm}\rangle = y_{ortho}\langle E_{ortho}\rangle + y_{para}\langle E_{para}\rangle \quad 24.34$$

$$\Delta H_{rxn} = \langle E_{ortho}\rangle - \langle E_{para}\rangle \quad 24.35$$

$$C_{v,ortho} = \frac{\partial\langle E_{ortho}\rangle}{\partial T} = \frac{3\sum_{oddJ}^{\infty} E_J{}^2(2J+1)e^{-\frac{E_J}{kT}}}{kT^2 Z_{ortho}} + \frac{\left(3\sum_{oddJ}^{\infty} E_J(2J+1)e^{-\frac{E_J}{kT}}\right)^2}{kT^2 Z_{ortho}{}^2} \quad 24.36$$

$$C_{v,para} = \frac{\partial\langle E_{para}\rangle}{\partial T} = \frac{3\sum_{evenJ}^{\infty} E_J{}^2(2J+1)e^{-\frac{E_J}{kT}}}{kT^2 Z_{para}} + \frac{\left(3\sum_{evenJ}^{\infty} E_J(2J+1)e^{-\frac{E_J}{kT}}\right)^2}{kT^2 Z_{para}{}^2} \quad 24.37$$

$$C_{v,eqm} = \frac{\partial\langle E_{eqm}\rangle}{\partial T}$$

The expressions for heat capacities, 24.34 – 24.36, are for the heat capacity contributions of the rotational degrees of freedom of the Hydrogen molecule. Since the molecule is diatomic, it has two rotational degrees of freedom and by the equipartition theorem, the rotational heat capacity approaches R, the ideal gas constant, as temperature approaches infinity. These expressions are not the complete heat capacities for Hydrogen, however, since the molecule also has three translational degrees of freedom which contribute 1.5R and one vibrational degree of freedom which contributes R. By the equipartition theorem, the total heat capacity should approach 3.5R as temperature goes to infinity.

Expressions for the rotational heat capacity for any composition of Hydrogen spin isomers can be obtained in the same manner:

$$\langle E_x \rangle = y'_{ortho} \langle E_{ortho} \rangle + y'_{para} \langle E_{para} \rangle$$

$$C_{v,x} = \frac{\partial \langle E_x \rangle}{\partial T}$$

24.3 Calculator Blocks Fortran Walkthrough

The heat of reaction as a function of temperature was calculated using equation 24.34 and the methods in section 24.2. In order to lower computational complexity in ASPEN, a polynomial curve was fitted to the values produced by equation 24.34.

Average heat of reaction across the exchanger was calculated using the following relation. In the Fortran subroutine, the heat of reaction was calculated using the Trapezoid rule, as Fortran can only perform numerical integration.

$$Q_{rxn,avg} = \frac{1}{T_{out} - T_{in}} \int_{T_{in}}^{T_{out}} HRxn(T)$$

The exothermic release of heat from ortho-para conversion was modeled as a Heat Leak in ASPEN PLUS, as shown below:

$$Q_{leak} = Q_{rxn,avg} * \Delta(Moles\ o - H_2)$$

This was then multiplied by the guessed change in the amount of ortho-Hydrogen between the inlet and the outlet of the hot Hydrogen stream. These guessed changes are reasonable because they represent around 80% of equilibrium conversion.

24.4 Green LH2 economic comparison to Natural Gas

Per million BTUs, liquid Hydrogen is

Fuel Type	***Cost per million BTUs (\$/10^6 BTUs)***
Natural Gas	\$2.13
Liquid Hydrogen	\$89

Table 24.2

Hydrogen Source	***Break even price (\$/kg)***
Grey Natural Gas SMR	\$4.50/kg
Green Water Electrolysis	\$9.20/kg

Table 24.3

24.5 Estimation of Neon in Troposphere

Surface atmospheric pressure (N/m^2)	101,325
Earth Radius (m)	6,378,100
Earth surface area (m^2)	511,201,962,310,545
Approximate weight of Earth's atmosphere (N)	51,797,538,831,116,000,000
acceleration due to gravity (N/kg)	9.81
Approximate mass of Earth's atmosphere (kg)	5,280,075,314,079,100,000
% of atmosphere mass comprised of troposphere	75%
Neon concentration in troposphere, by mass (ppm)	18
Approximate mass of Neon available in troposphere (kg)	71,993,826,907,469
(metric tons)	35,996,913,454
(metric tons, order of magnitude)	10,000,000,000

24.6 Cost of Liquefaction

24.6.1 Specific Liquefaction Cost (SLC) at 100% Electrolysis

In order to calculate the cost of liquefaction when 100% of the Hydrogen feedstock is sourced from electrolysis of water, the electrolyzers are removed from the capital cost. Specific liquefaction cost (SLC) is calculated using the equation below:

$$SLC = \frac{Total\ Annual\ Production\ Cost - Annual\ Electrolysis\ Energy\ Cost\ [\$]}{Production\ of\ LH_2\ per\ year\ [kg]}$$

$$SLC = \frac{\$119.4\ M - \$73.1\ M}{(45\ MTD) * \left(1000\frac{kg}{MT}\right) * \left(330\frac{days}{yr}\right)} = \$2.71/kg$$

24.6.2 Specific Liquefaction Cost (SLC) at 100% SMR

In order to calculate the cost of liquefaction when 100% of the Hydrogen feedstock is sourced from steam-methane reforming, the steam methane reformers are removed from the capital cost. Specific liquefaction cost (SLC) is calculated using the equation below:

$$SLC = \frac{Total\ Annual\ Production\ Cost - Annual\ LNG\ Feedstock\ Cost\ [\$]}{Production\ of\ LH_2\ per\ year\ [kg]}$$

$$SLC = \frac{\$81.1\ M - \$15.7\ M}{(45\ MTD) * \left(1000\frac{kg}{MT}\right) * \left(330\frac{days}{yr}\right)} = \$3.46/kg$$

24.7 Costing and Sizing of Process Equipment

24.7.1 Compressor Correlations

Screw compressors and centrifugal compressors were used in our process. The Hydrogen feed compressor and Hydrogen recycle compressor are screw compressors, and the Neon recycle compressor is a centrifugal compressor. All compressor correlations were taken from *Product and Process Design (Seider et al.)*. Sample calculations are provided below. Screw compressor correlation for F.O.B purchase cost is smooth, with no discontinuities, and so can be extrapolated past the prescribed range.

Type	***Range***	***Equation (F.O.B purchase cost, $, 2013)***
Screw	10 hp < P < 750 hp	$\exp(8.2496 + 0.7243 \times \ln(P))$
Centrifugal	200 hp < P < 30000 hp	$\exp(9.1553 + 0.63 \times \ln(P))$

Table 24.4

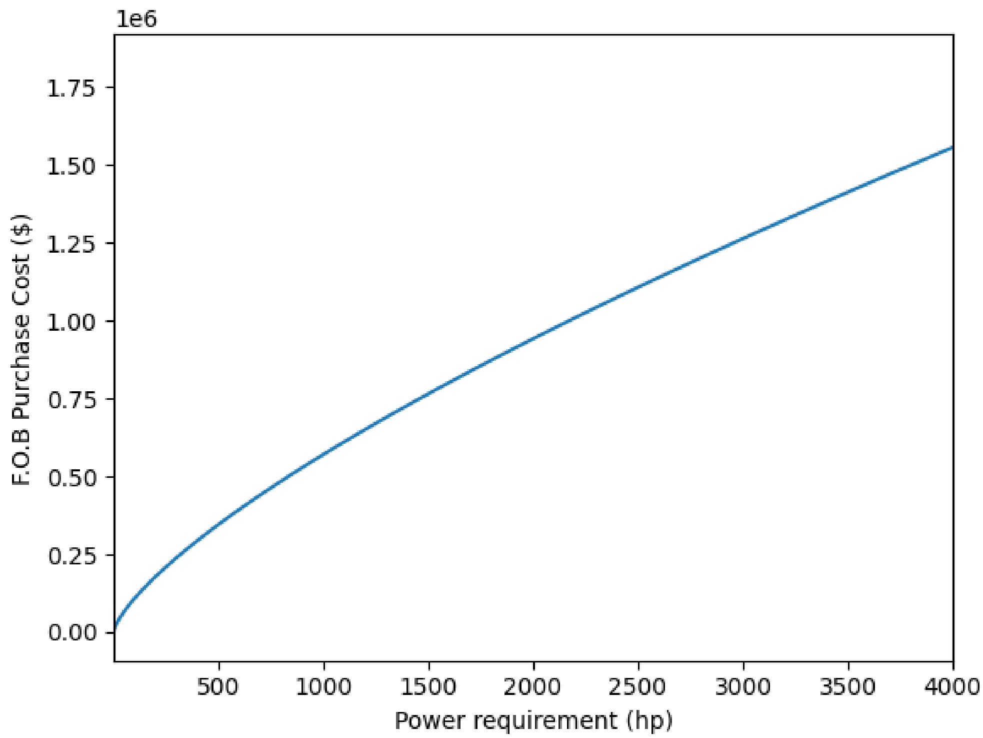
<u>Sample calculation for Hydrogen Feed Compressor:</u>

For each of these cases, our ASPEN flowsheet provides a power requirement for each compressor. These power requirements were scaled up by 10% for contingencies. This adjusted power requirement was plugged in to the correlations above.

$$P_{adj} = 1.1 \times P_{actual} = 1.1 \times \ (1312\ hp) = 1443\ hp$$

$$F.O.B\ purchase\ cost = exp(24.2496 + 0.7243 \times \ln(1443) = \$693{,}503$$

Bare module cost was found by scaling the F.O.B Purchase Cost by a Driver factor (F_d), a Material Factor (F_M), and the bare module factor for compressors (F_{BM} = 2.15). For this process, a driver factor of 1 was used for all cases. A Material factor of 2.5 (stainless steel) was used for all compressors in contact with Hydrogen. A material factor of 1 (Carbon steel) was used for the large Neon recycle centrifugal compressor.



While the valid range of the correlation is from 10 hp to 750 hp, the correlation itself is smooth, with no discontinuities past this point, and so it is used to estimate the F.O.B. purchase cost for screw compressors larger than 750 hp.

24.7.2 Turboexpander Correlations

Two turboexpanders and one dense fluid expander were used in our process. Both turboexpanders and the dense fluid expander were modeled using a heuristic provided by project author Adam Brostow. Sample calculations are provided below.

Type	***Range***	***Equation (Purchase cost w/ stainless steel, $, 2023)***
Turboexpander (Cryogenic)	Not given	$\frac{\$4000}{hp} required$

Table 24.5

<u>Sample calculation for High-level Turboexpander:</u>

For each of these cases, our ASPEN flowsheet provides a power recovered for each expander. These power requirements were **not** scaled up for contingencies, as is convention in cryogenics work.

$$Purchase\ \ cost\ (Material\ is\ stainless\ steel) = \frac{\$4000}{hp} \times (556\ hp) = \$2{,}224{,}000$$

Bare module cost was found by scaling the Purchase Cost (assuming stainless steel) by the bare module factor for compressors (F_{BM} = 3.3). To clarify, a material factor of 2.5 (stainless steel) was factored into the $4000/hp heuristic. All expanders require stainless steel material when they are operating at cryogenic temperatures.

24.7.3 KO Drum

KO Drum diameter and height were calculated using the Souders-Brown method. An aspect ratio of 3 was assumed for the drum. A demister is not used for our operation, so a vapor velocity factor (K_v) of 0.12 ft/s at 85% flooding was used, based on recommendation from project author Adam Brostow. A factor of 0.7 was applied to K_v in order to account for the mass of ferric oxide catalyst in the knockout drum impeding fluid movement in the drum. Ferric oxide catalyst will sit in a tray in the liquid of the Knockout Drum.

$$Maximum\ allowable\ vapor\ velocity\ v = K_v\sqrt{\frac{\rho_l}{\rho_v} - 1}$$

$$Drum\ cross\ sectional\ area\ A = \frac{\dot{V}}{v}$$

$$\dot{V}: vapor\ outlet\ volumetric\ flow\ rate$$

$$Drum\ diameter\ D = \sqrt{\frac{4A}{\pi}}$$

$$Drum\ height\ H = 3D$$

Vapor outlet volumetric flow rate, and vapor and liquid phase densities were taken from ASPEN PLUS stream results. Vessel thickness and vessel cost were calculated using correlations provided in *Product and Process Design (Seider et al.)*

Type	***Range***	***Design Correlation (Seider et al)***
Vessel thickness (t_p)	---	$t_p = P_d \times \frac{D_i}{2SE - 1.2 \times P_d}$
Vessel weight (W)	---	$W = \pi \times (D_i + t_p) \times (H + 0.8D_i) \times t_p$

Table 24.6

Values used in this calculation are listed here: design pressure of 52 psia, maximum allowable stress of 15,000 lb_f/in^2, fractional weld efficiency of 1. Vessel thickness of 0.3125 inches was calculated. Vessel weight (W) was calculated to be 282 lbs. A material factor F_M of 1.7 (stainless steel) was used.

Type	***Range***	***Equation $, 2013)***

Cost of Vessel (C_v)	---	$C_v = \exp\left(7.139 + 0.18255 \times \ln(W) + 0.02297 * \left(ln(W)\right)^2\right)$
Cost of Platforms and Ladders (C_{PL})	---	$C_{PL} = 410 * (D_i)^{0.7396} * (H)^{0.70684}$
Purchase Cost (C_p)	---	$C_p = F_M * C_v + C_{PL}$

Table 24.7

$$C_p = F_M * C_v + C_{PL} = \$13{,}838$$

This was multiplied by a bare module factor of 4.16 for vertical pressure vessels to yield the KO drum CBM.

24.7.4 PFHX (Plate-Frame Heat Exchanger) Correlations

UA values were taken from each MHeatX block in ASPEN Plus results. U ~ 150 Btu/lb-sqft-R was provided by an industrial consultant in cryogenics for Hydrogen liquefaction processes. All PFHX correlations were taken from *Product and Process Design (Seider et al.)*. These calculations match the elaborate heat exchanger calculations performed in an earlier section, and so can be used for costing purposes. For Blocks 26 and 30 where ferric oxide catalyst was packed into the heat exchanger, the area calculation is somewhat more involved and is covered in a previous section. A Material factor F_M of 1 (aluminum) was used for all cases.

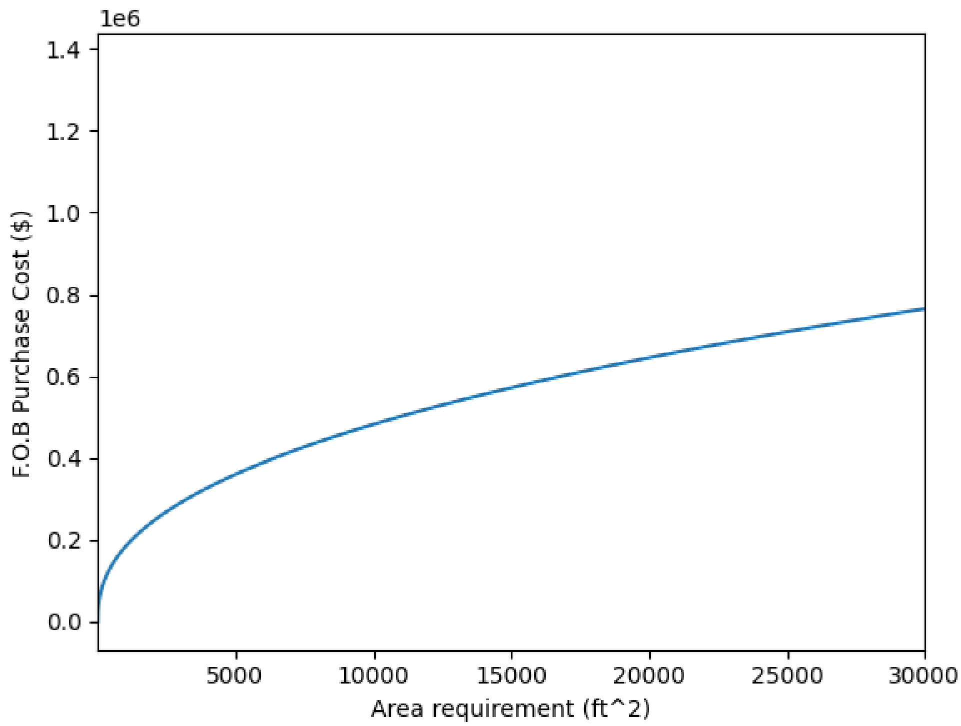
Type	***Range***	***Equation (F.O.B purchase cost, $, 2013)***
PFHX	150 ft^2 < A < 15,000 ft^2	$F.O.B\ purchase\ cost = 10070 * (A)^{0.42}$

Table 24.8

<u>Sample calculation for MHeatX-18</u>

$$F.O.B\ Purchase\ Cost = 10070 * (8306)^{0.42} = \$445{,}834$$

A bare module factor of 3.3 is taken for PFHXs, even though one is not provided in *Product and Process Design (Seider et al.)* PFHXs are assumed to have the same piping connections and instrumentation needs as Shell and Tube Heat Exchanger (F_{BM} = 3.3, *Seider et al*). A case could even be made for a higher bare module factor, given the intricate design of the BAHXs used in our process.



While the valid range of the correlation is from 150 ft^2 to 15,000 ft^2, the correlation itself is smooth, with no discontinuities past this point, and so can be used to estimate the F.O.B. purchase cost for Plate Frame heat exchangers (PFHXs) past A = 15,000 ft^2. However, A values over 15,000 ft^2 were subdivided into "N" smaller heat area values, representing "N" smaller heat exchangers. The smallest integer "N" was chosen, so that the resulting effective heat transfer areas of the "N" smaller heat exchangers was as close to 15,000 ft^2 as possible.

24.7.5 Cold Box & Storage vessel correlations

As suggested by Professor Bruce Vrana, the cold box was modeled as two concentric, vertical pressure vessels. The correlations used to price the cold box are identical to vertical pressure vessel sizing and costing correlations used in the KO Drum correlations section. Diameter of the cold box was taken to be 4 ft greater than the largest dimension of the largest heat exchanger (Block 18). Cold boxes in cryogenic plants typically have unit operations stacked vertically, so they are reasonably narrow. Chart Industries recently constructed a cold box with a diameter of 10 ft. and a height of 40 ft. for a process to liquefy 10 tons/day of Hydrogen. 10 tons/day is equivalent to 44 MTD, essentially the same as our process configuration. Therefore, our heuristic is reasonable. An aspect ratio of 3 was used to calculate cold box height for both inner and outer pressure vessels. The outer vessel diameter was taken to be 10% greater than the inner vessel diameter

$$Cold\ Box\ Diameter = 4\ ft + (8\ ft) = 12\ ft.$$

$$Cold\ Box\ Height = 3 * 12\ ft\ = 36\ ft$$

	Diameter (ft)	***Height (ft)***
Internal Pressure Vessel	12 ft.	36 ft
External Pressure Vessel	13.2 ft.	39.6 ft.

Calculations were performed according to Tables 24.6 and 24.7, using the subsequent values: design pressure of 52 psia, maximum allowable stress of 15,000 lb_f/in^2, fractional weld efficiency of 1. Vessel thickness of 0.3125 inches was calculated. Vessel weights and vessel thicknesses adjusted for hurricane loading are displayed below.

	Vessel Weight, W (lbs)	***Adjusted vessel thickness, t_p (in)***
Internal Pressure Vessel	35,274	0.5
External Pressure Vessel	42,668	0.5

According to above table, vessel purchase cost was calculated, as shown below. A material factor F_M of 1.7 (stainless steel) was used for both vessels.

$$C_p(Inner\ Pressure\ Vessel) = F_M * C_v + C_{PL} = \$212{,}218$$

$$C_p(Outer\ Pressure\ Vessel) = F_M * C_v + C_{PL} = \$241{,}328$$

This was multiplied by a bare module factor of 4.16 for vertical pressure vessels to yield the Cold Box Pressure Vessel CBMs. For both the cold box and the storage vessel, perlite insulation was packed in the annulus between both, at a price of $3.5/ft^3. The empty space between the internal and external pressure vessels was calculated by subtracting the volume of the inner pressure vessel from the volume of the outer pressure vessel. This volume was filled with perlite insulation and cost $3,500. The total bare module cost for the Cold Box was calculated by summing the Cold Box Pressure Vessel CBMs with the one-time purchase cost of perlite insulation.

This exact calculation was performed for the Hydrogen storage unit, except, the unit was assumed to be spherical, instead of vertical. Accordingly, a bare module factor of 3.05 was applied.

24.7.6 Electrolyzer Correlations

We used 800 Nm^3/hr (normal cubic meters per hour) alkaline electrolyzers from McPhy. McPhy's website did not contain purchase cost values, so we used the numbers given by a previous CBE senior design report from 2019 *(Synthesis of Green Hydrocarbons using the Air to Fuels Technology)* and extrapolated the price using the 2023 CE Index. This yielded a purchase cost of $2,200,000 per electrolyzer in 2023 dollars. 27 electrolyzers were purchased for our process.

24.7.7 80K Adsorber Correlations

No information was found on this online, so a purchase cost of $100,000 was estimated, based on feedback from an industrial consultant in cryogenics.

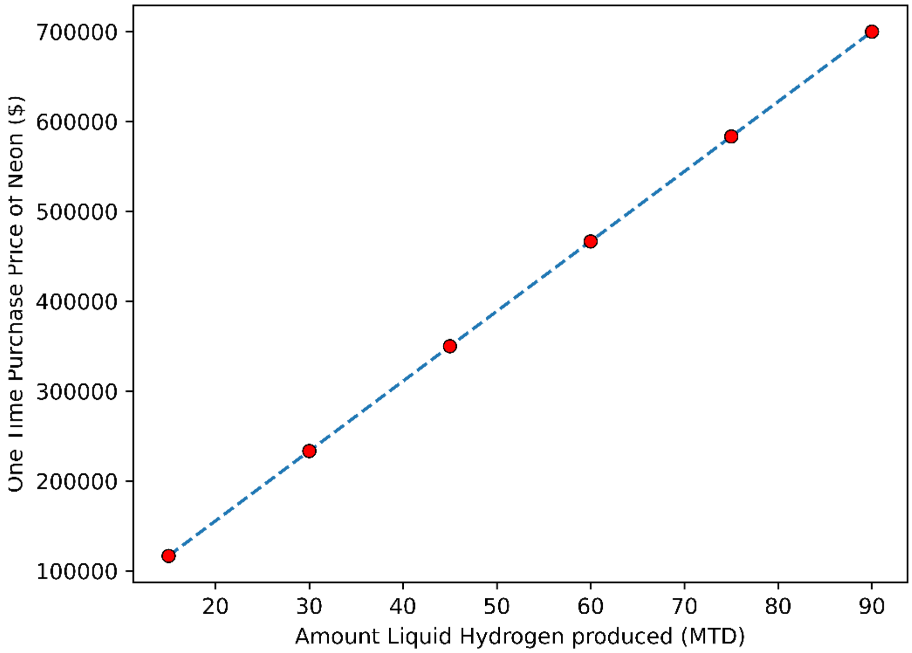
24.8 Documentation of Approximate Profitability Analysis

24.8.1 Extrapolation of One Time Purchase cost of Working Fluid from 45 MTD case

24.8.1.1 Neon

Amount of Liquid Hydrogen Produced (MTD)	***One-time purchase cost of Neon***
15	$116,667
30	$233,334
45	$350,000
60	$466,668
75	$583,335
90	$700,002

For approximate profitability analysis spreadsheets, a linear extrapolation is performed in order to predict the one-time purchase cost of Neon at process configurations where more/less than 45 MTD of liquid Hydrogen is produced per day. This extrapolation was performed for helium and Hydrogen in Sections 24.7.1.2 and 24.7.1.3.



24.8.1.2 Helium

As estimated in 24.1.3, the total mass of Neon required for 45 MTD process is 220 lb.

Amount of helium required is calculated using the following heuristic:

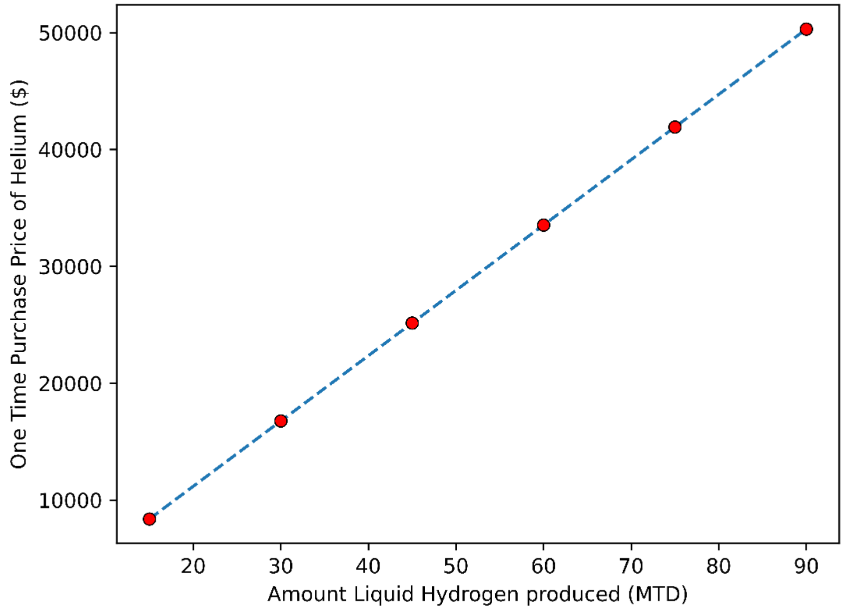
$$(220\ lb\ Ne)\left(20.18\frac{g\ Ne}{mol\ Ne}\right) = (\ x\ lb\ He)\left(4.003\frac{g\ He}{mol\ He}\right) --> x = 1109\ lb\ He$$

Unit price of helium is \$0.05/g

Below is the calculation for the one-time purchase cost for He for 45 MTD case

$$(1109\ lb\ He) * \left(\frac{\$0.05}{g\ He}\right) * \left(\frac{1000\ g\ He}{1\ kg\ He}\right) * \left(\frac{1\ kg}{2.2046\ lb}\right) = \mathbf{\$25,152}$$

The same linear extrapolation was performed as was done in Section 24.7.1.1



Amount of Liquid Hydrogen Produced (MTD)	***One-time purchase cost of Helium***
15	$8,384
30	$16,768
45	$25,152
60	$33,536
75	$41,920
90	$50,304

24.8.1.3 Hydrogen

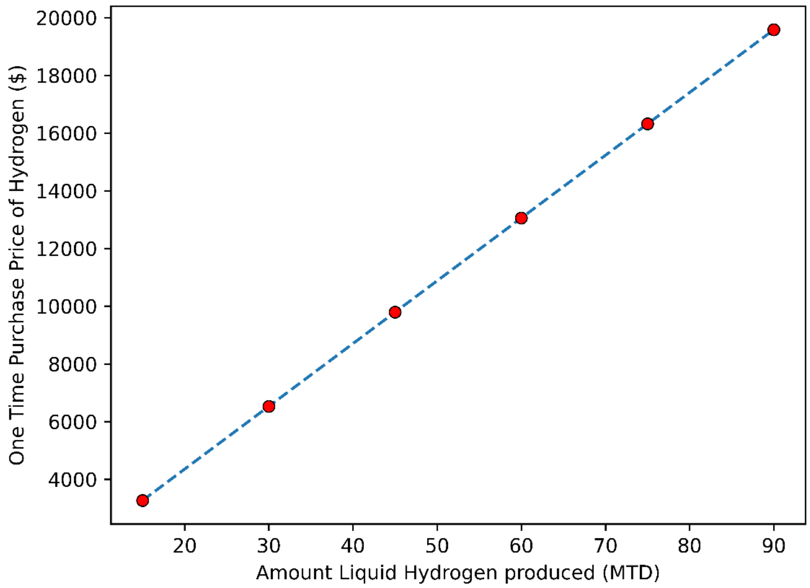
Amount of Hydrogen required is calculated using the following heuristic:

$$(220\ lb\ Ne)\left(20.18\frac{g\ Ne}{mol\ Ne}\right) = (\ x\ lb\ H_2)\left(2.016\frac{g\ H_2}{mol\ H_2}\right) --> x = 2202\ lb\ H_2$$

Unit price of green vapor Hydrogen is $9.80/kg (Our sales price of $13/kg – cost of liquefaction of $3.20/kg)

$$(2202\ lb\ H_2) * \left(\frac{\$9.80}{kg\ H_2}\right) * \left(\frac{1\ kg}{2.204\ lb}\right) = \$\mathbf{9791}$$

The same linear extrapolation was performed as was done in Section 24.7.1.1



Amount of Liquid Hydrogen Produced (MTD)	***One-time purchase cost of Hydrogen***
15	$3,264
30	$6,527
45	$9,791
60	$13,055
75	$16,318
90	$19,582

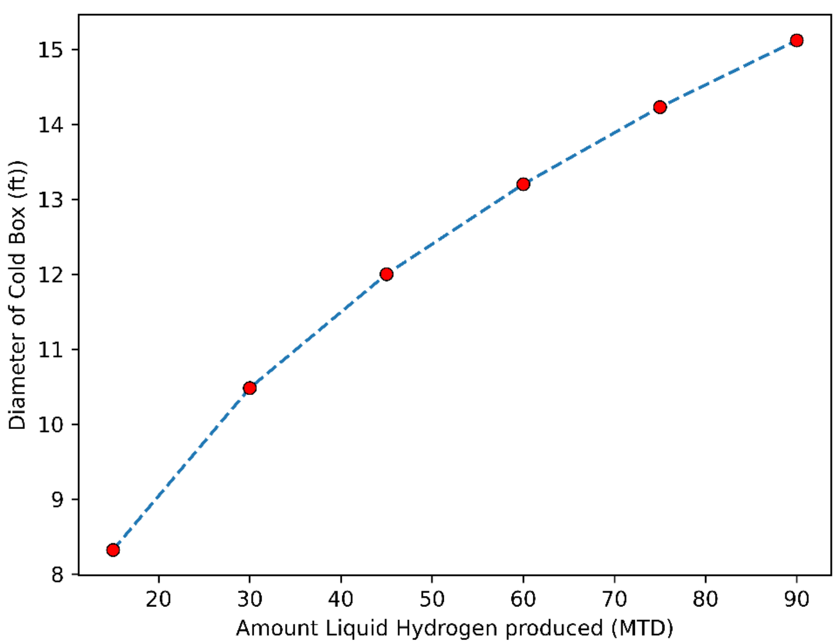
24.8.2 Extrapolation of Cold Box diameter from 45 MTD case

Amount of Liquid Hydrogen Produced (MTD)	*Cold Box Diameter (ft)*
15	8.32
30	10.48
<u>45</u>	<u>12</u>
60	13.2
75	14.2
90	15.1

Cold Box diameter was for 45 MTD case was calculated by adding 4 ft to largest dimension of largest heat exchanger. (Chart 15 TPD cold box, in an email). Extrapolated diameters calculated using the following relation:

$$D_{new} = \left(\frac{Amount\ LH_2\ Produced}{45}\right)^{\frac{1}{3}}$$

Height of the cold box is taken to be three times the diameter.



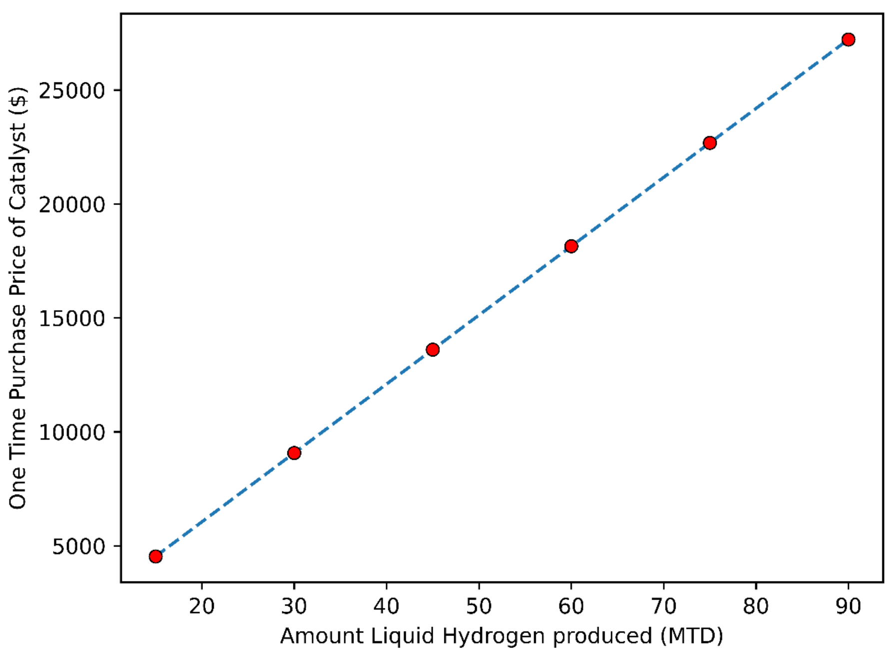
24.8.3 Extrapolation of Amount of Catalyst

Amount of Liquid Hydrogen Produced (MTD)	*One time purchase cost of Ferric Oxide catalyst ($)*
15	$4536
30	$9072
45	$13,608
60	$18,144
75	$22,680
90	$27,215

Ferric Oxide catalyst required for 45 MTD case was calculated by (BLANK METHOD) to be 1500 lb. Extrapolated one time catalyst purchase cost calculated using the following relation:

$$(1500\ lb\ FeO) * \left(\frac{\$0.02}{g}\right) * \left(\frac{1000\ g}{kg}\right) * \left(\frac{1\ kg}{2.204\ lb}\right) = \$\mathbf{13,608}$$

The same linear extrapolation was performed as was done in Section 24.7.1.1



24.8.4 Summary of Approximate Profitability analysis on Neon

	15 MTD	***30 MTD***	***45 MTD***	***60 MTD***	***75 MTD***	***90 MTD***
One time purchase Cost of ferric oxide catalyst	\$4,536	\$9,072	\$13,608	\$18,144	\$22,680	\$27,215
One time purchase cost of Neon refrigerant	\$116,667	\$233,334	\$350,000	\$466,668	\$583,335	\$700,002
Total Bare Module Cost	\$54.8 MM	\$101.1 MM	\$146.5 MM	\$190.9 MM	\$233.2 MM	\$278.5 MM
Profit	\$12.4 MM	\$27.9 MM	\$43.6 MM	\$59.5 MM	\$76.1 MM	\$91.8 MM
Total Capital Investment	\$100.6 MM	\$187.3 MM	\$262.6 MM	\$356.0 MM	\$436.3 MM	\$521.3 MM
ROI (3rd Year)	12.36%	14.88%	16.59%	16.72%	17.43%	17.61%

15 MTD, Ne, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)	CE Index (2013)	
CBM of Compressors	$6,932,581.97	$10,025,956.29	567	
CBM of Expanders	$2,799,994.37	$4,049,374.57		(=Rigorous Calculation, Guthrie method)
CBM of Heat Exchangers	$2,592,868.95	$3,749,828.11	CE Index (2023)	(=value that must be entered)
CBM of KO Drum	$20,106.20	$29,077.75	820	(=Final Value)
CBM of Electrolyzers	$20,536,463.41	$29,700,000.00		(=Seider et. al, 2013 convention)
CBM of Cold Box	$856,281.57	$1,238,361.36	Investment Site Factor (2023)	(=Intermediate calculation)
CBM of Spherical Storage Vessel	$4,041,466.77	$5,844,802.04	1	
CBM- All Process Equipment	---	$54,637,400.10		
One-time Purchase Cost of Catalyst	---	$4,536.00		
One-time Purchase Cost of Refrigerant	---	$116,667.00		
Total CBM	---	$54,758,603.10		

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$5,463,740.01	$5,463,740.01	$0.00	$54,637,400.10	$65,564,880.12
	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)	
Total Depreciable Capital: C(TDC)	$11,801,678.42	$65,564,880.12	$77,366,558.55		
	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$7,736,655.85	$3,477,831.17	$1,547,331.17	$77,366,558.55	$90,128,376.74
	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$3,706,984.13	$1,235,520.00	$5,506,629.85	$779.02	$10,449,912.99
	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$90,128,376.74	$10,449,912.99	$100,578,289.74		$14,850,000.00
	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$64,350,000.00	$51,918,013.93	$12,431,986.07	$4,972,794.43
					IRA-Adjustment to Tax ($/yr)
	Profit	Total Capital Investment: C(TCI)	ROI (%)		0
Return on Investment (ROI)	$12,431,986.07	$100,578,289.74	12.36%		Net Earnings ($/yr)
					$12,431,986.07

Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency
5228	15	6.2	2.768	44.4%

30 MTD, Ne, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)
CBM of Compressors	$10,621,547.62	$15,360,968.34
CBM of Expanders	$5,644,917.03	$8,163,724.80
CBM of Heat Exchangers	$4,253,043.99	$6,150,786.73
CBM of KO Drum	$34,992.28	$50,606.12
CBM of Electrolyzers	$41,072,926.83	$59,400,000.00
CBM of Cold Box	$1,160,976.68	$1,679,013.90
CBM of Spherical Storage Vessel	$6,939,727.37	$10,036,290.02
CBM- All Process Equipment	---	$100,841,389.91
One-time Purchase Cost of Catalyst	---	$9,072.00
One-time Purchase Cost of Refrigerant	---	$233,334.00
Total CBM	---	$101,083,795.91

CE Index (2013)
567
CE Index (2023)
820
Investment Site Factor (2023)
1

(=Rigorous Calculation, Guthrie method)
(=value that must be entered)
(=Final Value)
(=Seider et. al, 2013 convention)
(=Intermediate calculation)

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$10,084,138.99	$10,084,138.99	$0.00	$100,841,389.91	$121,009,667.89
	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)	
Total Depreciable Capital: C(TDC)	$21,781,740.22	$121,009,667.89	$142,791,408.11		
	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$14,279,140.81	$6,716,828.16	$2,855,828.16	$142,791,408.11	$166,643,205.25
	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$7,162,985.79	$2,471,040.00	$11,013,259.69	$1,558.05	$20,648,843.53
	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$166,643,205.25	$20,648,843.53	$187,292,048.78		$29,700,000.00
	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$128,700,000.00	$100,824,117.88	$27,875,882.12	$11,150,352.85
					IRA-Adjustment to Tax ($/yr)
	Profit	Total Capital Investment: C(TCI)	ROI (%)		0
Return on Investment (ROI)	$27,875,882.12	$187,292,048.78	14.88364418		Net Earnings ($/yr)
					$27,875,882.12

Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency
10457	30	6.2	2.768	44.4%

45 MTD, Ne, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)	CE Index (2013)		
CBM of Compressors	$14,424,928.58	$20,861,448.74	567		
CBM of Expanders	$8,476,020.34	$12,258,089.38		(=Rigorous Calculation, Guthrie method)	
CBM of Heat Exchangers	$5,042,642.34	$7,292,710.26	CE Index (2023)	(=value that must be entered)	
CBM of KO Drum	$68,182.45	$98,606.02	820	(=Final Value)	
CBM of Electrolyzers	$61,609,390.24	$89,100,000.00		(=Seider et. al, 2013 convention)	
CBM of Cold Box	$1,900,415.76	$2,748,396.69	Investment Site Factor (2023)	(=Intermediate calculation)	
CBM of Spherical Storage Vessels	$9,521,662.47	$13,770,305.51	1		
CBM- All Process Equipment	---	$146,129,556.58			
One-time Purchase Cost of Catalyst	---	$13,608.00			
One-time Purchase Cost of Refrigerant	---	$350,000.00			
Total CBM	---	$146,493,164.58			

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$14,612,955.66	$14,612,955.66	$0.00	$146,129,556.58	$175,355,467.90
	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)	
Total Depreciable Capital: C(TDC)	$31,563,984.22	$175,355,467.90	$206,919,452.12		
	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$20,691,945.21	0	$4,138,389.04	$206,919,452.12	$231,749,786.38
	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$10,598,352.12	$3,706,560.00	$16,519,889.54	$2,337.07	$30,827,138.73
	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$231,749,786.38	$30,827,138.73	$262,576,925.11		$44,550,000.00
	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$193,050,000.00	$149,482,587.87	$43,567,412.13	$17,426,964.85
					IRA-Adjustment to Tax ($/yr)
	Profit	Total Capital Investment: C(TCI)	ROI (%)		0
Return on Investment (ROI)	$43,567,412.13	$262,576,925.11	16.59224706		Net Earnings ($/yr)
					$43,567,412.13
Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency	
15684	45	6.2	2.768	44.4%	

60 MTD, Ne, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)	CE Index (2013)	
CBM of Compressors	$18,054,123.53	$26,110,019.92	567	
CBM of Expanders	$11,258,837.45	$16,282,622.06		(=Rigorous Calculation, Guthrie method)
CBM of Heat Exchangers	$6,629,746.33	$9,587,992.92	**CE Index (2023)**	(=value that must be entered)
CBM of KO Drum	$48,355.25	$69,931.76	820	(=Final Value)
CBM of Electrolyzers	$82,145,853.66	$118,800,000.00		(=Seider et. al, 2013 convention)
CBM of Cold Box	$1,588,828.68	$2,297,776.92	**Investment Site Factor (2023)**	(=Intermediate calculation)
CBM of Spherical Storage Vessel	$11,916,420.11	$17,233,623.45	1	
CBM- All Process Equipment	---	$190,381,967.04		
One-time Purchase Cost of Catalyst	---	$18,144.00		
One-time Purchase Cost of Refrigerant	---	$466,668.00		
Total CBM	---	$190,866,779.04		

	Site Preparation Cost: C(Site)	**Service Facilities Cost: C(Service)**	**Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)**	**Total Bare Module Cost: C(TBM)**	**Final C(DPI), Guthrie, ($/yr)**
Direct Permanent Investment: C(DPI)	$19,038,196.70	$19,038,196.70	$0.00	$190,381,967.04	$228,458,360.45
	Contingencies Cost: C(Cont)	**Direct Permanent Investment: C(DPI)**	**Final C(TDC), Guthrie, ($/yr)**	**Reportable C(TDC) (Including Refrigerant), ($/yr)**	
Total Depreciable Capital: C(TDC)	$41,122,504.88	$228,458,360.45	$269,580,865.33		
	Startup Cost: C(Startup)	**Royalties Cost: C(Royalties)**	**Land Cost: C(Land)**	**Total Depreciable Capital: C(TDC)**	**Final C(TPI), Guthrie, ($/yr)**
Total Permanent Investment: C(TPI)	$26,958,086.53	$13,113,617.31	$5,391,617.31	$269,580,865.33	$315,044,186.47
	Cash reserves	**Inventory**	**Accounts Receivable**	**Accounts Payable**	**Final C(WC), Guthrie, ($/yr)**
Working Capital: C(WC)	$14,010,780.59	$4,942,080.00	$22,026,519.38	$3,116.10	$40,982,496.07
	Total Permanent Investment: C(TPI)	**Working Capital: C(WC)**	**Final C(TCI), Guthrie, ($/yr)**		**IRA Tax Credit ($/yr)**
Total Capital Investment: C(TCI)	$315,044,186.47	$40,982,496.07	$356,026,682.54		$59,400,000.00
	Total Tax Rate (State + Federal)	**Annual Sales Revenue**	**Annual Production Cost**	**Gross Earnings ($/yr)**	**Tax ($/yr)**
Net Earnings (Profit)	0.4	$257,400,000.00	$197,865,792.58	$59,534,207.42	$23,813,682.97
					IRA-Adjustment to Tax ($/yr)
	Profit	**Total Capital Investment: C(TCI)**	**ROI (%)**		0
Return on Investment (ROI)	$59,534,207.42	$356,026,682.54	**16.72183865**		**Net Earnings ($/yr)**
					$59,534,207.42

Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency
20919	60	6.2	2.768	44.4%

75 MTD, Ne, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)	CE Index (2013)	
CBM of Compressors	$22,057,155.65	$31,899,237.45	567	
CBM of Expanders	$14,310,591.09	$20,696,092.93		(=Rigorous Calculation, Guthrie method)
CBM of Heat Exchangers	$8,095,302.73	$11,707,492.48	**CE Index (2023)**	(=value that must be entered)
CBM of KO Drum	$53,888.47	$77,933.94	820	(=Final Value)
CBM of Electrolyzers	$100,400,487.80	$145,200,000.00		(=Seider et. al, 2013 convention)
CBM of Cold Box	$1,761,251.54	$2,547,136.26	**Investment Site Factor (2023)**	(=Intermediate calculation)
CBM of Spherical Storage Vessel	$14,181,939.10	$20,510,035.38	1	
CBM- All Process Equipment	---	$232,637,928.44		
One-time Purchase Cost of Catalyst	---	$22,680.00		
One-time Purchase Cost of Refrigerant	---	$583,335.00		
Total CBM	---	$233,243,943.44		

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$23,263,792.84	$23,263,792.84	$0.00	$232,637,928.44	$279,165,514.12

	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)
Total Depreciable Capital: C(TDC)	$50,249,792.54	$279,165,514.12	$329,415,306.67	

	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$32,941,530.67	$16,240,806.13	$6,588,306.13	$329,415,306.67	$385,185,949.60

	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$17,376,726.88	$6,177,600.00	$27,533,149.23	$3,895.12	$51,091,371.23

	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$385,185,949.60	$51,091,371.23	$436,277,320.83		$74,250,000.00

	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$321,750,000.00	$245,691,188.75	$76,058,811.25	$30,423,524.50
					IRA-Adjustment to Tax ($/yr)
	Profit	**Total Capital Investment: C(TCI)**	**ROI (%)**		0
Return on Investment (ROI)	$76,058,811.25	$436,277,320.83	17.43359272		**Net Earnings ($/yr)**
					$76,058,811.25

Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency
26120	75	6.2	2.768	44.4%

90 MTD, Ne, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)
CBM of Compressors	$26,065,267.29	$37,695,801.01
CBM of Expanders	$17,172,710.04	$24,835,312.58
CBM of Heat Exchangers	$9,602,591.25	$13,887,345.38
CBM of KO Drum	$58,982.07	$85,300.35
CBM of Electrolyzers	$120,936,951.22	$174,900,000.00
CBM of Cold Box	$1,917,391.32	$2,772,946.89
CBM of Spherical Storage Vessel	$16,349,218.12	$23,644,371.89
CBM- All Process Equipment	---	$277,821,078.09
One-time Purchase Cost of Catalyst	---	$27,215.00
One-time Purchase Cost of Refrigerant	---	$700,002.00
Total CBM	---	$278,548,295.09

CE Index (2013)	
567	
	(=Rigorous Calculation, Guthrie method)
CE Index (2023)	(=value that must be entered)
820	(=Final Value)
	(=Seider et. al, 2013 convention)
Investment Site Factor (2023)	(=Intermediate calculation)
1	

	Site Preparation Cost: C(Site)	**Service Facilities Cost: C(Service)**	**Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)**	**Total Bare Module Cost: C(TBM)**	**Final C(DPI), Guthrie, ($/yr)**
Direct Permanent Investment: C(DPI)	$27,782,107.81	$27,782,107.81	$0.00	$277,821,078.09	$333,385,293.71
	Contingencies Cost: C(Cont)	**Direct Permanent Investment: C(DPI)**	**Final C(TDC), Guthrie, ($/yr)**	**Reportable C(TDC) (Including Refrigerant), ($/yr)**	
Total Depreciable Capital: C(TDC)	$60,009,352.87	$333,385,293.71	$393,394,646.58		
	Startup Cost: C(Startup)	**Royalties Cost: C(Royalties)**	**Land Cost: C(Land)**	**Total Depreciable Capital: C(TDC)**	**Final C(TPI), Guthrie, ($/yr)**
Total Permanent Investment: C(TPI)	$39,339,464.66	$19,450,892.93	$7,867,892.93	$393,394,646.58	$460,052,897.10
	Cash reserves	**Inventory**	**Accounts Receivable**	**Accounts Payable**	**Final C(WC), Guthrie, ($/yr)**
Working Capital: C(WC)	$20,809,565.58	$7,413,120.00	$33,039,779.07	$4,674.15	$61,267,138.80
	Total Permanent Investment: C(TPI)	**Working Capital: C(WC)**	**Final C(TCI), Guthrie, ($/yr)**		**IRA Tax Credit ($/yr)**
Total Capital Investment: C(TCI)	$460,052,897.10	$61,267,138.80	$521,320,035.90		$89,100,000.00
	Total Tax Rate (State + Federal)	**Annual Sales Revenue**	**Annual Production Cost**	**Gross Earnings ($/yr)**	**Tax ($/yr)**
Net Earnings (Profit)	0.4	$386,100,000.00	$294,319,325.97	$91,780,674.03	$36,712,269.61
					IRA-Adjustment to Tax ($/yr)
	Profit	**Total Capital Investment: C(TCI)**	**ROI (%)**		0
Return on Investment (ROI)	$91,780,674.03	$521,320,035.90	**17.60543768**		**Net Earnings ($/yr)**
					$91,780,674.03
Total Power requirement (hp)	**Production (tonnes LH2/day)**	**Specific power (kWh/kg LH2)**	**Minimum specific power (kWh/kg LH2)**	**Efficiency**	
31344	90	6.2	2.768	44.4%	

24.8.5 Summary of Approximate Profitability analysis on Helium

	15 MTD	*30 MTD*	*45 MTD*	*60 MTD*	*75 MTD*	*90 MTD*
One time purchase Cost of ferric oxide catalyst	$4,536	$9,072	$13,608	$18,144	$22,680	$27,215
One time purchase cost of Helium refrigerant	$8,384	$16,768	$25,152	$33,536	$41,920	$50,304
Total Bare Module Cost	$56.8 MM	$109.7 MM	$161.9 MM	$215.3 MM	$263.0 MM	$316.9 MM
Profit	$11.7 MM	$25.2 MM	$39.6 MM	$53.1 MM	$68.4 MM	$81.7 MM
Total Capital Investment	$104.2 MM	$201.8 MM	$298.6 MM	$397.0 MM	$486.3 MM	$585.7 MM
ROI (3^{rd} Year)	11.20%	12.50%	13.28%	13.38%	14.07%	13.95%

15 MTD, He, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)
CBM of Compressors	$8,153,147.04	$11,791,147.39
CBM of Expanders	$3,173,338.14	$4,589,307.37
CBM of Heat Exchangers	$2,509,747.44	$3,629,617.10
CBM of KO Drum	$29,370.50	$42,475.85
CBM of Electrolyzers	$20,536,463.41	$29,700,000.00
CBM of Cold Box	$856,281.57	$1,238,361.36
CBM of Spherical Storage Vessel	$4,041,466.77	$5,844,802.04
CBM- All Process Equipment	---	$56,835,711.11
One-time Purchase Cost of Catalyst	---	$4,536.00
One-time Purchase Cost of Refrigerant	---	$8,384.00
Total CBM	---	$56,848,631.11

CE Index (2013)
567

CE Index (2023)
820

Investment Site Factor (2023)
1

(=Rigorous Calculation, Guthrie method)
(=value that must be entered)
(=Final Value)
(=Seider et. al, 2013 convention)
(=Intermediate calculation)

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$5,683,571.11	$5,683,571.11	$0.00	$56,835,711.11	$68,202,853.33

	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)
Total Depreciable Capital: C(TDC)	$12,276,513.60	$68,202,853.33	$80,479,366.93	

	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$8,047,936.69	$3,540,087.34	$1,609,587.34	$80,479,366.93	$93,676,978.30

	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$3,770,658.08	$1,235,520.00	$5,506,629.85	$779.02	$10,513,586.95

	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$93,676,978.30	$10,513,586.95	$104,190,565.25		$14,850,000.00

	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$64,350,000.00	$52,682,132.00	$11,667,868.00	$4,667,147.20
					IRA-Adjustment to Tax ($/yr)
	Profit	Total Capital Investment: C(TCI)	ROI (%)		0
Return on Investment (ROI)	$11,667,868.00	$104,190,565.25	11.19858403		Net Earnings ($/yr)
					$11,667,868.00

Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency
5551	15	6.6	2.768	41.8%

30 MTD, He, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)	CE Index (2013)		
CBM of Compressors	$16,130,645.70	$23,328,270.68	567		
CBM of Expanders	$6,299,829.41	$9,110,864.40		(=Rigorous Calculation, Guthrie method)	
CBM of Heat Exchangers	$4,208,273.20	$6,086,038.84	CE Index (2023)	(=value that must be entered)	
CBM of KO Drum	$35,083.55	$50,738.12	820	(=Final Value)	
CBM of Electrolyzers	$41,072,926.83	$59,400,000.00		(=Seider et. al, 2013 convention)	
CBM of Cold Box	$1,160,976.68	$1,679,013.90	Investment Site Factor (2023)	(=Intermediate calculation)	
CBM of Spherical Storage Vessel	$6,939,727.37	$10,036,290.02	1		
CBM- All Process Equipment	---	$109,691,215.96			
One-time Purchase Cost of Catalyst	---	$9,072.00			
One-time Purchase Cost of Refrigerant	---	$16,768.00			
Total CBM	---	$109,717,055.96			

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$10,969,121.60	$10,969,121.60	$0.00	$109,691,215.96	$131,629,459.15
	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)	
Total Depreciable Capital: C(TDC)	$23,693,302.65	$131,629,459.15	$155,322,761.80		
	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$15,532,276.18	$6,967,455.24	$3,106,455.24	$155,322,761.80	$180,928,948.45
	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$7,384,322.11	$2,471,040.00	$11,013,259.69	$1,558.05	$20,870,179.85
	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$180,928,948.45	$20,870,179.85	$201,799,128.30		$29,700,000.00
	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$128,700,000.00	$103,480,259.97	$25,219,740.03	$10,087,896.01
					IRA-Adjustment to Tax ($/yr)
	Profit	Total Capital Investment: C(TCI)	ROI (%)		0
Return on Investment (ROI)	$25,219,740.03	$201,799,128.30	12.49744746		Net Earnings ($/yr)
					$25,219,740.03
Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency	
10965	30	6.5	2.7	41.3%	

45 MTD, He, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)
CBM of Compressors	$24,196,149.50	$34,992,667.71
CBM of Expanders	$9,459,585.19	$13,680,528.84
CBM of Heat Exchangers	$5,714,714.06	$8,264,665.84
CBM of KO Drum	$42,203.67	$61,035.30
CBM of Electrolyzers	$61,609,390.24	$89,100,000.00
CBM of Cold Box	$1,393,333.67	$2,015,050.46
CBM of Spherical Storage Vessel	$9,521,662.47	$13,770,305.51
CBM- All Process Equipment	---	$161,884,253.65
One-time Purchase Cost of Catalyst	---	$13,608.00
One-time Purchase Cost of Refrigerant	---	$25,152.00
Total CBM	---	$161,923,013.65

CE Index (2013)	
567	
	(=Rigorous Calculation, Guthrie method)
CE Index (2023)	(=value that must be entered)
820	(=Final Value)
	(=Seider et. al, 2013 convention)
Investment Site Factor (2023)	(=Intermediate calculation)
1	

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$16,188,425.36	$16,188,425.36	$0.00	$161,884,253.65	$194,261,104.38

	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)
Total Depreciable Capital: C(TDC)	$34,966,998.79	$194,261,104.38	$229,228,103.17	

	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$22,922,810.32	$10,419,189.43	$4,584,562.06	$229,228,103.17	$267,154,664.98

	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$11,031,479.72	$3,734,161.52	$16,642,907.65	$2,354.48	$31,410,903.36

	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$267,154,664.98	$31,410,903.36	$298,565,568.34		$44,881,749.00

	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$194,487,579.00	$154,846,367.39	$39,641,211.61	$15,856,484.64
					IRA-Adjustment to Tax ($/yr)
	Profit	**Total Capital Investment: C(TCI)**	**ROI (%)**		0
Return on Investment (ROI)	$39,641,211.61	$298,565,568.34	13.27722142		**Net Earnings ($/yr)**
					$39,641,211.61

Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency
16447	45.3351	6.5	2.768	42.6%

60 MTD, He, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)	CE Index (2013)	
CBM of Compressors	$32,612,938.10	$47,165,095.67	567	
CBM of Expanders	$12,787,699.80	$18,493,675.20		(=Rigorous Calculation, Guthrie method)
CBM of Heat Exchangers	$7,706,192.99	$11,144,758.83	CE Index (2023)	(=value that must be entered)
CBM of KO Drum	$48,355.25	$69,931.76	820	(=Final Value)
CBM of Electrolyzers	$82,145,853.66	$118,800,000.00		(=Seider et. al, 2013 convention)
CBM of Cold Box	$1,588,828.68	$2,297,776.92	Investment Site Factor (2023)	(=Intermediate calculation)
CBM of Spherical Storage Vessel	$11,916,420.11	$17,233,623.45	1	
CBM- All Process Equipment	---	$215,204,861.83		
One-time Purchase Cost of Catalyst	---	$18,144.00		
One-time Purchase Cost of Refrigerant	---	$33,536.00		
Total CBM	---	$215,256,541.83		

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$21,520,486.18	$21,520,486.18	$0.00	$215,204,861.83	$258,245,834.19
	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)	
Total Depreciable Capital: C(TDC)	$46,484,250.15	$258,245,834.19	$304,730,084.35		
	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$30,473,008.43	$13,874,104.85	$6,094,601.69	$304,730,084.35	$355,171,799.32
	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$14,686,042.05	$4,978,882.02	$22,190,543.53	$3,139.30	$41,858,606.90
	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$355,171,799.32	$41,858,606.90	$397,030,406.22		$59,842,332.00
	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$259,316,772.00	$206,190,641.29	$53,126,130.71	$21,250,452.28
					IRA-Adjustment to Tax ($/yr)
	Profit	Total Capital Investment: C(TCI)	ROI (%)		0
Return on Investment (ROI)	$53,126,130.71	$397,030,406.22	13.38087206		Net Earnings ($/yr)
					$53,126,130.71
Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency	
22206	60.4468	6.6	2.7	41.1%	

75 MTD, He, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)	CE Index (2013)	
CBM of Compressors	$40,326,683.62	$58,320,777.02	567	
CBM of Expanders	$15,763,932.62	$22,797,927.24		(=Rigorous Calculation, Guthrie method)
CBM of Heat Exchangers	$9,315,662.84	$13,472,387.17	**CE Index (2023)**	(=value that must be entered)
CBM of KO Drum	$53,888.47	$77,933.94	820	(=Final Value)
CBM of Electrolyzers	$100,400,487.80	$145,200,000.00		(=Seider et. al, 2013 convention)
CBM of Cold Box	$1,761,251.54	$2,547,136.26	**Investment Site Factor (2023)**	(=Intermediate calculation)
CBM of Spherical Storage Vessel	$14,181,939.10	$20,510,035.38	1	
CBM- All Process Equipment	---	$262,926,197.01		
One-time Purchase Cost of Catalyst	---	$22,680.00		
One-time Purchase Cost of Refrigerant	---	$41,920.00		
Total CBM	---	$262,990,797.01		

	Site Preparation Cost: C(Site)	**Service Facilities Cost: C(Service)**	**Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)**	**Total Bare Module Cost: C(TBM)**	**Final C(DPI), Guthrie, ($/yr)**
Direct Permanent Investment: C(DPI)	$26,292,619.70	$26,292,619.70	$0.00	$262,926,197.01	$315,511,436.41
	Contingencies Cost: C(Cont)	**Direct Permanent Investment: C(DPI)**	**Final C(TDC), Guthrie, ($/yr)**	**Reportable C(TDC) (Including Refrigerant), ($/yr)**	
Total Depreciable Capital: C(TDC)	$56,792,058.55	$315,511,436.41	$372,303,494.97		
	Startup Cost: C(Startup)	**Royalties Cost: C(Royalties)**	**Land Cost: C(Land)**	**Total Depreciable Capital: C(TDC)**	**Final C(TPI), Guthrie, ($/yr)**
Total Permanent Investment: C(TPI)	$37,230,349.50	$17,170,435.98	$7,446,069.90	$372,303,494.97	$434,150,350.34
	Cash reserves	**Inventory**	**Accounts Receivable**	**Accounts Payable**	**Final C(WC), Guthrie, ($/yr)**
Working Capital: C(WC)	$18,190,171.76	$6,223,594.29	$27,738,142.70	$3,924.12	$52,155,832.88
	Total Permanent Investment: C(TPI)	**Working Capital: C(WC)**	**Final C(TCI), Guthrie, ($/yr)**		**IRA Tax Credit ($/yr)**
Total Capital Investment: C(TCI)	$434,150,350.34	$52,155,832.88	$486,306,183.22		$74,802,816.00
	Total Tax Rate (State + Federal)	**Annual Sales Revenue**	**Annual Production Cost**	**Gross Earnings ($/yr)**	**Tax ($/yr)**
Net Earnings (Profit)	0.4	$324,145,536.00	$255,729,602.18	$68,415,933.82	$27,366,373.53
					IRA-Adjustment to Tax ($/yr)
	Profit	**Total Capital Investment: C(TCI)**	**ROI (%)**		0
Return on Investment (ROI)	$68,415,933.82	$486,306,183.22	**14.0684894**		**Net Earnings ($/yr)**
					$68,415,933.82
Total Power requirement (hp)	**Production (tonnes LH2/day)**	**Specific power (kWh/kg LH2)**	**Minimum specific power (kWh/kg LH2)**	**Efficiency**	
27412	75.5584	6.5	2.768	42.6%	

90 MTD, He, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)
CBM of Compressors	$48,920,033.48	$70,748,549.30
CBM of Expanders	$19,181,536.01	$27,740,493.00
CBM of Heat Exchangers	$11,692,106.20	$16,909,218.84
CBM of KO Drum	$58,982.07	$85,300.35
CBM of Electrolyzers	$120,936,951.22	$174,900,000.00
CBM of Cold Box	$1,917,391.32	$2,772,946.89
CBM of Spherical Storage Vessel	$16,349,218.12	$23,644,371.89
CBM- All Process Equipment	---	$316,800,880.26
One-time Purchase Cost of Catalyst	---	$27,215.00
One-time Purchase Cost of Refrigerant	---	$50,304.00
Total CBM	---	$316,878,399.26

CE Index (2013)
567

CE Index (2023)
820

Investment Site Factor (2023)
1

(=Rigorous Calculation, Guthrie method)

(=value that must be entered)

(=Final Value)

(=Seider et. al, 2013 convention)

(=Intermediate calculation)

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$31,680,088.03	$31,680,088.03	$0.00	$316,800,880.26	$380,161,056.32

	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)
Total Depreciable Capital: C(TDC)	$68,428,990.14	$380,161,056.32	$448,590,046.45	

	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$44,859,004.65	$20,641,042.80	$8,971,800.93	$448,590,046.45	$523,061,894.83

	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$21,863,352.15	$7,468,314.80	$33,285,778.58	$4,708.95	$62,622,154.47

	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$523,061,894.83	$62,622,154.47	$585,684,049.30		$89,763,399.00

	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$388,974,729.00	$307,297,301.79	$81,677,427.21	$32,670,970.88
					IRA-Adjustment to Tax ($/yr)
	Profit	**Total Capital Investment: C(TCI)**	**ROI (%)**		0
Return on Investment (ROI)	$81,677,427.21	$585,684,049.30	**13.94564652**		**Net Earnings ($/yr)**
					$81,677,427.21

Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency
33309	90.6701	6.6	2.768	42.1%

24.8.6 Summary of Approximate Profitability analysis on Hydrogen

	15 MTD	***30 MTD***	***45 MTD***	***60 MTD***	***75 MTD***	***90 MTD***
One time purchase Cost of ferric oxide catalyst	$4,536	$9,072	$13,608	$18,144	$22,680	$27,215
One time purchase cost of Hydrogen refrigerant	$3,264	$6,527	$9,791	$13,055	$16,318	$19,582
Total Bare Module Cost	$69.4 MM	$133.3 MM	$201.7 MM	$261.1 MM	$319.4 MM	$387.4 MM
Profit	$10.3 MM	$19.5 MM	$29.7 MM	$42.0 MM	$54.5 MM	$64.5 MM
Total Capital Investment	$125.2 MM	$240.5 MM	$363.6 MM	$472.0 MM	$578.6 MM	$701.1 MM
ROI (3rd Year)	8.22%	8.10%	8.17%	8.89%	9.42%	10.04%

15 MTD, H2, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)	CE Index (2013)	
CBM of Compressors	$15,751,246.63	$22,779,580.67	567	
CBM of Expanders	$2,817,956.93	$4,075,352.17		(=Rigorous Calculation, Guthrie method)
CBM of Heat Exchangers	$3,929,227.54	$5,682,480.75	CE Index (2023)	(=value that must be entered)
CBM of KO Drum	$30,860.07	$44,630.09	820	(=Final Value)
CBM of Electrolyzers	$20,536,463.41	$29,700,000.00		(=Seider et. al, 2013 convention)
CBM of Cold Box	$856,281.57	$1,238,361.36	Investment Site Factor (2023)	(=Intermediate calculation)
CBM of Spherical Storage Vessel	$4,041,466.77	$5,844,802.04	1	
CBM- All Process Equipment	---	$69,365,207.06		
One-time Purchase Cost of Catalyst	---	$4,536.00		
One-time Purchase Cost of Refrigerant	---	$3,264.00		
Total CBM	---	$69,373,007.06		

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$6,936,520.71	$6,936,520.71	$0.00	$69,365,207.06	$83,238,248.48
	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)	
Total Depreciable Capital: C(TDC)	$14,982,884.73	$83,238,248.48	$98,221,133.20		
	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$9,822,113.32	$3,996,119.47	$1,964,422.66	$98,221,133.20	$114,003,788.66
	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$4,133,643.40	$1,300,285.96	$5,795,287.38	$819.86	$11,230,036.61
	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$114,003,788.66	$11,230,036.61	$125,233,825.27		$15,628,437.00
	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$67,723,227.00	$57,427,737.80	$10,295,489.20	$4,118,195.68
					IRA-Adjustment to Tax ($/yr)
	Profit	Total Capital Investment: C(TCI)	ROI (%)		0
Return on Investment (ROI)	$10,295,489.20	$125,233,825.27	8.221013115		Net Earnings ($/yr)
					$10,295,489.20

Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency
4986	15.7863	5.7	2.7	47.8%

30 MTD, H2, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)	CE Index (2013)	
CBM of Compressors	$31,502,522.70	$45,559,203.91	567	
CBM of Expanders	$5,686,606.05	$8,224,015.80		(=Rigorous Calculation, Guthrie method)
CBM of Heat Exchangers	$5,788,465.76	$8,371,326.14	CE Index (2023)	(=value that must be entered)
CBM of KO Drum	$41,146.08	$59,505.80	820	(=Final Value)
CBM of Electrolyzers	$41,072,926.83	$59,400,000.00		(=Seider et. al, 2013 convention)
CBM of Cold Box	$1,160,976.68	$1,679,013.90	Investment Site Factor (2023)	(=Intermediate calculation)
CBM of Spherical Storage Vessel	$6,939,727.37	$10,036,290.02	1	
CBM- All Process Equipment	---	$133,329,355.57		
One-time Purchase Cost of Catalyst	---	$9,072.00		
One-time Purchase Cost of Refrigerant	---	$6,527.00		
Total CBM	---	$133,344,954.57		

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$13,332,935.56	$13,332,935.56	$0.00	$133,329,355.57	$159,995,226.69
	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)	
Total Depreciable Capital: C(TDC)	$28,799,140.80	$159,995,226.69	$188,794,367.49		
	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$18,879,436.75	$7,642,974.86	$3,775,887.35	$188,794,367.49	$219,092,666.45
	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$7,877,865.26	$2,474,936.01	$11,030,623.93	$1,560.51	$21,384,985.71
	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$219,092,666.45	$21,384,985.71	$240,477,652.15		$29,746,827.00
	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$128,902,917.00	$109,426,451.62	$19,476,465.38	$7,790,586.15
					IRA-Adjustment to Tax ($/yr)
	Profit	Total Capital Investment: C(TCI)	ROI (%)		0
Return on Investment (ROI)	$19,476,465.38	$240,477,652.15	8.09907499		Net Earnings ($/yr)
					$19,476,465.38

Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency
9972	30.0473	5.9	2.768	46.6%

45 MTD, H2, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)	CE Index (2013)	
CBM of Compressors	$47,253,851.10	$68,338,902.83	567	
CBM of Expanders	$8,529,917.74	$12,336,036.24		(=Rigorous Calculation, Guthrie method)
CBM of Heat Exchangers	$11,072,145.48	$16,012,626.62	**CE Index (2023)**	(=value that must be entered)
CBM of KO Drum	$49,989.54	$72,295.28	820	(=Final Value)
CBM of Electrolyzers	$61,609,390.24	$89,100,000.00		(=Seider et. al, 2013 convention)
CBM of Cold Box	$1,393,333.67	$2,015,050.46	**Investment Site Factor (2023)**	(=Intermediate calculation)
CBM of Spherical Storage Vessel	$9,521,662.47	$13,770,305.51	1	
CBM- All Process Equipment	---	$201,645,216.93		
One-time Purchase Cost of Catalyst	---	$13,608.00		
One-time Purchase Cost of Refrigerant	---	$9,791.00		
Total CBM	---	$201,668,615.93		

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$20,164,521.69	$20,164,521.69	$0.00	$201,645,216.93	$241,974,260.32

	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)
Total Depreciable Capital: C(TDC)	$43,555,366.86	$241,974,260.32	$285,529,627.18	

	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$28,552,962.72	$11,545,219.91	$5,710,592.54	$285,529,627.18	$331,338,402.35

	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$11,858,959.58	$3,734,161.52	$16,642,907.65	$2,354.48	$32,238,383.22

	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$331,338,402.35	$32,238,383.22	$363,576,785.57		$44,881,749.00

	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$194,487,579.00	$164,776,522.89	$29,711,056.11	$11,884,422.44
					IRA-Adjustment to Tax ($/yr)
	Profit	**Total Capital Investment: C(TCI)**	**ROI (%)**		0
Return on Investment (ROI)	$29,711,056.11	$363,576,785.57	**8.171879308**		**Net Earnings ($/yr)**
					$29,711,056.11

Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency
14958	45.3351	5.9	2.768	46.9%

60 MTD, H2, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)	CE Index (2013)	
CBM of Compressors	$63,005,058.49	$91,118,426.74	567	
CBM of Expanders	$11,373,221.22	$16,448,044.80		(=Rigorous Calculation, Guthrie method)
CBM of Heat Exchangers	$10,440,498.51	$15,099,133.65	CE Index (2023)	(=value that must be entered)
CBM of KO Drum	$57,564.54	$83,250.30	820	(=Final Value)
CBM of Electrolyzers	$82,145,853.66	$118,800,000.00		(=Seider et. al, 2013 convention)
CBM of Cold Box	$1,588,828.68	$2,297,776.92	Investment Site Factor (2023)	(=Intermediate calculation)
CBM of Spherical Storage Vessel	$11,916,420.11	$17,233,623.45	1	
CBM- All Process Equipment	---	$261,080,255.86		
One-time Purchase Cost of Catalyst	---	$18,144.00		
One-time Purchase Cost of Refrigerant	---	$13,055.00		
Total CBM	---	$261,111,454.86		

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$26,108,025.59	$26,108,025.59	$0.00	$261,080,255.86	$313,296,307.04
	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)	
Total Depreciable Capital: C(TDC)	$56,393,335.27	$313,296,307.04	$369,689,642.30		
	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$36,968,964.23	$15,173,296.01	$7,393,792.85	$369,689,642.30	$429,225,695.39
	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$15,616,604.52	$4,978,882.02	$22,190,543.53	$3,139.30	$42,789,169.37
	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$429,225,695.39	$42,789,169.37	$472,014,864.76		$59,842,332.00
	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$259,316,772.00	$217,357,837.63	$41,958,934.37	$16,783,573.75
					IRA-Adjustment to Tax ($/yr)
	Profit	Total Capital Investment: C(TCI)	ROI (%)		0
Return on Investment (ROI)	$41,958,934.37	$472,014,864.76	8.889324786		Net Earnings ($/yr)
					$41,958,934.37
Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency	
19944	60.4468	5.9	2.768	46.9%	

75 MTD, H2, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)
CBM of Compressors	$78,756,233.17	$113,897,903.35
CBM of Expanders	$14,216,523.79	$20,560,052.04
CBM of Heat Exchangers	$11,466,407.76	$16,582,811.92
CBM of KO Drum	$64,398.78	$93,134.04
CBM of Electrolyzers	$100,400,487.80	$145,200,000.00
CBM of Cold Box	$1,761,251.54	$2,547,136.26
CBM of Spherical Storage Vessel	$14,181,939.10	$20,510,035.38
CBM- All Process Equipment	---	$319,391,073.00
One-time Purchase Cost of Catalyst	---	$22,680.00
One-time Purchase Cost of Refrigerant	---	$16,318.00
Total CBM	---	$319,430,071.00

CE Index (2013)
567

CE Index (2023)
820

Investment Site Factor (2023)
1

(=Rigorous Calculation, Guthrie method)
(=value that must be entered)
(=Final Value)
(=Seider et. al, 2013 convention)
(=Intermediate calculation)

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$31,939,107.30	$31,939,107.30	$0.00	$319,391,073.00	$383,269,287.60

	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)
Total Depreciable Capital: C(TDC)	$68,988,471.77	$383,269,287.60	$452,257,759.36	

	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$45,225,775.94	$18,769,521.27	$9,045,155.19	$452,257,759.36	$525,298,211.75

	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$19,348,967.68	$6,223,594.29	$27,738,142.70	$3,924.12	$53,314,628.79

	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$525,298,211.75	$53,314,628.79	$578,612,840.54		$74,802,816.00

	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$324,145,536.00	$269,635,709.42	$54,509,826.58	$21,803,930.63
					IRA-Adjustment to Tax ($/yr)
	Profit	Total Capital Investment: C(TCI)	ROI (%)		0
Return on Investment (ROI)	$54,509,826.58	$578,612,840.54	9.420777205		Net Earnings ($/yr)
					$54,509,826.58

Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency
24930	75.5584	5.9	2.768	46.9%

90 MTD, H2, Results

Important Metrics	Initial Value ($)	CE Index Adjusted Price ($)	CE Index (2013)	
CBM of Compressors	$94,507,401.31	$136,677,370.50	567	
CBM of Expanders	$17,059,827.27	$24,672,060.60		(=Rigorous Calculation, Guthrie method)
CBM of Heat Exchangers	$17,057,412.57	$24,668,568.44	CE Index (2023)	(=value that must be entered)
CBM of KO Drum	$30,860.07	$44,630.09	820	(=Final Value)
CBM of Electrolyzers	$120,936,951.22	$174,900,000.00		(=Seider et. al, 2013 convention)
CBM of Cold Box	$1,917,391.32	$2,772,946.89	Investment Site Factor (2023)	(=Intermediate calculation)
CBM of Spherical Storage Vessel	$16,349,218.12	$23,644,371.89	1	
CBM- All Process Equipment	---	$387,379,948.40		
One-time Purchase Cost of Catalyst	---	$27,215.00		
One-time Purchase Cost of Refrigerant	---	$19,582.00		
Total CBM	---	$387,426,745.40		

	Site Preparation Cost: C(Site)	Service Facilities Cost: C(Service)	Allocated Cost for Utilities Plants/Related Facilities: C(Alloc)	Total Bare Module Cost: C(TBM)	Final C(DPI), Guthrie, ($/yr)
Direct Permanent Investment: C(DPI)	$38,737,994.84	$38,737,994.84	$0.00	$387,379,948.40	$464,855,938.08
	Contingencies Cost: C(Cont)	Direct Permanent Investment: C(DPI)	Final C(TDC), Guthrie, ($/yr)	Reportable C(TDC) (Including Refrigerant), ($/yr)	
Total Depreciable Capital: C(TDC)	$83,674,068.85	$464,855,938.08	$548,530,006.93		
	Startup Cost: C(Startup)	Royalties Cost: C(Royalties)	Land Cost: C(Land)	Total Depreciable Capital: C(TDC)	Final C(TPI), Guthrie, ($/yr)
Total Permanent Investment: C(TPI)	$54,853,000.69	$22,639,842.01	$10,970,600.14	$548,530,006.93	$636,993,449.77
	Cash reserves	Inventory	Accounts Receivable	Accounts Payable	Final C(WC), Guthrie, ($/yr)
Working Capital: C(WC)	$23,298,859.41	$7,468,314.80	$33,285,778.58	$4,708.95	$64,057,661.74
	Total Permanent Investment: C(TPI)	Working Capital: C(WC)	Final C(TCI), Guthrie, ($/yr)		IRA Tax Credit ($/yr)
Total Capital Investment: C(TCI)	$636,993,449.77	$64,057,661.74	$701,051,111.51		$89,763,399.00
	Total Tax Rate (State + Federal)	Annual Sales Revenue	Annual Production Cost	Gross Earnings ($/yr)	Tax ($/yr)
Net Earnings (Profit)	0.4	$388,974,729.00	$324,524,078.05	$64,450,650.95	$25,780,260.38
					IRA-Adjustment to Tax ($/yr)
	Profit	Total Capital Investment: C(TCI)	ROI (%)		0
Return on Investment (ROI)	$64,450,650.95	$701,051,111.51	9.193431106		Net Earnings ($/yr)
					$64,450,650.95

Total Power requirement (hp)	Production (tonnes LH2/day)	Specific power (kWh/kg LH2)	Minimum specific power (kWh/kg LH2)	Efficiency
29916	90.6701	5.9	2.768	46.9%

24.9 Water Availability

With a final proposed design of an electrolytic plant producing 45MTD of LH2 per day, water availability is an important consideration when selecting regions that could facilitate the plant's needed process water consumption of nearly 107,000 gallons per day. Upon investigation of water consumption regulations across the United States via the National Research Council's report on the United States Geological Survey (USGS), pertinent logistical information was extracted to detail preparation steps required across each state. While groundwater consumptive use permits (CUPs) are issued at a state level, they may be subservient to micro-regional commissions such as county and town regulation over freshwater and groundwater reservoirs. However, each state appears to have avenues to acquire permits for groundwater access above this plant's needs following geological surveys of the proposed plant location. Two regions considered to be extremes on regulatory constriction have been identified as potential targets due to high potential consumer availability discussed earlier in this report – the St. John's River Management District of the Florida Department of Environmental Protection with its proximity to Cape Canaveral and multiple air and sea ports, and the Santa Maria region of California with its proximity to Vanderburg Space Force Base, Mojave Space and Air Port and many close air and sea ports.

24.9.1 **Florida**

Within the Florida district, an application Form Number 40C-2.900(1)b with supplementary information must be submitted to the district office, whereupon reception the staff must respond with office consent within 90 days. Within said 90 days, the district must notify concerned stakeholders in the affected area of the plant, from which respondents have 14 days to send concerns on the proposed development. At the end of said 90 days, a regulatory meeting will be held by the Governing Board of the St. John's River Water Management District whereat the proposal may be rejected, approved, or delayed for further consideration by no more than an additional 90 days. Consumptive Use Permits last 20 years. Compliance reports must be filed to maintain permits each ten years. Should the plant have an event that decreases production and thus water consumption for a period of time, the threat of water permit constriction is null. "In order to incentivize conservation of water, if actual water use is less than permitted water use due to documented implementation of water conservation measures, the permitted allocation shall not be modified by the District due to these circumstances during the term of the permit."

Water may be secured by any of the following sources:

(a) River, creek, or other watercourse

(b) Lake, pond, or other impoundment

(c) Aquifer

(d) Water supplier

Due to the incremental increase in water requirement of our process over the first years' operation (i.e. from plant construction requiring relatively little water to a scaled production schedule whereby 100% design production is not reached immediately,) the permit filing will request annualized allotment increases to consumptive use proportionate to the increased need per annum until full production is reached. Project viability is dictated by several location-based criteria, including the capacity of water-providers to accommodate the added supply strain while projecting a regional population growth. With the volumetric requirements proposed in this plant, it is assumed that public providers are viable sources for water needs, and well drilling is built into costs.

Per Florida EPA guidelines, permit to the volume of water required for this plant mandates several deliverables on a yearly basis, including:

(1) A water conservation public education program, which may scale in complexity from public service announcements included in invoices to consumers, to public water conservation exhibits.

(2) An outdoor water use reduction program, mostly irrelevant to the operations and landscape of this plant.

(3) A product rate structure that "promotes the efficient use of water," which in this case will be incorporated into the supply chain costs of the product.

(4) A water loss reduction plan, by which the plant will be subject to water utilization audits to validate that the plant does not waste more than 10% of its water received. Should audits fail, remediation plans must be implemented.
(5) An indoor water use reduction program, which minimizes indoor water consumption to bare operational minima.

The costs of the above are considered negligible. The hydrological effects of the proposed plant' consumptive water use may not:

(1) "Significantly and adversely impact wetlands, lakes, rivers, or springs."
(2) "Reduce a flow or level below any minimum flow or level established by the District or the Department of Environmental Protection"
(3) "Contribute to significant saline water intrusion."
(4) "Contribute to flood damage."

Permits may be temporarily suspended in the event of:

(1) Drought.
(2) Water contamination crisis.
(3) Breakage of any of the rules listed above.

24.9.2 **California**

California water management and permitting is distributed at a municipal level but is appropriately generalized by the Sustainable Groundwater Management Act (SGMA). Only regulatory restrictions that are found to be more restrictive than Florida's are discussed in this section for brevity's sake, but a copy of relevant regulatory information may be found in Appendix.

Due to California's complex water distribution system and frequent droughts, water consumption follows more restrictive guidelines to prevent "overdrafting" groundwater. With a proposed plant outside of the North Coast Instream Flow Policy Area, a more precisely regulated region of California, certain leniencies are afforded in regulatory steps that may slow the otherwise extensive water utilization approval requirements.

Upon the reception of a permit application, a local SGMA affiliated bureau must perform the following analyses to reach an opinion on the consumptive requirements of the plant:

(1) Environmental review in accordance with Environmental Quality Act, through which an "Environmental Impact Report" must be procured with the State Water Board to review not only the direct water availability and environmental impacts, but also impacts on local industries such as fisheries and transportation hubs. Mitigation strategies are devised.

(2) Water Availability Analysis, similar to those required under Florida law, but more limited in availability due to stricter requirements on groundwater tapping and higher drought probability.

Water utilization in the Santa Maria – Santa Barbara area are approved by the Central Coast Regional Water Quality Control Board (CCRWQCB). Despite not publishing detailed water permits online at this time, the board details its heuristic and strategic agenda in the Water Quality Control Plan for the Central Coastal Basin, provided in Appendix, prioritizing the use of preexisting reservoirs over new groundwater tapping. Of the nearly 300 water sources in the region, under 15 are designated for use in high purity industrial processes, a significant limitation on the already limited water availability of the region. While no clear plant capacity constraints are discussed in this plan, it is not clear that a production plant of this scale would be approved on its own support without special intervention from the state or federal government. However, with recent pushes towards Liquid Hydrogen in the federal government, such as the passage of the Inflation Reduction Act and the DOE's benchmark moonshot of $1/kg for LH2 production, it appears possible that liquefaction plants would be given preferential treatment, though at a local water rights approval scale, it is unclear if approval would be garnered. Further, should approval be garnered, it would be unclear if water availability and costing would be affected by future anticipated changes to the Colorado River Compact or other overdrafting measures that may increase the cost of water in certain sectors. Lacking further substance in investigating Southern California's capacity in supporting a water-fueled Hydrogen liquefaction plant, a competitive analysis was performed to understand the current statewide approach to LH2. As of now, the state boasts 62 “light-duty retail Hydrogen refueling stations” on its website, and as of January signed a contract with Linde to construct a five-megawatt electrolyzer capable of producing , suggesting state special interests in supplying sufficient water to produce Hydrogen at a rate deemed economical by Linde. It appears, despite potentially prolonged water permit processes, that California and the Santa Maria region are open to permitting large quantities of water. Their existing contract with Linde, however, would prove competitive for this proposed plant if designed by a competitor

24.10 Project Statement

7. Green Hydrogen Liquefaction
(Recommended by Adam Brostow, Sr. Process Engineer, Cryo Technologies – Chart)

Background

As global climate change is progressing, hydrogen is becoming an important clean fuel. Blue hydrogen is produced by steam methane reforming with CO_2 capture. It is subject to the volatile natural gas (NG) market (the EU gets 40% of NG from Russia). Green hydrogen is typically produced by electrolysis of water. Electricity comes from renewable sources such as the sun or wind. Hydrogen can then be liquefied for transport or storage. The first liquid hydrogen (LH2) ship completed its voyage in 2021.

Fig. 1

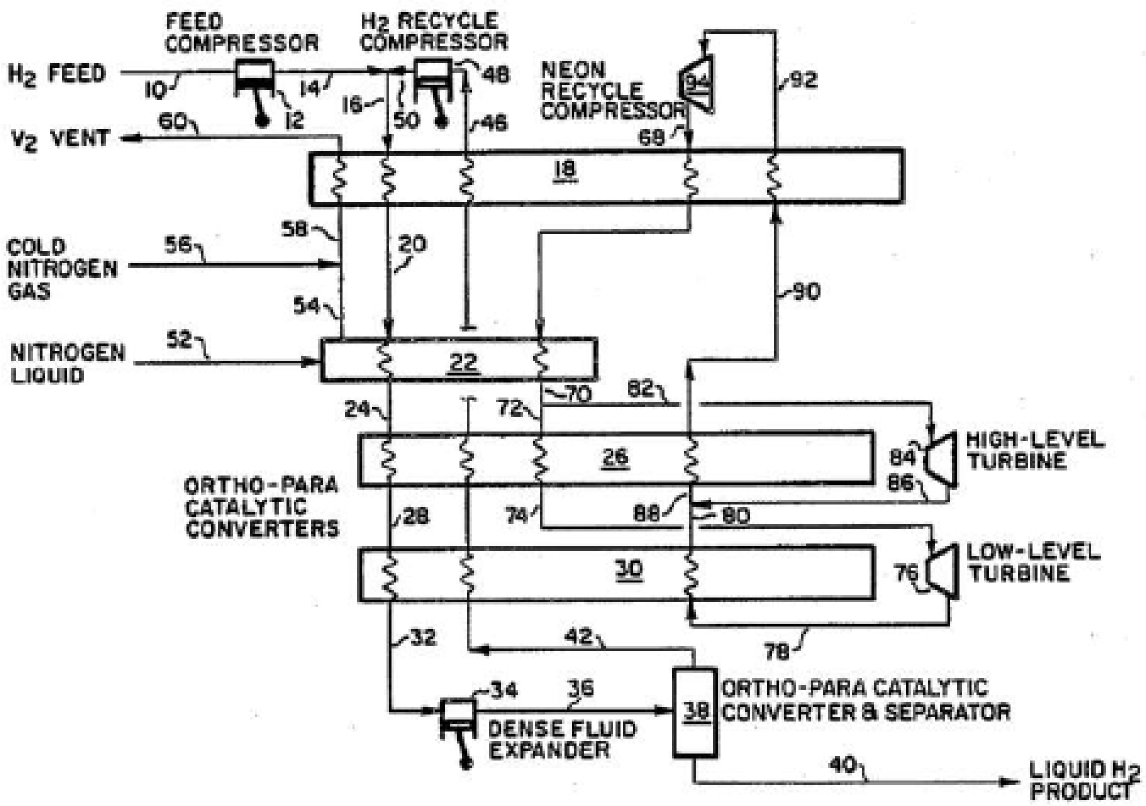


Fig. 1 shows a typical hydrogen liquefaction process. The H_2 is optionally compressed and liquefied. Normal hydrogen contains 75% ortho-hydrogen and 25% para-hydrogen, spin isomers. Liquid hydrogen contains at least 95% of para-hydrogen. The ortho-para (O-P) conversion is catalytic and exothermic. Compared to natural gas or nitrogen liquefaction, O-P conversion and temperature of about 20 K pose additional challenges in designing the liquefaction process. The heat exchangers used are brazed aluminum plate-and-fin (BAHXs).

Problem Statement

The group is to design a plant producing 15 MTD (metric tons per day) of LH2. The feed is precooled with liquid nitrogen (LIN). Most of the refrigeration is supplied by the so-called reverse-Brayton cycle, isentropic expansion of gas in two turboexpanders. The working fluid may be neon, helium, hydrogen, or the mixture of the above suggested in a recent paper (Cardella et al.).

The PFD follows Fig. 1. This is the NBA (next best alternative), coming from a relatively old patent. Students are encouraged to suggest improvements based on the newest developments mentioned in literature and their own new ideas. The problem is open-ended.

Basis of Design

Normal H_2 feed is at 100 deg. F, 265 psia. LH2 product is flashed to 52 psia and is to contain more than 95% para-hydrogen. The feed and refrigerant recycle compressors can be modeled in Aspen Plus as a 3-stage MCompr, with cooling water available at 90 deg. F, approach on inter/after-coolers of 10 deg. F, the pressure drop of inter/after-coolers of 5 psi, adiabatic (isentropic) efficiency of 85%. The turboexpanders have adiabatic efficiency of 85%. The dense fluid expander's adiabatic efficiency is 70%. A simple Joule-Thompson valve is to be considered in case the dense fluid expander is not operational. CAPEX-OPEX tradeoff is to be documented. There is no O-P conversion in the product phase separator. Precooling LIN is supplied saturated at 20 psia. It is throttled to atmospheric pressure (plus HX pressure drop) to provide refrigeration at about 77 K. Pressures such as feed booster compressor and recycle compressor discharge are to be optimized for minimal power consumption. The BAHX minimum approach is 3 deg. F, maximum 50 deg. F (due to thermal stresses). The cryogenic equipment is inside a vacuum cold box.

The group should perform techno-economic analysis of the base case (NBA) using pure neon and to consider using pure hydrogen and, time permitting, helium, neon-hydrogen, or neon-helium (nelium) mixture. They must consider ortho-para conversion (there is more than one way to do it). There is currently a shortage of neon due to war in Ukraine.

The group is to consider the power required to make LIN for pre-cooling, part of the life-cycle analysis. This information can come from literature, not a simulation. Green technologies require correct life cycle analysis as the benefits are often quoted out of context. What does the carbon footprint of green hydrogen look like if the liquefaction power comes from the grid (e.g., NG combined cycle)?

The group is encouraged to use SQP (sequential quadratic programming) optimization in Aspen Plus, but this is not required (instructions will follow). This is a useful skill.

References

LH2 patent (source of Fig. 1), Gaumer at al.:
https://patents.google.com/patent/US4765813A/en

LIN production patent, Brostow et al.:
https://patents.google.com/patent/US6298688B1/en

Economically viable large-scale hydrogen liquefaction, Cardella et al., 2017
https://doi.org/10.1088/1757-899X/171/1/012013

On green H_2:
https://www.cowi.com/insights/green-hydrogen-the-oldest-element-is-creating-a-new-world

Catalyst information:
https://www.molecularproducts.com/wp-content/uploads/2017/03/166_Rev-C_Technical-Datasheet_OP-Catalyst.pdf

24.11 ASPEN PLUS Input File

```
;
;Input Summary created by Aspen Plus Rel. 39.0 at 11:20:42 Tue Apr 18,
2023
;Directory C:\Users\guillerv\Downloads  Filename
C:\Users\guillerv\AppData\Local\Temp\~ap65e3.txt
;


DYNAMICS
    DYNAMICS RESULTS=ON

TITLE 'EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING'

IN-UNITS ENG SHORT-LENGTH=in

DEF-STREAMS CONVEN ALL

DIAGNOSTICS
    HISTORY SIM-LEVEL=3
    TERMINAL SIM-LEVEL=3
    MAX-PRINT SIM-LIMIT=9999

SIM-OPTIONS MASS-BAL-CHE=YES TLOWER=-456.0 FLASH-TOL=1.000000E-07

MODEL-OPTION

ACCOUNT-INFO USER-NAME="HARRISCF"

RUN-CONTROL MAX-TIME=60000.0 MAX-ERRORS=9999 MAX-FORT-ERR=9999

DATABANKS 'APV121 PURE38' / NOASPENPCD

PROP-SOURCES 'APV121 PURE38'

COMPONENTS
    NH2 H2 /
    PH2 H2-PARA /
    HE HE-4 /
    N2 N2 /
    NE NE

FORMULA NH2 H2 / PH2 H2-PARA / HE HE-4 / N2 N2 / NE NE

SOLVE
    RUN-MODE MODE=SIM

;STREAM 900
;    SUBSTREAM MIXED VFRAC=0.0 PRES=20 MASS-FLOW=3500
;    MOLE-FRAC PH2 1.0
;STREAM 906
;    SUBSTREAM MIXED VFRAC=1.0 PRES=17.5 MASS-FLOW=3500
;    MOLE-FRAC PH2 1.0
```

```
;    BLOCK HX1 IN=110 118 210 310 270 370 710 &
;             OUT=115 119 215 315 280 380 730
;    BLOCK TRLR IN=900 OUT=905 950
;    BLOCK TRLDIS IN=905 906 OUT=910
;    BLOCK TRLHSE IN=910 OUT=920

FLOWSHEET
    BLOCK 34,B IN=36,B1 OUT=36,B2
    BLOCK 34SPLIT IN=32 OUT=36,A1 36,B1
    BLOCK 38 IN=36,A2 36,B2 OUT=42 40
    BLOCK 84,1 IN=82 OUT=86
    BLOCK 76,1 IN=74 OUT=78
    BLOCK MIXN2 IN=54 T56 OUT=58
    BLOCK MIX2 IN=80 86 OUT=88
    BLOCK SPLIT IN=70 OUT=72 82
    BLOCK 34,A IN=36,A1 OUT=36,A2
    BLOCK 30 IN=78 42 28,1 OUT=32 46,-2 80
    BLOCK 26 IN=46,-2 72 88 24,1 OUT=90 74 46,+R 28,1
    BLOCK 22 IN=20L 68,1 T52 OUT=70 54 24
    BLOCK 94 IN=92 OUT=T68
    BLOCK 18 IN=16 68 90 58 46,+R OUT=46 92 20L 68,1 60
    BLOCK MIX IN=14 T50 OUT=16
    BLOCK 48 IN=46 OUT=50
    BLOCK 12 IN=10 OUT=14
    BLOCK JT2 IN=40LN2 OUT=FEEDN2
    BLOCK SPL IN=T20 OUT=25 55
    BLOCK MKP IN=10LN2 50LN2 OUT=15
    BLOCK KO IN=35 65 OUT=45 40LN2
    BLOCK JT IN=30LN2 OUT=35
    BLOCK BAXH IN=25 55 45 OUT=50LN2 60LN2 30LN2
    BLOCK EXP IN=60LN2 OUT=65
    BLOCK CMP IN=15 OUT=20
    BLOCK LINKO IN=FEEDN2 OUT=56 52

PROPERTIES REFPROP
    PROPERTIES REFPROP
    PROPERTIES PENG-ROB / BWRS

DEF-STREAMS CONVEN T20

PROP-SET ACFM VMX UNITS='cuft/min' SUBSTREAM=MIXED

PROP-SET CPCV CPCVMX SUBSTREAM=MIXED

PROP-SET KGFLW MASSFLMX UNITS='kg/hr' SUBSTREAM=MIXED

PROP-SET PRES PRES UNITS='bar' SUBSTREAM=MIXED

PROP-SET RHOL RHOMX UNITS='lb/cuft' SUBSTREAM=MIXED PHASE=L

PROP-SET RHOV RHOMX UNITS='lb/cuft' SUBSTREAM=MIXED PHASE=V

PROP-SET TEMPC TEMP UNITS='C' SUBSTREAM=MIXED
```

```
PROP-SET VISCL MUMX UNITS='lb/ft-hr' SUBSTREAM=MIXED PHASE=L

PROP-SET VISCV MUMX UNITS='lb/ft-hr' SUBSTREAM=MIXED PHASE=V

STREAM 10
    SUBSTREAM MIXED TEMP=90. PRES=265. MOLE-FLOW=2052.
    MOLE-FRAC NH2 1.

STREAM 10LN2
    SUBSTREAM MIXED TEMP=105. PRES=100. MOLE-FLOW=500.
    MASS-FRAC N2 1.0

STREAM 14

STREAM 16

STREAM 20L

STREAM 24,1
    SUBSTREAM MIXED TEMP=99. PRES=640. MOLE-FLOW=99.
    MOLE-FLOW PH2 1.

STREAM 32

STREAM 36,A2

STREAM 45
    SUBSTREAM MIXED TEMP=-427.0 PRES=45.0 MOLE-FLOW=2001.
    MOLE-FRAC HE 1.0

STREAM 46

STREAM 46,-2

STREAM 54

STREAM 68
    SUBSTREAM MIXED TEMP=110. PRES=150. MOLE-FLOW=9750.
    MOLE-FRAC NH2 0. / HE 0. / NE 1.

STREAM 68,1

STREAM 78

STREAM T20
    SUBSTREAM MIXED TEMP=100. PRES=999.0 MOLE-FLOW=200.
    MASS-FRAC N2 1.0

STREAM T50
    SUBSTREAM MIXED TEMP=110. PRES=650. MOLE-FLOW=304.23
    MOLE-FLOW NH2 1.

STREAM T52
    SUBSTREAM MIXED TEMP=-316. PRES=20. MOLE-FLOW=696.
```

```
    MOLE-FLOW N2 1.

STREAM T56
    SUBSTREAM MIXED TEMP=-318. PRES=17. MOLE-FLOW=804.
    MOLE-FLOW N2 1.

BLOCK MIX MIXER
    PARAM PRES=650.

BLOCK MIX2 MIXER
    PARAM

BLOCK MIXN2 MIXER
    PARAM

BLOCK MKP MIXER
    PARAM

BLOCK 34SPLIT FSPLIT
    FRAC 36,B1 1.

BLOCK SPL FSPLIT
    FRAC 55 .10

BLOCK SPLIT FSPLIT
    FRAC 82 0.47

BLOCK 38 FLASH2
    PARAM PRES=0. DUTY=0.

BLOCK KO FLASH2
    PARAM PRES=0.0 DUTY=0.0

BLOCK LINKO FLASH2
    PARAM PRES=0. DUTY=0.

BLOCK 18 MHEATX
    HOT-SIDE IN=16 OUT=20L TEMP=-290. PRES=-7. FREE-WATER=NO  &
        DPPARMOPT=NO
    HOT-SIDE IN=68 OUT=68,1 TEMP=-290. PRES=-6.19  &
        FREE-WATER=NO DPPARMOPT=NO
    COLD-SIDE IN=90 OUT=92 PRES=-2.65 FREE-WATER=NO  &
        DPPARMOPT=NO
    COLD-SIDE IN=58 OUT=60 PRES=-1.43 FREE-WATER=NO  &
        DPPARMOPT=NO
    COLD-SIDE IN=46,+R OUT=46 PRES=-0.47 FREE-WATER=NO  &
        DPPARMOPT=NO
    PARAM NPOINT=50 ADAPTIVE-GRI=YES TQTAB-NPOINT=10  &
        TQTAB-MAXNPT=10
    REPORT NOSTREAMS

BLOCK 22 MHEATX
    HOT-SIDE IN=20L OUT=24 TEMP=-305. PRES=-2.3 FREE-WATER=NO  &
        MAXIT=50 DPPARMOPT=NO
```

```
    HOT-SIDE IN=68,1 OUT=70 TEMP=-305. PRES=-8.3 FREE-WATER=NO  &
        MAXIT=50 DPPARMOPT=NO
    COLD-SIDE IN=T52 OUT=54 PRES=-1.13 FREE-WATER=NO MAXIT=50  &
        DPPARMOPT=NO
    PARAM NPOINT=50 ADAPTIVE-GRI=YES TQTAB-NPOINT=10  &
        TQTAB-MAXNPT=10

BLOCK 26 MHEATX
    COLD-SIDE IN=46,-2 OUT=46,+R PRES=-0.6 FREE-WATER=NO  &
        MAXIT=50 DPPARMOPT=NO
    HOT-SIDE IN=72 OUT=74 TEMP=-367. PRES=-2.94 NPHASE=2  &
        FREE-WATER=NO MAXIT=50 DPPARMOPT=NO
    COLD-SIDE IN=88 OUT=90 PRES=-1.94 NPHASE=2 FREE-WATER=NO  &
        MAXIT=50 DPPARMOPT=NO
    HOT-SIDE IN=24,1 OUT=28,1 TEMP=-367. PRES=-0.58  &
        FREE-WATER=NO DPPARMOPT=NO
    PARAM NPOINT=50 T-LOWER=-456. QLEAK=99. ADAPTIVE-GRI=YES  &
        TQTAB-NPOINT=10 TQTAB-MAXNPT=10

BLOCK 30 MHEATX
    COLD-SIDE IN=78 OUT=80 PRES=-1.53 NPHASE=2 FREE-WATER=NO  &
        MAXIT=50 DPPARMOPT=NO
    COLD-SIDE IN=42 OUT=46,-2 PRES=-0.36 FREE-WATER=NO  &
        DPPARMOPT=NO
    HOT-SIDE IN=28,1 OUT=32 TEMP=-404. PRES=-0.11  &
        FREE-WATER=NO DPPARMOPT=NO
    PARAM NPOINT=50 T-LOWER=-456. QLEAK=130000. ADAPTIVE-GRI=YES  &
        TQTAB-NPOINT=10 TQTAB-MAXNPT=10

BLOCK BAXH MHEATX
    HOT-SIDE IN=25 OUT=60LN2 TEMP=-150.0 PRES=-5.0  &
        FREE-WATER=NO DPPARMOPT=NO
    HOT-SIDE IN=55 OUT=30LN2 TEMP=-293.0 PRES=-5.0  &
        FREE-WATER=NO DPPARMOPT=NO
    COLD-SIDE IN=45 OUT=50LN2 FREE-WATER=NO DPPARMOPT=NO
    PARAM NPOINT=100 MAXIT=100 TOL=1.000000E-07 T-UPPER=1000.0  &
        T-LOWER=-456.0 ADD-POINTS=NO QLEAK=0.0 TQTAB-NPOINT=20  &
        TQTAB-MAXNPT=20 STREAM-RETN=NO
    REPORT STREAMS NOTQTABLES

BLOCK 34,B PUMP
    PARAM PRES=52. EFF=0.7 PUMP-TYPE=TURBINE

BLOCK 76,1 COMPR
    PARAM TYPE=ISENTROPIC PRES=27.4 SEFF=0.85 NPHASE=2  &
        SB-MAXIT=30 SB-TOL=0.0001 MODEL-TYPE=TURBINE
    PERFOR-PARAM CALC-SPEED=NO
    BLOCK-OPTION FREE-WATER=NO

BLOCK 84,1 COMPR
    PARAM TYPE=ISENTROPIC PRES=26.61 SEFF=0.85 NPHASE=2  &
        SB-MAXIT=30 SB-TOL=0.0001 MODEL-TYPE=TURBINE
    PERFOR-PARAM CALC-SPEED=NO
    BLOCK-OPTION FREE-WATER=NO
```

```
BLOCK EXP COMPR
    PARAM TYPE=ASME-ISENTROP PRES=50.0 SEFF=.850 NPHASE=2  &
        SB-MAXIT=30 SB-TOL=.00010 MODEL-TYPE=TURBINE
    PERFOR-PARAM CALC-SPEED=NO
    BLOCK-OPTION FREE-WATER=NO

BLOCK 12 MCOMPR
    PARAM NSTAGE=3 TYPE=ISENTROPIC PRES=650. SB-MAXIT=30  &
        SB-TOL=0.0001
    FEEDS 10 1
    PRODUCTS 14 3
    COOLER-SPECS 1 TEMP=110.

BLOCK 48 MCOMPR
    PARAM NSTAGE=3 TYPE=ISENTROPIC PRES=650. SB-MAXIT=30  &
        SB-TOL=0.0001
    FEEDS 46 1
    PRODUCTS 50 3
    COOLER-SPECS 1 TEMP=110.

BLOCK 94 MCOMPR
    PARAM NSTAGE=3 TYPE=ISENTROPIC PRES=150. SB-MAXIT=30  &
        SB-TOL=0.0001
    FEEDS 92 1
    PRODUCTS T68 3
    COOLER-SPECS 1 TEMP=110.

BLOCK CMP MCOMPR
    PARAM NSTAGE=3 TYPE=ASME-ISENTROPIC PRES=2000. SB-MAXIT=30  &
        SB-TOL=.00010
    FEEDS 15 1
    PRODUCTS 20 3
    COMPR-SPECS 1 SEFF=.850 / 2 SEFF=.850 / 3 SEFF=.850
    COOLER-SPECS 1 TEMP=100.0 PDROP=5.0 / 2 TEMP=100.0  &
        PDROP=5.0 / 3 TEMP=100.0 PDROP=5.0

;The vapor/liquid split that occurs here dicates what the size of the
recyle stream, and hence the loop will be.
;x = R/(M+R)
;

BLOCK 34,A VALVE
    PARAM P-OUT=52.

BLOCK JT VALVE
    PARAM P-OUT=50.0

BLOCK JT2 VALVE
    PARAM P-OUT=20.

EO-CONV-OPTI

CALCULATOR CMPP
```

```
    DEFINE P PARAMETER 1
    DEFINE PCMP BLOCK-VAR BLOCK=CMP VARIABLE=PRES  &
        SENTENCE=PARAM UOM="psia"
    DEFINE PTEAR STREAM-VAR STREAM=T20 SUBSTREAM=MIXED  &
        VARIABLE=PRES UOM="psia"
F        PCMP = P+5
F     PTEAR=P

    READ-VARS P
    WRITE-VARS PCMP PTEAR

CALCULATOR INIT
    DEFINE P PARAMETER 1
F     P=700
    WRITE-VARS P

;Correlation developed between Temperature [F] and Heat of Reaction (NH2
to PH2) [Btu/lb]   (In drive)
;y = (-9*10^-11)*x^5 + (-8*10^-8)*x^4 + (-2*10^-5)*x^3  + (0.0013)*x^2 +
(-0.1688)*x +29.258
;y = Heat of Reaction (NH2 to PH2) [Btu/lb]
;x = Temperature [F]
;
;Heat evolved from OP conversion = (CHG in Mass of NH2) * (Integral of
above correlation from TIN to TOUT)
;Integral calculated using Trapezoidal rule; "PN" are pieces of the above
correlation (done for space purposes)
;
;Next, conversion is set manually. 50% conversion is assumed
;Change in NH2 (mass) = (Molar change in NH2) * (2 lb/lbmol)
;STR = Molar flowrate of Stream 24,1 (Incoming hot H2 stream to HX)
[lbmol/hr]
;Molar change in NH2 = 0.5 * STR
;---> Change in NH2 (mass) = 0.5 * STR [lbmol/hr] * [2 lb/lbmol]
;
;
;

CALCULATOR OP1
    DEFINE Q BLOCK-VAR BLOCK=26 VARIABLE=QLEAK SENTENCE=PARAM  &
        UOM="Btu/hr"
    DEFINE TOUT STREAM-VAR STREAM=28,1 SUBSTREAM=MIXED  &
        VARIABLE=TEMP UOM="F"
    DEFINE TIN STREAM-VAR STREAM=24,1 SUBSTREAM=MIXED  &
        VARIABLE=TEMP UOM="F"
    DEFINE STR STREAM-VAR STREAM=24,1 SUBSTREAM=MIXED  &
        VARIABLE=MOLE-FLOW UOM="lbmol/hr"
F       P0 = 9*10**-6
F       P1 = 0.0045
F       P2 = -0.3074
F       P3 = 28.835
F
F       F0 = (P0*TIN**3)+(P1*TIN**2)+(P2*TIN)+(P3)
F
```

```
F       T1 = TIN + 1*(TOUT-TIN)/4
F       F1 = (P0*T1**3)+(P1*T1**2)+(P2*T1)+(P3)
F
F       T2 = TIN + 2*(TOUT-TIN)/4
F       F2 = (P0*T2**3)+(P1*T2**2)+(P2*T2)+(P3)
F
F       T3 = TIN + 3*(TOUT-TIN)/4
F       F3 = (P0*T3**3)+(P1*T3**2)+(P2*T3)+(P3)
F
F       T4 = TIN + 4*(TOUT-TIN)/4
F       F4 = (P0*T4**3)+(P1*T4**2)+(P2*T4)+(P3)
F
F       QCURVE = (0.5)*ABS(TOUT-TIN)/4*(F0+2*F1+2*F2+2*F3+F4)
F       AVGQ = 1/ABS(TOUT-TIN)*QCURVE
F
F       M0 = -2*10**-13
F       M1 = -1*10**-4
F       M2 = 9*10**-10
F       M3 = 6*10**-7
F       M4 = 0.0002
F       M5 = 0.7359
F       EI = 0.62
F       EO = 0.1
F
F       FCONV = 0.2
F       DELTOH2 = 2 * FCONV * (EI - EO) * STR
F       Q = AVGQ * DELTOH2
F
    READ-VARS TOUT TIN STR
    WRITE-VARS Q

;Correlation developed between Temperature [F] and Heat of Reaction (NH2
to PH2) [Btu/lb]   (In drive)
;y = (-9*10^-11)*x^5 + (-8*10^-8)*x^4 + (-2*10^-5)*x^3  + (0.0013)*x^2 +
(-0.1688)*x +29.258
;y = Heat of Reaction (NH2 to PH2) [Btu/lb]
;x = Temperature [F]
;
;Heat evolved from OP conversion = (CHG in Mass of NH2) * (Integral of
above correlation from TIN to TOUT)
;Integral calculated using Trapezoidal rule; "PN" are pieces of the above
correlation (done for space purposes)
;
;Next, conversion is set manually. 50% conversion is assumed
;Change in NH2 (mass) = (Molar change in NH2) * (2 lb/lbmol)
;STR = Molar flowrate of Stream 24,1 (Incoming hot H2 stream to HX)
[lbmol/hr]
;Molar change in NH2 = 0.5 * STR
;---> Change in NH2 (mass) = 0.5 * STR [lbmol/hr] * [2 lb/lbmol]
;
;

CALCULATOR OP2
    DEFINE Q BLOCK-VAR BLOCK=30 VARIABLE=QLEAK SENTENCE=PARAM  &
```

```
        UOM="Btu/hr"
    DEFINE TOUT STREAM-VAR STREAM=32 SUBSTREAM=MIXED  &
        VARIABLE=TEMP UOM="F"
    DEFINE TIN STREAM-VAR STREAM=28,1 SUBSTREAM=MIXED  &
        VARIABLE=TEMP UOM="F"
    DEFINE STR STREAM-VAR STREAM=28,1 SUBSTREAM=MIXED  &
        VARIABLE=MOLE-FLOW UOM="lbmol/hr"
F       P0 = 9*10**-6
F       P1 = 0.0045
F       P2 = -0.3074
F       P3 = 28.835
F
F       F0 = (P0*TIN**3)+(P1*TIN**2)+(P2*TIN)+(P3)
F
F       T1 = TIN + 1*(TOUT-TIN)/4
F       F1 = (P0*T1**3)+(P1*T1**2)+(P2*T1)+(P3)
F
F       T2 = TIN + 2*(TOUT-TIN)/4
F       F2 = (P0*T2**3)+(P1*T2**2)+(P2*T2)+(P3)
F
F       T3 = TIN + 3*(TOUT-TIN)/4
F       F3 = (P0*T3**3)+(P1*T3**2)+(P2*T3)+(P3)
F
F       T4 = TIN + 4*(TOUT-TIN)/4
F       F4 = (P0*T4**3)+(P1*T4**2)+(P2*T4)+(P3)
F
F       QCURVE = (0.5)*ABS(TOUT-TIN)/4*(F0+2*F1+2*F2+2*F3+F4)
F       AVGQ = 1/ABS(TOUT-TIN)*QCURVE
F
F       EI = 0.1
F       EO = 0
F
F       FCONV = 0.5
F       DELTOH2 = 2 * FCONV *(EI - EO) * STR
F       Q = AVGQ * DELTOH2
F
F
    READ-VARS TOUT TIN STR
    WRITE-VARS Q

TRANSFER T-1
    SET STREAM-VAR STREAM=24,1 SUBSTREAM=MIXED VARIABLE=TEMP  &
        UOM="F"
    EQUAL-TO STREAM-VAR STREAM=24 SUBSTREAM=MIXED VARIABLE=TEMP  &
        UOM="F"

TRANSFER T-2
    SET STREAM-VAR STREAM=24,1 SUBSTREAM=MIXED VARIABLE=PRES  &
        UOM="psia"
    EQUAL-TO STREAM-VAR STREAM=24 SUBSTREAM=MIXED VARIABLE=PRES  &
        UOM="psia"

TRANSFER T-3
    SET STREAM-VAR STREAM=24,1 SUBSTREAM=MIXED  &
```

```
        VARIABLE=MOLE-FLOW UOM="lbmol/hr"
    EQUAL-TO STREAM-VAR STREAM=24 SUBSTREAM=MIXED  &
        VARIABLE=MOLE-FLOW UOM="lbmol/hr"

CONSTRAINT DTHX
    VECTOR-DEF DT PROFILE BLOCK=BAXH VARIABLE=DTBASE  &
        SENTENCE=BASE-PROFILE UOM="F"
    PARAM VECTOR=YES
    SPEC "DT" GE "5"
    TOL-SPEC "0.1"

CONSTRAINT DTHX2
    VECTOR-DEF DT2 PROFILE BLOCK=BAXH VARIABLE=DTBASE  &
        SENTENCE=BASE-PROFILE UOM="F"
    PARAM VECTOR=YES
    SPEC "DT2" LE "200"
    TOL-SPEC "10"

CONSTRAINT DTHX18
    VECTOR-DEF DT22 PROFILE BLOCK=18 VARIABLE=DTBASE  &
        SENTENCE=BASE-PROFILE UOM="F"
    PARAM VECTOR=YES
    SPEC "DT22" LE "5"
    TOL-SPEC "0.1"

CONSTRAINT DTHX22
    VECTOR-DEF DT22 PROFILE BLOCK=22 VARIABLE=DTBASE  &
        SENTENCE=BASE-PROFILE UOM="F"
    PARAM VECTOR=YES
    SPEC "DT22" LE "5"
    TOL-SPEC "0.1"

CONSTRAINT DTHX26
    VECTOR-DEF DT26 PROFILE BLOCK=26 VARIABLE=DTBASE  &
        SENTENCE=BASE-PROFILE UOM="F"
    PARAM VECTOR=YES
    SPEC "DT26" LE "5"
    TOL-SPEC "0.1"

CONSTRAINT DTHX30
    VECTOR-DEF DT30 PROFILE BLOCK=30 VARIABLE=DTBASE  &
        SENTENCE=BASE-PROFILE UOM="F"
    PARAM VECTOR=YES
    SPEC "DT30" LE "5"
    TOL-SPEC "0.1"

CONSTRAINT FLOW
    DEFINE F2 STREAM-VAR STREAM=40LN2 SUBSTREAM=MIXED  &
        VARIABLE=MOLE-FLOW UOM="lbmol/hr"
    DEFINE F1 STREAM-VAR STREAM=10LN2 SUBSTREAM=MIXED  &
        VARIABLE=MOLE-FLOW UOM="lbmol/hr"
    SPEC "F2" EQ "F1"
    TOL-SPEC "0.1"
```

```
CONSTRAINT FLOW2
    DEFINE F2 STREAM-VAR STREAM=50 SUBSTREAM=MIXED  &
        VARIABLE=MOLE-FLOW UOM="lbmol/hr"
    DEFINE F1 STREAM-VAR STREAM=T50 SUBSTREAM=MIXED  &
        VARIABLE=MOLE-FLOW UOM="lbmol/hr"
    SPEC "F2" EQ "F1"
    TOL-SPEC "0.1"

CONSTRAINT FLOW3
    DEFINE F1 STREAM-VAR STREAM=68 SUBSTREAM=MIXED  &
        VARIABLE=MOLE-FLOW UOM="lbmol/hr"
    DEFINE F2 STREAM-VAR STREAM=T68 SUBSTREAM=MIXED  &
        VARIABLE=MOLE-FLOW UOM="lbmol/hr"
    SPEC "F1" EQ "F2"
    TOL-SPEC "0.1"

CONSTRAINT NTUHX
    DEFINE NTU BLOCK-VAR BLOCK=BAXH VARIABLE=CSNTU  &
        SENTENCE=RESULTS
    SPEC "NTU" LE "50"
    TOL-SPEC "1"

CONSTRAINT PRECYCLE
    DEFINE START STREAM-VAR STREAM=10LN2 SUBSTREAM=MIXED  &
        VARIABLE=PRES UOM="psia"
    DEFINE END STREAM-VAR STREAM=60 SUBSTREAM=MIXED  &
        VARIABLE=PRES UOM="psia"
    SPEC "START" EQ "END"
    TOL-SPEC "1"

CONSTRAINT TRECYCLE
    DEFINE START STREAM-VAR STREAM=10LN2 SUBSTREAM=MIXED  &
        VARIABLE=TEMP UOM="F"
    DEFINE END STREAM-VAR STREAM=60 SUBSTREAM=MIXED  &
        VARIABLE=TEMP UOM="F"
    SPEC "START" EQ "END"
    TOL-SPEC "1"

OPTIMIZATION OPT
    DEFINE PCMP BLOCK-VAR BLOCK=CMP VARIABLE=NET-WORK  &
        SENTENCE=RESULTS UOM="hp"
    DEFINE PEXP BLOCK-VAR BLOCK=EXP VARIABLE=BRAKE-POWER  &
        SENTENCE=RESULTS UOM="hp"
    DEFINE PRECOMP BLOCK-VAR BLOCK=12 VARIABLE=NET-WORK  &
        SENTENCE=RESULTS UOM="hp"
    DEFINE NEONCOMP BLOCK-VAR BLOCK=94 VARIABLE=NET-WORK  &
        SENTENCE=RESULTS UOM="hp"
    DEFINE RECYCOMP BLOCK-VAR BLOCK=48 VARIABLE=NET-WORK  &
        SENTENCE=RESULTS UOM="hp"
    DEFINE HIGHTURB BLOCK-VAR BLOCK=84,1 VARIABLE=NET-WORK  &
        SENTENCE=RESULTS UOM="hp"
    DEFINE LOWTURB BLOCK-VAR BLOCK=76,1 VARIABLE=NET-WORK  &
        SENTENCE=RESULTS UOM="hp"
    DEFINE DFE BLOCK-VAR BLOCK=34,B VARIABLE=NET-WORK  &
```

```
        SENTENCE=RESULTS UOM="hp"
    MINIMIZE  &
        "PCOMP + PEXP +PRECOMP+NEONCOMP+RECYCOMP+HIGHTURB+LOWTURB+DFE"
    CONSTRAINTS FLOW / NTUHX / FLOW2 / DTHX / DTHX2 /  &
        DTHX22 / DTHX26 / DTHX30 / DTHX18 / FLOW3
    VARY PARAMETER 1
    LIMITS "1" "30000" STEP-SIZE=1. MAX-STEP-SIZ=0.1
    VARY STREAM-VAR STREAM=T20 SUBSTREAM=MIXED  &
        VARIABLE=MOLE-FLOW UOM="lbmol/hr"
    LIMITS "10" "5000" STEP-SIZE=1. MAX-STEP-SIZ=0.1
    VARY BLOCK-VAR BLOCK=BAXH VARIABLE=TEMP SENTENCE=HOT-SIDE  &
        ID1=25 UOM="F"
    LIMITS "-300" "70" STEP-SIZE=0.1 MAX-STEP-SIZ=0.1
    VARY BLOCK-VAR BLOCK=SPL SENTENCE=FRAC VARIABLE=FRAC ID1=55
    LIMITS "0.001" "0.999" STEP-SIZE=0.0001 MAX-STEP-SIZ=0.1
    VARY BLOCK-VAR BLOCK=BAXH VARIABLE=TEMP SENTENCE=HOT-SIDE  &
        ID1=55 UOM="F"
    LIMITS "-300" "-260" STEP-SIZE=0.1 MAX-STEP-SIZ=0.1
    VARY STREAM-VAR STREAM=68 SUBSTREAM=MIXED VARIABLE=MOLE-FLOW  &
        UOM="lbmol/hr"
    LIMITS "1000" "5000*2" STEP-SIZE=1. MAX-STEP-SIZ=0.1
    VARY STREAM-VAR STREAM=10LN2 SUBSTREAM=MIXED  &
        VARIABLE=MOLE-FLOW UOM="lbmol/hr"
    LIMITS "0" "1000" STEP-SIZE=1. MAX-STEP-SIZ=0.1
    VARY STREAM-VAR STREAM=T50 SUBSTREAM=MIXED  &
        VARIABLE=MOLE-FLOW UOM="lbmol/hr"
    LIMITS "0" "1000" STEP-SIZE=0.1 MAX-STEP-SIZ=0.1

;    VARY BLOCK-VAR BLOCK=EXP2 SENTENCE=PARAM VARIABLE=PRES
;    LIMITS "20" "120" STEP-SIZE=0.1 MAX-STEP-SIZ=.05

CONV-OPTIONS
    PARAM TOL=.00010 TEAR-VAR=YES

CONVERGENCE OPT SQP
    OPTIMIZE OPTID=OPT
    PARAM MAXIT=100 MAXPASS=1500 NLIMIT=45 STEP-OPT=VALUE  &
        OPT-METHOD=BIEGLER CONST-ITER=5

REPORT INPUT

STREAM-REPOR WIDE MOLEFLOW MOLEFRAC PROPERTIES=ACFM CPCV VISCV  &
        RHOV VISCL RHOL TEMPC PRES KGFLW

PROPERTY-REP PCES

DISABLE
    BLOCK BAXH CMP EXP JT JT2 KO LINKO MKP SPL
    STREAM 10LN2 15 20 25 30LN2 35 40LN2 45 50LN2 52 55  &
        56 60LN2 65 T20
;
;
;
```

24.12 ASPEN PLUS Report File

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 ASPEN PLUS IS A TRADEMARK OF           HOTLINE:
 ASPEN TECHNOLOGY, INC.                 U.S.A. 888/996-7100
 781/221-6400                           EUROPE (44) 1189-226555

 PLATFORM: WIN-X64                                       APRIL 18, 2023
 VERSION: 39.0   Build 116                               TUESDAY
 INSTALLATION:                                           11:03:59 A.M.
```

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EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
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```
RUN CONTROL INFORMATION
-----------------------

THIS COPY OF ASPEN PLUS LICENSED TO UNIVERSITY OF PENNSYLVAN

TYPE OF RUN: NEW

INPUT FILE NAME: _3651doa.inm

OUTPUT PROBLEM DATA FILE NAME: _3651doa
LOCATED IN:


PDF SIZE USED FOR INPUT TRANSLATION:
  NUMBER OF FILE RECORDS (PSIZE) =     0
  NUMBER OF IN-CORE RECORDS      =   256
PSIZE NEEDED FOR SIMULATION    =   256

CALLING PROGRAM NAME:          apmain
LOCATED IN: C:\Program Files\AspenTech\Aspen Plus V12.1\Engine\\xeq

SIMULATION REQUESTED FOR ENTIRE FLOWSHEET
```

FLOWSHEET CONNECTIVITY BY STREAMS

STREAM	SOURCE	DEST	STREAM	SOURCE	DEST
T56	----	MIXN2	24,1	----	$26H04
T52	----	$22HTR	68	----	$18H02
T50	----	MIX	10	----	12
T20	----	SPL	10LN2	----	MKP
36,B2	34,B	38	36,A1	34SPLIT	34,A
36,B1	34SPLIT	34,B	42	38	$30HTR
40	38	----	86	84,1	MIX2
78	76,1	$30HTR	58	MIXN2	$18HTR
88	MIX2	$26HTR	72	SPLIT	$26H02
82	SPLIT	84,1	36,A2	34,A	38
32	$30H03	34SPLIT	46,-2	$30HTR	$26HTR
80	$30HTR	MIX2	90	$26HTR	$18HTR
74	$26H02	76,1	46,+R	$26HTR	$18HTR
28,1	$26H04	$30H03	70	$22H02	SPLIT
54	$22HTR	MIXN2	24	$22H01	----
T68	94	----	46	$18HTR	48
92	$18HTR	94	20L	$18H01	$22H01
68,1	$18H02	$22H02	60	$18HTR	----
16	MIX	$18H01	50	48	----
14	12	MIX	FEEDN2	JT2	LINKO
25	SPL	$BAXHH01	55	SPL	$BAXHH02
15	MKP	CMP	45	KO	$BAXHHTR
40LN2	KO	JT2	35	JT	KO
50LN2	$BAXHHTR	MKP	60LN2	$BAXHH01	EXP
30LN2	$BAXHH02	JT	65	EXP	KO
20	CMP	----	56	LINKO	----
52	LINKO	----	$18Q01	$18H01	$18HTR
$18Q02	$18H02	$18HTR	$22Q01	$22H01	$22HTR
$22Q02	$22H02	$22HTR	$26Q02	$26H02	$26HTR
$26Q04	$26H04	$26HTR	$30Q03	$30H03	$30HTR
$BAXHQ01	$BAXHH01	$BAXHHTR	$BAXHQ02	$BAXHH02	$BAXHHTR

FLOWSHEET CONNECTIVITY BY BLOCKS

BLOCK	INLETS	OUTLETS
34,B	36,B1	36,B2
34SPLIT	32	36,A1 36,B1
38	36,A2 36,B2	42 40
84,1	82	86
76,1	74	78
MIXN2	54 T56	58
MIX2	80 86	88
SPLIT	70	72 82

```
34,A          36,A1                          36,A2
94            92                             T68
MIX           14 T50                         16
48            46                             50
12            10                             14
JT2           40LN2                          FEEDN2
SPL           T20                            25 55
MKP           10LN2 50LN2                    15
```

```
 FLOWSHEET CONNECTIVITY BY BLOCKS (CONTINUED)
   KO           35 65                          45 40LN2
   JT           30LN2                          35
   EXP          60LN2                          65
   CMP          15                             20
   LINKO        FEEDN2                         56 52
   $18H01       16                             20L $18Q01
   $18H02       68                             68,1 $18Q02
   $18HTR       90 58 46,+R $18Q01 $18Q02      92 60 46
   $22H01       20L                            24 $22Q01
   $22H02       68,1                           70 $22Q02
   $22HTR       T52 $22Q01 $22Q02              54
   $26H02       72                             74 $26Q02
   $26H04       24,1                           28,1 $26Q04
   $26HTR       46,-2 88 $26Q02 $26Q04         46,+R 90
   $30H03       28,1                           32 $30Q03
   $30HTR       78 42 $30Q03                   80 46,-2
   $BAXHH01     25                             60LN2 $BAXHQ01
   $BAXHH02     55                             30LN2 $BAXHQ02
   $BAXHHTR     45 $BAXHQ01 $BAXHQ02           50LN2

 TRANSFER BLOCK: T-1
 -------------------
     EQUAL-TO : TEMPERATURE IN STREAM 24 SUBSTREAM MIXED
     SET      : TEMPERATURE IN STREAM 24,1 SUBSTREAM MIXED

 TRANSFER BLOCK: T-2
 -------------------
     EQUAL-TO : PRESSURE IN STREAM 24 SUBSTREAM MIXED
     SET      : PRESSURE IN STREAM 24,1 SUBSTREAM MIXED

 TRANSFER BLOCK: T-3
 -------------------
     EQUAL-TO : TOTAL MOLEFLOW IN STREAM 24 SUBSTREAM MIXED
     SET      : TOTAL MOLEFLOW IN STREAM 24,1 SUBSTREAM MIXED

 CALCULATOR BLOCK:   INIT
 ------------------------

   SAMPLED VARIABLES:
     P        : PARAMETER 1

   FORTRAN STATEMENTS:
         P=700

   WRITE VARIABLES: P

   VALUES OF ACCESSED FORTRAN VARIABLES ON MOST RECENT SIMULATION PASS:
     VARIABLE        VALUE READ      VALUE WRITTEN    UNITS
```

```
--------        ----------      -------------    -----
P                MISSING           700.000
```

```
 CALCULATOR BLOCK:   OP1
 -----------------------

   SAMPLED VARIABLES:
     Q        : SENTENCE=PARAM VARIABLE=QLEAK IN UOS BLOCK $26HTR
     TOUT     : TEMPERATURE IN STREAM 28,1 SUBSTREAM MIXED
     TIN      : TEMPERATURE IN STREAM 24,1 SUBSTREAM MIXED
     STR      : TOTAL MOLEFLOW IN STREAM 24,1 SUBSTREAM MIXED

   FORTRAN STATEMENTS:
           P0 = 9*10**-6
           P1 = 0.0045
           P2 = -0.3074
           P3 = 28.835

           F0 =  (P0*TIN**3)+(P1*TIN**2)+(P2*TIN)+(P3)

           T1 = TIN + 1*(TOUT-TIN)/4
           F1 =  (P0*T1**3)+(P1*T1**2)+(P2*T1)+(P3)

           T2 = TIN + 2*(TOUT-TIN)/4
           F2 =  (P0*T2**3)+(P1*T2**2)+(P2*T2)+(P3)

           T3 = TIN + 3*(TOUT-TIN)/4
           F3 =  (P0*T3**3)+(P1*T3**2)+(P2*T3)+(P3)

           T4 = TIN + 4*(TOUT-TIN)/4
           F4 =  (P0*T4**3)+(P1*T4**2)+(P2*T4)+(P3)

           QCURVE = (0.5)*ABS(TOUT-TIN)/4*(F0+2*F1+2*F2+2*F3+F4)
           AVGQ = 1/ABS(TOUT-TIN)*QCURVE

           M0 = -2*10**-13
           M1 = -1*10**-4
           M2 = 9*10**-10
           M3 = 6*10**-7
           M4 = 0.0002
           M5 = 0.7359
           EI = 0.62
           EO = 0.1

           FCONV = 0.2
           DELTOH2 = 2 * FCONV * (EI - EO) * STR
           Q = AVGQ * DELTOH2


   READ VARIABLES:  TOUT      TIN       STR

   WRITE VARIABLES: Q
```

```
VALUES OF ACCESSED FORTRAN VARIABLES ON MOST RECENT SIMULATION PASS:
  VARIABLE        VALUE READ      VALUE WRITTEN    UNITS
  --------        ----------      -------------    -----
  Q                99.0000          314531.        BTU/HR
```

```
CALCULATOR BLOCK:   OP1 (CONTINUED)
    TOUT            -367.000                        F
    TIN             -305.000                        F
    STR              2356.23                        LBMOL/HR

CALCULATOR BLOCK:   OP2
-----------------------

  SAMPLED VARIABLES:
    Q        : SENTENCE=PARAM VARIABLE=QLEAK IN UOS BLOCK $30HTR
    TOUT     : TEMPERATURE IN STREAM 32 SUBSTREAM MIXED
    TIN      : TEMPERATURE IN STREAM 28,1 SUBSTREAM MIXED
    STR      : TOTAL MOLEFLOW IN STREAM 28,1 SUBSTREAM MIXED

  FORTRAN STATEMENTS:
          P0 = 9*10**-6
          P1 = 0.0045
          P2 = -0.3074
          P3 = 28.835

          F0 = (P0*TIN**3)+(P1*TIN**2)+(P2*TIN)+(P3)

          T1 = TIN + 1*(TOUT-TIN)/4
          F1 = (P0*T1**3)+(P1*T1**2)+(P2*T1)+(P3)

          T2 = TIN + 2*(TOUT-TIN)/4
          F2 = (P0*T2**3)+(P1*T2**2)+(P2*T2)+(P3)

          T3 = TIN + 3*(TOUT-TIN)/4
          F3 = (P0*T3**3)+(P1*T3**2)+(P2*T3)+(P3)

          T4 = TIN + 4*(TOUT-TIN)/4
          F4 = (P0*T4**3)+(P1*T4**2)+(P2*T4)+(P3)

          QCURVE = (0.5)*ABS(TOUT-TIN)/4*(F0+2*F1+2*F2+2*F3+F4)
          AVGQ = 1/ABS(TOUT-TIN)*QCURVE

          EI = 0.1
          EO = 0

          FCONV = 0.5
          DELTOH2 = 2 * FCONV *(EI - EO) * STR
          Q = AVGQ * DELTOH2


  READ VARIABLES:  TOUT      TIN       STR

  WRITE VARIABLES: Q
```

```
VALUES OF ACCESSED FORTRAN VARIABLES ON MOST RECENT SIMULATION PASS:
  VARIABLE        VALUE READ      VALUE WRITTEN    UNITS
  --------        ----------      -------------    -----
  Q                130000.          192424.        BTU/HR
```

```
CALCULATOR BLOCK:   OP2 (CONTINUED)
    TOUT            -404.000                        F
    TIN             -367.000                        F
    STR              2356.23                        LBMOL/HR

COMPUTATIONAL SEQUENCE
----------------------

SEQUENCE USED WAS:
   INIT $18H02 12 MIX $18H01 $22H01 T-1 T-2 T-3 $26H04 OP1 $22H02 $22HTR
   MIXN2 SPLIT 84,1 $26H02 76,1 $30H03 OP2 34SPLIT 34,B *34,A 38 $30HTR
   MIX2 $26HTR $18HTR 94 48

OVERALL FLOWSHEET BALANCE
-------------------------

                     ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT         RELATIVE DIFF.
   CONVENTIONAL COMPONENTS (LBMOL/HR)
      NH2                      2356.23         2356.23          0.00000
      PH2                      2356.23         2356.23          0.00000
      HE                       0.00000         0.00000          0.00000
      N2                       2200.00         1500.00         0.318182
      NE                       9750.00         9750.00          0.00000
   TOTAL BALANCE
      MOLE(LBMOL/HR)           16662.5         15962.5         0.420106E-01
      MASS(LB/HR   )           267881.         248272.         0.732019E-01
      ENTHALPY(BTU/HR  )     -0.106772E+08   -0.116257E+08    0.815854E-01

                     ***  CO2 EQUIVALENT SUMMARY ***
   FEED STREAMS CO2E            0.00000      LB/HR
   PRODUCT STREAMS CO2E         0.00000      LB/HR
   NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
   UTILITIES CO2E PRODUCTION    0.00000      LB/HR
   TOTAL CO2E PRODUCTION        0.00000      LB/HR
```

COMPONENTS

ID	TYPE	ALIAS	NAME
NH2	C	H2	HYDROGEN
PH2	C	H2-PARA	PARA-HYDROGEN
HE	C	HE-4	HELIUM-4
N2	C	N2	NITROGEN
NE	C	NE	NEON

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
U-O-S BLOCK SECTION

```
 BLOCK:  18       MODEL: MHEATX
 ------------------------------

   HOT SIDE:    INLET STREAM     OUTLET STREAM
                ------------     -------------
                16               20L
                68               68,1

   COLD SIDE:   INLET STREAM     OUTLET STREAM
                ------------     -------------
                90               92
                58               60
                46,+R            46

   PROPERTIES FOR STREAM 16
   PROPERTY OPTION SET:   REFPROP

   PROPERTIES FOR STREAM 68
   PROPERTY OPTION SET:   REFPROP

   PROPERTIES FOR STREAM 90
   PROPERTY OPTION SET:   REFPROP

   PROPERTIES FOR STREAM 58
   PROPERTY OPTION SET:   REFPROP

   PROPERTIES FOR STREAM 46,+R
   PROPERTY OPTION SET:   REFPROP

                    ***  MASS AND ENERGY BALANCE  ***
                                IN              OUT        RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          23660.1         23660.1        0.153760E-15
       MASS(LB/HR   )          440887.         440887.        0.132024E-15
       ENTHALPY(BTU/HR  )    -0.222103E+08   -0.222103E+08    0.849750E-07
```

```
 BLOCK:  18       MODEL: MHEATX (CONTINUED)

                    ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E              0.00000      LB/HR
    PRODUCT STREAMS CO2E           0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION    0.00000      LB/HR
    UTILITIES CO2E PRODUCTION      0.00000      LB/HR
    TOTAL CO2E PRODUCTION          0.00000      LB/HR

                        ***  INPUT DATA  ***

   MAXIMUM NO. ITERATIONS                                          30
   CONVERGENCE TOLERANCE                                            0.000100000

   SPECIFICATIONS FOR STREAM 16      :
   TWO    PHASE  TP  FLASH
   SPECIFIED TEMPERATURE                F                            -290.000
   PRESSURE DROP                        PSI                             7.00000
   MAXIMUM NO. ITERATIONS                                              30
   CONVERGENCE TOLERANCE
0.100000-06

   SPECIFICATIONS FOR STREAM 68      :
   TWO    PHASE  TP  FLASH
   SPECIFIED TEMPERATURE                F                            -290.000
   PRESSURE DROP                        PSI                             6.19000
   MAXIMUM NO. ITERATIONS                                              30
   CONVERGENCE TOLERANCE
0.100000-06

   SPECIFICATIONS FOR STREAM 90      :
   TWO    PHASE      FLASH
   PRESSURE DROP                        PSI                             2.65000
   MAXIMUM NO. ITERATIONS                                              30
   CONVERGENCE TOLERANCE
0.100000-06

   SPECIFICATIONS FOR STREAM 58      :
   TWO    PHASE      FLASH
   PRESSURE DROP                        PSI                             1.43000
   MAXIMUM NO. ITERATIONS                                              30
   CONVERGENCE TOLERANCE
0.100000-06

   SPECIFICATIONS FOR STREAM 46,+R   :
   TWO    PHASE      FLASH
   PRESSURE DROP                        PSI                             0.47000
   MAXIMUM NO. ITERATIONS                                              30
```

```
   CONVERGENCE TOLERANCE
0.100000-06
```

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
U-O-S BLOCK SECTION

BLOCK: 18 MODEL: MHEATX (CONTINUED)

*** RESULTS ***

INLET STREAM	DUTY BTU/HR	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	OUTLET VAPOR FRAC
16	-0.62105E+07	-290.00	643.00	1.0000
68	-0.19540E+08	-290.00	143.81	1.0000
90	0.19759E+08	98.73	21.280	1.0000
58	0.50771E+07	98.73	15.570	1.0000
46,+R	0.91432E+06	98.73	50.570	1.0000

```
          ----------------------------------------
          |                                      |
 16       |                                      | 20L
--------->|        2356.2     LBMOL/HR           |--------->
   110.00 |                                      |  -290.00
          |                                      |
 68       |                                      | 68,1
--------->|        9750.0     LBMOL/HR           |--------->
   110.00 |                                      |  -290.00
          |                                      |
 92       |                                      | 90
<---------|        9750.0     LBMOL/HR           |<---------
    98.73 |                                      |  -308.75
          |                                      |
 60       |                                      | 58
<---------|        1500.0     LBMOL/HR           |<---------
    98.73 |                                      |  -318.17
          |                                      |
 46       |                                      | 46,+R
<---------|        303.90     LBMOL/HR           |<---------
    98.73 |                                      |  -308.75
          |                                      |
          ----------------------------------------
```

```
 BLOCK:  18       MODEL: MHEATX (CONTINUED)

                          ***  INTERNAL ANALYSIS  ***

  FLOW IS COUNTERCURRENT.
  DUTY                              0.25751E+08 BTU/HR
  UA                                0.12472E+07 BTU/HR-R
  AVERAGE LMTD (DUTY/UA)             20.647     F
  MIN TEMP APPROACH                  11.272     F
  HOT-SIDE TEMP APPROACH             11.272     F
  COLD-SIDE TEMP APPROACH            28.168     F
  HOT-SIDE NTU                       19.373
  COLD-SIDE NTU                      20.191

  TQ-TABLE(S) INTERPOLATED FOR AT LEAST ONE STREAM DURING INTERNAL ZONE
ANALYSIS.

  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       -290.00   -318.17     28.17
  0.2575E+06   -285.97   -318.24     32.26     30.17   8536.
0.2575E+06   8536.
  0.3863E+06   -283.96   -318.27     34.31     33.27   3870.
0.1288E+06  0.1241E+05
  0.5150E+06   -281.95   -318.30     36.35     35.32   3645.
0.1288E+06  0.1605E+05
  0.7777E+06   -277.84   -308.75     30.91     33.56   7829.
0.2627E+06  0.2388E+05         IN  90
  0.1030E+07   -273.90   -304.66     30.76     30.83   8183.
0.2523E+06  0.3206E+05
  0.1545E+07   -265.85   -296.30     30.45     30.60  0.1683E+05
0.5150E+06  0.4889E+05
  0.2060E+07   -257.80   -287.93     30.14     30.29  0.1700E+05
0.5150E+06  0.6590E+05
  0.2575E+07   -249.75   -279.57     29.83     29.98  0.1718E+05
0.5150E+06  0.8307E+05
  0.3090E+07   -241.66   -271.21     29.55     29.69  0.1735E+05
0.5150E+06  0.1004E+06
  0.3605E+07   -233.59   -262.85     29.26     29.40  0.1751E+05
0.5150E+06  0.1179E+06
  0.4120E+07   -225.52   -254.43     28.92     29.09  0.1771E+05
0.5150E+06  0.1356E+06
  0.4635E+07   -217.45   -246.05     28.60     28.76  0.1791E+05
0.5150E+06  0.1535E+06
```

0.5150E+07	-209.38	-237.67	28.29	28.45	0.1810E+05	0.5150E+06	0.1717E+06
0.5665E+07	-201.30	-229.28	27.98	28.14	0.1830E+05	0.5150E+06	0.1900E+06
0.6180E+07	-193.20	-220.90	27.69	27.84	0.1850E+05	0.5150E+06	0.2085E+06
0.6695E+07	-185.16	-212.51	27.35	27.52	0.1871E+05	0.5150E+06	0.2272E+06
0.7210E+07	-177.11	-204.13	27.02	27.18	0.1895E+05	0.5150E+06	0.2461E+06
0.7725E+07	-169.07	-195.75	26.68	26.85	0.1918E+05	0.5150E+06	0.2653E+06
0.8240E+07	-161.02	-187.36	26.34	26.51	0.1943E+05	0.5150E+06	0.2847E+06
0.8755E+07	-152.97	-178.97	26.01	26.17	0.1968E+05	0.5150E+06	0.3044E+06
0.9270E+07	-144.92	-170.58	25.66	25.83	0.1994E+05	0.5150E+06	0.3243E+06
0.9785E+07	-136.90	-162.19	25.30	25.48	0.2021E+05	0.5150E+06	0.3446E+06
0.1030E+08	-128.90	-153.80	24.91	25.10	0.2052E+05	0.5150E+06	0.3651E+06
0.1082E+08	-120.86	-145.42	24.55	24.73	0.2083E+05	0.5150E+06	0.3859E+06
0.1133E+08	-112.86	-137.02	24.16	24.36	0.2115E+05	0.5150E+06	0.4070E+06
0.1185E+08	-104.86	-128.63	23.77	23.96	0.2149E+05	0.5150E+06	0.4285E+06
0.1236E+08	-96.86	-120.23	23.37	23.57	0.2185E+05	0.5150E+06	0.4504E+06
0.1288E+08	-88.86	-111.83	22.97	23.17	0.2223E+05	0.5150E+06	0.4726E+06
0.1339E+08	-80.89	-103.43	22.54	22.75	0.2264E+05	0.5150E+06	0.4953E+06
0.1391E+08	-72.89	-95.03	22.14	22.34	0.2306E+05	0.5150E+06	0.5183E+06
0.1442E+08	-64.92	-86.62	21.71	21.92	0.2349E+05	0.5150E+06	0.5418E+06
0.1494E+08	-56.92	-78.22	21.30	21.50	0.2395E+05	0.5150E+06	0.5658E+06
0.1545E+08	-48.95	-69.80	20.85	21.08	0.2444E+05	0.5150E+06	0.5902E+06
0.1597E+08	-40.98	-61.39	20.41	20.63	0.2496E+05	0.5150E+06	0.6152E+06
0.1648E+08	-33.02	-52.98	19.96	20.18	0.2552E+05	0.5150E+06	0.6407E+06
0.1700E+08	-25.05	-44.56	19.51	19.74	0.2610E+05	0.5150E+06	0.6668E+06
0.1751E+08	-17.08	-36.15	19.07	19.29	0.2670E+05	0.5150E+06	0.6935E+06
0.1803E+08	-9.13	-27.73	18.60	18.83	0.2735E+05	0.5150E+06	0.7208E+06

BLOCK: 18   MODEL: MHEATX (CONTINUED)

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		
0.1854E+08	-1.18	-19.30	18.13	18.36	0.2805E+05	0.5150E+06	0.7489E+06		
0.1906E+08	6.77	-10.88	17.65	17.89	0.2879E+05	0.5150E+06	0.7777E+06		
0.1957E+08	14.73	-2.46	17.18	17.42	0.2957E+05	0.5150E+06	0.8072E+06		
0.2009E+08	22.68	5.97	16.71	16.95	0.3039E+05	0.5150E+06	0.8376E+06		
0.2060E+08	30.62	14.40	16.23	16.47	0.3127E+05	0.5150E+06	0.8689E+06		
0.2112E+08	38.57	22.83	15.74	15.98	0.3222E+05	0.5150E+06	0.9011E+06		
0.2163E+08	46.51	31.26	15.25	15.49	0.3324E+05	0.5150E+06	0.9344E+06		
0.2215E+08	54.45	39.69	14.76	15.00	0.3432E+05	0.5150E+06	0.9687E+06		
0.2266E+08	62.39	48.12	14.27	14.51	0.3548E+05	0.5150E+06	0.1004E+07		
0.2318E+08	70.33	56.55	13.78	14.02	0.3673E+05	0.5150E+06	0.1041E+07		
0.2369E+08	78.26	64.99	13.28	13.53	0.3808E+05	0.5150E+06	0.1079E+07		
0.2421E+08	86.20	73.42	12.78	13.02	0.3954E+05	0.5150E+06	0.1119E+07		
0.2472E+08	94.13	81.86	12.27	12.52	0.4112E+05	0.5150E+06	0.1160E+07		
0.2524E+08	102.07	90.29	11.77	12.02	0.4284E+05	0.5150E+06	0.1202E+07		
0.2575E+08	110.00	98.73	11.27	11.52	0.4470E+05	0.5150E+06	0.1247E+07	GBL	

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

BLOCK: 18 MODEL: MHEATX (CONTINUED)

*** TQ-TABLES FOR HOT SIDE STREAMS ***

STREAM: 16

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.6210E+07	110.0	650.0	1.000
0.5520E+07	67.87	649.2	1.000
0.4830E+07	25.54	648.4	1.000
0.4140E+07	-17.10	647.7	1.000
0.3450E+07	-60.18	646.9	1.000
0.2760E+07	-103.9	646.1	1.000
0.2070E+07	-148.5	645.3	1.000
0.1380E+07	-194.3	644.6	1.000
0.6901E+06	-241.6	643.8	1.000
0.000	-290.0	643.0	1.000

STREAM: 68

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.1954E+08	110.0	150.0	1.000
0.1737E+08	65.26	149.3	1.000
0.1520E+08	20.55	148.6	1.000
0.1303E+08	-24.14	147.9	1.000
0.1086E+08	-68.78	147.2	1.000
0.8685E+07	-113.4	146.6	1.000
0.6513E+07	-157.9	145.9	1.000
0.4342E+07	-202.2	145.2	1.000
0.2171E+07	-246.3	144.5	1.000
0.000	-290.0	143.8	1.000

BLOCK: 18 MODEL: MHEATX (CONTINUED)

*** TQ-TABLES FOR COLD SIDE STREAMS ***

STREAM: 90

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.000	-308.8	23.93	1.000
0.2195E+07	-263.7	23.64	1.000
0.4391E+07	-218.5	23.34	1.000
0.6586E+07	-173.2	23.05	1.000
0.8782E+07	-127.9	22.75	1.000
0.1098E+08	-82.60	22.46	1.000
0.1317E+08	-37.28	22.16	1.000
0.1537E+08	8.053	21.87	1.000
0.1756E+08	53.39	21.57	1.000
0.1976E+08	98.73	21.28	1.000

STREAM: 58

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.000	-318.2	17.00	0.8111
0.6346E+06	-318.3	16.82	0.9893
0.6727E+06	-318.3	16.81	1.000
0.1269E+07	-263.9	16.64	1.000
0.1904E+07	-204.1	16.46	1.000
0.2539E+07	-143.8	16.29	1.000
0.3173E+07	-83.31	16.11	1.000
0.3808E+07	-22.67	15.93	1.000
0.4442E+07	38.02	15.75	1.000
0.5077E+07	98.73	15.57	1.000

STREAM: 46,+R

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.000	-308.8	51.04	1.000
0.1016E+06	-256.8	50.99	1.000
0.2032E+06	-211.4	50.94	1.000
0.3048E+06	-168.6	50.88	1.000

0.4064E+06	-126.2	50.83	1.000
0.5080E+06	-83.15	50.78	1.000
0.6095E+06	-39.08	50.73	1.000
0.7111E+06	6.036	50.67	1.000
0.8127E+06	52.05	50.62	1.000

```
 BLOCK:  18       MODEL: MHEATX (CONTINUED)

   Q STREAM    TEMP        PRES        VFRAC

   BTU/HR      F           PSIA

   0.9143E+06   98.73       50.57       1.000

 BLOCK:  22       MODEL: MHEATX
 ------------------------------

   HOT SIDE:    INLET STREAM     OUTLET STREAM
                ------------     -------------
                20L              24
                68,1             70

   COLD SIDE:   INLET STREAM     OUTLET STREAM
                ------------     -------------
                T52              54

   PROPERTIES FOR STREAM 20L
   PROPERTY OPTION SET:   REFPROP

   PROPERTIES FOR STREAM 68,1
   PROPERTY OPTION SET:   REFPROP

   PROPERTIES FOR STREAM T52
   PROPERTY OPTION SET:   REFPROP

                       ***  MASS AND ENERGY BALANCE  ***
                                    IN              OUT         RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          12802.2         12802.2        0.142084E-
15
       MASS(LB/HR   )          220999.         220999.         0.00000
       ENTHALPY(BTU/HR  )    -0.271184E+08   -0.271184E+08   -0.137371E-
15
```

BLOCK: 22 MODEL: MHEATX (CONTINUED)

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	0.00000	LB/HR
PRODUCT STREAMS CO2E	0.00000	LB/HR
NET STREAMS CO2E PRODUCTION	0.00000	LB/HR
UTILITIES CO2E PRODUCTION	0.00000	LB/HR
TOTAL CO2E PRODUCTION	0.00000	LB/HR

*** INPUT DATA ***

SPECIFICATIONS FOR STREAM 20L :
TWO PHASE TP FLASH

SPECIFIED TEMPERATURE	F	-305.000
PRESSURE DROP	PSI	2.30000
MAXIMUM NO. ITERATIONS		50
CONVERGENCE TOLERANCE		0.100000-06

SPECIFICATIONS FOR STREAM 68,1 :
TWO PHASE TP FLASH

SPECIFIED TEMPERATURE	F	-305.000
PRESSURE DROP	PSI	8.30000
MAXIMUM NO. ITERATIONS		50
CONVERGENCE TOLERANCE		0.100000-06

SPECIFICATIONS FOR STREAM T52 :
TWO PHASE FLASH

PRESSURE DROP	PSI	1.13000
MAXIMUM NO. ITERATIONS		50
CONVERGENCE TOLERANCE		0.100000-06

*** RESULTS ***

INLET STREAM	DUTY BTU/HR	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	OUTLET VAPOR FRAC
20L	-0.21460E+06	-305.00	640.70	1.0000
68,1	-0.74531E+06	-305.00	135.51	1.0000
T52	0.95991E+06	-316.50	18.870	0.5863

BLOCK: 22   MODEL: MHEATX (CONTINUED)

```
          -----------------------------------------
          |                                       |
 20L      |                                       | 24
--------->|         2356.2     LBMOL/HR           |--------->
  -290.00 |                                       |  -305.00
          |                                       |
 68,1     |                                       | 70
--------->|         9750.0     LBMOL/HR           |--------->
  -290.00 |                                       |  -305.00
          |                                       |
 54       |                                       | T52
<---------|         696.00     LBMOL/HR           |<---------
  -316.50 |                                       |  -316.00
          |                                       |
          -----------------------------------------
```

```
 BLOCK:  22       MODEL: MHEATX (CONTINUED)

                          ***  INTERNAL ANALYSIS  ***

  FLOW IS COUNTERCURRENT.
  DUTY                             0.95991E+06 BTU/HR
  UA                                55466.     BTU/HR-R
  AVERAGE LMTD (DUTY/UA)            17.306     F
  MIN TEMP APPROACH                 10.619     F
  HOT-SIDE TEMP APPROACH            26.497     F
  COLD-SIDE TEMP APPROACH           11.000     F
  HOT-SIDE NTU                     0.86673
  COLD-SIDE NTU                    -.28732E-01

  TQ-TABLE(S) INTERPOLATED FOR AT LEAST ONE STREAM DURING INTERNAL ZONE
ANALYSIS.

  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       -305.00   -316.00     11.00
   2145.       -304.97   -315.78     10.81     10.90   196.7        2145.
196.7
   3217.       -304.95   -315.66     10.71     10.76   99.65        1072.
296.3
   4290.       -304.93   -315.55     10.62     10.67   100.5        1072.
396.9      GBL    BP  T52
  0.1920E+05   -304.70   -315.57     10.86     10.74   1388.
0.1491E+05   1785.
  0.3840E+05   -304.40   -315.58     11.18     11.02   1742.
0.1920E+05   3527.
  0.5759E+05   -304.10   -315.60     11.50     11.34   1693.
0.1920E+05   5219.
  0.7679E+05   -303.80   -315.62     11.82     11.66   1647.
0.1920E+05   6866.
  0.9599E+05   -303.51   -315.64     12.14     11.98   1603.
0.1920E+05   8469.
  0.1152E+06   -303.21   -315.66     12.45     12.29   1562.
0.1920E+05  0.1003E+05
  0.1344E+06   -302.91   -315.68     12.77     12.61   1522.
0.1920E+05  0.1155E+05
  0.1536E+06   -302.61   -315.70     13.09     12.93   1485.
0.1920E+05  0.1304E+05
  0.1728E+06   -302.31   -315.72     13.41     13.25   1449.
0.1920E+05  0.1449E+05
```

0.1920E+06	-302.01	-315.73	13.73	13.57	1415.	0.1920E+05	0.1590E+05
0.2112E+06	-301.71	-315.75	14.04	13.88	1383.	0.1920E+05	0.1728E+05
0.2304E+06	-301.41	-315.77	14.36	14.20	1352.	0.1920E+05	0.1864E+05
0.2496E+06	-301.11	-315.79	14.68	14.52	1322.	0.1920E+05	0.1996E+05
0.2688E+06	-300.81	-315.81	15.00	14.84	1294.	0.1920E+05	0.2125E+05
0.2880E+06	-300.51	-315.83	15.32	15.16	1267.	0.1920E+05	0.2252E+05
0.3072E+06	-300.21	-315.85	15.64	15.48	1241.	0.1920E+05	0.2376E+05
0.3264E+06	-299.91	-315.87	15.95	15.79	1215.	0.1920E+05	0.2497E+05
0.3456E+06	-299.61	-315.88	16.27	16.11	1191.	0.1920E+05	0.2617E+05
0.3648E+06	-299.31	-315.90	16.59	16.43	1168.	0.1920E+05	0.2733E+05
0.3840E+06	-299.01	-315.92	16.91	16.75	1146.	0.1920E+05	0.2848E+05
0.4032E+06	-298.71	-315.94	17.23	17.07	1125.	0.1920E+05	0.2961E+05
0.4224E+06	-298.41	-315.96	17.55	17.39	1104.	0.1920E+05	0.3071E+05
0.4416E+06	-298.11	-315.98	17.87	17.71	1084.	0.1920E+05	0.3179E+05
0.4608E+06	-297.81	-316.00	18.19	18.03	1065.	0.1920E+05	0.3286E+05
0.4800E+06	-297.51	-316.02	18.51	18.35	1046.	0.1920E+05	0.3390E+05
0.4992E+06	-297.21	-316.04	18.83	18.67	1029.	0.1920E+05	0.3493E+05
0.5184E+06	-296.91	-316.06	19.14	18.98	1011.	0.1920E+05	0.3594E+05
0.5375E+06	-296.61	-316.07	19.46	19.30	994.6	0.1920E+05	0.3694E+05
0.5567E+06	-296.31	-316.09	19.78	19.62	978.4	0.1920E+05	0.3792E+05
0.5759E+06	-296.01	-316.11	20.10	19.94	962.7	0.1920E+05	0.3888E+05
0.5951E+06	-295.71	-316.13	20.42	20.26	947.5	0.1920E+05	0.3983E+05
0.6143E+06	-295.41	-316.15	20.74	20.58	932.8	0.1920E+05	0.4076E+05
0.6335E+06	-295.11	-316.17	21.06	20.90	918.6	0.1920E+05	0.4168E+05
0.6527E+06	-294.81	-316.19	21.38	21.22	904.8	0.1920E+05	0.4258E+05
0.6719E+06	-294.51	-316.21	21.70	21.54	891.3	0.1920E+05	0.4348E+05

BLOCK: 22 MODEL: MHEATX (CONTINUED)

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	
0.6911E+06	-294.21	-316.23	22.02	21.86	878.3	0.1920E+05	0.4435E+05	
0.7103E+06	-293.91	-316.25	22.34	22.18	865.7	0.1920E+05	0.4522E+05	
0.7295E+06	-293.61	-316.27	22.66	22.50	853.4	0.1920E+05	0.4607E+05	
0.7487E+06	-293.31	-316.28	22.98	22.82	841.4	0.1920E+05	0.4691E+05	
0.7679E+06	-293.01	-316.30	23.30	23.14	829.8	0.1920E+05	0.4774E+05	
0.7871E+06	-292.71	-316.32	23.62	23.46	818.5	0.1920E+05	0.4856E+05	
0.8063E+06	-292.41	-316.34	23.94	23.78	807.4	0.1920E+05	0.4937E+05	
0.8255E+06	-292.11	-316.36	24.26	24.10	796.7	0.1920E+05	0.5017E+05	
0.8447E+06	-291.80	-316.38	24.58	24.42	786.3	0.1920E+05	0.5095E+05	
0.8639E+06	-291.50	-316.40	24.90	24.74	776.1	0.1920E+05	0.5173E+05	
0.8831E+06	-291.20	-316.42	25.22	25.06	766.2	0.1920E+05	0.5250E+05	
0.9023E+06	-290.90	-316.44	25.54	25.38	756.5	0.1920E+05	0.5325E+05	
0.9215E+06	-290.60	-316.46	25.86	25.70	747.1	0.1920E+05	0.5400E+05	
0.9407E+06	-290.30	-316.48	26.18	26.02	737.9	0.1920E+05	0.5474E+05	
0.9599E+06	-290.00	-316.50	26.50	26.34	728.9	0.1920E+05	0.5547E+05	

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

BLOCK: 22 MODEL: MHEATX (CONTINUED)

*** TQ-TABLES FOR HOT SIDE STREAMS ***

STREAM: 20L

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.2146E+06	-290.0	643.0	1.000
0.1908E+06	-291.7	642.7	1.000
0.1669E+06	-293.3	642.5	1.000
0.1431E+06	-295.0	642.2	1.000
0.1192E+06	-296.7	642.0	1.000
0.9538E+05	-298.4	641.7	1.000
0.7153E+05	-300.0	641.5	1.000
0.4769E+05	-301.7	641.2	1.000
0.2384E+05	-303.3	641.0	1.000
0.000	-305.0	640.7	1.000

STREAM: 68,1

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.7453E+06	-290.0	143.8	1.000
0.6625E+06	-291.7	142.9	1.000
0.5797E+06	-293.3	142.0	1.000
0.4969E+06	-295.0	141.0	1.000
0.4141E+06	-296.7	140.1	1.000
0.3312E+06	-298.3	139.2	1.000
0.2484E+06	-300.0	138.3	1.000
0.1656E+06	-301.7	137.4	1.000
0.8281E+05	-303.3	136.4	1.000
0.000	-305.0	135.5	1.000

*** TQ-TABLES FOR COLD SIDE STREAMS ***

STREAM: T52

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.000	-316.0	20.00	0.000

4290.	-315.6	20.00	0.000
0.1200E+06	-315.7	19.86	0.7120E-01
0.2400E+06	-315.8	19.72	0.1450
0.3600E+06	-315.9	19.58	0.2187
0.4800E+06	-316.0	19.44	0.2923
0.5999E+06	-316.1	19.29	0.3659
0.7199E+06	-316.3	19.15	0.4394
0.8399E+06	-316.4	19.01	0.5129
0.9599E+06	-316.5	18.87	0.5863

```
 BLOCK:  26       MODEL: MHEATX
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   HOT SIDE:    INLET STREAM     OUTLET STREAM
                ------------     -------------
                72               74
                24,1             28,1

   COLD SIDE:   INLET STREAM     OUTLET STREAM
                ------------     -------------
                46,-2            46,+R
                88               90

   PROPERTIES FOR STREAM 72
   PROPERTY OPTION SET:   REFPROP

   PROPERTIES FOR STREAM 24,1
   PROPERTY OPTION SET:   REFPROP

   PROPERTIES FOR STREAM 46,-2
   PROPERTY OPTION SET:   REFPROP

   PROPERTIES FOR STREAM 88
   PROPERTY OPTION SET:   REFPROP

                      ***  MASS AND ENERGY BALANCE  ***
                                  IN               OUT         RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           17577.6         17577.6         0.00000
       MASS(LB/HR   )           306393.         306393.        -0.189977E-
15
       ENTHALPY(BTU/HR  )      -0.393246E+08   -0.390100E+08   -0.799839E-
02

                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E             0.00000      LB/HR
    PRODUCT STREAMS CO2E          0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR
```

BLOCK: 26 MODEL: MHEATX (CONTINUED)

*** INPUT DATA ***

MAXIMUM NO. ITERATIONS		30
CONVERGENCE TOLERANCE		0.000100000
LOWER LIMIT ON TEMPERATURE	F	-456.000
HEAT LEAK	BTU/HR	314,531.

SPECIFICATIONS FOR STREAM 72 :

TWO PHASE TP FLASH		
SPECIFIED TEMPERATURE	F	-367.000
PRESSURE DROP	PSI	2.94000
MAXIMUM NO. ITERATIONS		50
CONVERGENCE TOLERANCE		0.100000-06

SPECIFICATIONS FOR STREAM 24,1 :

TWO PHASE TP FLASH		
SPECIFIED TEMPERATURE	F	-367.000
PRESSURE DROP	PSI	0.58000
MAXIMUM NO. ITERATIONS		30
CONVERGENCE TOLERANCE		0.100000-06

SPECIFICATIONS FOR STREAM 46,-2 :

TWO PHASE FLASH		
PRESSURE DROP	PSI	0.60000
MAXIMUM NO. ITERATIONS		50
CONVERGENCE TOLERANCE		0.100000-06

SPECIFICATIONS FOR STREAM 88 :

TWO PHASE FLASH		
PRESSURE DROP	PSI	1.94000
MAXIMUM NO. ITERATIONS		50
CONVERGENCE TOLERANCE		0.100000-06

*** RESULTS ***

INLET STREAM	DUTY BTU/HR	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	OUTLET VAPOR FRAC
72	-0.17255E+07	-367.00	132.57	1.0000
24,1	-0.10694E+07	-367.00	640.12	1.0000
46,-2	0.10278E+06	-308.75	51.040	1.0000
88	0.30067E+07	-308.75	23.930	1.0000

BLOCK: 26   MODEL: MHEATX (CONTINUED)

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          ------------------------------------------
          |                                        |
 72       |                                        | 74
--------->|        5167.5     LBMOL/HR             |--------->
  -305.00 |                                        |  -367.00
          |                                        |
 24,1     |                                        | 28,1
--------->|        2356.2     LBMOL/HR             |--------->
  -305.00 |                                        |  -367.00
          |                                        |
 46,+R    |                                        | 46,-2
<---------|        303.90     LBMOL/HR             |<---------
  -308.75 |                                        |  -370.86
          |                                        |
 90       |                                        | 88
<---------|        9750.0     LBMOL/HR             |<---------
  -308.75 |                                        |  -369.80
          |                                        |
          ------------------------------------------
```

```
 BLOCK:  26       MODEL: MHEATX (CONTINUED)

                          ***  INTERNAL ANALYSIS  ***

  FLOW IS COUNTERCURRENT.
  DUTY                             0.27949E+07 BTU/HR
  HEAT LEAK                        0.31453E+06 BTU/HR
  UA                               0.11992E+07 BTU/HR-R
  AVERAGE LMTD (DUTY/UA)            2.3306     F
  MIN TEMP APPROACH                 1.8236     F
  HOT-SIDE TEMP APPROACH            3.7511     F
  COLD-SIDE TEMP APPROACH           3.8619     F
  HOT-SIDE NTU                      26.603
  COLD-SIDE NTU                     23.954

  TQ-TABLE(S) INTERPOLATED FOR AT LEAST ONE STREAM DURING INTERNAL ZONE
ANALYSIS.

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

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

   0.000       0.000       -367.00   -370.86      3.86
   762.4       848.2       -366.99   -370.33      3.35      3.60   211.8
762.4       211.8
   1525.       1696.       -366.97   -369.80      2.83      3.08   247.3
762.4       459.1            IN  88
  0.2871E+05  0.3194E+05   -366.44   -369.22      2.78      2.81   9690.
0.2719E+05  0.1015E+05
  0.5590E+05  0.6219E+05   -365.90   -368.63      2.72      2.75   9882.
0.2719E+05  0.2003E+05
  0.1118E+06  0.1244E+06   -364.81   -367.42      2.61      2.67
0.2096E+05  0.5590E+05  0.4099E+05
  0.1677E+06  0.1866E+06   -363.71   -366.21      2.50      2.56
0.2188E+05  0.5590E+05  0.6286E+05
  0.2236E+06  0.2488E+06   -362.62   -365.01      2.39      2.44
0.2288E+05  0.5590E+05  0.8574E+05
  0.2795E+06  0.3109E+06   -361.52   -363.80      2.28      2.33
0.2398E+05  0.5590E+05  0.1097E+06
  0.3354E+06  0.3731E+06   -360.39   -362.59      2.20      2.24
0.2499E+05  0.5590E+05  0.1347E+06
  0.3913E+06  0.4353E+06   -359.24   -361.38      2.14      2.17
0.2578E+05  0.5590E+05  0.1605E+06
  0.4472E+06  0.4975E+06   -358.08   -360.16      2.08      2.11
0.2652E+05  0.5590E+05  0.1870E+06
```

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  0.5031E+06  0.5597E+06   -356.93   -358.95      2.02      2.05
0.2731E+05  0.5590E+05  0.2143E+06
  0.5590E+06  0.6219E+06   -355.78   -357.73      1.96      1.99
0.2815E+05  0.5590E+05  0.2425E+06
  0.6149E+06  0.6841E+06   -354.60   -356.52      1.92      1.94
0.2888E+05  0.5590E+05  0.2713E+06
  0.6708E+06  0.7463E+06   -353.41   -355.30      1.89      1.90
0.2937E+05  0.5590E+05  0.3007E+06
  0.7267E+06  0.8085E+06   -352.21   -354.08      1.87      1.88
0.2971E+05  0.5590E+05  0.3304E+06
  0.7826E+06  0.8707E+06   -351.01   -352.87      1.85      1.86
0.3000E+05  0.5590E+05  0.3604E+06
  0.8385E+06  0.9328E+06   -349.81   -351.65      1.84      1.85
0.3029E+05  0.5590E+05  0.3907E+06
  0.8944E+06  0.9950E+06   -348.61   -350.43      1.82      1.83
0.3054E+05  0.5590E+05  0.4213E+06  GBL
  0.9503E+06  0.1057E+07   -347.38   -349.21      1.83      1.83
0.3060E+05  0.5590E+05  0.4519E+06
  0.1006E+07  0.1119E+07   -346.15   -347.99      1.84      1.84
0.3045E+05  0.5590E+05  0.4823E+06
  0.1062E+07  0.1182E+07   -344.91   -346.77      1.86      1.85
0.3020E+05  0.5590E+05  0.5125E+06
  0.1118E+07  0.1244E+07   -343.67   -345.55      1.87      1.87
0.2994E+05  0.5590E+05  0.5424E+06
  0.1174E+07  0.1306E+07   -342.43   -344.32      1.89      1.88
0.2969E+05  0.5590E+05  0.5721E+06
  0.1230E+07  0.1368E+07   -341.18   -343.10      1.92      1.91
0.2933E+05  0.5590E+05  0.6015E+06
  0.1286E+07  0.1430E+07   -339.92   -341.88      1.95      1.94
0.2887E+05  0.5590E+05  0.6303E+06
  0.1342E+07  0.1493E+07   -338.66   -340.65      2.00      1.97
0.2831E+05  0.5590E+05  0.6586E+06
  0.1397E+07  0.1555E+07   -337.39   -339.43      2.04      2.02
0.2773E+05  0.5590E+05  0.6864E+06
  0.1453E+07  0.1617E+07   -336.13   -338.21      2.08      2.06
0.2717E+05  0.5590E+05  0.7135E+06
  0.1509E+07  0.1679E+07   -334.86   -336.98      2.12      2.10
0.2664E+05  0.5590E+05  0.7402E+06
  0.1565E+07  0.1741E+07   -333.59   -335.76      2.17      2.14
0.2606E+05  0.5590E+05  0.7662E+06
  0.1621E+07  0.1803E+07   -332.30   -334.53      2.23      2.20
0.2542E+05  0.5590E+05  0.7917E+06
  0.1677E+07  0.1866E+07   -331.02   -333.31      2.29      2.26
0.2476E+05  0.5590E+05  0.8164E+06
  0.1733E+07  0.1928E+07   -329.73   -332.08      2.35      2.32
0.2413E+05  0.5590E+05  0.8405E+06
  0.1789E+07  0.1990E+07   -328.45   -330.85      2.41      2.38
0.2353E+05  0.5590E+05  0.8641E+06
  0.1845E+07  0.2052E+07   -327.16   -329.63      2.47      2.44
0.2295E+05  0.5590E+05  0.8870E+06
  0.1901E+07  0.2114E+07   -325.87   -328.40      2.53      2.50
0.2237E+05  0.5590E+05  0.9094E+06
  0.1956E+07  0.2177E+07   -324.57   -327.17      2.60      2.57
0.2177E+05  0.5590E+05  0.9312E+06
```

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 BLOCK:  26       MODEL: MHEATX (CONTINUED)

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

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

  0.2012E+07  0.2239E+07   -323.27   -325.95      2.67      2.64
0.2119E+05  0.5590E+05  0.9523E+06
  0.2068E+07  0.2301E+07   -321.97   -324.72      2.74      2.71
0.2063E+05  0.5590E+05  0.9730E+06
  0.2124E+07  0.2363E+07   -320.68   -323.49      2.82      2.78
0.2011E+05  0.5590E+05  0.9931E+06
  0.2180E+07  0.2425E+07   -319.38   -322.26      2.89      2.85
0.1961E+05  0.5590E+05  0.1013E+07
  0.2236E+07  0.2488E+07   -318.07   -321.04      2.96      2.93
0.1911E+05  0.5590E+05  0.1032E+07
  0.2292E+07  0.2550E+07   -316.77   -319.81      3.04      3.00
0.1862E+05  0.5590E+05  0.1050E+07
  0.2348E+07  0.2612E+07   -315.46   -318.58      3.12      3.08
0.1815E+05  0.5590E+05  0.1069E+07
  0.2404E+07  0.2674E+07   -314.16   -317.35      3.20      3.16
0.1771E+05  0.5590E+05  0.1086E+07
  0.2460E+07  0.2736E+07   -312.85   -316.12      3.27      3.23
0.1728E+05  0.5590E+05  0.1104E+07
  0.2515E+07  0.2799E+07   -311.54   -314.89      3.35      3.31
0.1688E+05  0.5590E+05  0.1120E+07
  0.2571E+07  0.2861E+07   -310.23   -313.67      3.43      3.39
0.1649E+05  0.5590E+05  0.1137E+07
  0.2627E+07  0.2923E+07   -308.93   -312.44      3.51      3.47
0.1611E+05  0.5590E+05  0.1153E+07
  0.2683E+07  0.2985E+07   -307.62   -311.21      3.59      3.55
0.1574E+05  0.5590E+05  0.1169E+07
  0.2739E+07  0.3047E+07   -306.31   -309.98      3.67      3.63
0.1540E+05  0.5590E+05  0.1184E+07
  0.2795E+07  0.3109E+07   -305.00   -308.75      3.75      3.71
0.1506E+05  0.5590E+05  0.1199E+07


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

BLOCK: 26   MODEL: MHEATX (CONTINUED)

*** TQ-TABLES FOR HOT SIDE STREAMS ***

STREAM: 72

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.1726E+07	-305.0	135.5	1.000
0.1534E+07	-312.2	135.2	1.000
0.1342E+07	-319.3	134.9	1.000
0.1150E+07	-326.3	134.5	1.000
0.9586E+06	-333.4	134.2	1.000
0.7669E+06	-340.3	133.9	1.000
0.5752E+06	-347.2	133.6	1.000
0.3835E+06	-353.9	133.2	1.000
0.1917E+06	-360.5	132.9	1.000
0.000	-367.0	132.6	1.000

STREAM: 24,1

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.1069E+07	-305.0	640.7	1.000
0.9506E+06	-312.5	640.6	1.000
0.8318E+06	-319.9	640.6	1.000
0.7129E+06	-327.4	640.5	1.000
0.5941E+06	-334.8	640.4	1.000
0.4753E+06	-341.9	640.4	1.000
0.3565E+06	-348.8	640.3	1.000
0.2376E+06	-355.4	640.2	1.000
0.1188E+06	-361.4	640.2	1.000
0.000	-367.0	640.1	1.000

BLOCK: 26 MODEL: MHEATX (CONTINUED)

*** TQ-TABLES FOR COLD SIDE STREAMS ***

STREAM: 46,-2

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.000	-370.9	51.64	1.000
0.1142E+05	-363.7	51.57	1.000
0.2284E+05	-356.6	51.51	1.000
0.3426E+05	-349.5	51.44	1.000
0.4568E+05	-342.4	51.37	1.000
0.5710E+05	-335.5	51.31	1.000
0.6852E+05	-328.6	51.24	1.000
0.7994E+05	-321.8	51.17	1.000
0.9136E+05	-315.2	51.11	1.000
0.1028E+06	-308.8	51.04	1.000

STREAM: 88

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.000	-369.8	25.87	1.000
0.3341E+06	-363.1	25.65	1.000
0.6682E+06	-356.4	25.44	1.000
0.1002E+07	-349.6	25.22	1.000
0.1336E+07	-342.8	25.01	1.000
0.1670E+07	-336.0	24.79	1.000
0.2004E+07	-329.2	24.58	1.000
0.2339E+07	-322.4	24.36	1.000
0.2673E+07	-315.6	24.15	1.000
0.3007E+07	-308.8	23.93	1.000

BLOCK: 30 MODEL: MHEATX

	INLET STREAM	OUTLET STREAM
HOT SIDE:	28,1	32
COLD SIDE:	78	80

```
            42              46,-2

 PROPERTIES FOR STREAM 28,1
 PROPERTY OPTION SET:   REFPROP
```

```
 BLOCK:  30       MODEL: MHEATX (CONTINUED)

   PROPERTIES FOR STREAM 78
   PROPERTY OPTION SET:   REFPROP

   PROPERTIES FOR STREAM 42
   PROPERTY OPTION SET:   REFPROP

                      ***  MASS AND ENERGY BALANCE  ***
                               IN               OUT        RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           7827.63         7827.63      -0.116190E-15
       MASS(LB/HR   )           109641.         109641.        0.00000
       ENTHALPY(BTU/HR  )      -0.213513E+08   -0.211589E+08   -0.901227E-02

                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E             0.00000      LB/HR
    PRODUCT STREAMS CO2E          0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR

                          ***  INPUT DATA  ***

   MAXIMUM NO. ITERATIONS                                   30
   CONVERGENCE TOLERANCE                                     0.000100000
   LOWER LIMIT ON TEMPERATURE      F                     -456.000
   HEAT LEAK                       BTU/HR            192,424.

   SPECIFICATIONS FOR STREAM 28,1     :
   TWO    PHASE   TP  FLASH
   SPECIFIED TEMPERATURE                F                      -404.000
   PRESSURE DROP                        PSI                       0.11000
   MAXIMUM NO. ITERATIONS                                        30
   CONVERGENCE TOLERANCE                                  0.100000-06

   SPECIFICATIONS FOR STREAM 78       :
   TWO    PHASE      FLASH
   PRESSURE DROP                        PSI                       1.53000
   MAXIMUM NO. ITERATIONS                                        50
   CONVERGENCE TOLERANCE                                  0.100000-06
```

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
U-O-S BLOCK SECTION

BLOCK: 30 MODEL: MHEATX (CONTINUED)

SPECIFICATIONS FOR STREAM 42 :
TWO PHASE FLASH
PRESSURE DROP PSI 0.36000
MAXIMUM NO. ITERATIONS 30
CONVERGENCE TOLERANCE 0.100000-06

*** RESULTS ***

INLET STREAM	DUTY BTU/HR	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	OUTLET VAPOR FRAC
28,1	-0.83070E+06	-404.00	640.01	0.0000
78	0.94780E+06	-370.86	25.870	1.0000
42	75324.	-370.86	51.640	1.0000

```
          -------------------------------------------
          |                                         |
 28,1     |                                         | 32
--------->|      2356.2     LBMOL/HR                |--------->
  -367.00 |                                         |  -404.00
          |                                         |
 80       |                                         | 78
<---------|      5167.5     LBMOL/HR                |<---------
  -370.86 |                                         |  -405.06
          |                                         |
 46,-2    |                                         | 42
<---------|      303.90     LBMOL/HR                |<---------
  -370.86 |                                         |  -413.93
          |                                         |
          -------------------------------------------
```

BLOCK: 30 MODEL: MHEATX (CONTINUED)

*** INTERNAL ANALYSIS ***

FLOW IS COUNTERCURRENT.

DUTY	0.83070E+06	BTU/HR
HEAT LEAK	0.19242E+06	BTU/HR
UA	0.19321E+06	BTU/HR-R
AVERAGE LMTD (DUTY/UA)	4.2996	F
MIN TEMP APPROACH	1.9518	F
HOT-SIDE TEMP APPROACH	3.8619	F
COLD-SIDE TEMP APPROACH	9.9311	F
HOT-SIDE NTU	8.6055	
COLD-SIDE NTU	8.1331	

TQ-TABLE(S) INTERPOLATED FOR AT LEAST ONE STREAM DURING INTERNAL ZONE ANALYSIS.

Q HOT	Q COLD	T HOT	T COLD	DELTA T	LMTD	UA ZONE	Q ZONE	UA	PINCH POINT	STREAM IN/OUT/DEW/ BUBBLE POINT
BTU/HR	BTU/HR	F	F	F	F	BTU/HR-R	BTU/HR	BTU/HR-R		
0.000	0.000	-404.00	-413.93	9.93						
3641.	4484.	-403.78	-411.85	8.08	8.97	405.8	3641.	405.8		
7281.	8968.	-403.56	-409.75	6.19	7.09	513.4	3641.	919.1		
0.1092E+05	0.1345E+05	-403.33	-407.43	4.09	5.07	718.3	3641.	1637.		
0.1274E+05	0.1569E+05	-403.22	-406.27	3.04	3.54	514.1	1820.	2152.		
0.1365E+05	0.1681E+05	-403.17	-405.68	2.51	2.77	328.6	910.1	2480.		
0.1456E+05	0.1794E+05	-403.11	-405.06	1.95	2.22	409.7	910.1	2890.	GBL	IN 78
0.1482E+05	0.1825E+05	-403.10	-405.05	1.96	1.95	131.2	256.5	3021.		
0.1508E+05	0.1857E+05	-403.08	-405.04	1.96	1.96	130.9	256.5	3152.		
0.1559E+05	0.1920E+05	-403.05	-405.02	1.97	1.97	260.6	513.0	3413.		
0.1661E+05	0.2046E+05	-402.99	-404.98	2.00	1.99	516.7	1026.	3929.		
0.3323E+05	0.4093E+05	-401.97	-404.34	2.36	2.17	7642.	0.1661E+05	0.1157E+05		

0.4984E+05	0.6139E+05	-400.96	-403.69	2.73	2.54	6543.	0.1661E+05	0.1812E+05
0.6646E+05	0.8185E+05	-399.95	-403.04	3.09	2.90	5721.	0.1661E+05	0.2384E+05
0.8307E+05	0.1023E+06	-398.94	-402.39	3.45	3.27	5083.	0.1661E+05	0.2892E+05
0.9968E+05	0.1228E+06	-397.98	-401.74	3.76	3.60	4609.	0.1661E+05	0.3353E+05
0.1163E+06	0.1432E+06	-397.11	-401.08	3.98	3.87	4297.	0.1661E+05	0.3783E+05
0.1329E+06	0.1637E+06	-396.23	-400.41	4.19	4.08	4072.	0.1661E+05	0.4190E+05
0.1495E+06	0.1842E+06	-395.35	-399.75	4.40	4.29	3873.	0.1661E+05	0.4577E+05
0.1661E+06	0.2046E+06	-394.47	-399.08	4.60	4.50	3693.	0.1661E+05	0.4946E+05
0.1828E+06	0.2251E+06	-393.60	-398.41	4.81	4.71	3528.	0.1661E+05	0.5299E+05
0.1994E+06	0.2456E+06	-392.81	-397.74	4.93	4.87	3411.	0.1661E+05	0.5640E+05
0.2160E+06	0.2660E+06	-392.04	-397.06	5.02	4.97	3342.	0.1661E+05	0.5975E+05
0.2326E+06	0.2865E+06	-391.27	-396.38	5.11	5.06	3283.	0.1661E+05	0.6303E+05
0.2492E+06	0.3069E+06	-390.50	-395.69	5.19	5.15	3227.	0.1661E+05	0.6625E+05
0.2658E+06	0.3274E+06	-389.73	-395.01	5.28	5.24	3172.	0.1661E+05	0.6943E+05
0.2824E+06	0.3479E+06	-388.98	-394.33	5.34	5.31	3127.	0.1661E+05	0.7255E+05
0.2991E+06	0.3683E+06	-388.29	-393.64	5.35	5.35	3108.	0.1661E+05	0.7566E+05
0.3157E+06	0.3888E+06	-387.59	-392.94	5.35	5.35	3107.	0.1661E+05	0.7877E+05
0.3323E+06	0.4093E+06	-386.90	-392.25	5.35	5.35	3106.	0.1661E+05	0.8187E+05
0.3489E+06	0.4297E+06	-386.21	-391.56	5.35	5.35	3106.	0.1661E+05	0.8498E+05
0.3655E+06	0.4502E+06	-385.51	-390.86	5.35	5.35	3106.	0.1661E+05	0.8809E+05
0.3821E+06	0.4706E+06	-384.85	-390.16	5.31	5.33	3117.	0.1661E+05	0.9120E+05
0.3987E+06	0.4911E+06	-384.21	-389.46	5.26	5.28	3144.	0.1661E+05	0.9435E+05
0.4154E+06	0.5116E+06	-383.56	-388.76	5.20	5.23	3177.	0.1661E+05	0.9752E+05
0.4320E+06	0.5320E+06	-382.91	-388.06	5.15	5.17	3211.	0.1661E+05	0.1007E+06
0.4486E+06	0.5525E+06	-382.26	-387.35	5.09	5.12	3246.	0.1661E+05	0.1040E+06
0.4652E+06	0.5730E+06	-381.62	-386.65	5.03	5.06	3282.	0.1661E+05	0.1073E+06
0.4818E+06	0.5934E+06	-380.99	-385.94	4.95	4.99	3328.	0.1661E+05	0.1106E+06

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
U-O-S BLOCK SECTION

BLOCK: 30 MODEL: MHEATX (CONTINUED)

Q HOT BTU/HR	Q COLD BTU/HR	T HOT F	T COLD F	DELTA T F	LMTD F	UA ZONE BTU/HR-R	Q ZONE BTU/HR	UA BTU/HR-R	PINCH POINT	STREAM IN/OUT/DEW/ BUBBLE POINT
0.4984E+06	0.6139E+06	-380.36	-385.23	4.88	4.91	3381.	0.1661E+05	0.1140E+06		
0.5150E+06	0.6343E+06	-379.73	-384.52	4.80	4.84	3436.	0.1661E+05	0.1174E+06		
0.5317E+06	0.6548E+06	-379.09	-383.81	4.72	4.76	3494.	0.1661E+05	0.1209E+06		
0.5483E+06	0.6753E+06	-378.46	-383.10	4.64	4.68	3553.	0.1661E+05	0.1245E+06		
0.5649E+06	0.6957E+06	-377.83	-382.39	4.56	4.60	3613.	0.1661E+05	0.1281E+06		
0.5815E+06	0.7162E+06	-377.19	-381.67	4.48	4.52	3674.	0.1661E+05	0.1317E+06		
0.5981E+06	0.7367E+06	-376.55	-380.96	4.41	4.45	3736.	0.1661E+05	0.1355E+06		
0.6147E+06	0.7571E+06	-375.91	-380.24	4.33	4.37	3802.	0.1661E+05	0.1393E+06		
0.6313E+06	0.7776E+06	-375.27	-379.52	4.26	4.29	3869.	0.1661E+05	0.1431E+06		
0.6479E+06	0.7980E+06	-374.63	-378.81	4.18	4.22	3938.	0.1661E+05	0.1471E+06		
0.6646E+06	0.8185E+06	-373.96	-378.09	4.13	4.16	3996.	0.1661E+05	0.1511E+06		
0.6812E+06	0.8390E+06	-373.29	-377.37	4.08	4.11	4045.	0.1661E+05	0.1551E+06		
0.6978E+06	0.8594E+06	-372.62	-376.65	4.03	4.06	4095.	0.1661E+05	0.1592E+06		
0.7144E+06	0.8799E+06	-371.95	-375.93	3.98	4.01	4147.	0.1661E+05	0.1634E+06		
0.7310E+06	0.9004E+06	-371.28	-375.21	3.93	3.96	4200.	0.1661E+05	0.1676E+06		
0.7476E+06	0.9208E+06	-370.58	-374.48	3.90	3.92	4241.	0.1661E+05	0.1718E+06		
0.7642E+06	0.9413E+06	-369.86	-373.76	3.90	3.90	4261.	0.1661E+05	0.1761E+06		
0.7809E+06	0.9617E+06	-369.15	-373.04	3.89	3.89	4269.	0.1661E+05	0.1803E+06		
0.7975E+06	0.9822E+06	-368.43	-372.31	3.88	3.88	4279.	0.1661E+05	0.1846E+06		
0.8141E+06	0.1003E+07	-367.72	-371.59	3.87	3.87	4288.	0.1661E+05	0.1889E+06		

0.8307E+06 0.1023E+07 -367.00 -370.86 3.86 3.87 4297.
0.1661E+05 0.1932E+06

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

*** TQ-TABLES FOR HOT SIDE STREAMS ***

STREAM: 28,1

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.8307E+06	-367.0	640.1	1.000
0.7384E+06	-371.0	640.1	1.000
0.6461E+06	-374.7	640.1	1.000
0.5538E+06	-378.3	640.1	1.000
0.4615E+06	-381.8	640.1	1.000
0.3692E+06	-385.4	640.1	1.000
0.2769E+06	-389.2	640.0	1.000
0.1846E+06	-393.5	640.0	1.000
0.9230E+05	-398.4	640.0	1.000
0.000	-404.0	640.0	0.000

BLOCK: 30 MODEL: MHEATX (CONTINUED)

*** TQ-TABLES FOR COLD SIDE STREAMS ***

STREAM: 78

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.000	-405.1	27.40	1.000
0.1053E+06	-401.5	27.23	1.000
0.2106E+06	-397.9	27.06	1.000
0.3159E+06	-394.2	26.89	1.000
0.4212E+06	-390.4	26.72	1.000
0.5266E+06	-386.5	26.55	1.000
0.6319E+06	-382.7	26.38	1.000
0.7372E+06	-378.7	26.21	1.000
0.8425E+06	-374.8	26.04	1.000
0.9478E+06	-370.9	25.87	1.000

STREAM: 42

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.000	-413.9	52.00	1.000
8369.	-410.1	51.96	1.000
0.1674E+05	-405.7	51.92	1.000
0.2511E+05	-401.1	51.88	1.000
0.3348E+05	-396.3	51.84	1.000
0.4185E+05	-391.4	51.80	1.000
0.5022E+05	-386.3	51.76	1.000
0.5859E+05	-381.2	51.72	1.000
0.6695E+05	-376.1	51.68	1.000
0.7532E+05	-370.9	51.64	1.000

BLOCK: 12 MODEL: MCOMPR

INLET STREAMS: 10 TO STAGE 1
OUTLET STREAMS: 14 FROM STAGE 3
PROPERTY OPTION SET: REFPROP

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			

MOLE(LBMOL/HR)	2052.00	2052.00	0.00000
MASS(LB/HR)	4136.59	4136.59	0.00000
ENTHALPY(BTU/HR)	198616.	506416.	-0.607802

BLOCK: 12 MODEL: MCOMPR (CONTINUED)

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	0.00000	LB/HR
PRODUCT STREAMS CO2E	0.00000	LB/HR
NET STREAMS CO2E PRODUCTION	0.00000	LB/HR
UTILITIES CO2E PRODUCTION	0.00000	LB/HR
TOTAL CO2E PRODUCTION	0.00000	LB/HR

*** INPUT DATA ***

ISENTROPIC CENTRIFUGAL COMPRESSOR

NUMBER OF STAGES	3
FINAL PRESSURE, PSIA	650.000
DISTRIBUTION AMONG STAGES	EQUAL P-RATIO

COMPRESSOR SPECIFICATIONS PER STAGE

STAGE NUMBER	MECHANICAL EFFICIENCY	ISENTROPIC EFFICIENCY
1	1.000	0.7200
2	1.000	0.7200
3	1.000	0.7200

COOLER SPECIFICATIONS PER STAGE

STAGE NUMBER	PRESSURE DROP PSI	TEMPERATURE F
1	0.000	110.0
2	0.000	110.0
3	0.000	110.0

*** RESULTS ***

FINAL PRESSURE, PSIA	650.000
TOTAL WORK REQUIRED, HP	1,192.62
TOTAL COOLING DUTY , BTU/HR	-2,726,750.

*** PROFILE ***

COMPRESSOR PROFILE

STAGE NUMBER	OUTLET PRESSURE PSIA	PRESSURE RATIO	OUTLET TEMPERATURE F

1	357.4	1.349	158.6
2	482.0	1.349	181.0
3	650.0	1.349	181.1

BLOCK: 12 MODEL: MCOMPR (CONTINUED)

STAGE NUMBER	INDICATED HORSEPOWER HP	BRAKE HORSEPOWER HP
1	386.4	386.4
2	402.0	402.0
3	404.2	404.2

STAGE NUMBER	HEAD DEVELOPED FT-LBF/LB	VOLUMETRIC FLOW CUFT/HR	ISENTROPIC EFFICIENCY
1	0.1332E+06	0.4616E+05	0.7200
2	0.1385E+06	0.3559E+05	0.7200
3	0.1393E+06	0.2652E+05	0.7200

COOLER PROFILE

STAGE NUMBER	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	COOLING LOAD BTU/HR	VAPOR FRACTION
1	110.0	357.4	-.6934E+06	1.000
2	110.0	482.0	-.1015E+07	1.000
3	110.0	650.0	-.1018E+07	1.000

BLOCK: 12   MODEL: MCOMPR (CONTINUED)

BLOCK: 34,A   MODEL: VALVE

INLET STREAM: 36,A1
OUTLET STREAM: 36,A2
PROPERTY OPTION SET: REFPROP

****
*
*
*   ZERO FEED TO BLOCK
*
*
*

****

*** INPUT DATA ***

VALVE OUTLET PRESSURE	PSIA	52.0000
VALVE FLOW COEF CALC.		NO

```
BLOCK:  34,A     MODEL: VALVE (CONTINUED)

                          FLASH SPECIFICATIONS:
   NPHASE                                                    2
   MAX NUMBER OF ITERATIONS                                 30
   CONVERGENCE TOLERANCE                                     0.100000-06

BLOCK:  34,B     MODEL: PUMP
----------------------------
  INLET STREAM:          36,B1
  OUTLET STREAM:         36,B2
  PROPERTY OPTION SET:   REFPROP

                      ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT         RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)          2356.23         2356.23         0.00000
      MASS(LB/HR   )          4749.88         4749.88         0.00000
      ENTHALPY(BTU/HR  )   -0.866231E+07   -0.875250E+07    0.103048E-01

                      ***  CO2 EQUIVALENT SUMMARY ***
   FEED STREAMS CO2E            0.00000      LB/HR
   PRODUCT STREAMS CO2E         0.00000      LB/HR
   NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
   UTILITIES CO2E PRODUCTION    0.00000      LB/HR
   TOTAL CO2E PRODUCTION        0.00000      LB/HR

                          ***  INPUT DATA  ***
   EQUIPMENT TYPE: TURBINE
   OUTLET PRESSURE  PSIA                                    52.0000
   PUMP EFFICIENCY                                           0.70000
   DRIVER EFFICIENCY                                         1.00000

   FLASH SPECIFICATIONS:
   LIQUID PHASE CALCULATION
   NO FLASH PERFORMED
   MAXIMUM NUMBER OF ITERATIONS                              30
   TOLERANCE                                                 0.100000-06

                           ***  RESULTS  ***
   VOLUMETRIC FLOW RATE  CUFT/HR                         1,184.14
   PRESSURE CHANGE  PSI                                   -588.010
   NPSH AVAILABLE   FT-LBF/LB                          22,742.5
   FLUID POWER  HP                                         -50.6389
   BRAKE POWER  HP                                         -35.4472
   ELECTRICITY  KW                                         -26.4330
   PUMP EFFICIENCY USED                                      0.70000
```

NET WORK REQUIRED HP	-35.4472
HEAD DEVELOPED FT-LBF/LB	-21,109.0

```
 BLOCK:  34SPLIT  MODEL: FSPLIT
 ------------------------------
   INLET STREAM:          32
   OUTLET STREAMS:        36,A1       36,B1
   PROPERTY OPTION SET:   REFPROP

                      ***  MASS AND ENERGY BALANCE  ***
                                    IN                OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           2356.23         2356.23         0.00000
       MASS(LB/HR   )           4749.88         4749.88         0.00000
       ENTHALPY(BTU/HR  )      -0.866231E+07   -0.866231E+07     0.00000

                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E            0.00000      LB/HR
    PRODUCT STREAMS CO2E         0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
    UTILITIES CO2E PRODUCTION    0.00000      LB/HR
    TOTAL CO2E PRODUCTION        0.00000      LB/HR

                          ***  INPUT DATA  ***

  FRACTION OF FLOW                STRM=36,B1    FRAC=         1.00000

                           ***  RESULTS  ***

  STREAM= 36,A1          SPLIT=          0.0         KEY=  0    STREAM-
ORDER=   2
          36,B1                          1.00000           0
1

 BLOCK:  38       MODEL: FLASH2
 ------------------------------
   INLET STREAMS:         36,A2       36,B2
   OUTLET VAPOR STREAM:   42
   OUTLET LIQUID STREAM:  40
   PROPERTY OPTION SET:   REFPROP

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

```
 BLOCK:  38       MODEL: FLASH2 (CONTINUED)
    TOTAL BALANCE
       MOLE(LBMOL/HR)            2356.23         2356.23         0.00000
       MASS(LB/HR   )            4749.88         4749.88         0.00000
       ENTHALPY(BTU/HR  )      -0.875250E+07   -0.875250E+07   -0.276657E-
14

                   ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E             0.00000      LB/HR
    PRODUCT STREAMS CO2E          0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR

                       ***  INPUT DATA  ***
   TWO    PHASE  PQ  FLASH
   PRESSURE DROP         PSI                                   0.0
   SPECIFIED HEAT DUTY   BTU/HR                                0.0
   MAXIMUM NO. ITERATIONS                                     30
   CONVERGENCE TOLERANCE                                       0.100000-06

                        ***  RESULTS  ***
   OUTLET TEMPERATURE    F                                     -413.93
   OUTLET PRESSURE       PSIA                                   52.000
   VAPOR FRACTION                                              0.12898


   V-L PHASE EQUILIBRIUM :

      COMP              F(I)          X(I)          Y(I)          K(I)
      PH2               1.0000        1.0000        1.0000
1.0000

 BLOCK:  48       MODEL: MCOMPR
 ------------------------------
   INLET STREAMS:    46          TO STAGE   1
   OUTLET STREAMS:   50        FROM STAGE   3
   PROPERTY OPTION SET:   REFPROP

                   ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT         RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)            303.897         303.897         0.00000
       MASS(LB/HR   )            612.620         612.620         0.00000
       ENTHALPY(BTU/HR  )        55102.3         84741.2        -0.349758
```

BLOCK: 48 MODEL: MCOMPR (CONTINUED)

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	0.00000	LB/HR
PRODUCT STREAMS CO2E	0.00000	LB/HR
NET STREAMS CO2E PRODUCTION	0.00000	LB/HR
UTILITIES CO2E PRODUCTION	0.00000	LB/HR
TOTAL CO2E PRODUCTION	0.00000	LB/HR

*** INPUT DATA ***

ISENTROPIC CENTRIFUGAL COMPRESSOR	
NUMBER OF STAGES	3
FINAL PRESSURE, PSIA	650.000
DISTRIBUTION AMONG STAGES	EQUAL P-RATIO

COMPRESSOR SPECIFICATIONS PER STAGE

STAGE NUMBER	MECHANICAL EFFICIENCY	ISENTROPIC EFFICIENCY
1	1.000	0.7200
2	1.000	0.7200
3	1.000	0.7200

COOLER SPECIFICATIONS PER STAGE

STAGE NUMBER	PRESSURE DROP PSI	TEMPERATURE F
1	0.000	110.0
2	0.000	110.0
3	0.000	110.0

*** RESULTS ***

FINAL PRESSURE, PSIA	650.000
TOTAL WORK REQUIRED, HP	542.472
TOTAL COOLING DUTY , BTU/HR	-1,350,650.

*** PROFILE ***

COMPRESSOR PROFILE

STAGE NUMBER	OUTLET PRESSURE PSIA	PRESSURE RATIO	OUTLET TEMPERATURE F

1	118.5	2.342	309.4
2	277.5	2.342	325.4
3	650.0	2.342	325.8

BLOCK: 48 MODEL: MCOMPR (CONTINUED)

STAGE NUMBER	INDICATED HORSEPOWER HP	BRAKE HORSEPOWER HP
1	177.4	177.4
2	181.7	181.7
3	183.4	183.4

STAGE NUMBER	HEAD DEVELOPED FT-LBF/LB	VOLUMETRIC FLOW CUFT/HR	ISENTROPIC EFFICIENCY
1	0.4128E+06	0.3608E+05	0.7200
2	0.4229E+06	0.1576E+05	0.7200
3	0.4267E+06	6768.	0.7200

COOLER PROFILE

STAGE NUMBER	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	COOLING LOAD BTU/HR	VAPOR FRACTION
1	110.0	118.5	-.4265E+06	1.000
2	110.0	277.5	-.4610E+06	1.000
3	110.0	650.0	-.4632E+06	1.000

```
 BLOCK:  48       MODEL: MCOMPR (CONTINUED)

 BLOCK:  76,1     MODEL: COMPR
 -----------------------------
   INLET STREAM:          74
   OUTLET STREAM:         78
   PROPERTY OPTION SET:   REFPROP

                      ***  MASS AND ENERGY BALANCE  ***
                               IN               OUT         RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)            5167.50         5167.50         0.00000
       MASS(LB/HR   )            104279.         104279.         0.00000
       ENTHALPY(BTU/HR  )      -0.116154E+08   -0.124824E+08    0.694602E-01
```

BLOCK: 76,1   MODEL: COMPR (CONTINUED)

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	0.00000	LB/HR
PRODUCT STREAMS CO2E	0.00000	LB/HR
NET STREAMS CO2E PRODUCTION	0.00000	LB/HR
UTILITIES CO2E PRODUCTION	0.00000	LB/HR
TOTAL CO2E PRODUCTION	0.00000	LB/HR

*** INPUT DATA ***

ISENTROPIC TURBINE

OUTLET PRESSURE PSIA	27.4000
ISENTROPIC EFFICIENCY	0.85000
MECHANICAL EFFICIENCY	1.00000

*** RESULTS ***

INDICATED HORSEPOWER REQUIREMENT	HP	-340.756
BRAKE HORSEPOWER REQUIREMENT	HP	-340.756
NET WORK REQUIRED	HP	-340.756
POWER LOSSES	HP	0.0
ISENTROPIC HORSEPOWER REQUIREMENT	HP	-400.890
CALCULATED OUTLET TEMP F		-405.064
ISENTROPIC TEMPERATURE F		-406.866
EFFICIENCY (POLYTR/ISENTR) USED		0.85000
OUTLET VAPOR FRACTION		1.00000
HEAD DEVELOPED, FT-LBF/LB		-7,611.93
MECHANICAL EFFICIENCY USED		1.00000
INLET HEAT CAPACITY RATIO		1.87591
INLET VOLUMETRIC FLOW RATE , CUFT/HR		35,777.6
OUTLET VOLUMETRIC FLOW RATE, CUFT/HR		102,384.
INLET COMPRESSIBILITY FACTOR		0.92295
OUTLET COMPRESSIBILITY FACTOR		0.92642
AV. ISENT. VOL. EXPONENT		1.60435
AV. ISENT. TEMP EXPONENT		1.55464
AV. ACTUAL VOL. EXPONENT		1.49948
AV. ACTUAL TEMP EXPONENT		1.50484

```
 BLOCK:  84,1     MODEL: COMPR
 -----------------------------
   INLET STREAM:          82
   OUTLET STREAM:         86
   PROPERTY OPTION SET:   REFPROP

                      ***  MASS AND ENERGY BALANCE  ***
                               IN              OUT        RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          4582.50         4582.50         0.00000
       MASS(LB/HR   )          92473.5         92473.5         0.00000
       ENTHALPY(BTU/HR  )     -0.877025E+07   -0.101759E+08    0.138135

                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E             0.00000      LB/HR
    PRODUCT STREAMS CO2E          0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR

                           ***  INPUT DATA  ***

   ISENTROPIC TURBINE
    OUTLET PRESSURE  PSIA                                   26.6100
    ISENTROPIC EFFICIENCY                                    0.85000
    MECHANICAL EFFICIENCY                                    1.00000
```

BLOCK: 84,1 MODEL: COMPR (CONTINUED)

*** RESULTS ***

INDICATED HORSEPOWER REQUIREMENT	HP	-552.440
BRAKE HORSEPOWER REQUIREMENT	HP	-552.440
NET WORK REQUIRED	HP	-552.440
POWER LOSSES	HP	0.0
ISENTROPIC HORSEPOWER REQUIREMENT	HP	-649.929
CALCULATED OUTLET TEMP	F	-368.561
ISENTROPIC TEMPERATURE	F	-379.069
EFFICIENCY (POLYTR/ISENTR) USED		0.85000
OUTLET VAPOR FRACTION		1.00000
HEAD DEVELOPED,	FT-LBF/LB	-13,916.0
MECHANICAL EFFICIENCY USED		1.00000
INLET HEAT CAPACITY RATIO		1.72199
INLET VOLUMETRIC FLOW RATE ,	CUFT/HR	55,485.5
OUTLET VOLUMETRIC FLOW RATE,	CUFT/HR	165,737.
INLET COMPRESSIBILITY FACTOR		0.98852
OUTLET COMPRESSIBILITY FACTOR		0.98434
AV. ISENT. VOL. EXPONENT		1.68741
AV. ISENT. TEMP EXPONENT		1.66782
AV. ACTUAL VOL. EXPONENT		1.48752
AV. ACTUAL TEMP EXPONENT		1.48178

BLOCK: 94 MODEL: MCOMPR

INLET STREAMS: 92 TO STAGE 1
OUTLET STREAMS: T68 FROM STAGE 3
PROPERTY OPTION SET: REFPROP

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(LBMOL/HR)	9750.00	9750.00	0.00000
MASS(LB/HR)	196752.	196752.	0.00000
ENTHALPY(BTU/HR)	0.105554E+07	0.162553E+07	-0.350648

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	0.00000	LB/HR
PRODUCT STREAMS CO2E	0.00000	LB/HR
NET STREAMS CO2E PRODUCTION	0.00000	LB/HR
UTILITIES CO2E PRODUCTION	0.00000	LB/HR
TOTAL CO2E PRODUCTION	0.00000	LB/HR

BLOCK: 94   MODEL: MCOMPR (CONTINUED)

*** INPUT DATA ***

ISENTROPIC CENTRIFUGAL COMPRESSOR	
NUMBER OF STAGES	3
FINAL PRESSURE, PSIA	150.000
DISTRIBUTION AMONG STAGES	EQUAL P-RATIO

COMPRESSOR SPECIFICATIONS PER STAGE

STAGE NUMBER	MECHANICAL EFFICIENCY	ISENTROPIC EFFICIENCY
1	1.000	0.7200
2	1.000	0.7200
3	1.000	0.7200

COOLER SPECIFICATIONS PER STAGE

STAGE NUMBER	PRESSURE DROP PSI	TEMPERATURE F
1	0.000	110.0
2	0.000	110.0
3	0.000	110.0

*** RESULTS ***

FINAL PRESSURE, PSIA	150.000
TOTAL WORK REQUIRED, HP	13,367.4
TOTAL COOLING DUTY , BTU/HR	-0.334424+08

*** PROFILE ***

COMPRESSOR PROFILE

STAGE NUMBER	OUTLET PRESSURE PSIA	PRESSURE RATIO	OUTLET TEMPERATURE F
1	40.80	1.917	329.4
2	78.23	1.917	345.4
3	150.0	1.917	345.4

STAGE NUMBER	INDICATED HORSEPOWER HP	BRAKE HORSEPOWER HP

1	4392.	4392.
2	4484.	4484.
3	4491.	4491.

BLOCK: 94 MODEL: MCOMPR (CONTINUED)

STAGE NUMBER	HEAD DEVELOPED FT-LBF/LB	VOLUMETRIC FLOW CUFT/HR	ISENTROPIC EFFICIENCY
1	0.3182E+05	0.2747E+07	0.7200
2	0.3249E+05	0.1463E+07	0.7200
3	0.3254E+05	0.7638E+06	0.7200

COOLER PROFILE

STAGE NUMBER	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	COOLING LOAD BTU/HR	VAPOR FRACTION
1	110.0	40.80	-.1063E+08	1.000
2	110.0	78.23	-.1140E+08	1.000
3	110.0	150.0	-.1141E+08	1.000

```
 BLOCK:  MIX      MODEL: MIXER
 -----------------------------
   INLET STREAMS:         14          T50
   OUTLET STREAM:         16
   PROPERTY OPTION SET:   REFPROP

                      ***  MASS AND ENERGY BALANCE  ***
                                    IN                OUT        RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)            2356.23          2356.23          0.00000
       MASS(LB/HR   )            4749.88          4749.88         0.191478E-15
       ENTHALPY(BTU/HR  )        581497.          581497.          0.00000

                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E              0.00000      LB/HR
    PRODUCT STREAMS CO2E           0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION    0.00000      LB/HR
    UTILITIES CO2E PRODUCTION      0.00000      LB/HR
    TOTAL CO2E PRODUCTION          0.00000      LB/HR

                          ***  INPUT DATA  ***
   TWO    PHASE     FLASH
   MAXIMUM NO. ITERATIONS                                      30
   CONVERGENCE TOLERANCE                                        0.100000-06
   OUTLET PRESSURE   PSIA                                     650.000

 BLOCK:  MIX2     MODEL: MIXER
 -----------------------------
   INLET STREAMS:         80          86
   OUTLET STREAM:         88
   PROPERTY OPTION SET:   REFPROP

                      ***  MASS AND ENERGY BALANCE  ***
                                    IN                OUT        RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)            9750.00          9750.00          0.00000
       MASS(LB/HR   )            196752.          196752.          0.00000
       ENTHALPY(BTU/HR  )      -0.217105E+08   -0.217105E+08     0.00000
```

```
BLOCK:  MIX2     MODEL: MIXER (CONTINUED)

                     ***  CO2 EQUIVALENT SUMMARY ***
   FEED STREAMS CO2E            0.00000      LB/HR
   PRODUCT STREAMS CO2E         0.00000      LB/HR
   NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
   UTILITIES CO2E PRODUCTION    0.00000      LB/HR
   TOTAL CO2E PRODUCTION        0.00000      LB/HR

                         ***  INPUT DATA  ***
  TWO    PHASE     FLASH
  MAXIMUM NO. ITERATIONS                                    30
  CONVERGENCE TOLERANCE                                      0.100000-06
  OUTLET PRESSURE:  MINIMUM OF INLET STREAM PRESSURES

BLOCK:  MIXN2    MODEL: MIXER
-----------------------------
  INLET STREAMS:         54          T56
  OUTLET STREAM:         58
  PROPERTY OPTION SET:   REFPROP

                     ***  MASS AND ENERGY BALANCE  ***
                                IN               OUT          RELATIVE
DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)          1500.00         1500.00          0.00000
      MASS(LB/HR   )          42020.2         42020.2        -0.173154E-
15
      ENTHALPY(BTU/HR  )    -0.485432E+07   -0.485432E+07    0.383709E-
15

                     ***  CO2 EQUIVALENT SUMMARY ***
   FEED STREAMS CO2E            0.00000      LB/HR
   PRODUCT STREAMS CO2E         0.00000      LB/HR
   NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
   UTILITIES CO2E PRODUCTION    0.00000      LB/HR
   TOTAL CO2E PRODUCTION        0.00000      LB/HR

                         ***  INPUT DATA  ***
  TWO    PHASE     FLASH
  MAXIMUM NO. ITERATIONS                                    30
  CONVERGENCE TOLERANCE                                      0.100000-06
  OUTLET PRESSURE:  MINIMUM OF INLET STREAM PRESSURES
```

```
 BLOCK:  SPLIT    MODEL: FSPLIT
 ------------------------------
   INLET STREAM:          70
   OUTLET STREAMS:        72          82
   PROPERTY OPTION SET:   REFPROP

                      ***  MASS AND ENERGY BALANCE  ***
                                   IN                OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)            9750.00         9750.00          0.00000
       MASS(LB/HR   )            196752.         196752.          0.00000
       ENTHALPY(BTU/HR  )      -0.186601E+08   -0.186601E+08     0.00000

                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E              0.00000      LB/HR
    PRODUCT STREAMS CO2E           0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION    0.00000      LB/HR
    UTILITIES CO2E PRODUCTION      0.00000      LB/HR
    TOTAL CO2E PRODUCTION          0.00000      LB/HR

                          ***  INPUT DATA  ***

  FRACTION OF FLOW                 STRM=82       FRAC=         0.47000

                           ***  RESULTS  ***

  STREAM= 72            SPLIT=          0.53000     KEY=  0    STREAM-
ORDER=   2
         82                             0.47000           0
1
```

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
STREAM SECTION

```
 10 14 16 20L 24
 ---------------

 24,1 28,1 32 36,A1 36,A2
 ------------------------

 STREAM ID                10         14         16         20L        24
24,1       28,1       32         36,A1      36,A2
 FROM :                   ----       12         MIX        18         22
----       26         30         34SPLIT    34,A
 TO   :                   12         MIX        18         22         ----
26         30         34SPLIT    34,A       38

 SUBSTREAM: MIXED
 PHASE:                   VAPOR      VAPOR      VAPOR      VAPOR      VAPOR
VAPOR      VAPOR      LIQUID     MISSING    MISSING
 COMPONENTS: LBMOL/HR
   NH2                  2052.0000  2052.0000  2356.2300  2356.2300
2356.2300     0.0        0.0        0.0        0.0        0.0
   PH2                     0.0        0.0        0.0        0.0        0.0
2356.2300  2356.2300  2356.2300     0.0        0.0
   HE                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   N2                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   NE                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
 COMPONENTS: MOLE FRAC
   NH2                     1.0000     1.0000     1.0000     1.0000
1.0000     0.0        0.0        0.0        0.0        0.0
   PH2                     0.0        0.0        0.0        0.0        0.0
1.0000     1.0000     1.0000     0.0        0.0
   HE                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   N2                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   NE                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
 TOTAL FLOW:
   LBMOL/HR             2052.0000  2052.0000  2356.2300  2356.2300
2356.2300  2356.2300  2356.2300  2356.2300     0.0        0.0
   LB/HR                4136.5858  4136.5858  4749.8769  4749.8769
4749.8769  4749.8769  4749.8769  4749.8769     0.0        0.0
   CUFT/HR              4.6161+04  1.9792+04  2.2727+04  6646.0883
5983.6255  5981.3118  2732.9350  1184.1382     0.0        0.0
 STATE VARIABLES:
   TEMP   F               90.0000   110.0000   110.0000  -290.0000  -
305.0000  -305.0000  -367.0000  -404.0000    MISSING    MISSING
```

```
   PRES   PSIA          265.0000   650.0000   650.0000   643.0000
640.7000   640.7000   640.1200   640.0100    MISSING    MISSING
   VFRAC                  1.0000     1.0000     1.0000     1.0000
1.0000     1.0000     1.0000     0.0        MISSING    MISSING
   LFRAC                  0.0        0.0        0.0        0.0        0.0
0.0        0.0        1.0000    MISSING    MISSING
   SFRAC                  0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0       MISSING    MISSING
 ENTHALPY:
   BTU/LBMOL             96.7912   246.7914   246.7914 -2388.9665 -
2480.0449 -2869.9237 -3323.7867 -3676.3429    MISSING    MISSING
   BTU/LB                48.0144   122.4236   122.4236 -1185.0738 -
1230.2542 -1423.6580 -1648.8019 -1823.6913    MISSING    MISSING
   BTU/HR              1.9862+05  5.0642+05  5.8150+05 -5.6290+06 -
5.8436+06 -6.7622+06 -7.8316+06 -8.6623+06    MISSING    MISSING
 ENTROPY:
   BTU/LBMOL-R           -5.5864    -7.1305    -7.1305   -14.9551   -
15.5102   -19.7620   -23.5627   -28.3217    MISSING    MISSING
   BTU/LB-R              -2.7712    -3.5372    -3.5372    -7.4187    -
7.6940    -9.8032   -11.6885   -14.0493    MISSING    MISSING
 DENSITY:
   LBMOL/CUFT          4.4453-02     0.1037     0.1037     0.3545
0.3938     0.3939     0.8622     1.9898    MISSING    MISSING
   LB/CUFT             8.9612-02     0.2090     0.2090     0.7147
0.7938     0.7941     1.7380     4.0113    MISSING    MISSING
 AVG MW                   2.0159     2.0159     2.0159     2.0159
2.0159     2.0159     2.0159     2.0159    MISSING    MISSING

 MIXED    SUBSTREAM PROPERTIES:

 ***  ALL PHASES   ***
 VMX      CUFT/MIN      769.3513   329.8737   378.7808   110.7681
99.7271    99.6885    45.5489    19.7356    MISSING    MISSING
 CPCVMX                   1.4080     1.4094     1.4094     1.7813
1.8561     1.7110     2.9355     2.1077    MISSING    MISSING
 TEMP     C              32.2222    43.3333    43.3333  -178.8889  -
187.2222  -187.2222  -221.6667  -242.2222    MISSING    MISSING
 PRES     BAR            18.2711    44.8159    44.8159    44.3333
44.1747    44.1747    44.1347    44.1271    MISSING    MISSING
 MASSFLMX KG/HR        1876.3237  1876.3237  2154.5079  2154.5079
2154.5079  2154.5079  2154.5079  2154.5079    MISSING    MISSING

 *** VAPOR PHASE   ***
```

10 14 16 20L 24

24,1 28,1 32 36,A1 36,A2 (CONTINUED)

```
 STREAM ID              10         14         16         20L        24
24,1       28,1       32         36,A1      36,A2

 MUMX     LB/FT-HR     2.1916-02  2.2516-02  2.2516-02  1.0445-02  9.9667-
03  9.9673-03  9.3391-03    MISSING    MISSING    MISSING
 RHOMX    LB/CUFT      8.9612-02     0.2090     0.2090     0.7147
0.7938     0.7941     1.7380    MISSING    MISSING    MISSING

 *** LIQUID PHASE  ***
 MUMX     LB/FT-HR       MISSING    MISSING    MISSING    MISSING
MISSING    MISSING    MISSING  2.2316-02    MISSING    MISSING
 RHOMX    LB/CUFT        MISSING    MISSING    MISSING    MISSING
MISSING    MISSING    MISSING     4.0113    MISSING    MISSING
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              EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
                             STREAM SECTION

 36,B1 36,B2 40 42 46
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 46,+R 46,-2 50 54 58
 --------------------

 STREAM ID               36,B1      36,B2      40         42         46
46,+R      46,-2      50         54         58
 FROM :                  34SPLIT    34,B       38         38         18
26         30         48         22         MIXN2
 TO   :                  34,B       38         ----       30         48
18         26         ----       MIXN2      18

 SUBSTREAM: MIXED
 PHASE:                  LIQUID     LIQUID     LIQUID     VAPOR      VAPOR
VAPOR      VAPOR      VAPOR      MIXED      MIXED
 COMPONENTS: LBMOL/HR
   NH2                    0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   PH2                 2356.2300  2356.2300  2052.3331   303.8969
303.8969   303.8969   303.8969   303.8969    0.0        0.0
   HE                     0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   N2                     0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0      696.0000  1500.0000
   NE                     0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
 COMPONENTS: MOLE FRAC
   NH2                    0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   PH2                    1.0000     1.0000     1.0000     1.0000
1.0000     1.0000     1.0000     1.0000     0.0        0.0
   HE                     0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   N2                     0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        1.0000     1.0000
   NE                     0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
 TOTAL FLOW:
   LBMOL/HR            2356.2300  2356.2300  2052.3331   303.8969
303.8969   303.8969   303.8969   303.8969   696.0000  1500.0000
   LB/HR               4749.8769  4749.8769  4137.2572   612.6198
612.6198   612.6198   612.6198   612.6198  1.9497+04  4.2020+04
   CUFT/HR             1184.1382  1344.8425  1038.4020  2245.1674
3.6083+04  9606.2010  5439.0909  2931.1843  3.1678+04  1.0364+05
 STATE VARIABLES:
   TEMP   F            -404.0000  -408.5663  -413.9311  -413.9311
98.7281  -308.7511  -370.8619   110.0000  -316.4972  -318.1681
```

```
   PRES   PSIA          640.0100    52.0000    52.0000    52.0000
50.5700    51.0400    51.6400   650.0000    18.8700    17.0000
   VFRAC                  0.0        0.0        0.0        1.0000
1.0000     1.0000     1.0000     1.0000     0.5863     0.8111
   LFRAC                  1.0000     1.0000     1.0000     0.0        0.0
0.0        0.0        0.0        0.4137     0.1889
   SFRAC                  0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
 ENTHALPY:
   BTU/LBMOL          -3676.3429 -3714.6215 -3759.2218 -3413.4183
181.3191 -2827.3448 -3165.5588   278.8485 -3756.7813 -3236.2125
   BTU/LB             -1823.6913 -1842.6799 -1864.8044 -1693.2646
89.9454 -1402.5363 -1570.3111   138.3259  -134.1062  -115.5234
   BTU/HR             -8.6623+06 -8.7525+06 -7.7152+06 -1.0373+06
5.5102+04 -8.5922+05 -9.6200+05  8.4741+04 -2.6147+06 -4.8543+06
 ENTROPY:
   BTU/LBMOL-R          -28.3217   -27.9575   -28.8756   -21.3152    -
4.9619   -14.5223   -17.4193    -9.9070   -16.7144   -12.9189
   BTU/LB-R             -14.0493   -13.8686   -14.3241   -10.5736    -
2.4614    -7.2039    -8.6410    -4.9145    -0.5967    -0.4612
 DENSITY:
   LBMOL/CUFT             1.9898     1.7520     1.9764     0.1354  8.4221-
03  3.1635-02  5.5873-02     0.1037  2.1971-02  1.4473-02
   LB/CUFT                4.0113     3.5319     3.9843     0.2729  1.6978-
02  6.3773-02     0.1126     0.2090     0.6155     0.4054
 AVG MW                   2.0159     2.0159     2.0159     2.0159
2.0159     2.0159     2.0159     2.0159    28.0135    28.0135

 MIXED    SUBSTREAM PROPERTIES:

 ***  ALL PHASES   ***
 VMX      CUFT/MIN       19.7356    22.4140    17.3067    37.4195
601.3896   160.1034    90.6515    48.8531   527.9597  1727.3706
 CPCVMX                   2.1077     3.5892     2.3131     2.3497
1.3878     1.5411     1.7380     1.3938     1.6899     1.5713
 TEMP     C            -242.2222  -244.7591  -247.7395  -247.7395
37.0712  -189.3062  -223.8122    43.3333  -193.6096  -194.5378
 PRES     BAR            44.1271     3.5853     3.5853     3.5853
3.4867     3.5191     3.5605    44.8159     1.3010     1.1721
 MASSFLMX KG/HR        2154.5079  2154.5079  1876.6283   277.8797
277.8797   277.8797   277.8797   277.8797  8843.8637  1.9060+04

 *** VAPOR PHASE   ***
```

36,B1 36,B2 40 42 46

46,+R 46,-2 50 54 58 (CONTINUED)

```
 STREAM ID              36,B1      36,B2      40         42         46
46,+R      46,-2      50         54         58

 MUMX     LB/FT-HR       MISSING    MISSING    MISSING  3.1536-03  2.2133-
02  8.7867-03  5.8310-03  2.2516-02  1.3584-02  1.3407-02
 RHOMX    LB/CUFT        MISSING    MISSING    MISSING     0.2729  1.6978-
02  6.3773-02     0.1126     0.2090     0.3627     0.3294

 *** LIQUID PHASE  ***
 MUMX     LB/FT-HR     2.2316-02  1.6434-02  2.1971-02    MISSING
MISSING    MISSING    MISSING    MISSING     0.3572     0.3700
 RHOMX    LB/CUFT         4.0113     3.5319     3.9843    MISSING
MISSING    MISSING    MISSING    MISSING    49.6969    49.9640
```

```
 60 68 68,1 70 72
 ----------------

 74 78 80 82 86
 --------------

 STREAM ID                60         68         68,1       70         72
74         78         80         82         86
 FROM :                   18         ----       18         22         SPLIT
26         76,1       30         SPLIT      84,1
 TO   :                   ----       18         22         SPLIT      26
76,1       30         MIX2       84,1       MIX2

 SUBSTREAM: MIXED
 PHASE:                   VAPOR      VAPOR      VAPOR      VAPOR      VAPOR
VAPOR      VAPOR      VAPOR      VAPOR      VAPOR
 COMPONENTS: LBMOL/HR
   NH2                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   PH2                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   HE                       0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   N2                    1500.0000     0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   NE                       0.0     9750.0000  9750.0000  9750.0000
5167.5000  5167.5000  5167.5000  5167.5000  4582.5000  4582.5000
 COMPONENTS: MOLE FRAC
   NH2                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   PH2                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   HE                       0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   N2                       1.0000     0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   NE                       0.0        1.0000     1.0000     1.0000
1.0000     1.0000     1.0000     1.0000     1.0000     1.0000
 TOTAL FLOW:
   LBMOL/HR              1500.0000  9750.0000  9750.0000  9750.0000
5167.5000  5167.5000  5167.5000  5167.5000  4582.5000  4582.5000
   LB/HR                 4.2020+04  1.9675+05  1.9675+05  1.9675+05
1.0428+05  1.0428+05  1.0428+05  1.0428+05  9.2473+04  9.2473+04
   CUFT/HR               5.7725+05  3.9923+05  1.2254+05  1.1805+05
6.2569+04  3.5778+04  1.0238+05  1.8724+05  5.5486+04  1.6574+05
 STATE VARIABLES:
   TEMP   F                98.7281   110.0000  -290.0000  -305.0000  -
305.0000  -367.0000  -405.0641  -370.8619  -305.0000  -368.5612
```

```
   PRES   PSIA           15.5700   150.0000   143.8100   135.5100
135.5100   132.5700    27.4000    25.8700   135.5100    26.6100
   VFRAC                  1.0000     1.0000     1.0000     1.0000
1.0000     1.0000     1.0000     1.0000     1.0000     1.0000
   LFRAC                  0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   SFRAC                  0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
 ENTHALPY:
   BTU/LBMOL            148.5153   166.7215 -1837.4151 -1913.8569 -
1913.8569 -2247.7764 -2415.5619 -2232.1455 -1913.8569 -2220.5991
   BTU/LB                 5.3016     8.2618   -91.0526   -94.8407   -
94.8407  -111.3880  -119.7026  -110.6134   -94.8407  -110.0412
   BTU/HR              2.2277+05  1.6255+06 -1.7915+07 -1.8660+07 -
9.8899+06 -1.1615+07 -1.2482+07 -1.1535+07 -8.7702+06 -1.0176+07
 ENTROPY:
   BTU/LBMOL-R            0.1569    -4.3213   -10.3223   -10.6770   -
10.6770   -13.4072   -12.8498   -10.1189   -10.6770   -10.0457
   BTU/LB-R            5.5998-03    -0.2141    -0.5115    -0.5291    -
0.5291    -0.6644    -0.6368    -0.5014    -0.5291    -0.4978
 DENSITY:
   LBMOL/CUFT          2.5985-03  2.4422-02  7.9564-02  8.2589-02  8.2589-
02     0.1444  5.0472-02  2.7599-02  8.2589-02  2.7649-02
   LB/CUFT             7.2794-02     0.4928     1.6056     1.6666
1.6666     2.9146     1.0185     0.5569     1.6666     0.5580
 AVG MW                 28.0135    20.1797    20.1797    20.1797
20.1797    20.1797    20.1797    20.1797    20.1797    20.1797

 MIXED    SUBSTREAM PROPERTIES:

 ***  ALL PHASES   ***
 VMX      CUFT/MIN     9620.8635  6653.8733  2042.3845  1967.5726
1042.8135   596.2936  1706.4021  3120.5970   924.7591  2762.2764
 CPCVMX                  1.4011     1.6688     1.7126     1.7209
1.7209     1.8750     1.8172     1.7045     1.7209     1.7033
 TEMP     C             37.0712    43.3333  -178.8889  -187.2222  -
187.2222  -221.6667  -242.8134  -223.8122  -187.2222  -222.5340
 PRES     BAR            1.0735    10.3421     9.9154     9.3431
9.3431     9.1404     1.8892     1.7837     9.3431     1.8347
 MASSFLMX KG/HR        1.9060+04  8.9245+04  8.9245+04  8.9245+04
4.7300+04  4.7300+04  4.7300+04  4.7300+04  4.1945+04  4.1945+04

 *** VAPOR PHASE   ***
```

60 68 68,1 70 72

74 78 80 82 86 (CONTINUED)

STREAM ID		60	68	68,1	70	72	74	78	80	82	86
MUMX	LB/FT-HR	4.4411-02	7.9913-02	3.3199-02	3.0752-02	3.0752-02	1.9827-02	1.1324-02	1.8177-02	3.0752-02	1.8642-02
RHOMX	LB/CUFT	7.2794-02	0.4928	1.6056	1.6666	1.6666	2.9146	1.0185	0.5569	1.6666	0.5580
*** LIQUID PHASE ***											
MUMX	LB/FT-HR	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING
RHOMX	LB/CUFT	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
STREAM SECTION

88 90 92 T50 T52

T56 T68

STREAM ID	88	90	92	T50	T52	T56	T68
FROM :	MIX2	26	18	----	----	----	94
TO :	26	18	94	MIX	22	MIXN2	----
SUBSTREAM: MIXED							
PHASE:	VAPOR	VAPOR	VAPOR	VAPOR	LIQUID	VAPOR	VAPOR
COMPONENTS: LBMOL/HR							
NH2	0.0	0.0	0.0	304.2300	0.0	0.0	0.0
PH2	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HE	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N2	0.0	0.0	0.0	0.0	696.0000	804.0000	0.0
NE	9750.0000	9750.0000	9750.0000	0.0	0.0	0.0	9750.0000
COMPONENTS: MOLE FRAC							
NH2	0.0	0.0	0.0	1.0000	0.0	0.0	0.0
PH2	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HE	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N2	0.0	0.0	0.0	0.0	1.0000	1.0000	0.0
NE	1.0000	1.0000	1.0000	0.0	0.0	0.0	1.0000
TOTAL FLOW:							
LBMOL/HR	9750.0000	9750.0000	9750.0000	304.2300	696.0000	804.0000	9750.0000
LB/HR	1.9675+05	1.9675+05	1.9675+05	613.2912	1.9497+04	2.2523+04	1.9675+05
CUFT/HR	3.5770+05	6.5834+05	2.7475+06	2934.4288	392.9524	6.8478+04	3.9923+05
STATE VARIABLES:							
TEMP F	-369.8032	-308.7511	98.7281	110.0000	-316.0000	-318.0000	110.0000

```
   PRES   PSIA           25.8700    23.9300    21.2800   650.0000
20.0000    17.0000   150.0000
   VFRAC                  1.0000     1.0000     1.0000     1.0000     0.0
1.0000     1.0000
   LFRAC                  0.0        0.0        0.0        0.0
1.0000     0.0        0.0
   SFRAC                  0.0        0.0        0.0        0.0        0.0
0.0        0.0
 ENTHALPY:
   BTU/LBMOL          -2226.7187 -1918.3407   108.2610   246.7914 -
5135.9614 -2785.5709   166.7215
   BTU/LB              -110.3445   -95.0629     5.3648   122.4236  -
183.3389   -99.4368     8.2618
   BTU/HR             -2.1711+07 -1.8704+07  1.0555+06  7.5081+04 -
3.5746+06 -2.2396+06  1.6255+06
 ENTROPY:
   BTU/LBMOL-R          -10.0582    -7.2832    -0.5388    -7.1305   -
26.3483    -9.7342    -4.3213
   BTU/LB-R              -0.4984    -0.3609 -2.6698-02    -3.5372    -
0.9406    -0.3475    -0.2141
 DENSITY:
   LBMOL/CUFT          2.7258-02  1.4810-02  3.5487-03     0.1037
1.7712  1.1741-02  2.4422-02
   LB/CUFT                0.5501     0.2989  7.1612-02     0.2090
49.6177     0.3289     0.4928
 AVG MW                  20.1797    20.1797    20.1797     2.0159
28.0135    28.0135    20.1797

 MIXED    SUBSTREAM PROPERTIES:

 ***  ALL PHASES   ***
 VMX      CUFT/MIN     5961.6228  1.0972+04  4.5791+04    48.9071
6.5492  1141.2973  6653.8729
 CPCVMX                   1.7035     1.6766     1.6670     1.4094
1.9198     1.4640     1.6688
 TEMP     C            -223.2240  -189.3062    37.0712    43.3333  -
193.3333  -194.4444    43.3333
 PRES     BAR             1.7837     1.6499     1.4672    44.8159
1.3790     1.1721    10.3421
 MASSFLMX KG/HR        8.9245+04  8.9245+04  8.9245+04   278.1842
8843.8637  1.0216+04  8.9245+04

 *** VAPOR PHASE   ***
```

88 90 92 T50 T52

T56 T68 (CONTINUED)

STREAM ID		88	90	92	T50	T52	T56	T68
MUMX	LB/FT-HR	1.8389-02	2.9685-02	7.8775-02	2.2516-02	MISSING	1.3422-02	7.9913-02
RHOMX	LB/CUFT	0.5501	0.2989	7.1612-02	0.2090	MISSING	0.3289	0.4928
*** LIQUID PHASE ***								
MUMX	LB/FT-HR	MISSING	MISSING	MISSING	MISSING	0.3535	MISSING	MISSING
RHOMX	LB/CUFT	MISSING	MISSING	MISSING	MISSING	49.6177	MISSING	MISSING

BLOCK STATUS

```
*************************************************************************
**
 *
*
 * Calculations were completed with warnings
*
 *
*
 * The following Unit Operation blocks were
*
 * completed with warnings:
*
 *   34,A
*
 *
*
 * All Transfer blocks were completed normally
*
 *
*
 * All streams were flashed normally
*
 *
*
 * All Convergence blocks were completed normally
*
 *
*
 * All Calculator blocks were completed normally
*
 *
*

*************************************************************************
**
```

TABLE OF CONTENTS

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING

TABLE OF CONTENTS

```
                EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
                              RUN CONTROL SECTION
RUN CONTROL INFORMATION
-----------------------
THIS COPY OF ASPEN PLUS LICENSED TO UNIVERSITY OF PENNSYLVAN
TYPE OF RUN: NEW
INPUT FILE NAME: _3651doa.inm
OUTPUT PROBLEM DATA FILE NAME: _3651doa
LOCATED IN:
PDF SIZE USED FOR INPUT TRANSLATION:
  NUMBER OF FILE RECORDS (PSIZE) =     0
  NUMBER OF IN-CORE RECORDS      =   256
PSIZE NEEDED FOR SIMULATION    =   256
CALLING PROGRAM NAME:          apmain
LOCATED IN: C:\Program Files\AspenTech\Aspen Plus V12.1\Engine\\xeq
SIMULATION REQUESTED FOR ENTIRE FLOWSHEET
```

FLOWSHEET CONNECTIVITY BY STREAMS

STREAM	SOURCE	DEST	STREAM	SOURCE	DEST
T56	----	MIXN2	24,1	----	$26H04
T52	----	$22HTR	68	----	$18H02
T50	----	MIX	10	----	12
T20	----	SPL	10LN2	----	MKP
36,B2	34,B	38	36,A1	34SPLIT	34,A
36,B1	34SPLIT	34,B	42	38	$30HTR
40	38	----	86	84,1	MIX2
78	76,1	$30HTR	58	MIXN2	$18HTR
88	MIX2	$26HTR	72	SPLIT	$26H02
82	SPLIT	84,1	36,A2	34,A	38
32	$30H03	34SPLIT	46,-2	$30HTR	$26HTR
80	$30HTR	MIX2	90	$26HTR	$18HTR
74	$26H02	76,1	46,+R	$26HTR	$18HTR
28,1	$26H04	$30H03	70	$22H02	SPLIT
54	$22HTR	MIXN2	24	$22H01	----
T68	94	----	46	$18HTR	48
92	$18HTR	94	20L	$18H01	$22H01
68,1	$18H02	$22H02	60	$18HTR	----
16	MIX	$18H01	50	48	----
14	12	MIX	FEEDN2	JT2	LINKO
25	SPL	$BAXHH01	55	SPL	$BAXHH02
15	MKP	CMP	45	KO	$BAXHHTR
40LN2	KO	JT2	35	JT	KO
50LN2	$BAXHHTR	MKP	60LN2	$BAXHH01	EXP
30LN2	$BAXHH02	JT	65	EXP	KO
20	CMP	----	56	LINKO	----
52	LINKO	----	$18Q01	$18H01	$18HTR
$18Q02	$18H02	$18HTR	$22Q01	$22H01	$22HTR
$22Q02	$22H02	$22HTR	$26Q02	$26H02	$26HTR
$26Q04	$26H04	$26HTR	$30Q03	$30H03	$30HTR
$BAXHQ01	$BAXHH01	$BAXHHTR	$BAXHQ02	$BAXHH02	$BAXHHTR

FLOWSHEET CONNECTIVITY BY BLOCKS

BLOCK	INLETS	OUTLETS
34,B	36,B1	36,B2
34SPLIT	32	36,A1 36,B1
38	36,A2 36,B2	42 40
84,1	82	86
76,1	74	78
MIXN2	54 T56	58
MIX2	80 86	88
SPLIT	70	72 82
34,A	36,A1	36,A2
94	92	T68
MIX	14 T50	16
48	46	50
12	10	14

JT2	40LN2	FEEDN2
SPL	T20	25 55
MKP	10LN2 50LN2	15

```
 FLOWSHEET CONNECTIVITY BY BLOCKS (CONTINUED)
   KO           35 65                          45 40LN2
   JT           30LN2                          35
   EXP          60LN2                          65
   CMP          15                             20
   LINKO        FEEDN2                         56 52
   $18H01       16                             20L $18Q01
   $18H02       68                             68,1 $18Q02
   $18HTR       90 58 46,+R $18Q01 $18Q02      92 60 46
   $22H01       20L                            24 $22Q01
   $22H02       68,1                           70 $22Q02
   $22HTR       T52 $22Q01 $22Q02              54
   $26H02       72                             74 $26Q02
   $26H04       24,1                           28,1 $26Q04
   $26HTR       46,-2 88 $26Q02 $26Q04         46,+R 90
   $30H03       28,1                           32 $30Q03
   $30HTR       78 42 $30Q03                   80 46,-2
   $BAXHH01     25                             60LN2 $BAXHQ01
   $BAXHH02     55                             30LN2 $BAXHQ02
   $BAXHHTR     45 $BAXHQ01 $BAXHQ02           50LN2
 TRANSFER BLOCK: T-1
 -------------------
     EQUAL-TO : TEMPERATURE IN STREAM 24 SUBSTREAM MIXED
     SET      : TEMPERATURE IN STREAM 24,1 SUBSTREAM MIXED
 TRANSFER BLOCK: T-2
 -------------------
     EQUAL-TO : PRESSURE IN STREAM 24 SUBSTREAM MIXED
     SET      : PRESSURE IN STREAM 24,1 SUBSTREAM MIXED
 TRANSFER BLOCK: T-3
 -------------------
     EQUAL-TO : TOTAL MOLEFLOW IN STREAM 24 SUBSTREAM MIXED
     SET      : TOTAL MOLEFLOW IN STREAM 24,1 SUBSTREAM MIXED
 CALCULATOR BLOCK:   INIT
 ------------------------
   SAMPLED VARIABLES:
     P        : PARAMETER 1
   FORTRAN STATEMENTS:
         P=700
   WRITE VARIABLES: P
   VALUES OF ACCESSED FORTRAN VARIABLES ON MOST RECENT SIMULATION PASS:
     VARIABLE        VALUE READ      VALUE WRITTEN    UNITS
     --------        ----------      -------------    -----
     P                MISSING          700.000
```

```
CALCULATOR BLOCK:   OP1
-----------------------
  SAMPLED VARIABLES:
    Q        : SENTENCE=PARAM VARIABLE=QLEAK IN UOS BLOCK $26HTR
    TOUT     : TEMPERATURE IN STREAM 28,1 SUBSTREAM MIXED
    TIN      : TEMPERATURE IN STREAM 24,1 SUBSTREAM MIXED
    STR      : TOTAL MOLEFLOW IN STREAM 24,1 SUBSTREAM MIXED
  FORTRAN STATEMENTS:
          P0 = 9*10**-6
          P1 = 0.0045
          P2 = -0.3074
          P3 = 28.835

          F0 = (P0*TIN**3)+(P1*TIN**2)+(P2*TIN)+(P3)

          T1 = TIN + 1*(TOUT-TIN)/4
          F1 = (P0*T1**3)+(P1*T1**2)+(P2*T1)+(P3)

          T2 = TIN + 2*(TOUT-TIN)/4
          F2 = (P0*T2**3)+(P1*T2**2)+(P2*T2)+(P3)

          T3 = TIN + 3*(TOUT-TIN)/4
          F3 = (P0*T3**3)+(P1*T3**2)+(P2*T3)+(P3)

          T4 = TIN + 4*(TOUT-TIN)/4
          F4 = (P0*T4**3)+(P1*T4**2)+(P2*T4)+(P3)

          QCURVE = (0.5)*ABS(TOUT-TIN)/4*(F0+2*F1+2*F2+2*F3+F4)
          AVGQ = 1/ABS(TOUT-TIN)*QCURVE

          M0 = -2*10**-13
          M1 = -1*10**-4
          M2 = 9*10**-10
          M3 = 6*10**-7
          M4 = 0.0002
          M5 = 0.7359
          EI = 0.62
          EO = 0.1

          FCONV = 0.2
          DELTOH2 = 2 * FCONV * (EI - EO) * STR
          Q = AVGQ * DELTOH2

  READ VARIABLES:  TOUT      TIN       STR
  WRITE VARIABLES: Q
  VALUES OF ACCESSED FORTRAN VARIABLES ON MOST RECENT SIMULATION PASS:
    VARIABLE        VALUE READ      VALUE WRITTEN    UNITS
    --------        ----------      -------------    -----
    Q                99.0000          314531.        BTU/HR
```

```
CALCULATOR BLOCK:   OP1 (CONTINUED)
    TOUT            -367.000                        F
    TIN             -305.000                        F
    STR              2356.23                        LBMOL/HR
CALCULATOR BLOCK:   OP2
-----------------------
  SAMPLED VARIABLES:
    Q        : SENTENCE=PARAM VARIABLE=QLEAK IN UOS BLOCK $30HTR
    TOUT     : TEMPERATURE IN STREAM 32 SUBSTREAM MIXED
    TIN      : TEMPERATURE IN STREAM 28,1 SUBSTREAM MIXED
    STR      : TOTAL MOLEFLOW IN STREAM 28,1 SUBSTREAM MIXED
  FORTRAN STATEMENTS:
          P0 = 9*10**-6
          P1 = 0.0045
          P2 = -0.3074
          P3 = 28.835

          F0 = (P0*TIN**3)+(P1*TIN**2)+(P2*TIN)+(P3)

          T1 = TIN + 1*(TOUT-TIN)/4
          F1 = (P0*T1**3)+(P1*T1**2)+(P2*T1)+(P3)

          T2 = TIN + 2*(TOUT-TIN)/4
          F2 = (P0*T2**3)+(P1*T2**2)+(P2*T2)+(P3)

          T3 = TIN + 3*(TOUT-TIN)/4
          F3 = (P0*T3**3)+(P1*T3**2)+(P2*T3)+(P3)

          T4 = TIN + 4*(TOUT-TIN)/4
          F4 = (P0*T4**3)+(P1*T4**2)+(P2*T4)+(P3)

          QCURVE = (0.5)*ABS(TOUT-TIN)/4*(F0+2*F1+2*F2+2*F3+F4)
          AVGQ = 1/ABS(TOUT-TIN)*QCURVE

          EI = 0.1
          EO = 0

          FCONV = 0.5
          DELTOH2 = 2 * FCONV *(EI - EO) * STR
          Q = AVGQ * DELTOH2


  READ VARIABLES:  TOUT      TIN       STR
  WRITE VARIABLES: Q
  VALUES OF ACCESSED FORTRAN VARIABLES ON MOST RECENT SIMULATION PASS:
    VARIABLE        VALUE READ      VALUE WRITTEN    UNITS
    --------        ----------      -------------    -----
    Q                130000.          192424.        BTU/HR
```

```
 CALCULATOR BLOCK:   OP2 (CONTINUED)
     TOUT           -404.000                        F
     TIN            -367.000                        F
     STR             2356.23                        LBMOL/HR
 COMPUTATIONAL SEQUENCE
 ----------------------
 SEQUENCE USED WAS:
    INIT $18H02 12 MIX $18H01 $22H01 T-1 T-2 T-3 $26H04 OP1 $22H02 $22HTR
    MIXN2 SPLIT 84,1 $26H02 76,1 $30H03 OP2 34SPLIT 34,B *34,A 38 $30HTR
    MIX2 $26HTR $18HTR 94 48
 OVERALL FLOWSHEET BALANCE
 -------------------------
                      ***  MASS AND ENERGY BALANCE  ***
                                    IN                OUT          RELATIVE
DIFF.
    CONVENTIONAL COMPONENTS (LBMOL/HR)
       NH2                       2356.23          2356.23           0.00000
       PH2                       2356.23          2356.23           0.00000
       HE                        0.00000          0.00000           0.00000
       N2                        2200.00          1500.00          0.318182
       NE                        9750.00          9750.00           0.00000
    TOTAL BALANCE
       MOLE(LBMOL/HR)            16662.5          15962.5          0.420106E-
01
       MASS(LB/HR   )            267881.          248272.          0.732019E-
01
       ENTHALPY(BTU/HR  )      -0.106772E+08   -0.116257E+08    0.815854E-
01
                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E             0.00000      LB/HR
    PRODUCT STREAMS CO2E          0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR
```

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
PHYSICAL PROPERTIES SECTION

COMPONENTS

ID	TYPE	ALIAS	NAME
NH2	C	H2	HYDROGEN
PH2	C	H2-PARA	PARA-HYDROGEN
HE	C	HE-4	HELIUM-4
N2	C	N2	NITROGEN
NE	C	NE	NEON

```
              EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
                            U-O-S BLOCK SECTION
 BLOCK:  18       MODEL: MHEATX
 ------------------------------
   HOT SIDE:    INLET STREAM     OUTLET STREAM
                ------------     -------------
                16               20L
                68               68,1
   COLD SIDE:   INLET STREAM     OUTLET STREAM
                ------------     -------------
                90               92
                58               60
                46,+R            46
   PROPERTIES FOR STREAM 16
   PROPERTY OPTION SET:   REFPROP
   PROPERTIES FOR STREAM 68
   PROPERTY OPTION SET:   REFPROP
   PROPERTIES FOR STREAM 90
   PROPERTY OPTION SET:   REFPROP
   PROPERTIES FOR STREAM 58
   PROPERTY OPTION SET:   REFPROP
   PROPERTIES FOR STREAM 46,+R
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                                   IN               OUT          RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)            23660.1         23660.1        0.153760E-
15
       MASS(LB/HR   )            440887.         440887.        0.132024E-
15
       ENTHALPY(BTU/HR  )      -0.222103E+08   -0.222103E+08    0.849750E-
07
```

```
 BLOCK:  18       MODEL: MHEATX (CONTINUED)
                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E             0.00000      LB/HR
    PRODUCT STREAMS CO2E          0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR
                          ***  INPUT DATA  ***
   MAXIMUM NO. ITERATIONS                                  30
   CONVERGENCE TOLERANCE                                    0.000100000
   SPECIFICATIONS FOR STREAM 16      :
   TWO    PHASE  TP  FLASH
   SPECIFIED TEMPERATURE              F                       -290.000
   PRESSURE DROP                      PSI                        7.00000
   MAXIMUM NO. ITERATIONS                                      30
   CONVERGENCE TOLERANCE
0.100000-06
   SPECIFICATIONS FOR STREAM 68      :
   TWO    PHASE  TP  FLASH
   SPECIFIED TEMPERATURE              F                       -290.000
   PRESSURE DROP                      PSI                        6.19000
   MAXIMUM NO. ITERATIONS                                      30
   CONVERGENCE TOLERANCE
0.100000-06
   SPECIFICATIONS FOR STREAM 90      :
   TWO    PHASE      FLASH
   PRESSURE DROP                      PSI                        2.65000
   MAXIMUM NO. ITERATIONS                                      30
   CONVERGENCE TOLERANCE
0.100000-06
   SPECIFICATIONS FOR STREAM 58      :
   TWO    PHASE      FLASH
   PRESSURE DROP                      PSI                        1.43000
   MAXIMUM NO. ITERATIONS                                      30
   CONVERGENCE TOLERANCE
0.100000-06
   SPECIFICATIONS FOR STREAM 46,+R   :
   TWO    PHASE      FLASH
   PRESSURE DROP                      PSI                        0.47000
   MAXIMUM NO. ITERATIONS                                      30
   CONVERGENCE TOLERANCE
0.100000-06
```

```
 BLOCK:  18       MODEL: MHEATX (CONTINUED)
                        ***  RESULTS  ***
  INLET                  OUTLET       OUTLET       OUTLET
  STREAM    DUTY         TEMPERATURE  PRESSURE     VAPOR FRAC
            BTU/HR       F            PSIA
  16        -0.62105E+07   -290.00      643.00        1.0000
  68        -0.19540E+08   -290.00      143.81        1.0000
  90         0.19759E+08     98.73      21.280        1.0000
  58         0.50771E+07     98.73      15.570        1.0000
  46,+R      0.91432E+06     98.73      50.570        1.0000
           ------------------------------------------
           |                                        |
  16       |                                        | 20L
 --------->|         2356.2     LBMOL/HR            |--------->
   110.00  |                                        |  -290.00
           |                                        |
  68       |                                        | 68,1
 --------->|         9750.0     LBMOL/HR            |--------->
   110.00  |                                        |  -290.00
           |                                        |
  92       |                                        | 90
 <---------|         9750.0     LBMOL/HR            |<---------
    98.73  |                                        |  -308.75
           |                                        |
  60       |                                        | 58
 <---------|         1500.0     LBMOL/HR            |<---------
    98.73  |                                        |  -318.17
           |                                        |
  46       |                                        | 46,+R
 <---------|         303.90     LBMOL/HR            |<---------
    98.73  |                                        |  -308.75
           |                                        |
           ------------------------------------------
```

```
 ASPEN PLUS   PLAT: WIN-X64   VER: 39.0                    04/18/2023  PAGE
22
                EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
                              U-O-S BLOCK SECTION
 BLOCK:  18       MODEL: MHEATX (CONTINUED)
                          ***  INTERNAL ANALYSIS  ***
  FLOW IS COUNTERCURRENT.
  DUTY                            0.25751E+08 BTU/HR
  UA                              0.12472E+07 BTU/HR-R
  AVERAGE LMTD (DUTY/UA)           20.647     F
  MIN TEMP APPROACH                11.272     F
  HOT-SIDE TEMP APPROACH           11.272     F
  COLD-SIDE TEMP APPROACH          28.168     F
  HOT-SIDE NTU                     19.373
  COLD-SIDE NTU                    20.191
  TQ-TABLE(S) INTERPOLATED FOR AT LEAST ONE STREAM DURING INTERNAL ZONE
ANALYSIS.
  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       -290.00   -318.17     28.17
  0.2575E+06   -285.97   -318.24     32.26     30.17   8536.
0.2575E+06   8536.
  0.3863E+06   -283.96   -318.27     34.31     33.27   3870.
0.1288E+06  0.1241E+05
  0.5150E+06   -281.95   -318.30     36.35     35.32   3645.
0.1288E+06  0.1605E+05
  0.7777E+06   -277.84   -308.75     30.91     33.56   7829.
0.2627E+06  0.2388E+05         IN  90
  0.1030E+07   -273.90   -304.66     30.76     30.83   8183.
0.2523E+06  0.3206E+05
  0.1545E+07   -265.85   -296.30     30.45     30.60  0.1683E+05
0.5150E+06  0.4889E+05
  0.2060E+07   -257.80   -287.93     30.14     30.29  0.1700E+05
0.5150E+06  0.6590E+05
  0.2575E+07   -249.75   -279.57     29.83     29.98  0.1718E+05
0.5150E+06  0.8307E+05
  0.3090E+07   -241.66   -271.21     29.55     29.69  0.1735E+05
0.5150E+06  0.1004E+06
  0.3605E+07   -233.59   -262.85     29.26     29.40  0.1751E+05
0.5150E+06  0.1179E+06
  0.4120E+07   -225.52   -254.43     28.92     29.09  0.1771E+05
0.5150E+06  0.1356E+06
  0.4635E+07   -217.45   -246.05     28.60     28.76  0.1791E+05
0.5150E+06  0.1535E+06
  0.5150E+07   -209.38   -237.67     28.29     28.45  0.1810E+05
0.5150E+06  0.1717E+06
  0.5665E+07   -201.30   -229.28     27.98     28.14  0.1830E+05
0.5150E+06  0.1900E+06
  0.6180E+07   -193.20   -220.90     27.69     27.84  0.1850E+05
0.5150E+06  0.2085E+06
```

0.6695E+07	-185.16	-212.51	27.35	27.52	0.1871E+05	0.5150E+06	0.2272E+06
0.7210E+07	-177.11	-204.13	27.02	27.18	0.1895E+05	0.5150E+06	0.2461E+06
0.7725E+07	-169.07	-195.75	26.68	26.85	0.1918E+05	0.5150E+06	0.2653E+06
0.8240E+07	-161.02	-187.36	26.34	26.51	0.1943E+05	0.5150E+06	0.2847E+06
0.8755E+07	-152.97	-178.97	26.01	26.17	0.1968E+05	0.5150E+06	0.3044E+06
0.9270E+07	-144.92	-170.58	25.66	25.83	0.1994E+05	0.5150E+06	0.3243E+06
0.9785E+07	-136.90	-162.19	25.30	25.48	0.2021E+05	0.5150E+06	0.3446E+06
0.1030E+08	-128.90	-153.80	24.91	25.10	0.2052E+05	0.5150E+06	0.3651E+06
0.1082E+08	-120.86	-145.42	24.55	24.73	0.2083E+05	0.5150E+06	0.3859E+06
0.1133E+08	-112.86	-137.02	24.16	24.36	0.2115E+05	0.5150E+06	0.4070E+06
0.1185E+08	-104.86	-128.63	23.77	23.96	0.2149E+05	0.5150E+06	0.4285E+06
0.1236E+08	-96.86	-120.23	23.37	23.57	0.2185E+05	0.5150E+06	0.4504E+06
0.1288E+08	-88.86	-111.83	22.97	23.17	0.2223E+05	0.5150E+06	0.4726E+06
0.1339E+08	-80.89	-103.43	22.54	22.75	0.2264E+05	0.5150E+06	0.4953E+06
0.1391E+08	-72.89	-95.03	22.14	22.34	0.2306E+05	0.5150E+06	0.5183E+06
0.1442E+08	-64.92	-86.62	21.71	21.92	0.2349E+05	0.5150E+06	0.5418E+06
0.1494E+08	-56.92	-78.22	21.30	21.50	0.2395E+05	0.5150E+06	0.5658E+06
0.1545E+08	-48.95	-69.80	20.85	21.08	0.2444E+05	0.5150E+06	0.5902E+06
0.1597E+08	-40.98	-61.39	20.41	20.63	0.2496E+05	0.5150E+06	0.6152E+06
0.1648E+08	-33.02	-52.98	19.96	20.18	0.2552E+05	0.5150E+06	0.6407E+06
0.1700E+08	-25.05	-44.56	19.51	19.74	0.2610E+05	0.5150E+06	0.6668E+06
0.1751E+08	-17.08	-36.15	19.07	19.29	0.2670E+05	0.5150E+06	0.6935E+06
0.1803E+08	-9.13	-27.73	18.60	18.83	0.2735E+05	0.5150E+06	0.7208E+06

```
               EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
                            U-O-S BLOCK SECTION
 BLOCK:  18       MODEL: MHEATX (CONTINUED)
  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
  0.1854E+08     -1.18    -19.30     18.13     18.36  0.2805E+05
0.5150E+06  0.7489E+06
  0.1906E+08      6.77    -10.88     17.65     17.89  0.2879E+05
0.5150E+06  0.7777E+06
  0.1957E+08     14.73     -2.46     17.18     17.42  0.2957E+05
0.5150E+06  0.8072E+06
  0.2009E+08     22.68      5.97     16.71     16.95  0.3039E+05
0.5150E+06  0.8376E+06
  0.2060E+08     30.62     14.40     16.23     16.47  0.3127E+05
0.5150E+06  0.8689E+06
  0.2112E+08     38.57     22.83     15.74     15.98  0.3222E+05
0.5150E+06  0.9011E+06
  0.2163E+08     46.51     31.26     15.25     15.49  0.3324E+05
0.5150E+06  0.9344E+06
  0.2215E+08     54.45     39.69     14.76     15.00  0.3432E+05
0.5150E+06  0.9687E+06
  0.2266E+08     62.39     48.12     14.27     14.51  0.3548E+05
0.5150E+06  0.1004E+07
  0.2318E+08     70.33     56.55     13.78     14.02  0.3673E+05
0.5150E+06  0.1041E+07
  0.2369E+08     78.26     64.99     13.28     13.53  0.3808E+05
0.5150E+06  0.1079E+07
  0.2421E+08     86.20     73.42     12.78     13.02  0.3954E+05
0.5150E+06  0.1119E+07
  0.2472E+08     94.13     81.86     12.27     12.52  0.4112E+05
0.5150E+06  0.1160E+07
  0.2524E+08    102.07     90.29     11.77     12.02  0.4284E+05
0.5150E+06  0.1202E+07
  0.2575E+08    110.00     98.73     11.27     11.52  0.4470E+05
0.5150E+06  0.1247E+07  GBL
  GBL = GLOBAL     LOC = LOCAL     DP  = DEW POINT     BP  = BUBBLE POINT
```

BLOCK: 18 MODEL: MHEATX (CONTINUED)

*** TQ-TABLES FOR HOT SIDE STREAMS ***

STREAM: 16

Q STREAM BTU/HR	TEMP F	PRES PSIA	VFRAC
0.6210E+07	110.0	650.0	1.000
0.5520E+07	67.87	649.2	1.000
0.4830E+07	25.54	648.4	1.000
0.4140E+07	-17.10	647.7	1.000
0.3450E+07	-60.18	646.9	1.000
0.2760E+07	-103.9	646.1	1.000
0.2070E+07	-148.5	645.3	1.000
0.1380E+07	-194.3	644.6	1.000
0.6901E+06	-241.6	643.8	1.000
0.000	-290.0	643.0	1.000

STREAM: 68

Q STREAM BTU/HR	TEMP F	PRES PSIA	VFRAC
0.1954E+08	110.0	150.0	1.000
0.1737E+08	65.26	149.3	1.000
0.1520E+08	20.55	148.6	1.000
0.1303E+08	-24.14	147.9	1.000
0.1086E+08	-68.78	147.2	1.000
0.8685E+07	-113.4	146.6	1.000
0.6513E+07	-157.9	145.9	1.000
0.4342E+07	-202.2	145.2	1.000
0.2171E+07	-246.3	144.5	1.000
0.000	-290.0	143.8	1.000

BLOCK: 18   MODEL: MHEATX (CONTINUED)

*** TQ-TABLES FOR COLD SIDE STREAMS ***

STREAM: 90

Q STREAM BTU/HR	TEMP F	PRES PSIA	VFRAC
0.000	-308.8	23.93	1.000
0.2195E+07	-263.7	23.64	1.000
0.4391E+07	-218.5	23.34	1.000
0.6586E+07	-173.2	23.05	1.000
0.8782E+07	-127.9	22.75	1.000
0.1098E+08	-82.60	22.46	1.000
0.1317E+08	-37.28	22.16	1.000
0.1537E+08	8.053	21.87	1.000
0.1756E+08	53.39	21.57	1.000
0.1976E+08	98.73	21.28	1.000

STREAM: 58

Q STREAM BTU/HR	TEMP F	PRES PSIA	VFRAC
0.000	-318.2	17.00	0.8111
0.6346E+06	-318.3	16.82	0.9893
0.6727E+06	-318.3	16.81	1.000
0.1269E+07	-263.9	16.64	1.000
0.1904E+07	-204.1	16.46	1.000
0.2539E+07	-143.8	16.29	1.000
0.3173E+07	-83.31	16.11	1.000
0.3808E+07	-22.67	15.93	1.000
0.4442E+07	38.02	15.75	1.000
0.5077E+07	98.73	15.57	1.000

STREAM: 46,+R

Q STREAM BTU/HR	TEMP F	PRES PSIA	VFRAC
0.000	-308.8	51.04	1.000
0.1016E+06	-256.8	50.99	1.000
0.2032E+06	-211.4	50.94	1.000
0.3048E+06	-168.6	50.88	1.000
0.4064E+06	-126.2	50.83	1.000
0.5080E+06	-83.15	50.78	1.000
0.6095E+06	-39.08	50.73	1.000
0.7111E+06	6.036	50.67	1.000
0.8127E+06	52.05	50.62	1.000

```
 BLOCK:  18       MODEL: MHEATX (CONTINUED)
   Q STREAM    TEMP        PRES        VFRAC
   BTU/HR      F           PSIA
   0.9143E+06   98.73       50.57       1.000
 BLOCK:  22       MODEL: MHEATX
 ------------------------------
   HOT SIDE:    INLET STREAM     OUTLET STREAM
                ------------     -------------
                20L              24
                68,1             70
   COLD SIDE:   INLET STREAM     OUTLET STREAM
                ------------     -------------
                T52              54
   PROPERTIES FOR STREAM 20L
   PROPERTY OPTION SET:   REFPROP
   PROPERTIES FOR STREAM 68,1
   PROPERTY OPTION SET:   REFPROP
   PROPERTIES FOR STREAM T52
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                                 IN              OUT         RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           12802.2         12802.2        0.142084E-
15
       MASS(LB/HR   )           220999.         220999.         0.00000
       ENTHALPY(BTU/HR  )      -0.271184E+08   -0.271184E+08   -0.137371E-
15
```

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING

U-O-S BLOCK SECTION

BLOCK: 22 MODEL: MHEATX (CONTINUED)

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	0.00000	LB/HR
PRODUCT STREAMS CO2E	0.00000	LB/HR
NET STREAMS CO2E PRODUCTION	0.00000	LB/HR
UTILITIES CO2E PRODUCTION	0.00000	LB/HR
TOTAL CO2E PRODUCTION	0.00000	LB/HR

*** INPUT DATA ***

SPECIFICATIONS FOR STREAM 20L :

TWO PHASE TP FLASH

SPECIFIED TEMPERATURE	F	-305.000
PRESSURE DROP	PSI	2.30000
MAXIMUM NO. ITERATIONS		50
CONVERGENCE TOLERANCE		0.100000-06

SPECIFICATIONS FOR STREAM 68,1 :

TWO PHASE TP FLASH

SPECIFIED TEMPERATURE	F	-305.000
PRESSURE DROP	PSI	8.30000
MAXIMUM NO. ITERATIONS		50
CONVERGENCE TOLERANCE		0.100000-06

SPECIFICATIONS FOR STREAM T52 :

TWO PHASE FLASH

PRESSURE DROP	PSI	1.13000
MAXIMUM NO. ITERATIONS		50
CONVERGENCE TOLERANCE		0.100000-06

*** RESULTS ***

INLET STREAM	DUTY BTU/HR	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	OUTLET VAPOR FRAC
20L	-0.21460E+06	-305.00	640.70	1.0000
68,1	-0.74531E+06	-305.00	135.51	1.0000
T52	0.95991E+06	-316.50	18.870	0.5863

```
              EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
                           U-O-S BLOCK SECTION
 BLOCK:  22       MODEL: MHEATX (CONTINUED)
          ------------------------------------------
          |                                        |
  20L     |                                        | 24
--------->|       2356.2     LBMOL/HR              |--------->
  -290.00 |                                        |  -305.00
          |                                        |
  68,1    |                                        | 70
--------->|       9750.0     LBMOL/HR              |--------->
  -290.00 |                                        |  -305.00
          |                                        |
  54      |                                        | T52
<---------|       696.00     LBMOL/HR              |<---------
  -316.50 |                                        |  -316.00
          |                                        |
          ------------------------------------------
```

```
              EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
                            U-O-S BLOCK SECTION
 BLOCK:  22       MODEL: MHEATX (CONTINUED)
                        ***  INTERNAL ANALYSIS  ***
  FLOW IS COUNTERCURRENT.
  DUTY                               0.95991E+06 BTU/HR
  UA                                  55466.     BTU/HR-R
  AVERAGE LMTD (DUTY/UA)              17.306     F
  MIN TEMP APPROACH                   10.619     F
  HOT-SIDE TEMP APPROACH              26.497     F
  COLD-SIDE TEMP APPROACH             11.000     F
  HOT-SIDE NTU                       0.86673
  COLD-SIDE NTU                      -.28732E-01
  TQ-TABLE(S) INTERPOLATED FOR AT LEAST ONE STREAM DURING INTERNAL ZONE
ANALYSIS.
  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       -305.00   -316.00     11.00
   2145.       -304.97   -315.78     10.81     10.90   196.7       2145.
196.7
   3217.       -304.95   -315.66     10.71     10.76   99.65       1072.
296.3
   4290.       -304.93   -315.55     10.62     10.67   100.5       1072.
396.9      GBL    BP  T52
  0.1920E+05   -304.70   -315.57     10.86     10.74   1388.
0.1491E+05   1785.
  0.3840E+05   -304.40   -315.58     11.18     11.02   1742.
0.1920E+05   3527.
  0.5759E+05   -304.10   -315.60     11.50     11.34   1693.
0.1920E+05   5219.
  0.7679E+05   -303.80   -315.62     11.82     11.66   1647.
0.1920E+05   6866.
  0.9599E+05   -303.51   -315.64     12.14     11.98   1603.
0.1920E+05   8469.
  0.1152E+06   -303.21   -315.66     12.45     12.29   1562.
0.1920E+05  0.1003E+05
  0.1344E+06   -302.91   -315.68     12.77     12.61   1522.
0.1920E+05  0.1155E+05
  0.1536E+06   -302.61   -315.70     13.09     12.93   1485.
0.1920E+05  0.1304E+05
  0.1728E+06   -302.31   -315.72     13.41     13.25   1449.
0.1920E+05  0.1449E+05
  0.1920E+06   -302.01   -315.73     13.73     13.57   1415.
0.1920E+05  0.1590E+05
  0.2112E+06   -301.71   -315.75     14.04     13.88   1383.
0.1920E+05  0.1728E+05
  0.2304E+06   -301.41   -315.77     14.36     14.20   1352.
0.1920E+05  0.1864E+05
```

```
  0.2496E+06   -301.11   -315.79     14.68     14.52   1322.
0.1920E+05  0.1996E+05
  0.2688E+06   -300.81   -315.81     15.00     14.84   1294.
0.1920E+05  0.2125E+05
  0.2880E+06   -300.51   -315.83     15.32     15.16   1267.
0.1920E+05  0.2252E+05
  0.3072E+06   -300.21   -315.85     15.64     15.48   1241.
0.1920E+05  0.2376E+05
  0.3264E+06   -299.91   -315.87     15.95     15.79   1215.
0.1920E+05  0.2497E+05
  0.3456E+06   -299.61   -315.88     16.27     16.11   1191.
0.1920E+05  0.2617E+05
  0.3648E+06   -299.31   -315.90     16.59     16.43   1168.
0.1920E+05  0.2733E+05
  0.3840E+06   -299.01   -315.92     16.91     16.75   1146.
0.1920E+05  0.2848E+05
  0.4032E+06   -298.71   -315.94     17.23     17.07   1125.
0.1920E+05  0.2961E+05
  0.4224E+06   -298.41   -315.96     17.55     17.39   1104.
0.1920E+05  0.3071E+05
  0.4416E+06   -298.11   -315.98     17.87     17.71   1084.
0.1920E+05  0.3179E+05
  0.4608E+06   -297.81   -316.00     18.19     18.03   1065.
0.1920E+05  0.3286E+05
  0.4800E+06   -297.51   -316.02     18.51     18.35   1046.
0.1920E+05  0.3390E+05
  0.4992E+06   -297.21   -316.04     18.83     18.67   1029.
0.1920E+05  0.3493E+05
  0.5184E+06   -296.91   -316.06     19.14     18.98   1011.
0.1920E+05  0.3594E+05
  0.5375E+06   -296.61   -316.07     19.46     19.30   994.6
0.1920E+05  0.3694E+05
  0.5567E+06   -296.31   -316.09     19.78     19.62   978.4
0.1920E+05  0.3792E+05
  0.5759E+06   -296.01   -316.11     20.10     19.94   962.7
0.1920E+05  0.3888E+05
  0.5951E+06   -295.71   -316.13     20.42     20.26   947.5
0.1920E+05  0.3983E+05
  0.6143E+06   -295.41   -316.15     20.74     20.58   932.8
0.1920E+05  0.4076E+05
  0.6335E+06   -295.11   -316.17     21.06     20.90   918.6
0.1920E+05  0.4168E+05
  0.6527E+06   -294.81   -316.19     21.38     21.22   904.8
0.1920E+05  0.4258E+05
  0.6719E+06   -294.51   -316.21     21.70     21.54   891.3
0.1920E+05  0.4348E+05
```

BLOCK: 22   MODEL: MHEATX (CONTINUED)
```
  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
  0.6911E+06   -294.21   -316.23     22.02     21.86   878.3
0.1920E+05  0.4435E+05
  0.7103E+06   -293.91   -316.25     22.34     22.18   865.7
0.1920E+05  0.4522E+05
  0.7295E+06   -293.61   -316.27     22.66     22.50   853.4
0.1920E+05  0.4607E+05
  0.7487E+06   -293.31   -316.28     22.98     22.82   841.4
0.1920E+05  0.4691E+05
  0.7679E+06   -293.01   -316.30     23.30     23.14   829.8
0.1920E+05  0.4774E+05
  0.7871E+06   -292.71   -316.32     23.62     23.46   818.5
0.1920E+05  0.4856E+05
  0.8063E+06   -292.41   -316.34     23.94     23.78   807.4
0.1920E+05  0.4937E+05
  0.8255E+06   -292.11   -316.36     24.26     24.10   796.7
0.1920E+05  0.5017E+05
  0.8447E+06   -291.80   -316.38     24.58     24.42   786.3
0.1920E+05  0.5095E+05
  0.8639E+06   -291.50   -316.40     24.90     24.74   776.1
0.1920E+05  0.5173E+05
  0.8831E+06   -291.20   -316.42     25.22     25.06   766.2
0.1920E+05  0.5250E+05
  0.9023E+06   -290.90   -316.44     25.54     25.38   756.5
0.1920E+05  0.5325E+05
  0.9215E+06   -290.60   -316.46     25.86     25.70   747.1
0.1920E+05  0.5400E+05
  0.9407E+06   -290.30   -316.48     26.18     26.02   737.9
0.1920E+05  0.5474E+05
  0.9599E+06   -290.00   -316.50     26.50     26.34   728.9
0.1920E+05  0.5547E+05
  GBL = GLOBAL     LOC = LOCAL     DP  = DEW POINT     BP  = BUBBLE POINT
```

BLOCK: 22   MODEL: MHEATX (CONTINUED)

*** TQ-TABLES FOR HOT SIDE STREAMS ***

STREAM: 20L

Q STREAM BTU/HR	TEMP F	PRES PSIA	VFRAC
0.2146E+06	-290.0	643.0	1.000
0.1908E+06	-291.7	642.7	1.000
0.1669E+06	-293.3	642.5	1.000
0.1431E+06	-295.0	642.2	1.000
0.1192E+06	-296.7	642.0	1.000
0.9538E+05	-298.4	641.7	1.000
0.7153E+05	-300.0	641.5	1.000
0.4769E+05	-301.7	641.2	1.000
0.2384E+05	-303.3	641.0	1.000
0.000	-305.0	640.7	1.000

STREAM: 68,1

Q STREAM BTU/HR	TEMP F	PRES PSIA	VFRAC
0.7453E+06	-290.0	143.8	1.000
0.6625E+06	-291.7	142.9	1.000
0.5797E+06	-293.3	142.0	1.000
0.4969E+06	-295.0	141.0	1.000
0.4141E+06	-296.7	140.1	1.000
0.3312E+06	-298.3	139.2	1.000
0.2484E+06	-300.0	138.3	1.000
0.1656E+06	-301.7	137.4	1.000
0.8281E+05	-303.3	136.4	1.000
0.000	-305.0	135.5	1.000

*** TQ-TABLES FOR COLD SIDE STREAMS ***

STREAM: T52

Q STREAM BTU/HR	TEMP F	PRES PSIA	VFRAC
0.000	-316.0	20.00	0.000
4290.	-315.6	20.00	0.000
0.1200E+06	-315.7	19.86	0.7120E-01
0.2400E+06	-315.8	19.72	0.1450
0.3600E+06	-315.9	19.58	0.2187
0.4800E+06	-316.0	19.44	0.2923
0.5999E+06	-316.1	19.29	0.3659
0.7199E+06	-316.3	19.15	0.4394
0.8399E+06	-316.4	19.01	0.5129
0.9599E+06	-316.5	18.87	0.5863

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
U-O-S BLOCK SECTION

```
 BLOCK:  26       MODEL: MHEATX
 ------------------------------
   HOT SIDE:    INLET STREAM     OUTLET STREAM
                ------------     -------------
                72               74
                24,1             28,1
   COLD SIDE:   INLET STREAM     OUTLET STREAM
                ------------     -------------
                46,-2            46,+R
                88               90
   PROPERTIES FOR STREAM 72
   PROPERTY OPTION SET:   REFPROP
   PROPERTIES FOR STREAM 24,1
   PROPERTY OPTION SET:   REFPROP
   PROPERTIES FOR STREAM 46,-2
   PROPERTY OPTION SET:   REFPROP
   PROPERTIES FOR STREAM 88
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                                   IN              OUT         RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           17577.6         17577.6         0.00000
       MASS(LB/HR   )           306393.         306393.       -0.189977E-
15
       ENTHALPY(BTU/HR  )      -0.393246E+08   -0.390100E+08   -0.799839E-
02
                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E            0.00000      LB/HR
    PRODUCT STREAMS CO2E         0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
    UTILITIES CO2E PRODUCTION    0.00000      LB/HR
    TOTAL CO2E PRODUCTION        0.00000      LB/HR
```

```
 BLOCK:  26       MODEL: MHEATX (CONTINUED)
                          ***  INPUT DATA  ***
   MAXIMUM NO. ITERATIONS                                      30
   CONVERGENCE TOLERANCE                                        0.000100000
   LOWER LIMIT ON TEMPERATURE      F                         -456.000
   HEAT LEAK                       BTU/HR                 314,531.
   SPECIFICATIONS FOR STREAM 72      :
   TWO    PHASE   TP  FLASH
   SPECIFIED TEMPERATURE                F                          -367.000
   PRESSURE DROP                        PSI                           2.94000
   MAXIMUM NO. ITERATIONS                                            50
   CONVERGENCE TOLERANCE
0.100000-06
   SPECIFICATIONS FOR STREAM 24,1    :
   TWO    PHASE   TP  FLASH
   SPECIFIED TEMPERATURE                F                          -367.000
   PRESSURE DROP                        PSI                           0.58000
   MAXIMUM NO. ITERATIONS                                            30
   CONVERGENCE TOLERANCE
0.100000-06
   SPECIFICATIONS FOR STREAM 46,-2   :
   TWO    PHASE       FLASH
   PRESSURE DROP                        PSI                           0.60000
   MAXIMUM NO. ITERATIONS                                            50
   CONVERGENCE TOLERANCE
0.100000-06
   SPECIFICATIONS FOR STREAM 88      :
   TWO    PHASE       FLASH
   PRESSURE DROP                        PSI                           1.94000
   MAXIMUM NO. ITERATIONS                                            50
   CONVERGENCE TOLERANCE
0.100000-06
                          ***  RESULTS  ***
  INLET                  OUTLET       OUTLET        OUTLET
  STREAM    DUTY         TEMPERATURE  PRESSURE      VAPOR FRAC
            BTU/HR       F            PSIA
  72        -0.17255E+07   -367.00      132.57         1.0000
  24,1      -0.10694E+07   -367.00      640.12         1.0000
  46,-2      0.10278E+06   -308.75      51.040         1.0000
  88         0.30067E+07   -308.75      23.930         1.0000
```

```
 BLOCK:  26       MODEL: MHEATX (CONTINUED)
            ------------------------------------------
            |                                        |
  72        |                                        | 74
 ---------->|        5167.5     LBMOL/HR             |---------->
   -305.00  |                                        |  -367.00
            |                                        |
  24,1      |                                        | 28,1
 ---------->|        2356.2     LBMOL/HR             |---------->
   -305.00  |                                        |  -367.00
            |                                        |
  46,+R     |                                        | 46,-2
 <----------|        303.90     LBMOL/HR             |<----------
   -308.75  |                                        |  -370.86
            |                                        |
  90        |                                        | 88
 <----------|        9750.0     LBMOL/HR             |<----------
   -308.75  |                                        |  -369.80
            |                                        |
            ------------------------------------------
```

```
 BLOCK:  26       MODEL: MHEATX (CONTINUED)
                          ***  INTERNAL ANALYSIS  ***
  FLOW IS COUNTERCURRENT.
  DUTY                               0.27949E+07 BTU/HR
  HEAT LEAK                          0.31453E+06 BTU/HR
  UA                                 0.11992E+07 BTU/HR-R
  AVERAGE LMTD (DUTY/UA)              2.3306     F
  MIN TEMP APPROACH                   1.8236     F
  HOT-SIDE TEMP APPROACH              3.7511     F
  COLD-SIDE TEMP APPROACH             3.8619     F
  HOT-SIDE NTU                        26.603
  COLD-SIDE NTU                       23.954
  TQ-TABLE(S) INTERPOLATED FOR AT LEAST ONE STREAM DURING INTERNAL ZONE
ANALYSIS.
  Q HOT       Q COLD        T HOT     T COLD    DELTA T   LMTD    UA ZONE
Q ZONE      UA          PINCH  STREAM IN/OUT/DEW/

POINT  BUBBLE POINT
  BTU/HR      BTU/HR          F         F         F         F     BTU/HR-R
BTU/HR      BTU/HR-R
   0.000       0.000       -367.00   -370.86      3.86
   762.4       848.2       -366.99   -370.33      3.35      3.60   211.8
762.4       211.8
   1525.       1696.       -366.97   -369.80      2.83      3.08   247.3
762.4       459.1             IN  88
  0.2871E+05  0.3194E+05   -366.44   -369.22      2.78      2.81   9690.
0.2719E+05  0.1015E+05
  0.5590E+05  0.6219E+05   -365.90   -368.63      2.72      2.75   9882.
0.2719E+05  0.2003E+05
  0.1118E+06  0.1244E+06   -364.81   -367.42      2.61      2.67
0.2096E+05  0.5590E+05  0.4099E+05
  0.1677E+06  0.1866E+06   -363.71   -366.21      2.50      2.56
0.2188E+05  0.5590E+05  0.6286E+05
  0.2236E+06  0.2488E+06   -362.62   -365.01      2.39      2.44
0.2288E+05  0.5590E+05  0.8574E+05
  0.2795E+06  0.3109E+06   -361.52   -363.80      2.28      2.33
0.2398E+05  0.5590E+05  0.1097E+06
  0.3354E+06  0.3731E+06   -360.39   -362.59      2.20      2.24
0.2499E+05  0.5590E+05  0.1347E+06
  0.3913E+06  0.4353E+06   -359.24   -361.38      2.14      2.17
0.2578E+05  0.5590E+05  0.1605E+06
  0.4472E+06  0.4975E+06   -358.08   -360.16      2.08      2.11
0.2652E+05  0.5590E+05  0.1870E+06
  0.5031E+06  0.5597E+06   -356.93   -358.95      2.02      2.05
0.2731E+05  0.5590E+05  0.2143E+06
  0.5590E+06  0.6219E+06   -355.78   -357.73      1.96      1.99
0.2815E+05  0.5590E+05  0.2425E+06
  0.6149E+06  0.6841E+06   -354.60   -356.52      1.92      1.94
0.2888E+05  0.5590E+05  0.2713E+06
```

```
  0.6708E+06  0.7463E+06   -353.41   -355.30      1.89      1.90
0.2937E+05  0.5590E+05  0.3007E+06
  0.7267E+06  0.8085E+06   -352.21   -354.08      1.87      1.88
0.2971E+05  0.5590E+05  0.3304E+06
  0.7826E+06  0.8707E+06   -351.01   -352.87      1.85      1.86
0.3000E+05  0.5590E+05  0.3604E+06
  0.8385E+06  0.9328E+06   -349.81   -351.65      1.84      1.85
0.3029E+05  0.5590E+05  0.3907E+06
  0.8944E+06  0.9950E+06   -348.61   -350.43      1.82      1.83
0.3054E+05  0.5590E+05  0.4213E+06  GBL
  0.9503E+06  0.1057E+07   -347.38   -349.21      1.83      1.83
0.3060E+05  0.5590E+05  0.4519E+06
  0.1006E+07  0.1119E+07   -346.15   -347.99      1.84      1.84
0.3045E+05  0.5590E+05  0.4823E+06
  0.1062E+07  0.1182E+07   -344.91   -346.77      1.86      1.85
0.3020E+05  0.5590E+05  0.5125E+06
  0.1118E+07  0.1244E+07   -343.67   -345.55      1.87      1.87
0.2994E+05  0.5590E+05  0.5424E+06
  0.1174E+07  0.1306E+07   -342.43   -344.32      1.89      1.88
0.2969E+05  0.5590E+05  0.5721E+06
  0.1230E+07  0.1368E+07   -341.18   -343.10      1.92      1.91
0.2933E+05  0.5590E+05  0.6015E+06
  0.1286E+07  0.1430E+07   -339.92   -341.88      1.95      1.94
0.2887E+05  0.5590E+05  0.6303E+06
  0.1342E+07  0.1493E+07   -338.66   -340.65      2.00      1.97
0.2831E+05  0.5590E+05  0.6586E+06
  0.1397E+07  0.1555E+07   -337.39   -339.43      2.04      2.02
0.2773E+05  0.5590E+05  0.6864E+06
  0.1453E+07  0.1617E+07   -336.13   -338.21      2.08      2.06
0.2717E+05  0.5590E+05  0.7135E+06
  0.1509E+07  0.1679E+07   -334.86   -336.98      2.12      2.10
0.2664E+05  0.5590E+05  0.7402E+06
  0.1565E+07  0.1741E+07   -333.59   -335.76      2.17      2.14
0.2606E+05  0.5590E+05  0.7662E+06
  0.1621E+07  0.1803E+07   -332.30   -334.53      2.23      2.20
0.2542E+05  0.5590E+05  0.7917E+06
  0.1677E+07  0.1866E+07   -331.02   -333.31      2.29      2.26
0.2476E+05  0.5590E+05  0.8164E+06
  0.1733E+07  0.1928E+07   -329.73   -332.08      2.35      2.32
0.2413E+05  0.5590E+05  0.8405E+06
  0.1789E+07  0.1990E+07   -328.45   -330.85      2.41      2.38
0.2353E+05  0.5590E+05  0.8641E+06
  0.1845E+07  0.2052E+07   -327.16   -329.63      2.47      2.44
0.2295E+05  0.5590E+05  0.8870E+06
  0.1901E+07  0.2114E+07   -325.87   -328.40      2.53      2.50
0.2237E+05  0.5590E+05  0.9094E+06
  0.1956E+07  0.2177E+07   -324.57   -327.17      2.60      2.57
0.2177E+05  0.5590E+05  0.9312E+06
```

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 BLOCK:  26       MODEL: MHEATX (CONTINUED)
  Q HOT       Q COLD        T HOT     T COLD    DELTA T   LMTD    UA ZONE
Q ZONE      UA          PINCH  STREAM IN/OUT/DEW/

POINT  BUBBLE POINT
  BTU/HR      BTU/HR          F         F         F         F     BTU/HR-R
BTU/HR      BTU/HR-R
  0.2012E+07  0.2239E+07   -323.27   -325.95      2.67      2.64
0.2119E+05  0.5590E+05  0.9523E+06
  0.2068E+07  0.2301E+07   -321.97   -324.72      2.74      2.71
0.2063E+05  0.5590E+05  0.9730E+06
  0.2124E+07  0.2363E+07   -320.68   -323.49      2.82      2.78
0.2011E+05  0.5590E+05  0.9931E+06
  0.2180E+07  0.2425E+07   -319.38   -322.26      2.89      2.85
0.1961E+05  0.5590E+05  0.1013E+07
  0.2236E+07  0.2488E+07   -318.07   -321.04      2.96      2.93
0.1911E+05  0.5590E+05  0.1032E+07
  0.2292E+07  0.2550E+07   -316.77   -319.81      3.04      3.00
0.1862E+05  0.5590E+05  0.1050E+07
  0.2348E+07  0.2612E+07   -315.46   -318.58      3.12      3.08
0.1815E+05  0.5590E+05  0.1069E+07
  0.2404E+07  0.2674E+07   -314.16   -317.35      3.20      3.16
0.1771E+05  0.5590E+05  0.1086E+07
  0.2460E+07  0.2736E+07   -312.85   -316.12      3.27      3.23
0.1728E+05  0.5590E+05  0.1104E+07
  0.2515E+07  0.2799E+07   -311.54   -314.89      3.35      3.31
0.1688E+05  0.5590E+05  0.1120E+07
  0.2571E+07  0.2861E+07   -310.23   -313.67      3.43      3.39
0.1649E+05  0.5590E+05  0.1137E+07
  0.2627E+07  0.2923E+07   -308.93   -312.44      3.51      3.47
0.1611E+05  0.5590E+05  0.1153E+07
  0.2683E+07  0.2985E+07   -307.62   -311.21      3.59      3.55
0.1574E+05  0.5590E+05  0.1169E+07
  0.2739E+07  0.3047E+07   -306.31   -309.98      3.67      3.63
0.1540E+05  0.5590E+05  0.1184E+07
  0.2795E+07  0.3109E+07   -305.00   -308.75      3.75      3.71
0.1506E+05  0.5590E+05  0.1199E+07
  GBL = GLOBAL     LOC = LOCAL     DP  = DEW POINT     BP  = BUBBLE POINT
```

BLOCK: 26 MODEL: MHEATX (CONTINUED)

*** TQ-TABLES FOR HOT SIDE STREAMS ***

STREAM: 72

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.1726E+07	-305.0	135.5	1.000
0.1534E+07	-312.2	135.2	1.000
0.1342E+07	-319.3	134.9	1.000
0.1150E+07	-326.3	134.5	1.000
0.9586E+06	-333.4	134.2	1.000
0.7669E+06	-340.3	133.9	1.000
0.5752E+06	-347.2	133.6	1.000
0.3835E+06	-353.9	133.2	1.000
0.1917E+06	-360.5	132.9	1.000
0.000	-367.0	132.6	1.000

STREAM: 24,1

Q STREAM	TEMP	PRES	VFRAC
BTU/HR	F	PSIA	
0.1069E+07	-305.0	640.7	1.000
0.9506E+06	-312.5	640.6	1.000
0.8318E+06	-319.9	640.6	1.000
0.7129E+06	-327.4	640.5	1.000
0.5941E+06	-334.8	640.4	1.000
0.4753E+06	-341.9	640.4	1.000
0.3565E+06	-348.8	640.3	1.000
0.2376E+06	-355.4	640.2	1.000
0.1188E+06	-361.4	640.2	1.000
0.000	-367.0	640.1	1.000

BLOCK: 26 MODEL: MHEATX (CONTINUED)

*** TQ-TABLES FOR COLD SIDE STREAMS ***

STREAM: 46,-2

Q STREAM BTU/HR	TEMP F	PRES PSIA	VFRAC
0.000	-370.9	51.64	1.000
0.1142E+05	-363.7	51.57	1.000
0.2284E+05	-356.6	51.51	1.000
0.3426E+05	-349.5	51.44	1.000
0.4568E+05	-342.4	51.37	1.000
0.5710E+05	-335.5	51.31	1.000
0.6852E+05	-328.6	51.24	1.000
0.7994E+05	-321.8	51.17	1.000
0.9136E+05	-315.2	51.11	1.000
0.1028E+06	-308.8	51.04	1.000

STREAM: 88

Q STREAM BTU/HR	TEMP F	PRES PSIA	VFRAC
0.000	-369.8	25.87	1.000
0.3341E+06	-363.1	25.65	1.000
0.6682E+06	-356.4	25.44	1.000
0.1002E+07	-349.6	25.22	1.000
0.1336E+07	-342.8	25.01	1.000
0.1670E+07	-336.0	24.79	1.000
0.2004E+07	-329.2	24.58	1.000
0.2339E+07	-322.4	24.36	1.000
0.2673E+07	-315.6	24.15	1.000
0.3007E+07	-308.8	23.93	1.000

BLOCK: 30 MODEL: MHEATX

	INLET STREAM	OUTLET STREAM
HOT SIDE:	28,1	32
COLD SIDE:	78	80
	42	46,-2

PROPERTIES FOR STREAM 28,1
PROPERTY OPTION SET: REFPROP

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 BLOCK:  30       MODEL: MHEATX (CONTINUED)
   PROPERTIES FOR STREAM 78
   PROPERTY OPTION SET:   REFPROP
   PROPERTIES FOR STREAM 42
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                                   IN               OUT         RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           7827.63         7827.63       -0.116190E-
15
       MASS(LB/HR   )           109641.         109641.         0.00000
       ENTHALPY(BTU/HR  )      -0.213513E+08   -0.211589E+08   -0.901227E-
02
                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E            0.00000      LB/HR
    PRODUCT STREAMS CO2E         0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
    UTILITIES CO2E PRODUCTION    0.00000      LB/HR
    TOTAL CO2E PRODUCTION        0.00000      LB/HR
                          ***  INPUT DATA  ***
   MAXIMUM NO. ITERATIONS                                   30
   CONVERGENCE TOLERANCE                                     0.000100000
   LOWER LIMIT ON TEMPERATURE      F                      -456.000
   HEAT LEAK                       BTU/HR              192,424.
   SPECIFICATIONS FOR STREAM 28,1    :
   TWO    PHASE   TP  FLASH
   SPECIFIED TEMPERATURE                F                       -404.000
   PRESSURE DROP                        PSI                        0.11000
   MAXIMUM NO. ITERATIONS                                        30
   CONVERGENCE TOLERANCE
0.100000-06
   SPECIFICATIONS FOR STREAM 78      :
   TWO    PHASE       FLASH
   PRESSURE DROP                        PSI                        1.53000
   MAXIMUM NO. ITERATIONS                                        50
   CONVERGENCE TOLERANCE
0.100000-06
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              EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
                          U-O-S BLOCK SECTION
 BLOCK:  30       MODEL: MHEATX (CONTINUED)
   SPECIFICATIONS FOR STREAM 42      :
   TWO    PHASE      FLASH
   PRESSURE DROP                        PSI                          0.36000
   MAXIMUM NO. ITERATIONS                                          30
   CONVERGENCE TOLERANCE
0.100000-06
                          ***  RESULTS  ***
  INLET                 OUTLET       OUTLET       OUTLET
  STREAM    DUTY        TEMPERATURE  PRESSURE     VAPOR FRAC
            BTU/HR      F            PSIA
  28,1      -0.83070E+06   -404.00     640.01        0.0000
  78         0.94780E+06   -370.86     25.870        1.0000
  42          75324.       -370.86     51.640        1.0000
          ------------------------------------------
          |                                        |
  28,1    |                                        | 32
 --------->|        2356.2     LBMOL/HR            |--------->
   -367.00 |                                       |  -404.00
          |                                        |
  80      |                                        | 78
 <---------|        5167.5     LBMOL/HR            |<---------
   -370.86 |                                       |  -405.06
          |                                        |
  46,-2   |                                        | 42
 <---------|        303.90     LBMOL/HR            |<---------
   -370.86 |                                       |  -413.93
          |                                        |
          ------------------------------------------
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               EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
                            U-O-S BLOCK SECTION
 BLOCK:  30       MODEL: MHEATX (CONTINUED)
                        ***  INTERNAL ANALYSIS  ***
  FLOW IS COUNTERCURRENT.
  DUTY                             0.83070E+06 BTU/HR
  HEAT LEAK                        0.19242E+06 BTU/HR
  UA                               0.19321E+06 BTU/HR-R
  AVERAGE LMTD (DUTY/UA)            4.2996     F
  MIN TEMP APPROACH                 1.9518     F
  HOT-SIDE TEMP APPROACH            3.8619     F
  COLD-SIDE TEMP APPROACH           9.9311     F
  HOT-SIDE NTU                      8.6055
  COLD-SIDE NTU                     8.1331
  TQ-TABLE(S) INTERPOLATED FOR AT LEAST ONE STREAM DURING INTERNAL ZONE
ANALYSIS.
  Q HOT       Q COLD        T HOT     T COLD    DELTA T   LMTD    UA ZONE
Q ZONE      UA          PINCH  STREAM IN/OUT/DEW/

POINT  BUBBLE POINT
  BTU/HR      BTU/HR          F         F         F         F     BTU/HR-R
BTU/HR      BTU/HR-R
   0.000       0.000       -404.00   -413.93      9.93
   3641.       4484.       -403.78   -411.85      8.08      8.97   405.8
3641.       405.8
   7281.       8968.       -403.56   -409.75      6.19      7.09   513.4
3641.       919.1
  0.1092E+05  0.1345E+05   -403.33   -407.43      4.09      5.07   718.3
3641.       1637.
  0.1274E+05  0.1569E+05   -403.22   -406.27      3.04      3.54   514.1
1820.       2152.
  0.1365E+05  0.1681E+05   -403.17   -405.68      2.51      2.77   328.6
910.1       2480.
  0.1456E+05  0.1794E+05   -403.11   -405.06      1.95      2.22   409.7
910.1       2890.      GBL    IN  78
  0.1482E+05  0.1825E+05   -403.10   -405.05      1.96      1.95   131.2
256.5       3021.
  0.1508E+05  0.1857E+05   -403.08   -405.04      1.96      1.96   130.9
256.5       3152.
  0.1559E+05  0.1920E+05   -403.05   -405.02      1.97      1.97   260.6
513.0       3413.
  0.1661E+05  0.2046E+05   -402.99   -404.98      2.00      1.99   516.7
1026.       3929.
  0.3323E+05  0.4093E+05   -401.97   -404.34      2.36      2.17   7642.
0.1661E+05  0.1157E+05
  0.4984E+05  0.6139E+05   -400.96   -403.69      2.73      2.54   6543.
0.1661E+05  0.1812E+05
  0.6646E+05  0.8185E+05   -399.95   -403.04      3.09      2.90   5721.
0.1661E+05  0.2384E+05
  0.8307E+05  0.1023E+06   -398.94   -402.39      3.45      3.27   5083.
0.1661E+05  0.2892E+05
```

0.9968E+05	0.1228E+06	-397.98	-401.74	3.76	3.60	4609.	0.1661E+05	0.3353E+05
0.1163E+06	0.1432E+06	-397.11	-401.08	3.98	3.87	4297.	0.1661E+05	0.3783E+05
0.1329E+06	0.1637E+06	-396.23	-400.41	4.19	4.08	4072.	0.1661E+05	0.4190E+05
0.1495E+06	0.1842E+06	-395.35	-399.75	4.40	4.29	3873.	0.1661E+05	0.4577E+05
0.1661E+06	0.2046E+06	-394.47	-399.08	4.60	4.50	3693.	0.1661E+05	0.4946E+05
0.1828E+06	0.2251E+06	-393.60	-398.41	4.81	4.71	3528.	0.1661E+05	0.5299E+05
0.1994E+06	0.2456E+06	-392.81	-397.74	4.93	4.87	3411.	0.1661E+05	0.5640E+05
0.2160E+06	0.2660E+06	-392.04	-397.06	5.02	4.97	3342.	0.1661E+05	0.5975E+05
0.2326E+06	0.2865E+06	-391.27	-396.38	5.11	5.06	3283.	0.1661E+05	0.6303E+05
0.2492E+06	0.3069E+06	-390.50	-395.69	5.19	5.15	3227.	0.1661E+05	0.6625E+05
0.2658E+06	0.3274E+06	-389.73	-395.01	5.28	5.24	3172.	0.1661E+05	0.6943E+05
0.2824E+06	0.3479E+06	-388.98	-394.33	5.34	5.31	3127.	0.1661E+05	0.7255E+05
0.2991E+06	0.3683E+06	-388.29	-393.64	5.35	5.35	3108.	0.1661E+05	0.7566E+05
0.3157E+06	0.3888E+06	-387.59	-392.94	5.35	5.35	3107.	0.1661E+05	0.7877E+05
0.3323E+06	0.4093E+06	-386.90	-392.25	5.35	5.35	3106.	0.1661E+05	0.8187E+05
0.3489E+06	0.4297E+06	-386.21	-391.56	5.35	5.35	3106.	0.1661E+05	0.8498E+05
0.3655E+06	0.4502E+06	-385.51	-390.86	5.35	5.35	3106.	0.1661E+05	0.8809E+05
0.3821E+06	0.4706E+06	-384.85	-390.16	5.31	5.33	3117.	0.1661E+05	0.9120E+05
0.3987E+06	0.4911E+06	-384.21	-389.46	5.26	5.28	3144.	0.1661E+05	0.9435E+05
0.4154E+06	0.5116E+06	-383.56	-388.76	5.20	5.23	3177.	0.1661E+05	0.9752E+05
0.4320E+06	0.5320E+06	-382.91	-388.06	5.15	5.17	3211.	0.1661E+05	0.1007E+06
0.4486E+06	0.5525E+06	-382.26	-387.35	5.09	5.12	3246.	0.1661E+05	0.1040E+06
0.4652E+06	0.5730E+06	-381.62	-386.65	5.03	5.06	3282.	0.1661E+05	0.1073E+06
0.4818E+06	0.5934E+06	-380.99	-385.94	4.95	4.99	3328.	0.1661E+05	0.1106E+06

BLOCK: 30 MODEL: MHEATX (CONTINUED)

Q HOT	Q COLD	T HOT	T COLD	DELTA T	LMTD	UA ZONE	Q ZONE	UA	PINCH POINT	STREAM IN/OUT/DEW/ BUBBLE POINT
BTU/HR	BTU/HR	F	F	F	F	BTU/HR-R	BTU/HR	BTU/HR-R		
0.4984E+06	0.6139E+06	-380.36	-385.23	4.88	4.91	3381.	0.1661E+05	0.1140E+06		
0.5150E+06	0.6343E+06	-379.73	-384.52	4.80	4.84	3436.	0.1661E+05	0.1174E+06		
0.5317E+06	0.6548E+06	-379.09	-383.81	4.72	4.76	3494.	0.1661E+05	0.1209E+06		
0.5483E+06	0.6753E+06	-378.46	-383.10	4.64	4.68	3553.	0.1661E+05	0.1245E+06		
0.5649E+06	0.6957E+06	-377.83	-382.39	4.56	4.60	3613.	0.1661E+05	0.1281E+06		
0.5815E+06	0.7162E+06	-377.19	-381.67	4.48	4.52	3674.	0.1661E+05	0.1317E+06		
0.5981E+06	0.7367E+06	-376.55	-380.96	4.41	4.45	3736.	0.1661E+05	0.1355E+06		
0.6147E+06	0.7571E+06	-375.91	-380.24	4.33	4.37	3802.	0.1661E+05	0.1393E+06		
0.6313E+06	0.7776E+06	-375.27	-379.52	4.26	4.29	3869.	0.1661E+05	0.1431E+06		
0.6479E+06	0.7980E+06	-374.63	-378.81	4.18	4.22	3938.	0.1661E+05	0.1471E+06		
0.6646E+06	0.8185E+06	-373.96	-378.09	4.13	4.16	3996.	0.1661E+05	0.1511E+06		
0.6812E+06	0.8390E+06	-373.29	-377.37	4.08	4.11	4045.	0.1661E+05	0.1551E+06		
0.6978E+06	0.8594E+06	-372.62	-376.65	4.03	4.06	4095.	0.1661E+05	0.1592E+06		
0.7144E+06	0.8799E+06	-371.95	-375.93	3.98	4.01	4147.	0.1661E+05	0.1634E+06		
0.7310E+06	0.9004E+06	-371.28	-375.21	3.93	3.96	4200.	0.1661E+05	0.1676E+06		
0.7476E+06	0.9208E+06	-370.58	-374.48	3.90	3.92	4241.	0.1661E+05	0.1718E+06		
0.7642E+06	0.9413E+06	-369.86	-373.76	3.90	3.90	4261.	0.1661E+05	0.1761E+06		
0.7809E+06	0.9617E+06	-369.15	-373.04	3.89	3.89	4269.	0.1661E+05	0.1803E+06		
0.7975E+06	0.9822E+06	-368.43	-372.31	3.88	3.88	4279.	0.1661E+05	0.1846E+06		
0.8141E+06	0.1003E+07	-367.72	-371.59	3.87	3.87	4288.	0.1661E+05	0.1889E+06		
0.8307E+06	0.1023E+07	-367.00	-370.86	3.86	3.87	4297.	0.1661E+05	0.1932E+06		

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

*** TQ-TABLES FOR HOT SIDE STREAMS ***

STREAM: 28,1

Q STREAM BTU/HR	TEMP F	PRES PSIA	VFRAC
0.8307E+06	-367.0	640.1	1.000
0.7384E+06	-371.0	640.1	1.000
0.6461E+06	-374.7	640.1	1.000
0.5538E+06	-378.3	640.1	1.000
0.4615E+06	-381.8	640.1	1.000
0.3692E+06	-385.4	640.1	1.000
0.2769E+06	-389.2	640.0	1.000
0.1846E+06	-393.5	640.0	1.000
0.9230E+05	-398.4	640.0	1.000
0.000	-404.0	640.0	0.000

BLOCK: 30 MODEL: MHEATX (CONTINUED)

*** TQ-TABLES FOR COLD SIDE STREAMS ***

STREAM: 78

Q STREAM BTU/HR	TEMP F	PRES PSIA	VFRAC
0.000	-405.1	27.40	1.000
0.1053E+06	-401.5	27.23	1.000
0.2106E+06	-397.9	27.06	1.000
0.3159E+06	-394.2	26.89	1.000
0.4212E+06	-390.4	26.72	1.000
0.5266E+06	-386.5	26.55	1.000
0.6319E+06	-382.7	26.38	1.000
0.7372E+06	-378.7	26.21	1.000
0.8425E+06	-374.8	26.04	1.000
0.9478E+06	-370.9	25.87	1.000

STREAM: 42

Q STREAM BTU/HR	TEMP F	PRES PSIA	VFRAC
0.000	-413.9	52.00	1.000
8369.	-410.1	51.96	1.000
0.1674E+05	-405.7	51.92	1.000
0.2511E+05	-401.1	51.88	1.000
0.3348E+05	-396.3	51.84	1.000
0.4185E+05	-391.4	51.80	1.000
0.5022E+05	-386.3	51.76	1.000
0.5859E+05	-381.2	51.72	1.000
0.6695E+05	-376.1	51.68	1.000
0.7532E+05	-370.9	51.64	1.000

BLOCK: 12 MODEL: MCOMPR

INLET STREAMS: 10 TO STAGE 1
OUTLET STREAMS: 14 FROM STAGE 3
PROPERTY OPTION SET: REFPROP

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
TOTAL BALANCE			
MOLE(LBMOL/HR)	2052.00	2052.00	0.00000
MASS(LB/HR)	4136.59	4136.59	0.00000
ENTHALPY(BTU/HR)	198616.	506416.	-0.607802

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING

U-O-S BLOCK SECTION

BLOCK: 12   MODEL: MCOMPR (CONTINUED)

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	0.00000	LB/HR
PRODUCT STREAMS CO2E	0.00000	LB/HR
NET STREAMS CO2E PRODUCTION	0.00000	LB/HR
UTILITIES CO2E PRODUCTION	0.00000	LB/HR
TOTAL CO2E PRODUCTION	0.00000	LB/HR

*** INPUT DATA ***

ISENTROPIC CENTRIFUGAL COMPRESSOR

NUMBER OF STAGES	3
FINAL PRESSURE, PSIA	650.000
DISTRIBUTION AMONG STAGES	EQUAL P-RATIO

COMPRESSOR SPECIFICATIONS PER STAGE

STAGE NUMBER	MECHANICAL EFFICIENCY	ISENTROPIC EFFICIENCY
1	1.000	0.7200
2	1.000	0.7200
3	1.000	0.7200

COOLER SPECIFICATIONS PER STAGE

STAGE NUMBER	PRESSURE DROP PSI	TEMPERATURE F
1	0.000	110.0
2	0.000	110.0
3	0.000	110.0

*** RESULTS ***

FINAL PRESSURE, PSIA	650.000
TOTAL WORK REQUIRED, HP	1,192.62
TOTAL COOLING DUTY , BTU/HR	-2,726,750.

*** PROFILE ***

COMPRESSOR PROFILE

STAGE NUMBER	OUTLET PRESSURE PSIA	PRESSURE RATIO	OUTLET TEMPERATURE F
1	357.4	1.349	158.6
2	482.0	1.349	181.0
3	650.0	1.349	181.1

BLOCK: 12 MODEL: MCOMPR (CONTINUED)

STAGE NUMBER	INDICATED HORSEPOWER HP	BRAKE HORSEPOWER HP
1	386.4	386.4
2	402.0	402.0
3	404.2	404.2

STAGE NUMBER	HEAD DEVELOPED FT-LBF/LB	VOLUMETRIC FLOW CUFT/HR	ISENTROPIC EFFICIENCY
1	0.1332E+06	0.4616E+05	0.7200
2	0.1385E+06	0.3559E+05	0.7200
3	0.1393E+06	0.2652E+05	0.7200

COOLER PROFILE

STAGE NUMBER	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	COOLING LOAD BTU/HR	VAPOR FRACTION
1	110.0	357.4	-.6934E+06	1.000
2	110.0	482.0	-.1015E+07	1.000
3	110.0	650.0	-.1018E+07	1.000

```
 BLOCK:  12       MODEL: MCOMPR (CONTINUED)
 BLOCK:  34,A     MODEL: VALVE
 -----------------------------
   INLET STREAM:          36,A1
   OUTLET STREAM:         36,A2
   PROPERTY OPTION SET:   REFPROP

***********************************************************************
     *
*
     *     ZERO FEED TO BLOCK
*
     *
*

***********************************************************************
                          ***  INPUT DATA  ***
    VALVE OUTLET PRESSURE     PSIA                            52.0000
    VALVE FLOW COEF CALC.                                    NO
```

```
 BLOCK:  34,A     MODEL: VALVE (CONTINUED)
                      FLASH SPECIFICATIONS:
   NPHASE                                                2
   MAX NUMBER OF ITERATIONS                             30
   CONVERGENCE TOLERANCE                                 0.100000-06
 BLOCK:  34,B     MODEL: PUMP
 ----------------------------
   INLET STREAM:          36,B1
   OUTLET STREAM:         36,B2
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                               IN               OUT        RELATIVE DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)          2356.23         2356.23         0.00000
      MASS(LB/HR   )          4749.88         4749.88         0.00000
      ENTHALPY(BTU/HR  )    -0.866231E+07   -0.875250E+07    0.103048E-01
                      ***  CO2 EQUIVALENT SUMMARY ***
   FEED STREAMS CO2E            0.00000      LB/HR
   PRODUCT STREAMS CO2E         0.00000      LB/HR
   NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
   UTILITIES CO2E PRODUCTION    0.00000      LB/HR
   TOTAL CO2E PRODUCTION        0.00000      LB/HR
                          ***  INPUT DATA  ***
   EQUIPMENT TYPE: TURBINE
   OUTLET PRESSURE  PSIA                                52.0000
   PUMP EFFICIENCY                                       0.70000
   DRIVER EFFICIENCY                                     1.00000
   FLASH SPECIFICATIONS:
   LIQUID PHASE CALCULATION
   NO FLASH PERFORMED
   MAXIMUM NUMBER OF ITERATIONS                          30
   TOLERANCE                                             0.100000-06
                           ***  RESULTS  ***
   VOLUMETRIC FLOW RATE  CUFT/HR                     1,184.14
   PRESSURE CHANGE  PSI                               -588.010
   NPSH AVAILABLE   FT-LBF/LB                      22,742.5
   FLUID POWER  HP                                     -50.6389
   BRAKE POWER  HP                                     -35.4472
   ELECTRICITY  KW                                     -26.4330
   PUMP EFFICIENCY USED                                  0.70000
   NET WORK REQUIRED  HP                               -35.4472
   HEAD DEVELOPED FT-LBF/LB                        -21,109.0
```

```
 BLOCK:  34SPLIT  MODEL: FSPLIT
 ------------------------------
   INLET STREAM:          32
   OUTLET STREAMS:        36,A1       36,B1
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                                IN                OUT         RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)            2356.23         2356.23          0.00000
       MASS(LB/HR   )            4749.88         4749.88          0.00000
       ENTHALPY(BTU/HR  )      -0.866231E+07   -0.866231E+07     0.00000
                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E            0.00000      LB/HR
    PRODUCT STREAMS CO2E         0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR
                          ***  INPUT DATA  ***
  FRACTION OF FLOW               STRM=36,B1    FRAC=         1.00000
                           ***  RESULTS  ***
  STREAM= 36,A1          SPLIT=          0.0          KEY=  0    STREAM-
ORDER=   2
          36,B1                          1.00000           0
1
 BLOCK:  38       MODEL: FLASH2
 ------------------------------
   INLET STREAMS:         36,A2       36,B2
   OUTLET VAPOR STREAM:   42
   OUTLET LIQUID STREAM:  40
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                                IN                OUT         RELATIVE
DIFF.
```

```
 BLOCK:  38       MODEL: FLASH2 (CONTINUED)
    TOTAL BALANCE
       MOLE(LBMOL/HR)          2356.23         2356.23         0.00000
       MASS(LB/HR   )          4749.88         4749.88         0.00000
       ENTHALPY(BTU/HR  )     -0.875250E+07   -0.875250E+07   -0.276657E-
14
                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E            0.00000      LB/HR
    PRODUCT STREAMS CO2E         0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
    UTILITIES CO2E PRODUCTION    0.00000      LB/HR
    TOTAL CO2E PRODUCTION        0.00000      LB/HR
                          ***  INPUT DATA  ***
   TWO    PHASE  PQ  FLASH
   PRESSURE DROP        PSI                              0.0
   SPECIFIED HEAT DUTY  BTU/HR                           0.0
   MAXIMUM NO. ITERATIONS                               30
   CONVERGENCE TOLERANCE                                 0.100000-06
                           ***  RESULTS  ***
   OUTLET TEMPERATURE    F                              -413.93
   OUTLET PRESSURE       PSIA                            52.000
   VAPOR FRACTION                                        0.12898
   V-L PHASE EQUILIBRIUM :
      COMP              F(I)          X(I)          Y(I)          K(I)
      PH2               1.0000        1.0000        1.0000
1.0000
 BLOCK:  48       MODEL: MCOMPR
 ------------------------------
   INLET STREAMS:    46          TO STAGE   1
   OUTLET STREAMS:   50        FROM STAGE   3
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                                IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          303.897         303.897         0.00000
       MASS(LB/HR   )          612.620         612.620         0.00000
       ENTHALPY(BTU/HR  )      55102.3         84741.2        -0.349758
```

BLOCK: 48 MODEL: MCOMPR (CONTINUED)

*** CO2 EQUIVALENT SUMMARY ***

FEED STREAMS CO2E	0.00000	LB/HR
PRODUCT STREAMS CO2E	0.00000	LB/HR
NET STREAMS CO2E PRODUCTION	0.00000	LB/HR
UTILITIES CO2E PRODUCTION	0.00000	LB/HR
TOTAL CO2E PRODUCTION	0.00000	LB/HR

*** INPUT DATA ***

ISENTROPIC CENTRIFUGAL COMPRESSOR

NUMBER OF STAGES	3
FINAL PRESSURE, PSIA	650.000
DISTRIBUTION AMONG STAGES	EQUAL P-RATIO

COMPRESSOR SPECIFICATIONS PER STAGE

STAGE NUMBER	MECHANICAL EFFICIENCY	ISENTROPIC EFFICIENCY
1	1.000	0.7200
2	1.000	0.7200
3	1.000	0.7200

COOLER SPECIFICATIONS PER STAGE

STAGE NUMBER	PRESSURE DROP PSI	TEMPERATURE F
1	0.000	110.0
2	0.000	110.0
3	0.000	110.0

*** RESULTS ***

FINAL PRESSURE, PSIA	650.000
TOTAL WORK REQUIRED, HP	542.472
TOTAL COOLING DUTY , BTU/HR	-1,350,650.

*** PROFILE ***

COMPRESSOR PROFILE

STAGE NUMBER	OUTLET PRESSURE PSIA	PRESSURE RATIO	OUTLET TEMPERATURE F
1	118.5	2.342	309.4
2	277.5	2.342	325.4
3	650.0	2.342	325.8

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING

U-O-S BLOCK SECTION

BLOCK: 48 MODEL: MCOMPR (CONTINUED)

STAGE NUMBER	INDICATED HORSEPOWER HP	BRAKE HORSEPOWER HP
1	177.4	177.4
2	181.7	181.7
3	183.4	183.4

STAGE NUMBER	HEAD DEVELOPED FT-LBF/LB	VOLUMETRIC FLOW CUFT/HR	ISENTROPIC EFFICIENCY
1	0.4128E+06	0.3608E+05	0.7200
2	0.4229E+06	0.1576E+05	0.7200
3	0.4267E+06	6768.	0.7200

COOLER PROFILE

STAGE NUMBER	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	COOLING LOAD BTU/HR	VAPOR FRACTION
1	110.0	118.5	-.4265E+06	1.000
2	110.0	277.5	-.4610E+06	1.000
3	110.0	650.0	-.4632E+06	1.000

```
 BLOCK:  48       MODEL: MCOMPR (CONTINUED)
 BLOCK:  76,1     MODEL: COMPR
 -----------------------------
   INLET STREAM:          74
   OUTLET STREAM:         78
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                                 IN              OUT         RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          5167.50         5167.50        0.00000
       MASS(LB/HR   )          104279.         104279.        0.00000
       ENTHALPY(BTU/HR  )    -0.116154E+08   -0.124824E+08    0.694602E-01
```

```
 BLOCK:  76,1     MODEL: COMPR (CONTINUED)
                    ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E             0.00000      LB/HR
    PRODUCT STREAMS CO2E          0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR
                         ***  INPUT DATA  ***
   ISENTROPIC TURBINE
    OUTLET PRESSURE  PSIA                                       27.4000
    ISENTROPIC EFFICIENCY                                        0.85000
    MECHANICAL EFFICIENCY                                        1.00000
                         ***  RESULTS  ***
    INDICATED  HORSEPOWER REQUIREMENT  HP                     -340.756
    BRAKE      HORSEPOWER REQUIREMENT  HP                     -340.756
    NET WORK REQUIRED                  HP                     -340.756
    POWER LOSSES                       HP                        0.0
    ISENTROPIC HORSEPOWER REQUIREMENT  HP                     -400.890
    CALCULATED OUTLET TEMP  F                                 -405.064
    ISENTROPIC TEMPERATURE  F                                 -406.866
    EFFICIENCY (POLYTR/ISENTR) USED                              0.85000
    OUTLET VAPOR FRACTION                                        1.00000
    HEAD DEVELOPED,       FT-LBF/LB                         -7,611.93
    MECHANICAL EFFICIENCY USED                                   1.00000
    INLET HEAT CAPACITY RATIO                                    1.87591
    INLET VOLUMETRIC FLOW RATE , CUFT/HR                    35,777.6
    OUTLET VOLUMETRIC FLOW RATE, CUFT/HR                   102,384.
    INLET  COMPRESSIBILITY FACTOR                                0.92295
    OUTLET COMPRESSIBILITY FACTOR                                0.92642
    AV. ISENT. VOL. EXPONENT                                     1.60435
    AV. ISENT. TEMP EXPONENT                                     1.55464
    AV. ACTUAL VOL. EXPONENT                                     1.49948
    AV. ACTUAL TEMP EXPONENT                                     1.50484
```

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
U-O-S BLOCK SECTION

```
 BLOCK:  84,1     MODEL: COMPR
 -----------------------------
   INLET STREAM:          82
   OUTLET STREAM:         86
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                                 IN                OUT        RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           4582.50         4582.50         0.00000
       MASS(LB/HR   )           92473.5         92473.5         0.00000
       ENTHALPY(BTU/HR  )     -0.877025E+07   -0.101759E+08    0.138135
                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E            0.00000      LB/HR
    PRODUCT STREAMS CO2E         0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
    UTILITIES CO2E PRODUCTION    0.00000      LB/HR
    TOTAL CO2E PRODUCTION        0.00000      LB/HR
                           ***  INPUT DATA  ***
   ISENTROPIC TURBINE
    OUTLET PRESSURE  PSIA                                      26.6100
    ISENTROPIC EFFICIENCY                                       0.85000
    MECHANICAL EFFICIENCY                                       1.00000
```

```
 BLOCK:  84,1     MODEL: COMPR (CONTINUED)
                          ***  RESULTS  ***
   INDICATED  HORSEPOWER REQUIREMENT  HP                  -552.440
   BRAKE      HORSEPOWER REQUIREMENT  HP                  -552.440
   NET WORK REQUIRED                  HP                  -552.440
   POWER LOSSES                       HP                     0.0
   ISENTROPIC HORSEPOWER REQUIREMENT  HP                  -649.929
   CALCULATED OUTLET TEMP  F                              -368.561
   ISENTROPIC TEMPERATURE  F                              -379.069
   EFFICIENCY (POLYTR/ISENTR) USED                           0.85000
   OUTLET VAPOR FRACTION                                     1.00000
   HEAD DEVELOPED,        FT-LBF/LB                    -13,916.0
   MECHANICAL EFFICIENCY USED                                1.00000
   INLET HEAT CAPACITY RATIO                                 1.72199
   INLET VOLUMETRIC FLOW RATE , CUFT/HR                  55,485.5
   OUTLET VOLUMETRIC FLOW RATE, CUFT/HR                 165,737.
   INLET  COMPRESSIBILITY FACTOR                             0.98852
   OUTLET COMPRESSIBILITY FACTOR                             0.98434
   AV. ISENT. VOL. EXPONENT                                  1.68741
   AV. ISENT. TEMP EXPONENT                                  1.66782
   AV. ACTUAL VOL. EXPONENT                                  1.48752
   AV. ACTUAL TEMP EXPONENT                                  1.48178
 BLOCK:  94       MODEL: MCOMPR
 ------------------------------
   INLET STREAMS:    92          TO STAGE   1
   OUTLET STREAMS:   T68       FROM STAGE   3
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                                   IN              OUT        RELATIVE
DIFF.
   TOTAL BALANCE
      MOLE(LBMOL/HR)           9750.00         9750.00         0.00000
      MASS(LB/HR   )           196752.         196752.         0.00000
      ENTHALPY(BTU/HR  )      0.105554E+07    0.162553E+07   -0.350648
                      ***  CO2 EQUIVALENT SUMMARY ***
   FEED STREAMS CO2E            0.00000      LB/HR
   PRODUCT STREAMS CO2E         0.00000      LB/HR
   NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
   UTILITIES CO2E PRODUCTION    0.00000      LB/HR
   TOTAL CO2E PRODUCTION        0.00000      LB/HR
```

BLOCK: 94 MODEL: MCOMPR (CONTINUED)

*** INPUT DATA ***

ISENTROPIC CENTRIFUGAL COMPRESSOR

NUMBER OF STAGES	3
FINAL PRESSURE, PSIA	150.000
DISTRIBUTION AMONG STAGES	EQUAL P-RATIO

COMPRESSOR SPECIFICATIONS PER STAGE

STAGE NUMBER	MECHANICAL EFFICIENCY	ISENTROPIC EFFICIENCY
1	1.000	0.7200
2	1.000	0.7200
3	1.000	0.7200

COOLER SPECIFICATIONS PER STAGE

STAGE NUMBER	PRESSURE DROP PSI	TEMPERATURE F
1	0.000	110.0
2	0.000	110.0
3	0.000	110.0

*** RESULTS ***

FINAL PRESSURE, PSIA	150.000
TOTAL WORK REQUIRED, HP	13,367.4
TOTAL COOLING DUTY , BTU/HR	-0.334424+08

*** PROFILE ***

COMPRESSOR PROFILE

STAGE NUMBER	OUTLET PRESSURE PSIA	PRESSURE RATIO	OUTLET TEMPERATURE F
1	40.80	1.917	329.4
2	78.23	1.917	345.4
3	150.0	1.917	345.4

STAGE NUMBER	INDICATED HORSEPOWER HP	BRAKE HORSEPOWER HP
1	4392.	4392.
2	4484.	4484.
3	4491.	4491.

BLOCK: 94 MODEL: MCOMPR (CONTINUED)

STAGE NUMBER	HEAD DEVELOPED FT-LBF/LB	VOLUMETRIC FLOW CUFT/HR	ISENTROPIC EFFICIENCY
1	0.3182E+05	0.2747E+07	0.7200
2	0.3249E+05	0.1463E+07	0.7200
3	0.3254E+05	0.7638E+06	0.7200

COOLER PROFILE

STAGE NUMBER	OUTLET TEMPERATURE F	OUTLET PRESSURE PSIA	COOLING LOAD BTU/HR	VAPOR FRACTION
1	110.0	40.80	-.1063E+08	1.000
2	110.0	78.23	-.1140E+08	1.000
3	110.0	150.0	-.1141E+08	1.000

```
 BLOCK:  MIX      MODEL: MIXER
 -----------------------------
   INLET STREAMS:         14          T50
   OUTLET STREAM:         16
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                                IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          2356.23         2356.23         0.00000
       MASS(LB/HR   )          4749.88         4749.88       0.191478E-
15
       ENTHALPY(BTU/HR  )      581497.         581497.         0.00000
                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E            0.00000      LB/HR
    PRODUCT STREAMS CO2E         0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
    UTILITIES CO2E PRODUCTION    0.00000      LB/HR
    TOTAL CO2E PRODUCTION        0.00000      LB/HR
                          ***  INPUT DATA  ***
   TWO    PHASE      FLASH
   MAXIMUM NO. ITERATIONS                                   30
   CONVERGENCE TOLERANCE                                     0.100000-06
   OUTLET PRESSURE   PSIA                                 650.000
 BLOCK:  MIX2     MODEL: MIXER
 -----------------------------
   INLET STREAMS:         80          86
   OUTLET STREAM:         88
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                                IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          9750.00         9750.00         0.00000
       MASS(LB/HR   )          196752.         196752.         0.00000
       ENTHALPY(BTU/HR  )    -0.217105E+08   -0.217105E+08     0.00000
```

```
 BLOCK:  MIX2     MODEL: MIXER (CONTINUED)
                     ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E             0.00000      LB/HR
    PRODUCT STREAMS CO2E          0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR
                         ***  INPUT DATA  ***
   TWO    PHASE      FLASH
   MAXIMUM NO. ITERATIONS                                        30
   CONVERGENCE TOLERANCE                                          0.100000-06
   OUTLET PRESSURE:  MINIMUM OF INLET STREAM PRESSURES
 BLOCK:  MIXN2    MODEL: MIXER
 -----------------------------
   INLET STREAMS:         54          T56
   OUTLET STREAM:         58
   PROPERTY OPTION SET:   REFPROP
                     ***  MASS AND ENERGY BALANCE  ***
                                IN               OUT        RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           1500.00         1500.00         0.00000
       MASS(LB/HR   )           42020.2         42020.2       -0.173154E-15
       ENTHALPY(BTU/HR  )      -0.485432E+07   -0.485432E+07    0.383709E-15
                     ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E             0.00000      LB/HR
    PRODUCT STREAMS CO2E          0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION   0.00000      LB/HR
    UTILITIES CO2E PRODUCTION     0.00000      LB/HR
    TOTAL CO2E PRODUCTION         0.00000      LB/HR
                         ***  INPUT DATA  ***
   TWO    PHASE      FLASH
   MAXIMUM NO. ITERATIONS                                        30
   CONVERGENCE TOLERANCE                                          0.100000-06
   OUTLET PRESSURE:  MINIMUM OF INLET STREAM PRESSURES
```

```
              EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
                           U-O-S BLOCK SECTION
 BLOCK:  SPLIT    MODEL: FSPLIT
 ------------------------------
   INLET STREAM:          70
   OUTLET STREAMS:        72          82
   PROPERTY OPTION SET:   REFPROP
                      ***  MASS AND ENERGY BALANCE  ***
                                IN               OUT        RELATIVE DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          9750.00         9750.00         0.00000
       MASS(LB/HR   )          196752.         196752.         0.00000
       ENTHALPY(BTU/HR  )    -0.186601E+08   -0.186601E+08     0.00000
                      ***  CO2 EQUIVALENT SUMMARY ***
    FEED STREAMS CO2E            0.00000      LB/HR
    PRODUCT STREAMS CO2E         0.00000      LB/HR
    NET STREAMS CO2E PRODUCTION  0.00000      LB/HR
    UTILITIES CO2E PRODUCTION    0.00000      LB/HR
    TOTAL CO2E PRODUCTION        0.00000      LB/HR
                          ***  INPUT DATA  ***
  FRACTION OF FLOW                STRM=82       FRAC=         0.47000
                           ***  RESULTS  ***
  STREAM= 72           SPLIT=          0.53000     KEY=  0    STREAM-ORDER=   2
          82                           0.47000           0
1
```

```
               EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
                                STREAM SECTION
 10 14 16 20L 24
 ---------------
 24,1 28,1 32 36,A1 36,A2
 ------------------------
 STREAM ID                10          14          16          20L         24
24,1       28,1       32         36,A1      36,A2
 FROM :                   ----        12          MIX         18          22
----       26         30         34SPLIT    34,A
 TO   :                   12          MIX         18          22          ----
26         30         34SPLIT    34,A       38
 SUBSTREAM: MIXED
 PHASE:                   VAPOR       VAPOR       VAPOR       VAPOR       VAPOR
VAPOR      VAPOR      LIQUID     MISSING    MISSING
 COMPONENTS: LBMOL/HR
   NH2                  2052.0000  2052.0000  2356.2300  2356.2300
2356.2300     0.0        0.0        0.0        0.0        0.0
   PH2                     0.0        0.0        0.0        0.0        0.0
2356.2300  2356.2300  2356.2300     0.0        0.0
   HE                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   N2                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   NE                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
 COMPONENTS: MOLE FRAC
   NH2                     1.0000     1.0000     1.0000     1.0000
1.0000     0.0        0.0        0.0        0.0        0.0
   PH2                     0.0        0.0        0.0        0.0        0.0
1.0000     1.0000     1.0000     0.0        0.0
   HE                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   N2                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   NE                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
 TOTAL FLOW:
   LBMOL/HR             2052.0000  2052.0000  2356.2300  2356.2300
2356.2300  2356.2300  2356.2300  2356.2300     0.0        0.0
   LB/HR                4136.5858  4136.5858  4749.8769  4749.8769
4749.8769  4749.8769  4749.8769  4749.8769     0.0        0.0
   CUFT/HR              4.6161+04  1.9792+04  2.2727+04  6646.0883
5983.6255  5981.3118  2732.9350  1184.1382     0.0        0.0
 STATE VARIABLES:
   TEMP   F               90.0000   110.0000   110.0000  -290.0000  -
305.0000  -305.0000  -367.0000  -404.0000    MISSING    MISSING
   PRES   PSIA           265.0000   650.0000   650.0000   643.0000
640.7000   640.7000   640.1200   640.0100    MISSING    MISSING
   VFRAC                   1.0000     1.0000     1.0000     1.0000
1.0000     1.0000     1.0000     0.0       MISSING    MISSING
```

```
   LFRAC                  0.0        0.0        0.0        0.0        0.0
0.0        0.0        1.0000    MISSING    MISSING
   SFRAC                  0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0       MISSING    MISSING
 ENTHALPY:
   BTU/LBMOL             96.7912   246.7914   246.7914 -2388.9665 -
2480.0449 -2869.9237 -3323.7867 -3676.3429    MISSING    MISSING
   BTU/LB                48.0144   122.4236   122.4236 -1185.0738 -
1230.2542 -1423.6580 -1648.8019 -1823.6913    MISSING    MISSING
   BTU/HR              1.9862+05  5.0642+05  5.8150+05 -5.6290+06 -
5.8436+06 -6.7622+06 -7.8316+06 -8.6623+06    MISSING    MISSING
 ENTROPY:
   BTU/LBMOL-R           -5.5864    -7.1305    -7.1305   -14.9551   -
15.5102   -19.7620   -23.5627   -28.3217    MISSING    MISSING
   BTU/LB-R              -2.7712    -3.5372    -3.5372    -7.4187    -
7.6940    -9.8032   -11.6885   -14.0493    MISSING    MISSING
 DENSITY:
   LBMOL/CUFT          4.4453-02     0.1037     0.1037     0.3545
0.3938     0.3939     0.8622     1.9898    MISSING    MISSING
   LB/CUFT             8.9612-02     0.2090     0.2090     0.7147
0.7938     0.7941     1.7380     4.0113    MISSING    MISSING
 AVG MW                   2.0159     2.0159     2.0159     2.0159
2.0159     2.0159     2.0159     2.0159    MISSING    MISSING
 MIXED    SUBSTREAM PROPERTIES:
 ***  ALL PHASES   ***
 VMX      CUFT/MIN      769.3513   329.8737   378.7808   110.7681
99.7271    99.6885    45.5489    19.7356    MISSING    MISSING
 CPCVMX                   1.4080     1.4094     1.4094     1.7813
1.8561     1.7110     2.9355     2.1077    MISSING    MISSING
 TEMP     C              32.2222    43.3333    43.3333  -178.8889  -
187.2222  -187.2222  -221.6667  -242.2222    MISSING    MISSING
 PRES     BAR            18.2711    44.8159    44.8159    44.3333
44.1747    44.1747    44.1347    44.1271    MISSING    MISSING
 MASSFLMX KG/HR        1876.3237  1876.3237  2154.5079  2154.5079
2154.5079  2154.5079  2154.5079  2154.5079    MISSING    MISSING
 *** VAPOR PHASE   ***
```

10 14 16 20L 24
24,1 28,1 32 36,A1 36,A2 (CONTINUED)

STREAM ID		10	14	16	20L	24	24,1	28,1	32	36,A1	36,A2
MUMX	LB/FT-HR	2.1916-02	2.2516-02	2.2516-02	1.0445-02	9.9667-03	9.9673-03	9.3391-03	MISSING	MISSING	MISSING
RHOMX	LB/CUFT	8.9612-02	0.2090	0.2090	0.7147	0.7938	0.7941	1.7380	MISSING	MISSING	MISSING
*** LIQUID PHASE ***											
MUMX	LB/FT-HR	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	2.2316-02	MISSING	MISSING
RHOMX	LB/CUFT	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	4.0113	MISSING	MISSING

EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING

STREAM SECTION

36,B1 36,B2 40 42 46

46,+R 46,-2 50 54 58

STREAM ID	36,B1	36,B2	40	42	46	46,+R	46,-2	50	54	58
FROM :	34SPLIT	34,B	38	38	18	26	30	48	22	MIXN2
TO :	34,B	38	----	30	48	18	26	----	MIXN2	18
SUBSTREAM: MIXED										
PHASE:	LIQUID	LIQUID	LIQUID	VAPOR	VAPOR	VAPOR	VAPOR	VAPOR	MIXED	MIXED
COMPONENTS: LBMOL/HR										
NH2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PH2	2356.2300	2356.2300	2052.3331	303.8969	303.8969	303.8969	303.8969	303.8969	0.0	0.0
HE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	696.0000	1500.0000
NE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
COMPONENTS: MOLE FRAC										
NH2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PH2	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0	0.0
HE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.0000	1.0000
NE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL FLOW:										
LBMOL/HR	2356.2300	2356.2300	2052.3331	303.8969	303.8969	303.8969	303.8969	303.8969	696.0000	1500.0000
LB/HR	4749.8769	4749.8769	4137.2572	612.6198	612.6198	612.6198	612.6198	612.6198	1.9497+04	4.2020+04
CUFT/HR	1184.1382	1344.8425	1038.4020	2245.1674	3.6083+04	9606.2010	5439.0909	2931.1843	3.1678+04	1.0364+05
STATE VARIABLES:										
TEMP F	-404.0000	-408.5663	-413.9311	-413.9311	98.7281	-308.7511	-370.8619	110.0000	-316.4972	-318.1681
PRES PSIA	640.0100	52.0000	52.0000	52.0000	50.5700	51.0400	51.6400	650.0000	18.8700	17.0000
VFRAC	0.0	0.0	0.0	1.0000	1.0000	1.0000	1.0000	1.0000	0.5863	0.8111

```
   LFRAC                  1.0000     1.0000     1.0000     0.0        0.0
0.0        0.0        0.0        0.4137     0.1889
   SFRAC                  0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
 ENTHALPY:
   BTU/LBMOL          -3676.3429 -3714.6215 -3759.2218 -3413.4183
181.3191 -2827.3448 -3165.5588   278.8485 -3756.7813 -3236.2125
   BTU/LB             -1823.6913 -1842.6799 -1864.8044 -1693.2646
89.9454 -1402.5363 -1570.3111   138.3259  -134.1062  -115.5234
   BTU/HR             -8.6623+06 -8.7525+06 -7.7152+06 -1.0373+06
5.5102+04 -8.5922+05 -9.6200+05  8.4741+04 -2.6147+06 -4.8543+06
 ENTROPY:
   BTU/LBMOL-R          -28.3217   -27.9575   -28.8756   -21.3152    -
4.9619   -14.5223   -17.4193    -9.9070   -16.7144   -12.9189
   BTU/LB-R             -14.0493   -13.8686   -14.3241   -10.5736    -
2.4614    -7.2039    -8.6410    -4.9145    -0.5967    -0.4612
 DENSITY:
   LBMOL/CUFT             1.9898     1.7520     1.9764     0.1354  8.4221-
03  3.1635-02  5.5873-02     0.1037  2.1971-02  1.4473-02
   LB/CUFT                4.0113     3.5319     3.9843     0.2729  1.6978-
02  6.3773-02     0.1126     0.2090     0.6155     0.4054
 AVG MW                   2.0159     2.0159     2.0159     2.0159
2.0159     2.0159     2.0159     2.0159    28.0135    28.0135
 MIXED    SUBSTREAM PROPERTIES:
 ***  ALL PHASES   ***
 VMX      CUFT/MIN       19.7356    22.4140    17.3067    37.4195
601.3896   160.1034    90.6515    48.8531   527.9597  1727.3706
 CPCVMX                   2.1077     3.5892     2.3131     2.3497
1.3878     1.5411     1.7380     1.3938     1.6899     1.5713
 TEMP     C            -242.2222  -244.7591  -247.7395  -247.7395
37.0712  -189.3062  -223.8122    43.3333  -193.6096  -194.5378
 PRES     BAR            44.1271     3.5853     3.5853     3.5853
3.4867     3.5191     3.5605    44.8159     1.3010     1.1721
 MASSFLMX KG/HR        2154.5079  2154.5079  1876.6283   277.8797
277.8797   277.8797   277.8797   277.8797  8843.8637  1.9060+04
 *** VAPOR PHASE   ***
```

36,B1 36,B2 40 42 46
46,+R 46,-2 50 54 58 (CONTINUED)

STREAM ID		36,B1	36,B2	40	42	46	46,+R	46,-2	50	54	58
MUMX	LB/FT-HR	MISSING	MISSING	MISSING	3.1536-03	2.2133-02	8.7867-03	5.8310-03	2.2516-02	1.3584-02	1.3407-02
RHOMX	LB/CUFT	MISSING	MISSING	MISSING	0.2729	1.6978-02	6.3773-02	0.1126	0.2090	0.3627	0.3294
*** LIQUID PHASE ***											
MUMX	LB/FT-HR	2.2316-02	1.6434-02	2.1971-02	MISSING	MISSING	MISSING	MISSING	MISSING	0.3572	0.3700
RHOMX	LB/CUFT	4.0113	3.5319	3.9843	MISSING	MISSING	MISSING	MISSING	MISSING	49.6969	49.9640

```
               EQUIP DESIGN HMB - 15 TPD - NO TRAILER LOADING
                                STREAM SECTION
 60 68 68,1 70 72
 ----------------
 74 78 80 82 86
 --------------
 STREAM ID                60         68         68,1       70         72
74         78         80         82         86
 FROM :                   18         ----       18         22         SPLIT
26         76,1       30         SPLIT      84,1
 TO   :                   ----       18         22         SPLIT      26
76,1       30         MIX2       84,1       MIX2
 SUBSTREAM: MIXED
 PHASE:                   VAPOR      VAPOR      VAPOR      VAPOR      VAPOR
VAPOR      VAPOR      VAPOR      VAPOR      VAPOR
 COMPONENTS: LBMOL/HR
   NH2                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   PH2                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   HE                       0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   N2                    1500.0000     0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   NE                       0.0     9750.0000  9750.0000  9750.0000
5167.5000  5167.5000  5167.5000  5167.5000  4582.5000  4582.5000
 COMPONENTS: MOLE FRAC
   NH2                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   PH2                      0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   HE                       0.0        0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   N2                       1.0000     0.0        0.0        0.0        0.0
0.0        0.0        0.0        0.0        0.0
   NE                       0.0        1.0000     1.0000     1.0000
1.0000     1.0000     1.0000     1.0000     1.0000     1.0000
 TOTAL FLOW:
   LBMOL/HR              1500.0000  9750.0000  9750.0000  9750.0000
5167.5000  5167.5000  5167.5000  5167.5000  4582.5000  4582.5000
   LB/HR                 4.2020+04  1.9675+05  1.9675+05  1.9675+05
1.0428+05  1.0428+05  1.0428+05  1.0428+05  9.2473+04  9.2473+04
   CUFT/HR               5.7725+05  3.9923+05  1.2254+05  1.1805+05
6.2569+04  3.5778+04  1.0238+05  1.8724+05  5.5486+04  1.6574+05
 STATE VARIABLES:
   TEMP   F                98.7281   110.0000  -290.0000  -305.0000  -
305.0000  -367.0000  -405.0641  -370.8619  -305.0000  -368.5612
   PRES   PSIA             15.5700   150.0000   143.8100   135.5100
135.5100   132.5700    27.4000    25.8700   135.5100    26.6100
   VFRAC                    1.0000     1.0000     1.0000     1.0000
1.0000     1.0000     1.0000     1.0000     1.0000     1.0000
```

```
   LFRAC                   0.0         0.0         0.0         0.0         0.0
0.0         0.0         0.0         0.0         0.0
   SFRAC                   0.0         0.0         0.0         0.0         0.0
0.0         0.0         0.0         0.0         0.0
 ENTHALPY:
   BTU/LBMOL             148.5153   166.7215 -1837.4151 -1913.8569 -
1913.8569 -2247.7764 -2415.5619 -2232.1455 -1913.8569 -2220.5991
   BTU/LB                  5.3016     8.2618   -91.0526   -94.8407   -
94.8407  -111.3880  -119.7026  -110.6134   -94.8407  -110.0412
   BTU/HR                2.2277+05  1.6255+06 -1.7915+07 -1.8660+07 -
9.8899+06 -1.1615+07 -1.2482+07 -1.1535+07 -8.7702+06 -1.0176+07
 ENTROPY:
   BTU/LBMOL-R             0.1569    -4.3213   -10.3223   -10.6770   -
10.6770   -13.4072   -12.8498   -10.1189   -10.6770   -10.0457
   BTU/LB-R              5.5998-03   -0.2141    -0.5115    -0.5291    -
0.5291    -0.6644    -0.6368    -0.5014    -0.5291    -0.4978
 DENSITY:
   LBMOL/CUFT            2.5985-03  2.4422-02  7.9564-02  8.2589-02  8.2589-
02     0.1444  5.0472-02  2.7599-02  8.2589-02  2.7649-02
   LB/CUFT               7.2794-02     0.4928     1.6056     1.6666
1.6666     2.9146     1.0185     0.5569     1.6666     0.5580
 AVG MW                   28.0135    20.1797    20.1797    20.1797
20.1797    20.1797    20.1797    20.1797    20.1797    20.1797
 MIXED    SUBSTREAM PROPERTIES:
 ***  ALL PHASES   ***
 VMX      CUFT/MIN     9620.8635  6653.8733  2042.3845  1967.5726
1042.8135   596.2936  1706.4021  3120.5970   924.7591  2762.2764
 CPCVMX                    1.4011     1.6688     1.7126     1.7209
1.7209     1.8750     1.8172     1.7045     1.7209     1.7033
 TEMP     C               37.0712    43.3333  -178.8889  -187.2222  -
187.2222  -221.6667  -242.8134  -223.8122  -187.2222  -222.5340
 PRES     BAR              1.0735    10.3421     9.9154     9.3431
9.3431     9.1404     1.8892     1.7837     9.3431     1.8347
 MASSFLMX KG/HR          1.9060+04  8.9245+04  8.9245+04  8.9245+04
4.7300+04  4.7300+04  4.7300+04  4.7300+04  4.1945+04  4.1945+04
 *** VAPOR PHASE   ***
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60 68 68,1 70 72
74 78 80 82 86 (CONTINUED)

STREAM ID		60	68	68,1	70	72	74	78	80	82	86
MUMX	LB/FT-HR	4.4411-02	7.9913-02	3.3199-02	3.0752-02	3.0752-02	1.9827-02	1.1324-02	1.8177-02	3.0752-02	1.8642-02
RHOMX	LB/CUFT	7.2794-02	0.4928	1.6056	1.6666	1.6666	2.9146	1.0185	0.5569	1.6666	0.5580
*** LIQUID PHASE ***											
MUMX	LB/FT-HR	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING
RHOMX	LB/CUFT	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING	MISSING

88 90 92 T50 T52

T56 T68

STREAM ID	88	90	92	T50	T52	T56	T68
FROM :	MIX2	26	18	----	----	----	94
TO :	26	18	94	MIX	22	MIXN2	----
SUBSTREAM: MIXED							
PHASE:	VAPOR	VAPOR	VAPOR	VAPOR	LIQUID	VAPOR	VAPOR
COMPONENTS: LBMOL/HR							
NH2	0.0	0.0	0.0	304.2300	0.0	0.0	0.0
PH2	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HE	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N2	0.0	0.0	0.0	0.0	696.0000	804.0000	0.0
NE	9750.0000	9750.0000	9750.0000	0.0	0.0	0.0	9750.0000
COMPONENTS: MOLE FRAC							
NH2	0.0	0.0	0.0	1.0000	0.0	0.0	0.0
PH2	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HE	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N2	0.0	0.0	0.0	0.0	1.0000	1.0000	0.0
NE	1.0000	1.0000	1.0000	0.0	0.0	0.0	1.0000
TOTAL FLOW:							
LBMOL/HR	9750.0000	9750.0000	9750.0000	304.2300	696.0000	804.0000	9750.0000
LB/HR	1.9675+05	1.9675+05	1.9675+05	613.2912	1.9497+04	2.2523+04	1.9675+05
CUFT/HR	3.5770+05	6.5834+05	2.7475+06	2934.4288	392.9524	6.8478+04	3.9923+05
STATE VARIABLES:							
TEMP F	-369.8032	-308.7511	98.7281	110.0000	-316.0000	-318.0000	110.0000
PRES PSIA	25.8700	23.9300	21.2800	650.0000	20.0000	17.0000	150.0000
VFRAC	1.0000	1.0000	1.0000	1.0000	0.0	1.0000	1.0000

```
   LFRAC                  0.0        0.0        0.0        0.0
1.0000     0.0        0.0
   SFRAC                  0.0        0.0        0.0        0.0        0.0
0.0        0.0
 ENTHALPY:
   BTU/LBMOL          -2226.7187 -1918.3407   108.2610   246.7914 -
5135.9614 -2785.5709   166.7215
   BTU/LB              -110.3445   -95.0629     5.3648   122.4236  -
183.3389   -99.4368     8.2618
   BTU/HR             -2.1711+07 -1.8704+07  1.0555+06  7.5081+04 -
3.5746+06 -2.2396+06  1.6255+06
 ENTROPY:
   BTU/LBMOL-R          -10.0582    -7.2832    -0.5388    -7.1305   -
26.3483    -9.7342    -4.3213
   BTU/LB-R              -0.4984    -0.3609 -2.6698-02    -3.5372    -
0.9406    -0.3475    -0.2141
 DENSITY:
   LBMOL/CUFT          2.7258-02  1.4810-02  3.5487-03     0.1037
1.7712  1.1741-02  2.4422-02
   LB/CUFT                0.5501     0.2989  7.1612-02     0.2090
49.6177     0.3289     0.4928
 AVG MW                  20.1797    20.1797    20.1797     2.0159
28.0135    28.0135    20.1797
 MIXED    SUBSTREAM PROPERTIES:
 ***  ALL PHASES   ***
 VMX      CUFT/MIN     5961.6228  1.0972+04  4.5791+04    48.9071
6.5492  1141.2973  6653.8729
 CPCVMX                   1.7035     1.6766     1.6670     1.4094
1.9198     1.4640     1.6688
 TEMP     C            -223.2240  -189.3062    37.0712    43.3333  -
193.3333  -194.4444    43.3333
 PRES     BAR             1.7837     1.6499     1.4672    44.8159
1.3790     1.1721    10.3421
 MASSFLMX KG/HR        8.9245+04  8.9245+04  8.9245+04   278.1842
8843.8637  1.0216+04  8.9245+04
 *** VAPOR PHASE   ***
```

```
 88 90 92 T50 T52
 T56 T68 (CONTINUED)
 STREAM ID              88         90         92         T50        T52
T56        T68
 MUMX     LB/FT-HR     1.8389-02  2.9685-02  7.8775-02  2.2516-02
MISSING  1.3422-02  7.9913-02
 RHOMX    LB/CUFT         0.5501     0.2989  7.1612-02     0.2090
MISSING     0.3289     0.4928
 *** LIQUID PHASE  ***
 MUMX     LB/FT-HR       MISSING    MISSING    MISSING    MISSING
0.3535    MISSING    MISSING
 RHOMX    LB/CUFT        MISSING    MISSING    MISSING    MISSING
49.6177    MISSING    MISSING
```

```
 BLOCK STATUS
 ------------

***************************************************************************
**
 *
*
 * Calculations were completed with warnings
*
 *
*
 * The following Unit Operation blocks were
*
 * completed with warnings:
*
 *   34,A
*
 *
*
 * All Transfer blocks were completed normally
*
 *
*
 * All streams were flashed normally
*
 *
*
 * All Convergence blocks were completed normally
*
 *
*
 * All Calculator blocks were completed normally
*
 *
*

***************************************************************************
**
```

24.13 Ionex® (Hydrous Ferric Oxide Catalyst) Data Sheet

TECHNICAL DATASHEET

Ionex® – Type O-P Catalyst

Hydrous Ferric Oxide

A specially manufactured catalyst used in hydrogen liquefaction and storage systems.

molecular

Description

Ionex® –Type O-P Catalyst is a specially manufactured catalyst used in hydrogen liquefaction and storage systems. This material efficiently catalyzes the conversion of ortho hydrogen to para hydrogen. Due to its large surface area and particle size uniformity, rapid conversion and uniform flow characteristics are assured.

Typical properties		Test method
Chemical formula	Fe_2O_3	
Apparent density	1.32 ± .07 g/cc	ASTM D2854
Moisture content, as packed	<2%	ASTM D2867
Space Velocity	1200 cc H_2/min./cc of ferric oxide	

Standard mesh size (US Sieve)	Test method
30 x 50	ASTM D2862

Particle size distribution	
Oversize maximum	5%
Nominal mesh size	90% minimum
Undersize maximum	5%

Packaging

Packaging is dependent on quantity. Orders in excess of 200 lbs. are generally packaged in 30- to 55-gallon steel drums under an inert atmosphere. Product is shipped in a highly active state and must remain in its original packaging until its intended use.

Activity criteria

The catalyst will achieve 46.5% or higher of para hydrogen after 16 hours of reactivation @ 160 C with dry hydrogen flow, when fed an equilibrium mixture of 75% ortho and 25% para hydrogen @1.36 atmospheres and 77 K, and a flow rate of 1200cc (STP) per minutes per cc of catalyst. Contact Molecular Products for typical performance data.

Note: This spec sheet indicates physical properties that are standard and typical. Molecular Products Inc. will meet specifications as required.

Molecular Products Inc.

6837 Winchester Circle
Boulder, CO 80301

T (303) 666 4400
F (303) 665 0563

E infousa@molprod.com
W www.molecularproducts.com

Doc. No. 166 Rev. C

24.14 Safety Data Sheets

24.14.1 Water

LabChem
performance through chemistry

Water
Safety Data Sheet
according to Federal Register / Vol. 77, No. 58 / Monday, March 26, 2012 / Rules and Regulations
Issue date: 11/15/2013 Revision date: 06/26/2020 Supersedes: 06/12/2018 Version: 1.4

SECTION 1: Identification

1.1. Identification

Product form : Substance
Substance name : Water
CAS-No. : 7732-18-5
Product code : LC26750
Formula : H2O

1.2. Recommended use and restrictions on use

Use of the substance/mixture : For laboratory and manufacturing use only.
Recommended use : Laboratory chemicals
Restrictions on use : Not for food, drug or household use

1.3. Supplier

LabChem, Inc.
1010 Jackson's Pointe Ct.
Zelienople, PA 16063 - USA
T 412-826-5230 - F 724-473-0647
info@labchem.com - www.labchem.com

1.4. Emergency telephone number

Emergency number : CHEMTREC: 1-800-424-9300 or +1-703-741-5970

SECTION 2: Hazard(s) identification

2.1. Classification of the substance or mixture

GHS US classification
Not classified

2.2. GHS Label elements, including precautionary statements

Not classified as a hazardous chemical.
Other hazards not contributing to the classification : None.

2.4. Unknown acute toxicity (GHS US)

Not applicable

SECTION 3: Composition/Information on ingredients

3.1. Substances

Substance type : Mono-constituent

Name	Product identifier	%	GHS US classification
Water (Main constituent)	(CAS-No.) 7732-18-5	100	Not classified

Full text of hazard classes and H-statements : see section 16

3.2. Mixtures

Not applicable

SECTION 4: First-aid measures

4.1. Description of first aid measures

First-aid measures general : If you feel unwell, seek medical advice (show the label where possible).
First-aid measures after inhalation : Allow affected person to breathe fresh air. Allow the victim to rest. Adverse effects not expected from this product.
First-aid measures after skin contact : Adverse effects not expected from this product. Take off contaminated clothing.
First-aid measures after eye contact : Adverse effects not expected from this product.
First-aid measures after ingestion : Do NOT induce vomiting. Adverse effects not expected from this product.

4.2. Most important symptoms and effects (acute and delayed)

Potential Adverse human health effects and symptoms : Based on available data, the classification criteria are not met.

Symptoms/effects : Not expected to present a significant hazard under anticipated conditions of normal use.

4.3. Immediate medical attention and special treatment, if necessary

Treat symptomatically.

SECTION 5: Fire-fighting measures

5.1. Suitable (and unsuitable) extinguishing media

Suitable extinguishing media : Foam. Dry powder. Carbon dioxide. Water spray. Sand.

5.2. Specific hazards arising from the chemical

Fire hazard : Not flammable.

5.3. Special protective equipment and precautions for fire-fighters

Firefighting instructions : Use water spray or fog for cooling exposed containers. Exercise caution when fighting any chemical fire.

Protection during firefighting : Do not enter fire area without proper protective equipment, including respiratory protection.

SECTION 6: Accidental release measures

6.1. Personal precautions, protective equipment and emergency procedures

6.1.1. For non-emergency personnel

Emergency procedures : Evacuate unnecessary personnel.

6.1.2. For emergency responders

Protective equipment : Equip cleanup crew with proper protection.

Emergency procedures : Ventilate area.

6.2. Environmental precautions

Prevent entry to sewers and public waters. Notify authorities if liquid enters sewers or public waters.

6.3. Methods and material for containment and cleaning up

Methods for cleaning up : Soak up spills with inert solids, such as clay or diatomaceous earth as soon as possible.

6.4. Reference to other sections

See Heading 8. Exposure controls and personal protection.

SECTION 7: Handling and storage

7.1. Precautions for safe handling

Precautions for safe handling : Wash hands and other exposed areas with mild soap and water before eating, drinking or smoking and when leaving work.

7.2. Conditions for safe storage, including any incompatibilities

Storage conditions : Keep container closed when not in use.

Incompatible products : Metallic sodium.

Incompatible materials : Sources of ignition. Direct sunlight.

SECTION 8: Exposure controls/personal protection

8.1. Control parameters

Water (7732-18-5)
No additional information available

8.2. Appropriate engineering controls

Appropriate engineering controls : Provide adequate general and local exhaust ventilation.

8.3. Individual protection measures/Personal protective equipment

Personal protective equipment:

Safety glasses.

Eye protection:

Chemical goggles or safety glasses

Respiratory protection:

None necessary.

Personal protective equipment symbol(s):

Other information:

Do not eat, drink or smoke during use.

SECTION 9: Physical and chemical properties

9.1. Information on basic physical and chemical properties

Physical state	: Liquid
Color	: Colorless
Odor	: None.
Odor threshold	: No data available
pH	: 7
Melting point	: 0 °C
Freezing point	: No data available
Boiling point	: 100 °C
Critical temperature	: 374.1 °C
Critical pressure	: 218.3 atm
Flash point	: No data available
Relative evaporation rate (butyl acetate=1)	: No data available
Flammability (solid, gas)	: Non flammable.
Vapor pressure	: 17.535 mm Hg
Vapor pressure at 50 °C	: 92.51 mm Hg
Relative vapor density at 20 °C	: No data available
Relative density	: 1
Specific gravity / density	: 0.99823 g/ml
Molecular mass	: 18 g/mol
Solubility	: Soluble in acetic acid. Soluble in acetone. Soluble in ammonia. Soluble in ammonium chloride. Soluble in ethanol. Soluble in glycerol. Soluble in hydrochloric acid. Soluble in methanol. Soluble in nitric acid. Soluble in sulfuric acid. Soluble in sodium hydroxide solution. Soluble in propylene glycol.
Log Pow	: No data available
Auto-ignition temperature	: No data available
Decomposition temperature	: No data available
Viscosity, kinematic	: 1.004 mm^2/s
Viscosity, dynamic	: 1.002 cP
Explosion limits	: No data available
Explosive properties	: Not applicable.
Oxidizing properties	: None.

9.2. Other information

VOC content	: 0 %

SECTION 10: Stability and reactivity

10.1. Reactivity

No additional information available

10.2. Chemical stability

Stable under normal conditions.

10.3. Possibility of hazardous reactions

Not established.

10.4. Conditions to avoid

Extremely high or low temperatures.

10.5. Incompatible materials

Metallic sodium.

10.6. Hazardous decomposition products

Hydrogen. oxygen.

SECTION 11: Toxicological information

11.1. Information on toxicological effects

Acute toxicity (oral) : Not classified
Acute toxicity (dermal) : Not classified
Acute toxicity (inhalation) : Not classified

Water (7732-18-5)	
LD50 oral rat	≥ 90000 mg/kg
ATE US (oral)	90000 mg/kg body weight

Skin corrosion/irritation : Not classified
pH: 7
Serious eye damage/irritation : Not classified
pH: 7
Respiratory or skin sensitization : Not classified
Germ cell mutagenicity : Not classified
Carcinogenicity : Not classified (Based on available data, the classification criteria are not met)
Reproductive toxicity : Not classified
STOT-single exposure : Not classified
STOT-repeated exposure : Not classified
Aspiration hazard : Not classified
Viscosity, kinematic : 1.004 mm^2/s
Likely routes of exposure : Skin and eye contact.
Potential Adverse human health effects and symptoms : Based on available data, the classification criteria are not met.
Symptoms/effects : Not expected to present a significant hazard under anticipated conditions of normal use.

SECTION 12: Ecological information

12.1. Toxicity

No additional information available

12.2. Persistence and degradability

Water (7732-18-5)	
Persistence and degradability	Not established.

12.3. Bioaccumulative potential

Water (7732-18-5)	
Bioaccumulative potential	Not established.

12.4. Mobility in soil

No additional information available

12.5. Other adverse effects

Other information : No other effects known.

SECTION 13: Disposal considerations

13.1. Disposal methods

Waste disposal recommendations : Dispose in a safe manner in accordance with local/national regulations.

SECTION 14: Transport information

Department of Transportation (DOT)

In accordance with DOT

Not regulated

Transport by sea

Not regulated

Air transport

Not regulated

SECTION 15: Regulatory information

15.1. US Federal regulations

Water (7732-18-5)
Listed on the United States TSCA (Toxic Substances Control Act) inventory

All components of this product are listed, or excluded from listing, on the United States Environmental Protection Agency Toxic Substances Control Act (TSCA) inventory

15.2. International regulations

CANADA

Water (7732-18-5)
Listed on the Canadian DSL (Domestic Substances List)

EU-Regulations

No additional information available

National regulations

No additional information available

15.3. US State regulations

California Proposition 65 - This product does not contain any substances known to the state of California to cause cancer, developmental and/or reproductive harm

SECTION 16: Other information

according to Federal Register / Vol. 77, No. 58 / Monday, March 26, 2012 / Rules and Regulations

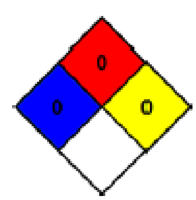
Revision date : 06/26/2020

Other information : None.

NFPA health hazard : 0 - Materials that, under emergency conditions, would offer no hazard beyond that of ordinary combustible materials.

NFPA fire hazard : 0 - Materials that will not burn under typical fire conditions, including intrinsically noncombustible materials such as concrete, stone, and sand.

NFPA reactivity : 0 - Material that in themselves are normally stable, even under fire conditions.



Hazard Rating

Health	:	0 Minimal Hazard - No significant risk to health
Flammability	:	0 Minimal Hazard - Materials that will not burn
Physical	:	0 Minimal Hazard - Materials that are normally stable, even under fire conditions, and will NOT react with water, polymerize, decompose, condense, or self-react. Non-Explosives.
Personal protection	:	A A - Safety glasses

SDS US LabChem

24.14.2 Hydrogen

SAFETY DATA SHEET

Hydrogen

Airgas
an Air Liquide company

Section 1. Identification

GHS product identifier : Hydrogen

Chemical name : hydrogen

Other means of identification : Dihydrogen; o-Hydrogen; p-Hydrogen; Molecular hydrogen; H2; UN 1049

Product type : Gas.

Product use : Synthetic/Analytical chemistry.

Synonym : Dihydrogen; o-Hydrogen; p-Hydrogen; Molecular hydrogen; H2; UN 1049

SDS # : 001026

Supplier's details : Airgas USA, LLC and its affiliates
259 North Radnor-Chester Road
Suite 100
Radnor, PA 19087-5283
1-610-687-5253

24-hour telephone : 1-866-734-3438

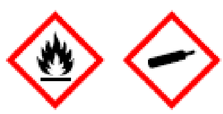
Section 2. Hazards identification

OSHA/HCS status : This material is considered hazardous by the OSHA Hazard Communication Standard (29 CFR 1910.1200).

Classification of the substance or mixture : FLAMMABLE GASES - Category 1
GASES UNDER PRESSURE - Compressed gas

GHS label elements

Hazard pictograms :



Signal word : Danger

Hazard statements : Extremely flammable gas.
Contains gas under pressure; may explode if heated.
May displace oxygen and cause rapid suffocation.
Burns with invisible flame.
May form explosive mixtures with air.

Precautionary statements

General : Read and follow all Safety Data Sheets (SDS'S) before use. Read label before use. Keep out of reach of children. If medical advice is needed, have product container or label at hand. Close valve after each use and when empty. Use equipment rated for cylinder pressure. Do not open valve until connected to equipment prepared for use. Use a back flow preventative device in the piping. Use only equipment of compatible materials of construction. Approach suspected leak area with caution.

Prevention : Keep away from heat, hot surfaces, sparks, open flames and other ignition sources. No smoking.

Response : Leaking gas fire: Do not extinguish, unless leak can be stopped safely. In case of leakage, eliminate all ignition sources.

Storage : Protect from sunlight. Store in a well-ventilated place.

Disposal : Not applicable.

Hazards not otherwise classified : In addition to any other important health or physical hazards, this product may displace oxygen and cause rapid suffocation.

Section 3. Composition/information on ingredients

Substance/mixture : Substance

Chemical name : hydrogen

Other means of identification : Dihydrogen; o-Hydrogen; p-Hydrogen; Molecular hydrogen; H2; UN 1049

Product code : 001026

CAS number/other identifiers

CAS number : 1333-74-0

Ingredient name	%	CAS number
hydrogen	100	1333-74-0

Any concentration shown as a range is to protect confidentiality or is due to batch variation.

There are no additional ingredients present which, within the current knowledge of the supplier and in the concentrations applicable, are classified as hazardous to health or the environment and hence require reporting in this section.

Occupational exposure limits, if available, are listed in Section 8.

Section 4. First aid measures

Description of necessary first aid measures

Eye contact : Immediately flush eyes with plenty of water, occasionally lifting the upper and lower eyelids. Check for and remove any contact lenses. Continue to rinse for at least 10 minutes. Get medical attention if irritation occurs.

Inhalation : Remove victim to fresh air and keep at rest in a position comfortable for breathing. If not breathing, if breathing is irregular or if respiratory arrest occurs, provide artificial respiration or oxygen by trained personnel. It may be dangerous to the person providing aid to give mouth-to-mouth resuscitation. Get medical attention if adverse health effects persist or are severe. If unconscious, place in recovery position and get medical attention immediately. Maintain an open airway. Loosen tight clothing such as a collar, tie, belt or waistband.

Skin contact : Flush contaminated skin with plenty of water. Remove contaminated clothing and shoes. To avoid the risk of static discharges and gas ignition, soak contaminated clothing thoroughly with water before removing it. Get medical attention if symptoms occur. Wash clothing before reuse. Clean shoes thoroughly before reuse.

Ingestion : As this product is a gas, refer to the inhalation section.

Most important symptoms/effects, acute and delayed

Potential acute health effects

Eye contact : Contact with rapidly expanding gas may cause burns or frostbite.

Inhalation : No known significant effects or critical hazards.

Skin contact : Contact with rapidly expanding gas may cause burns or frostbite.

Frostbite : Try to warm up the frozen tissues and seek medical attention.

Ingestion : As this product is a gas, refer to the inhalation section.

Over-exposure signs/symptoms

Eye contact : No specific data.

Inhalation : No specific data.

Skin contact : No specific data.

Ingestion : No specific data.

Indication of immediate medical attention and special treatment needed, if necessary

Notes to physician : Treat symptomatically. Contact poison treatment specialist immediately if large quantities have been ingested or inhaled.

Specific treatments : No specific treatment.

Section 4. First aid measures

Protection of first-aiders : No action shall be taken involving any personal risk or without suitable training. It may be dangerous to the person providing aid to give mouth-to-mouth resuscitation.

See toxicological information (Section 11)

Section 5. Fire-fighting measures

Extinguishing media

Suitable extinguishing media : Use an extinguishing agent suitable for the surrounding fire.

Unsuitable extinguishing media : None known.

Specific hazards arising from the chemical : Contains gas under pressure. Extremely flammable gas. In a fire or if heated, a pressure increase will occur and the container may burst, with the risk of a subsequent explosion.

Hazardous thermal decomposition products : No specific data.

Special protective actions for fire-fighters : Promptly isolate the scene by removing all persons from the vicinity of the incident if there is a fire. No action shall be taken involving any personal risk or without suitable training. Contact supplier immediately for specialist advice. Move containers from fire area if this can be done without risk. Use water spray to keep fire-exposed containers cool. If involved in fire, shut off flow immediately if it can be done without risk. If this is impossible, withdraw from area and allow fire to burn. Fight fire from protected location or maximum possible distance. Eliminate all ignition sources if safe to do so.

Special protective equipment for fire-fighters : Fire-fighters should wear appropriate protective equipment and self-contained breathing apparatus (SCBA) with a full face-piece operated in positive pressure mode.

Section 6. Accidental release measures

Personal precautions, protective equipment and emergency procedures

For non-emergency personnel : Accidental releases pose a serious fire or explosion hazard. No action shall be taken involving any personal risk or without suitable training. Evacuate surrounding areas. Keep unnecessary and unprotected personnel from entering. Shut off all ignition sources. No flares, smoking or flames in hazard area. Avoid breathing gas. Provide adequate ventilation. Wear appropriate respirator when ventilation is inadequate. Put on appropriate personal protective equipment.

For emergency responders : If specialized clothing is required to deal with the spillage, take note of any information in Section 8 on suitable and unsuitable materials. See also the information in "For non-emergency personnel".

Environmental precautions : Ensure emergency procedures to deal with accidental gas releases are in place to avoid contamination of the environment. Inform the relevant authorities if the product has caused environmental pollution (sewers, waterways, soil or air).

Methods and materials for containment and cleaning up

Small spill : Immediately contact emergency personnel. Stop leak if without risk. Use spark-proof tools and explosion-proof equipment.

Large spill : Immediately contact emergency personnel. Stop leak if without risk. Use spark-proof tools and explosion-proof equipment. Note: see Section 1 for emergency contact information and Section 13 for waste disposal.

Section 7. Handling and storage

Precautions for safe handling

Protective measures : Put on appropriate personal protective equipment (see Section 8). Contains gas under pressure. Avoid breathing gas. Use only with adequate ventilation. Wear appropriate respirator when ventilation is inadequate. Do not enter storage areas and confined spaces unless adequately ventilated. Do not puncture or incinerate container. Use equipment rated for cylinder pressure. Close valve after each use and when empty. Protect cylinders from physical damage; do not drag, roll, slide, or drop. Use a suitable hand truck for cylinder movement.
Use only non-sparking tools. Avoid contact with eyes, skin and clothing. Empty containers retain product residue and can be hazardous. Store and use away from heat, sparks, open flame or any other ignition source. Use explosion-proof electrical (ventilating, lighting and material handling) equipment.

Advice on general occupational hygiene : Eating, drinking and smoking should be prohibited in areas where this material is handled, stored and processed. Workers should wash hands and face before eating, drinking and smoking. Remove contaminated clothing and protective equipment before entering eating areas. See also Section 8 for additional information on hygiene measures.

Conditions for safe storage, including any incompatibilities : Store in accordance with local regulations. Store in a segregated and approved area. Store away from direct sunlight in a dry, cool and well-ventilated area, away from incompatible materials (see Section 10). Eliminate all ignition sources. Cylinders should be stored upright, with valve protection cap in place, and firmly secured to prevent falling or being knocked over. Cylinder temperatures should not exceed 52 °C (125 °F). Keep container tightly closed and sealed until ready for use. See Section 10 for incompatible materials before handling or use.

Section 8. Exposure controls/personal protection

Control parameters

Occupational exposure limits

Ingredient name	Exposure limits
hydrogen	**California PEL for Chemical Contaminants (*Table AC-1*) (United States). Oxygen Depletion [Asphyxiant].** **ACGIH TLV (United States, 3/2019). Oxygen Depletion [Asphyxiant]. Explosive potential.**

Appropriate engineering controls : Use only with adequate ventilation. Use process enclosures, local exhaust ventilation or other engineering controls to keep worker exposure to airborne contaminants below any recommended or statutory limits. The engineering controls also need to keep gas, vapor or dust concentrations below any lower explosive limits. Use explosion-proof ventilation equipment.

Environmental exposure controls : Emissions from ventilation or work process equipment should be checked to ensure they comply with the requirements of environmental protection legislation. In some cases, fume scrubbers, filters or engineering modifications to the process equipment will be necessary to reduce emissions to acceptable levels.

Individual protection measures

Hygiene measures : Wash hands, forearms and face thoroughly after handling chemical products, before eating, smoking and using the lavatory and at the end of the working period. Appropriate techniques should be used to remove potentially contaminated clothing. Wash contaminated clothing before reusing. Ensure that eyewash stations and safety showers are close to the workstation location.

Section 8. Exposure controls/personal protection

Eye/face protection	: Safety eyewear complying with an approved standard should be used when a risk assessment indicates this is necessary to avoid exposure to liquid splashes, mists, gases or dusts. If contact is possible, the following protection should be worn, unless the assessment indicates a higher degree of protection: safety glasses with side-shields.
Skin protection	
Hand protection	: Chemical-resistant, impervious gloves complying with an approved standard should be worn at all times when handling chemical products if a risk assessment indicates this is necessary. Considering the parameters specified by the glove manufacturer, check during use that the gloves are still retaining their protective properties. It should be noted that the time to breakthrough for any glove material may be different for different glove manufacturers. In the case of mixtures, consisting of several substances, the protection time of the gloves cannot be accurately estimated.
Body protection	: Personal protective equipment for the body should be selected based on the task being performed and the risks involved and should be approved by a specialist before handling this product. When there is a risk of ignition from static electricity, wear anti-static protective clothing. For the greatest protection from static discharges, clothing should include anti-static overalls, boots and gloves.
Other skin protection	: Appropriate footwear and any additional skin protection measures should be selected based on the task being performed and the risks involved and should be approved by a specialist before handling this product.
Respiratory protection	: Based on the hazard and potential for exposure, select a respirator that meets the appropriate standard or certification. Respirators must be used according to a respiratory protection program to ensure proper fitting, training, and other important aspects of use. Respirator selection must be based on known or anticipated exposure levels, the hazards of the product and the safe working limits of the selected respirator.

Section 9. Physical and chemical properties

Appearance	
Physical state	: Gas
Color	: Colorless.
Odor	: Odorless.
Odor threshold	: Not available.
pH	: Not available.
Melting point	: -259.15°C (-434.5°F)
Boiling point	: -253°C (-423.4°F)
Critical temperature	: -240.15°C (-400.3°F)
Flash point	: Not available.
Evaporation rate	: Not available.
Flammability (solid, gas)	: Extremely flammable in the presence of the following materials or conditions: oxidizing materials.
Lower and upper explosive (flammable) limits	: Lower: 4% Upper: 76%
Vapor pressure	: Not available.
Vapor density	: 0.07 (Air = 1) Liquid Density@BP: 4.43 lb/ft3 (70.96 kg/m3)
Specific Volume (ft^3/lb)	: 12.0482
Gas Density (lb/ft^3)	: 0.083
Relative density	: Not applicable.
Solubility	: Not available.
Solubility in water	: Not available.
Partition coefficient: n-octanol/water	: Not available.
Auto-ignition temperature	: 500 to 571°C (932 to 1059.8°F)
Decomposition temperature	: Not available.

Section 9. Physical and chemical properties

Viscosity	: Not applicable.
Flow time (ISO 2431)	: Not available.
Molecular weight	: 2.02 g/mole
Aerosol product	
Heat of combustion	: -116486080 J/kg

Section 10. Stability and reactivity

Reactivity	: No specific test data related to reactivity available for this product or its ingredients.
Chemical stability	: The product is stable.
Possibility of hazardous reactions	: Under normal conditions of storage and use, hazardous reactions will not occur.
Conditions to avoid	: Avoid all possible sources of ignition (spark or flame). Do not pressurize, cut, weld, braze, solder, drill, grind or expose containers to heat or sources of ignition.
Incompatible materials	: Oxidizers
Hazardous decomposition products	: Under normal conditions of storage and use, hazardous decomposition products should not be produced.
Hazardous polymerization	: Under normal conditions of storage and use, hazardous polymerization will not occur.

Section 11. Toxicological information

Information on toxicological effects

Acute toxicity

Not available.

Irritation/Corrosion

Not available.

Sensitization

Not available.

Mutagenicity

Not available.

Carcinogenicity

Not available.

Reproductive toxicity

Not available.

Teratogenicity

Not available.

Specific target organ toxicity (single exposure)

Not available.

Specific target organ toxicity (repeated exposure)

Not available.

Aspiration hazard

Not available.

Section 11. Toxicological information

Information on the likely routes of exposure : Not available.

Potential acute health effects

- **Eye contact** : Contact with rapidly expanding gas may cause burns or frostbite.
- **Inhalation** : No known significant effects or critical hazards.
- **Skin contact** : Contact with rapidly expanding gas may cause burns or frostbite.
- **Ingestion** : As this product is a gas, refer to the inhalation section.

Symptoms related to the physical, chemical and toxicological characteristics

- **Eye contact** : No specific data.
- **Inhalation** : No specific data.
- **Skin contact** : No specific data.
- **Ingestion** : No specific data.

Delayed and immediate effects and also chronic effects from short and long term exposure

Short term exposure

- **Potential immediate effects** : Not available.
- **Potential delayed effects** : Not available.

Long term exposure

- **Potential immediate effects** : Not available.
- **Potential delayed effects** : Not available.

Potential chronic health effects

Not available.

- **General** : No known significant effects or critical hazards.
- **Carcinogenicity** : No known significant effects or critical hazards.
- **Mutagenicity** : No known significant effects or critical hazards.
- **Teratogenicity** : No known significant effects or critical hazards.
- **Developmental effects** : No known significant effects or critical hazards.
- **Fertility effects** : No known significant effects or critical hazards.

Numerical measures of toxicity

Acute toxicity estimates

Not available.

Section 12. Ecological information

Toxicity

Not available.

Persistence and degradability

Not available.

Bioaccumulative potential

Not available.

Section 12. Ecological information

Mobility in soil

Soil/water partition coefficient (K_{OC}) : Not available.

Other adverse effects : No known significant effects or critical hazards.

Section 13. Disposal considerations

Disposal methods : The generation of waste should be avoided or minimized wherever possible. Disposal of this product, solutions and any by-products should at all times comply with the requirements of environmental protection and waste disposal legislation and any regional local authority requirements. Dispose of surplus and non-recyclable products via a licensed waste disposal contractor. Waste should not be disposed of untreated to the sewer unless fully compliant with the requirements of all authorities with jurisdiction. Empty Airgas-owned pressure vessels should be returned to Airgas. Waste packaging should be recycled. Incineration or landfill should only be considered when recycling is not feasible. This material and its container must be disposed of in a safe way. Empty containers or liners may retain some product residues. Do not puncture or incinerate container.

Section 14. Transport information

	DOT	TDG	Mexico	IMDG	IATA
UN number	UN1049	UN1049	UN1049	UN1049	UN1049
UN proper shipping name	HYDROGEN, COMPRESSED	HYDROGEN, COMPRESSED	HYDROGEN COMPRESSED	HYDROGEN, COMPRESSED	HYDROGEN, COMPRESSED
Transport hazard class(es)	2.1	2.1	2.1	2.1	2.1
Packing group	-	-	-	-	-
Environmental hazards	No.	No.	No.	No.	No.

"Refer to CFR 49 (or authority having jurisdiction) to determine the information required for shipment of the product."

Additional information

DOT Classification : **Limited quantity** Yes.
Quantity limitation Passenger aircraft/rail: Forbidden. Cargo aircraft: 150 kg.

TDG Classification : Product classified as per the following sections of the Transportation of Dangerous Goods Regulations: 2.13-2.17 (Class 2).
Explosive Limit and Limited Quantity Index 0.125
ERAP Index 3000
Passenger Carrying Vessel Index Forbidden
Passenger Carrying Road or Rail Index Forbidden

IATA : **Quantity limitation** Passenger and Cargo Aircraft: Forbidden. Cargo Aircraft Only: 150 kg.

Special precautions for user : **Transport within user's premises:** always transport in closed containers that are upright and secure. Ensure that persons transporting the product know what to do in the event of an accident or spillage.

Section 14. Transport information

Transport in bulk according to IMO instruments : Not available.

Section 15. Regulatory information

U.S. Federal regulations : **TSCA 8(a) CDR Exempt/Partial exemption**: This material is listed or exempted.

Clean Air Act (CAA) 112 regulated flammable substances: hydrogen

Clean Air Act Section 112 (b) Hazardous Air Pollutants (HAPs) : Not listed

Clean Air Act Section 602 Class I Substances : Not listed

Clean Air Act Section 602 Class II Substances : Not listed

DEA List I Chemicals (Precursor Chemicals) : Not listed

DEA List II Chemicals (Essential Chemicals) : Not listed

SARA 302/304

Composition/information on ingredients

No products were found.

SARA 304 RQ : Not applicable.

SARA 311/312

Classification : Refer to Section 2: Hazards Identification of this SDS for classification of substance.

State regulations

Massachusetts : This material is listed.

New York : This material is not listed.

New Jersey : This material is listed.

Pennsylvania : This material is listed.

California Prop. 65

This product does not require a Safe Harbor warning under California Prop. 65.

International regulations

Chemical Weapon Convention List Schedules I, II & III Chemicals

Not listed.

Montreal Protocol

Not listed.

Stockholm Convention on Persistent Organic Pollutants

Not listed.

Rotterdam Convention on Prior Informed Consent (PIC)

Not listed.

UNECE Aarhus Protocol on POPs and Heavy Metals

Not listed.

Inventory list

Australia : This material is listed or exempted.

Canada : This material is listed or exempted.

China : This material is listed or exempted.

Europe : This material is listed or exempted.

Section 15. Regulatory information

Japan : **Japan inventory (ENCS)**: Not determined.
Japan inventory (ISHL): Not determined.

New Zealand : This material is listed or exempted.

Philippines : This material is listed or exempted.

Republic of Korea : This material is listed or exempted.

Taiwan : This material is listed or exempted.

Thailand : Not determined.

Turkey : Not determined.

United States : This material is active or exempted.

Viet Nam : This material is listed or exempted.

Section 16. Other information

Hazardous Material Information System (U.S.A.)

Health	/	1
Flammability		4
Physical hazards		3

Caution: HMIS® ratings are based on a 0-4 rating scale, with 0 representing minimal hazards or risks, and 4 representing significant hazards or risks. Although HMIS® ratings and the associated label are not required on SDSs or products leaving a facility under 29 CFR 1910.1200, the preparer may choose to provide them. HMIS® ratings are to be used with a fully implemented HMIS® program. HMIS® is a registered trademark and service mark of the American Coatings Association, Inc.

The customer is responsible for determining the PPE code for this material. For more information on HMIS® Personal Protective Equipment (PPE) codes, consult the HMIS® Implementation Manual.

National Fire Protection Association (U.S.A.)

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Copyright ©2001, National Fire Protection Association, Quincy, MA 02269. This warning system is intended to be interpreted and applied only by properly trained individuals to identify fire, health and reactivity hazards of chemicals. The user is referred to certain limited number of chemicals with recommended classifications in NFPA 49 and NFPA 325, which would be used as a guideline only. Whether the chemicals are classified by NFPA or not, anyone using the 704 systems to classify chemicals does so at their own risk.

Procedure used to derive the classification

Classification	Justification
FLAMMABLE GASES - Category 1 GASES UNDER PRESSURE - Compressed gas	Expert judgment According to package

History

Date of printing : 11/15/2020

Date of issue/Date of revision : 11/15/2020

Date of previous issue : 9/27/2018

Version : 1.01

Section 16. Other information

Key to abbreviations : ATE = Acute Toxicity Estimate
BCF = Bioconcentration Factor
GHS = Globally Harmonized System of Classification and Labelling of Chemicals
IATA = International Air Transport Association
IBC = Intermediate Bulk Container
IMDG = International Maritime Dangerous Goods
LogPow = logarithm of the octanol/water partition coefficient
MARPOL = International Convention for the Prevention of Pollution From Ships, 1973 as modified by the Protocol of 1978. ("Marpol" = marine pollution)
UN = United Nations

References : Not available.

Notice to reader

To the best of our knowledge, the information contained herein is accurate. However, neither the above-named supplier, nor any of its subsidiaries, assumes any liability whatsoever for the accuracy or completeness of the information contained herein.
Final determination of suitability of any material is the sole responsibility of the user. All materials may present unknown hazards and should be used with caution. Although certain hazards are described herein, we cannot guarantee that these are the only hazards that exist.

24.14.3 Oxygen

SAFETY DATA SHEET

Oxygen

Airgas
an Air Liquide company

Section 1. Identification

GHS product identifier : Oxygen

Chemical name : oxygen

Other means of identification : Molecular oxygen; Oxygen molecule; Pure oxygen; O2; UN 1072; Dioxygen; Oxygen USP, Aviator's Breathing Oxygen (ABO)

Product type : Gas.

Product use : Synthetic/Analytical chemistry.

Synonym : Molecular oxygen; Oxygen molecule; Pure oxygen; O2; UN 1072; Dioxygen; Oxygen USP, Aviator's Breathing Oxygen (ABO)

SDS # : 001043

Supplier's details : Airgas USA, LLC and its affiliates
259 North Radnor-Chester Road
Suite 100
Radnor, PA 19087-5283
1-610-687-5253

24-hour telephone : 1-866-734-3438

Section 2. Hazards identification

OSHA/HCS status : This material is considered hazardous by the OSHA Hazard Communication Standard (29 CFR 1910.1200).

Classification of the substance or mixture : OXIDIZING GASES - Category 1
GASES UNDER PRESSURE - Compressed gas

GHS label elements

Hazard pictograms :

Signal word : Danger

Hazard statements : May cause or intensify fire; oxidizer.
Contains gas under pressure; may explode if heated.

Precautionary statements

General : Read and follow all Safety Data Sheets (SDS'S) before use. Read label before use. Keep out of reach of children. If medical advice is needed, have product container or label at hand. Close valve after each use and when empty. Use equipment rated for cylinder pressure. Do not open valve until connected to equipment prepared for use. Use a back flow preventative device in the piping. Use only equipment of compatible materials of construction. Open valve slowly. Use only with equipment cleaned for Oxygen service.

Prevention : Keep away from clothing and other combustible materials. Keep reduction valves, valves and fittings free from oil and grease.

Response : In case of fire: Stop leak if safe to do so.

Storage : Protect from sunlight. Store in a well-ventilated place.

Disposal : Not applicable.

Hazards not otherwise classified : None known.

Section 3. Composition/information on ingredients

Substance/mixture : Substance

Chemical name : oxygen

Other means of identification : Molecular oxygen; Oxygen molecule; Pure oxygen; O2; UN 1072; Dioxygen; Oxygen USP, Aviator's Breathing Oxygen (ABO)

Product code : 001043

CAS number/other identifiers

CAS number : 7782-44-7

Ingredient name	%	CAS number
oxygen	100	7782-44-7

Any concentration shown as a range is to protect confidentiality or is due to batch variation.

There are no additional ingredients present which, within the current knowledge of the supplier and in the concentrations applicable, are classified as hazardous to health or the environment and hence require reporting in this section.

Occupational exposure limits, if available, are listed in Section 8.

Section 4. First aid measures

Description of necessary first aid measures

Eye contact : Immediately flush eyes with plenty of water, occasionally lifting the upper and lower eyelids. Check for and remove any contact lenses. Continue to rinse for at least 10 minutes. Get medical attention.

Inhalation : Remove victim to fresh air and keep at rest in a position comfortable for breathing. If not breathing, if breathing is irregular or if respiratory arrest occurs, provide artificial respiration or oxygen by trained personnel. It may be dangerous to the person providing aid to give mouth-to-mouth resuscitation. Get medical attention if adverse health effects persist or are severe. If unconscious, place in recovery position and get medical attention immediately. Maintain an open airway. Loosen tight clothing such as a collar, tie, belt or waistband.

Skin contact : Flush contaminated skin with plenty of water. Remove contaminated clothing and shoes. Get medical attention if symptoms occur. Wash clothing before reuse. Clean shoes thoroughly before reuse.

Ingestion : As this product is a gas, refer to the inhalation section.

Most important symptoms/effects, acute and delayed

Potential acute health effects

Eye contact : Contact with rapidly expanding gas may cause burns or frostbite.

Inhalation : No known significant effects or critical hazards.

Skin contact : Contact with rapidly expanding gas may cause burns or frostbite.

Frostbite : Try to warm up the frozen tissues and seek medical attention.

Ingestion : As this product is a gas, refer to the inhalation section.

Over-exposure signs/symptoms

Eye contact : No specific data.

Inhalation : No specific data.

Skin contact : No specific data.

Ingestion : No specific data.

Indication of immediate medical attention and special treatment needed, if necessary

Notes to physician : Treat symptomatically. Contact poison treatment specialist immediately if large quantities have been ingested or inhaled.

Specific treatments : No specific treatment.

Section 4. First aid measures

Protection of first-aiders : No action shall be taken involving any personal risk or without suitable training. It may be dangerous to the person providing aid to give mouth-to-mouth resuscitation.

See toxicological information (Section 11)

Section 5. Fire-fighting measures

Extinguishing media

Suitable extinguishing media : Use an extinguishing agent suitable for the surrounding fire.

Unsuitable extinguishing media : None known.

Specific hazards arising from the chemical : Contains gas under pressure. Oxidizing material. This material increases the risk of fire and may aid combustion. Contact with combustible material may cause fire. In a fire or if heated, a pressure increase will occur and the container may burst or explode.

Hazardous thermal decomposition products : No specific data.

Special protective actions for fire-fighters : Promptly isolate the scene by removing all persons from the vicinity of the incident if there is a fire. No action shall be taken involving any personal risk or without suitable training. Contact supplier immediately for specialist advice. Move containers from fire area if this can be done without risk. Use water spray to keep fire-exposed containers cool. If involved in fire, shut off flow immediately if it can be done without risk.

Special protective equipment for fire-fighters : Fire-fighters should wear appropriate protective equipment and self-contained breathing apparatus (SCBA) with a full face-piece operated in positive pressure mode.

Section 6. Accidental release measures

Personal precautions, protective equipment and emergency procedures

For non-emergency personnel : No action shall be taken involving any personal risk or without suitable training. Evacuate surrounding areas. Keep unnecessary and unprotected personnel from entering. Shut off all ignition sources. No flares, smoking or flames in hazard area. Avoid breathing gas. Provide adequate ventilation. Wear appropriate respirator when ventilation is inadequate. Put on appropriate personal protective equipment.

For emergency responders : If specialized clothing is required to deal with the spillage, take note of any information in Section 8 on suitable and unsuitable materials. See also the information in "For non-emergency personnel".

Environmental precautions : Ensure emergency procedures to deal with accidental gas releases are in place to avoid contamination of the environment. Inform the relevant authorities if the product has caused environmental pollution (sewers, waterways, soil or air).

Methods and materials for containment and cleaning up

Small spill : Immediately contact emergency personnel. Stop leak if without risk. Use spark-proof tools and explosion-proof equipment.

Large spill : Immediately contact emergency personnel. Stop leak if without risk. Use spark-proof tools and explosion-proof equipment. Note: see Section 1 for emergency contact information and Section 13 for waste disposal.

Section 7. Handling and storage

Precautions for safe handling

Section 8. Exposure controls/personal protection

Body protection	: Personal protective equipment for the body should be selected based on the task being performed and the risks involved and should be approved by a specialist before handling this product.
Other skin protection	: Appropriate footwear and any additional skin protection measures should be selected based on the task being performed and the risks involved and should be approved by a specialist before handling this product.
Respiratory protection	: Based on the hazard and potential for exposure, select a respirator that meets the appropriate standard or certification. Respirators must be used according to a respiratory protection program to ensure proper fitting, training, and other important aspects of use. Respirator selection must be based on known or anticipated exposure levels, the hazards of the product and the safe working limits of the selected respirator.

Section 9. Physical and chemical properties

Appearance	
Physical state	: Gas. [Compressed gas.]
Color	: Colorless. Blue.
Odor	: Odorless.
Odor threshold	: Not available.
pH	: Not available.
Melting point	: -218.4°C (-361.1°F)
Boiling point	: -183°C (-297.4°F)
Critical temperature	: -118.15°C (-180.7°F)
Flash point	: [Product does not sustain combustion.]
Evaporation rate	: Not available.
Flammability (solid, gas)	: Extremely flammable in the presence of the following materials or conditions: reducing materials, combustible materials and organic materials.
Lower and upper explosive (flammable) limits	: Not available.
Vapor pressure	: Not available.
Vapor density	: 1.1 (Air = 1)
Specific Volume (ft ^{3}/lb)	: 12.0482
Gas Density (lb/ft ^{3})	: 0.083
Relative density	: Not applicable.
Solubility	: Not available.
Solubility in water	: Not available.
Partition coefficient: n-octanol/water	: 0.65
Auto-ignition temperature	: Not available.
Decomposition temperature	: Not available.
Viscosity	: Not applicable.
Flow time (ISO 2431)	: Not available.
Molecular weight	: 32 g/mole

Section 10. Stability and reactivity

Reactivity	: No specific test data related to reactivity available for this product or its ingredients.
Chemical stability	: The product is stable.
Possibility of hazardous reactions	: Hazardous reactions or instability may occur under certain conditions of storage or use. Conditions may include the following: contact with combustible materials Reactions may include the following: risk of causing fire

Section 10. Stability and reactivity

Conditions to avoid : No specific data.

Incompatible materials : Highly reactive or incompatible with the following materials:
combustible materials
reducing materials
grease
oil

Hazardous decomposition products : Under normal conditions of storage and use, hazardous decomposition products should not be produced.

Hazardous polymerization : Under normal conditions of storage and use, hazardous polymerization will not occur.

Section 11. Toxicological information

Information on toxicological effects

Acute toxicity

Not available.

Irritation/Corrosion

Not available.

Sensitization

Not available.

Mutagenicity

Not available.

Carcinogenicity

Not available.

Reproductive toxicity

Not available.

Teratogenicity

Not available.

Specific target organ toxicity (single exposure)

Not available.

Specific target organ toxicity (repeated exposure)

Not available.

Aspiration hazard

Not available.

Information on the likely routes of exposure : Not available.

Potential acute health effects

Eye contact : Contact with rapidly expanding gas may cause burns or frostbite.
Inhalation : No known significant effects or critical hazards.
Skin contact : Contact with rapidly expanding gas may cause burns or frostbite.
Ingestion : As this product is a gas, refer to the inhalation section.

Symptoms related to the physical, chemical and toxicological characteristics

Section 11. Toxicological information

Eye contact : No specific data.
Inhalation : No specific data.
Skin contact : No specific data.
Ingestion : No specific data.

Delayed and immediate effects and also chronic effects from short and long term exposure

Short term exposure

Potential immediate effects : Not available.
Potential delayed effects : Not available.

Long term exposure

Potential immediate effects : Not available.
Potential delayed effects : Not available.

Potential chronic health effects

Not available.

General : No known significant effects or critical hazards.
Carcinogenicity : No known significant effects or critical hazards.
Mutagenicity : No known significant effects or critical hazards.
Teratogenicity : No known significant effects or critical hazards.
Developmental effects : No known significant effects or critical hazards.
Fertility effects : No known significant effects or critical hazards.

Numerical measures of toxicity

Acute toxicity estimates

Not available.

Section 12. Ecological information

Toxicity

Not available.

Persistence and degradability

Not available.

Bioaccumulative potential

Product/ingredient name	$LogP_{ow}$	BCF	Potential
oxygen	0.65	-	low

Mobility in soil

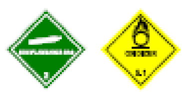
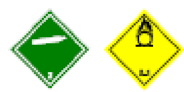
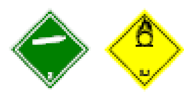
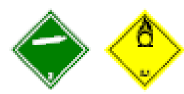
Soil/water partition coefficient (K_{oc}) : Not available.

Other adverse effects : No known significant effects or critical hazards.

Section 13. Disposal considerations

Disposal methods : The generation of waste should be avoided or minimized wherever possible. Disposal of this product, solutions and any by-products should at all times comply with the requirements of environmental protection and waste disposal legislation and any regional local authority requirements. Dispose of surplus and non-recyclable products via a licensed waste disposal contractor. Waste should not be disposed of untreated to the sewer unless fully compliant with the requirements of all authorities with jurisdiction. Empty Airgas-owned pressure vessels should be returned to Airgas. Waste packaging should be recycled. Incineration or landfill should only be considered when recycling is not feasible. This material and its container must be disposed of in a safe way. Empty containers or liners may retain some product residues. Do not puncture or incinerate container.

Section 14. Transport information

	DOT	TDG	Mexico	IMDG	IATA
UN number	UN1072	UN1072	UN1072	UN1072	UN1072
UN proper shipping name	OXYGEN, COMPRESSED	OXYGEN, COMPRESSED	OXYGEN, COMPRESSED	OXYGEN, COMPRESSED	OXYGEN, COMPRESSED
Transport hazard class(es)	2.2 (5.1) 	2.2 	2.2 (5.1) 	2.2 (5.1) 	2.2 (5.1) 
Packing group	-	-	-	-	-
Environmental hazards	No.	No.	No.	No.	No.

"Refer to CFR 49 (or authority having jurisdiction) to determine the information required for shipment of the product."

Additional information

DOT Classification : **Limited quantity** Yes.
Quantity limitation Passenger aircraft/rail: 75 kg. Cargo aircraft: 150 kg.
Special provisions A52

TDG Classification : Product classified as per the following sections of the Transportation of Dangerous Goods Regulations: 2.13-2.17 (Class 2), 2.23-2.25 (Class 5).
Explosive Limit and Limited Quantity Index 0.125
ERAP Index 3000
Passenger Carrying Vessel Index 50
Passenger Carrying Road or Rail Index 75
Special provisions 42

IATA : **Quantity limitation** Passenger and Cargo Aircraft: 75 kg. Cargo Aircraft Only: 150 kg.

Special precautions for user : **Transport within user's premises:** always transport in closed containers that are upright and secure. Ensure that persons transporting the product know what to do in the event of an accident or spillage.

Transport in bulk according to IMO instruments : Not available.

Section 15. Regulatory information

U.S. Federal regulations	:	**TSCA 8(a) CDR Exempt/Partial exemption**: This material is listed or exempted.
Clean Air Act Section 112 (b) Hazardous Air Pollutants (HAPs)	:	Not listed
Clean Air Act Section 602 Class I Substances	:	Not listed
Clean Air Act Section 602 Class II Substances	:	Not listed
DEA List I Chemicals (Precursor Chemicals)	:	Not listed
DEA List II Chemicals (Essential Chemicals)	:	Not listed

SARA 302/304

Composition/information on ingredients

No products were found.

SARA 304 RQ : Not applicable.

SARA 311/312

Classification : Refer to Section 2: Hazards Identification of this SDS for classification of substance.

State regulations

Massachusetts	:	This material is listed.
New York	:	This material is not listed.
New Jersey	:	This material is listed.
Pennsylvania	:	This material is listed.

California Prop. 65

This product does not require a Safe Harbor warning under California Prop. 65.

International regulations

Chemical Weapon Convention List Schedules I, II & III Chemicals

Not listed.

Montreal Protocol

Not listed.

Stockholm Convention on Persistent Organic Pollutants

Not listed.

Rotterdam Convention on Prior Informed Consent (PIC)

Not listed.

UNECE Aarhus Protocol on POPs and Heavy Metals

Not listed.

Inventory list

Australia	:	This material is listed or exempted.
Canada	:	This material is listed or exempted.
China	:	This material is listed or exempted.
Europe	:	This material is listed or exempted.
Japan	:	**Japan inventory (ENCS)**: Not determined. **Japan inventory (ISHL)**: Not determined.
New Zealand	:	This material is listed or exempted.
Philippines	:	This material is listed or exempted.

Section 15. Regulatory information

Republic of Korea	:	This material is listed or exempted.
Taiwan	:	This material is listed or exempted.
Thailand	:	Not determined.
Turkey	:	Not determined.
United States	:	This material is active or exempted.
Viet Nam	:	This material is listed or exempted.

Section 16. Other information

Hazardous Material Information System (U.S.A.)

Health	/	0
Flammability		0
Physical hazards		3

Caution: HMIS® ratings are based on a 0-4 rating scale, with 0 representing minimal hazards or risks, and 4 representing significant hazards or risks. Although HMIS® ratings and the associated label are not required on SDSs or products leaving a facility under 29 CFR 1910.1200, the preparer may choose to provide them. HMIS® ratings are to be used with a fully implemented HMIS® program. HMIS® is a registered trademark and service mark of the American Coatings Association, Inc.

The customer is responsible for determining the PPE code for this material. For more information on HMIS® Personal Protective Equipment (PPE) codes, consult the HMIS® Implementation Manual.

National Fire Protection Association (U.S.A.)

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Copyright ©2001, National Fire Protection Association, Quincy, MA 02269. This warning system is intended to be interpreted and applied only by properly trained individuals to identify fire, health and reactivity hazards of chemicals. The user is referred to certain limited number of chemicals with recommended classifications in NFPA 49 and NFPA 325, which would be used as a guideline only. Whether the chemicals are classified by NFPA or not, anyone using the 704 systems to classify chemicals does so at their own risk.

Procedure used to derive the classification

Classification	Justification
OXIDIZING GASES - Category 1 GASES UNDER PRESSURE - Compressed gas	Expert judgment According to package

History

Date of printing	:	9/22/2020
Date of issue/Date of revision	:	9/22/2020
Date of previous issue	:	2/3/2018
Version	:	1
Key to abbreviations	:	ATE = Acute Toxicity Estimate BCF = Bioconcentration Factor GHS = Globally Harmonized System of Classification and Labelling of Chemicals IATA = International Air Transport Association IBC = Intermediate Bulk Container IMDG = International Maritime Dangerous Goods LogPow = logarithm of the octanol/water partition coefficient

Section 16. Other information

MARPOL = International Convention for the Prevention of Pollution From Ships, 1973 as modified by the Protocol of 1978. ("Marpol" = marine pollution)
UN = United Nations

References : Not available.

Notice to reader

To the best of our knowledge, the information contained herein is accurate. However, neither the above-named supplier, nor any of its subsidiaries, assumes any liability whatsoever for the accuracy or completeness of the information contained herein.
Final determination of suitability of any material is the sole responsibility of the user. All materials may present unknown hazards and should be used with caution. Although certain hazards are described herein, we cannot guarantee that these are the only hazards that exist.

24.14.4 Neon

SAFETY DATA SHEET

Neon

Airgas
an Air Liquide company

Section 1. Identification

GHS product identifier	:	Neon
Chemical name	:	neon
Other means of identification	:	Ne; UN 1065
Product type	:	Gas.
Product use	:	Synthetic/Analytical chemistry.
Synonym	:	Ne; UN 1065
SDS #	:	001038
Supplier's details	:	Airgas USA, LLC and its affiliates 259 North Radnor-Chester Road Suite 100 Radnor, PA 19087-5283 1-610-687-5253
24-hour telephone	:	1-866-734-3438

Section 2. Hazards identification

OSHA/HCS status	:	This material is considered hazardous by the OSHA Hazard Communication Standard (29 CFR 1910.1200).
Classification of the substance or mixture	:	GASES UNDER PRESSURE - Compressed gas

GHS label elements

Hazard pictograms :

Signal word	:	Warning
Hazard statements	:	Contains gas under pressure; may explode if heated. May displace oxygen and cause rapid suffocation.

Precautionary statements

General	:	Read and follow all Safety Data Sheets (SDS'S) before use. Read label before use. Keep out of reach of children. If medical advice is needed, have product container or label at hand. Close valve after each use and when empty. Use equipment rated for cylinder pressure. Do not open valve until connected to equipment prepared for use. Use a back flow preventative device in the piping. Use only equipment of compatible materials of construction.
Prevention	:	Not applicable.
Response	:	Not applicable.
Storage	:	Protect from sunlight. Store in a well-ventilated place.
Disposal	:	Not applicable.
Hazards not otherwise classified	:	In addition to any other important health or physical hazards, this product may displace oxygen and cause rapid suffocation.

Section 3. Composition/information on ingredients

Substance/mixture : Substance

Chemical name : neon

Other means of identification : Ne; UN 1065

Product code : 001038

CAS number/other identifiers

CAS number : 7440-01-9

Ingredient name	%	CAS number
neon	100	7440-01-9

Any concentration shown as a range is to protect confidentiality or is due to batch variation.

There are no additional ingredients present which, within the current knowledge of the supplier and in the concentrations applicable, are classified as hazardous to health or the environment and hence require reporting in this section.

Occupational exposure limits, if available, are listed in Section 8.

Section 4. First aid measures

Description of necessary first aid measures

Eye contact : Immediately flush eyes with plenty of water, occasionally lifting the upper and lower eyelids. Check for and remove any contact lenses. Continue to rinse for at least 10 minutes. Get medical attention if irritation occurs.

Inhalation : Remove victim to fresh air and keep at rest in a position comfortable for breathing. If not breathing, if breathing is irregular or if respiratory arrest occurs, provide artificial respiration or oxygen by trained personnel. It may be dangerous to the person providing aid to give mouth-to-mouth resuscitation. Get medical attention if adverse health effects persist or are severe. If unconscious, place in recovery position and get medical attention immediately. Maintain an open airway. Loosen tight clothing such as a collar, tie, belt or waistband.

Skin contact : Flush contaminated skin with plenty of water. Remove contaminated clothing and shoes. Get medical attention if symptoms occur. Wash clothing before reuse. Clean shoes thoroughly before reuse.

Ingestion : As this product is a gas, refer to the inhalation section.

Most important symptoms/effects, acute and delayed

Potential acute health effects

Eye contact : Contact with rapidly expanding gas may cause burns or frostbite.

Inhalation : No known significant effects or critical hazards.

Skin contact : Contact with rapidly expanding gas may cause burns or frostbite.

Frostbite : Try to warm up the frozen tissues and seek medical attention.

Ingestion : As this product is a gas, refer to the inhalation section.

Over-exposure signs/symptoms

Eye contact : No specific data.

Inhalation : No specific data.

Skin contact : No specific data.

Ingestion : No specific data.

Indication of immediate medical attention and special treatment needed, if necessary

Notes to physician : Treat symptomatically. Contact poison treatment specialist immediately if large quantities have been ingested or inhaled.

Specific treatments : No specific treatment.

Section 4. First aid measures

Protection of first-aiders : No action shall be taken involving any personal risk or without suitable training. It may be dangerous to the person providing aid to give mouth-to-mouth resuscitation.

See toxicological information (Section 11)

Section 5. Fire-fighting measures

Extinguishing media

Suitable extinguishing media : Use an extinguishing agent suitable for the surrounding fire.

Unsuitable extinguishing media : None known.

Specific hazards arising from the chemical : Contains gas under pressure. In a fire or if heated, a pressure increase will occur and the container may burst or explode.

Hazardous thermal decomposition products : No specific data.

Special protective actions for fire-fighters : Promptly isolate the scene by removing all persons from the vicinity of the incident if there is a fire. No action shall be taken involving any personal risk or without suitable training. Contact supplier immediately for specialist advice. Move containers from fire area if this can be done without risk. Use water spray to keep fire-exposed containers cool.

Special protective equipment for fire-fighters : Fire-fighters should wear appropriate protective equipment and self-contained breathing apparatus (SCBA) with a full face-piece operated in positive pressure mode.

Section 6. Accidental release measures

Personal precautions, protective equipment and emergency procedures

For non-emergency personnel : No action shall be taken involving any personal risk or without suitable training. Evacuate surrounding areas. Keep unnecessary and unprotected personnel from entering. Avoid breathing gas. Provide adequate ventilation. Wear appropriate respirator when ventilation is inadequate. Put on appropriate personal protective equipment.

For emergency responders : If specialized clothing is required to deal with the spillage, take note of any information in Section 8 on suitable and unsuitable materials. See also the information in "For non-emergency personnel".

Environmental precautions : Ensure emergency procedures to deal with accidental gas releases are in place to avoid contamination of the environment. Inform the relevant authorities if the product has caused environmental pollution (sewers, waterways, soil or air).

Methods and materials for containment and cleaning up

Small spill : Immediately contact emergency personnel. Stop leak if without risk.

Large spill : Immediately contact emergency personnel. Stop leak if without risk. Note: see Section 1 for emergency contact information and Section 13 for waste disposal.

Section 7. Handling and storage

Precautions for safe handling

Protective measures : Put on appropriate personal protective equipment (see Section 8). Contains gas under pressure. Avoid breathing gas. Do not puncture or incinerate container. Use equipment rated for cylinder pressure. Close valve after each use and when empty. Protect cylinders from physical damage; do not drag, roll, slide, or drop. Use a suitable hand truck for cylinder movement.
Avoid contact with eyes, skin and clothing. Empty containers retain product residue and can be hazardous.

Section 7. Handling and storage

Advice on general occupational hygiene : Eating, drinking and smoking should be prohibited in areas where this material is handled, stored and processed. Workers should wash hands and face before eating, drinking and smoking. Remove contaminated clothing and protective equipment before entering eating areas. See also Section 8 for additional information on hygiene measures.

Conditions for safe storage, including any incompatibilities : Store in accordance with local regulations. Store in a segregated and approved area. Store away from direct sunlight in a dry, cool and well-ventilated area, away from incompatible materials (see Section 10). Cylinders should be stored upright, with valve protection cap in place, and firmly secured to prevent falling or being knocked over. Cylinder temperatures should not exceed 52 °C (125 °F). Keep container tightly closed and sealed until ready for use. See Section 10 for incompatible materials before handling or use.

Section 8. Exposure controls/personal protection

Control parameters

Occupational exposure limits

Ingredient name	Exposure limits
neon	**ACGIH TLV (United States, 3/2019). Oxygen Depletion [Asphyxiant].**

Appropriate engineering controls : Good general ventilation should be sufficient to control worker exposure to airborne contaminants.

Environmental exposure controls : Emissions from ventilation or work process equipment should be checked to ensure they comply with the requirements of environmental protection legislation. In some cases, fume scrubbers, filters or engineering modifications to the process equipment will be necessary to reduce emissions to acceptable levels.

Individual protection measures

Hygiene measures : Wash hands, forearms and face thoroughly after handling chemical products, before eating, smoking and using the lavatory and at the end of the working period. Appropriate techniques should be used to remove potentially contaminated clothing. Wash contaminated clothing before reusing. Ensure that eyewash stations and safety showers are close to the workstation location.

Eye/face protection : Safety eyewear complying with an approved standard should be used when a risk assessment indicates this is necessary to avoid exposure to liquid splashes, mists, gases or dusts. If contact is possible, the following protection should be worn, unless the assessment indicates a higher degree of protection: safety glasses with side-shields.

Skin protection

Hand protection : Chemical-resistant, impervious gloves complying with an approved standard should be worn at all times when handling chemical products if a risk assessment indicates this is necessary. Considering the parameters specified by the glove manufacturer, check during use that the gloves are still retaining their protective properties. It should be noted that the time to breakthrough for any glove material may be different for different glove manufacturers. In the case of mixtures, consisting of several substances, the protection time of the gloves cannot be accurately estimated.

Body protection : Personal protective equipment for the body should be selected based on the task being performed and the risks involved and should be approved by a specialist before handling this product.

Other skin protection : Appropriate footwear and any additional skin protection measures should be selected based on the task being performed and the risks involved and should be approved by a specialist before handling this product.

Section 8. Exposure controls/personal protection

Respiratory protection	: Based on the hazard and potential for exposure, select a respirator that meets the appropriate standard or certification. Respirators must be used according to a respiratory protection program to ensure proper fitting, training, and other important aspects of use. Respirator selection must be based on known or anticipated exposure levels, the hazards of the product and the safe working limits of the selected respirator.

Section 9. Physical and chemical properties

Appearance	
Physical state	: Gas. [Compressed gas.]
Color	: Colorless.
Odor	: Odorless.
Odor threshold	: Not available.
pH	: Not available.
Melting point	: -248.7°C (-415.7°F)
Boiling point	: -246.1°C (-411°F)
Critical temperature	: -228.7°C (-379.7°F)
Flash point	: [Product does not sustain combustion.]
Evaporation rate	: Not available.
Flammability (solid, gas)	: Not available.
Lower and upper explosive (flammable) limits	: Not available.
Vapor pressure	: Not available.
Vapor density	: 0.69 (Air = 1)
Specific Volume (ft^3/lb)	: 19.2307
Gas Density (lb/ft^3)	: 0.0522
Relative density	: Not applicable.
Solubility	: Not available.
Solubility in water	: Not available.
Partition coefficient: n-octanol/water	: 0.28
Auto-ignition temperature	: Not available.
Decomposition temperature	: Not available.
Viscosity	: Not applicable.
Flow time (ISO 2431)	: Not available.
Molecular weight	: 20.18 g/mole

Section 10. Stability and reactivity

Reactivity	: No specific test data related to reactivity available for this product or its ingredients.
Chemical stability	: The product is stable.
Possibility of hazardous reactions	: Under normal conditions of storage and use, hazardous reactions will not occur.
Conditions to avoid	: No specific data.
Incompatible materials	: No specific data.
Hazardous decomposition products	: Under normal conditions of storage and use, hazardous decomposition products should not be produced.

Section 10. Stability and reactivity

Hazardous polymerization : Under normal conditions of storage and use, hazardous polymerization will not occur.

Section 11. Toxicological information

Information on toxicological effects

Acute toxicity

Not available.

Irritation/Corrosion

Not available.

Sensitization

Not available.

Mutagenicity

Not available.

Carcinogenicity

Not available.

Reproductive toxicity

Not available.

Teratogenicity

Not available.

Specific target organ toxicity (single exposure)

Not available.

Specific target organ toxicity (repeated exposure)

Not available.

Aspiration hazard

Not available.

Information on the likely routes of exposure : Not available.

Potential acute health effects

Eye contact : Contact with rapidly expanding gas may cause burns or frostbite.
Inhalation : No known significant effects or critical hazards.
Skin contact : Contact with rapidly expanding gas may cause burns or frostbite.
Ingestion : As this product is a gas, refer to the inhalation section.

Symptoms related to the physical, chemical and toxicological characteristics

Eye contact : No specific data.
Inhalation : No specific data.
Skin contact : No specific data.
Ingestion : No specific data.

Delayed and immediate effects and also chronic effects from short and long term exposure

Short term exposure

Potential immediate effects : Not available.
Potential delayed effects : Not available.

Long term exposure

Section 11. Toxicological information

Potential immediate effects : Not available.

Potential delayed effects : Not available.

Potential chronic health effects

Not available.

General : No known significant effects or critical hazards.

Carcinogenicity : No known significant effects or critical hazards.

Mutagenicity : No known significant effects or critical hazards.

Teratogenicity : No known significant effects or critical hazards.

Developmental effects : No known significant effects or critical hazards.

Fertility effects : No known significant effects or critical hazards.

Numerical measures of toxicity

Acute toxicity estimates

Not available.

Section 12. Ecological information

Toxicity

Not available.

Persistence and degradability

Not available.

Bioaccumulative potential

Product/ingredient name	$LogP_{ow}$	BCF	Potential
neon	0.28	-	low

Mobility in soil

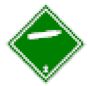
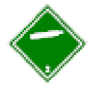
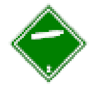
Soil/water partition coefficient (K_{OC}) : Not available.

Other adverse effects : No known significant effects or critical hazards.

Section 13. Disposal considerations

Disposal methods : The generation of waste should be avoided or minimized wherever possible. Disposal of this product, solutions and any by-products should at all times comply with the requirements of environmental protection and waste disposal legislation and any regional local authority requirements. Dispose of surplus and non-recyclable products via a licensed waste disposal contractor. Waste should not be disposed of untreated to the sewer unless fully compliant with the requirements of all authorities with jurisdiction. Empty Airgas-owned pressure vessels should be returned to Airgas. Waste packaging should be recycled. Incineration or landfill should only be considered when recycling is not feasible. This material and its container must be disposed of in a safe way. Empty containers or liners may retain some product residues. Do not puncture or incinerate container.

Section 14. Transport information

	DOT	TDG	Mexico	IMDG	IATA
UN number	UN1065	UN1065	UN1065	UN1065	UN1065
UN proper shipping name	NEON, COMPRESSED	NEON, COMPRESSED	NEON, COMPRESSED	NEON, COMPRESSED	NEON, COMPRESSED
Transport hazard class(es)	2.2 	2.2 	2.2 	2.2 	2.2 
Packing group	-	-	-	-	-
Environmental hazards	No.	No.	No.	No.	No.

"Refer to CFR 49 (or authority having jurisdiction) to determine the information required for shipment of the product."

Additional information

DOT Classification : **Limited quantity** Yes.
Quantity limitation Passenger aircraft/rail: 75 kg. Cargo aircraft: 150 kg.

TDG Classification : Product classified as per the following sections of the Transportation of Dangerous Goods Regulations: 2.13-2.17 (Class 2).
Explosive Limit and Limited Quantity Index 0.125
Passenger Carrying Road or Rail Index 75

IATA : **Quantity limitation** Passenger and Cargo Aircraft: 75 kg. Cargo Aircraft Only: 150 kg.

Special precautions for user : **Transport within user's premises:** always transport in closed containers that are upright and secure. Ensure that persons transporting the product know what to do in the event of an accident or spillage.

Transport in bulk according to IMO instruments : Not available.

Section 15. Regulatory information

U.S. Federal regulations : **TSCA 8(a) CDR Exempt/Partial exemption**: Not determined

Clean Air Act Section 112 (b) Hazardous Air Pollutants (HAPs) : Not listed

Clean Air Act Section 602 Class I Substances : Not listed

Clean Air Act Section 602 Class II Substances : Not listed

DEA List I Chemicals (Precursor Chemicals) : Not listed

DEA List II Chemicals (Essential Chemicals) : Not listed

SARA 302/304

Composition/information on ingredients

No products were found.

SARA 304 RQ : Not applicable.

Section 15. Regulatory information

SARA 311/312

Classification : Refer to Section 2: Hazards Identification of this SDS for classification of substance.

State regulations

Massachusetts	: This material is listed.
New York	: This material is not listed.
New Jersey	: This material is listed.
Pennsylvania	: This material is listed.

California Prop. 65

This product does not require a Safe Harbor warning under California Prop. 65.

International regulations

Chemical Weapon Convention List Schedules I, II & III Chemicals

Not listed.

Montreal Protocol

Not listed.

Stockholm Convention on Persistent Organic Pollutants

Not listed.

Rotterdam Convention on Prior Informed Consent (PIC)

Not listed.

UNECE Aarhus Protocol on POPs and Heavy Metals

Not listed.

Inventory list

Australia	: This material is listed or exempted.
Canada	: This material is listed or exempted.
China	: This material is listed or exempted.
Europe	: This material is listed or exempted.
Japan	: **Japan inventory (ENCS)**: Not determined. **Japan inventory (ISHL)**: Not determined.
New Zealand	: This material is listed or exempted.
Philippines	: This material is listed or exempted.
Republic of Korea	: This material is listed or exempted.
Taiwan	: This material is listed or exempted.
Thailand	: Not determined.
Turkey	: Not determined.
United States	: This material is active or exempted.
Viet Nam	: This material is listed or exempted.

Section 16. Other information

Hazardous Material Information System (U.S.A.)

Health	/	0
Flammability		0
Physical hazards		3

Section 16. Other information

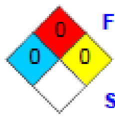
Caution: HMIS® ratings are based on a 0-4 rating scale, with 0 representing minimal hazards or risks, and 4 representing significant hazards or risks. Although HMIS® ratings and the associated label are not required on SDSs or products leaving a facility under 29 CFR 1910.1200, the preparer may choose to provide them. HMIS® ratings are to be used with a fully implemented HMIS® program. HMIS® is a registered trademark and service mark of the American Coatings Association, Inc.

The customer is responsible for determining the PPE code for this material. For more information on HMIS® Personal Protective Equipment (PPE) codes, consult the HMIS® Implementation Manual.

National Fire Protection Association (U.S.A.)

Flammability

Health



Instability/Reactivity

Special

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Copyright ©2001, National Fire Protection Association, Quincy, MA 02269. This warning system is intended to be interpreted and applied only by properly trained individuals to identify fire, health and reactivity hazards of chemicals. The user is referred to certain limited number of chemicals with recommended classifications in NFPA 49 and NFPA 325, which would be used as a guideline only. Whether the chemicals are classified by NFPA or not, anyone using the 704 systems to classify chemicals does so at their own risk.

Procedure used to derive the classification

Classification	Justification
GASES UNDER PRESSURE - Compressed gas	On basis of test data

History

Date of printing : 1/4/2021

Date of issue/Date of revision : 1/4/2021

Date of previous issue : 6/16/2016

Version : 0.02

Key to abbreviations : ATE = Acute Toxicity Estimate
BCF = Bioconcentration Factor
GHS = Globally Harmonized System of Classification and Labelling of Chemicals
IATA = International Air Transport Association
IBC = Intermediate Bulk Container
IMDG = International Maritime Dangerous Goods
LogPow = logarithm of the octanol/water partition coefficient
MARPOL = International Convention for the Prevention of Pollution From Ships, 1973 as modified by the Protocol of 1978. ("Marpol" = marine pollution)
UN = United Nations

References : Not available.

Notice to reader

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