Combined Cycle Hydrogen-Fired Turbine Power Plant Preliminary Design Report

Miranda Bailey, Colby Jones, and Jacob Young

20 May 2021

OVERVIEW

Hydrogen combustion is a promising technology in the global transition away from fossil fuels and hydrocarbon fuels, including natural gas, which release carbon dioxide upon combustion. According to the *World Energy Outlook*, “low-carbon hydrogen is rising in importance in many energy transition strategies, and several countries are accelerating efforts to scale up infrastructure, demand, and expertise”.[1] By the mid-2020’s, contracts for 150 cubic kilometers of liquified natural gas are set to expire, and it is expected that the economic gains from switching from coal to natural gas will be exhausted by that time; therefore, hydrogen will be a competitive replacement for natural gas by mid-decade.[1] Additionally, hydrogen is a valuable transition tool to meet future goals for net zero emissions.

Discussed below is the design and simulation of a combined cycle hydrogen turbine power plant, which is intended to act as preliminary design for a plant that could supplement solar and wind energy demands using hydrogen as a fuel, which produces zero carbon-based emissions. In this design, hydrogen is formed via electrolysis of water during times of high energy production from wind and solar sources. As a result, a reserve of energy is produced, which can be accessed by reintroduction of oxygen to reverse the reaction, either by combustion of hydrogen with air or by electrolysis in a hydrogen fuel cell. Figure 1 depicts the design of the unit operations in the combined cycle hydrogen gas turbine plan. Aspen HYSYS was used to simulate the process and investigate various design parameters.

Over the next decade, renewable energy sources, particularly solar, are expected to grow rapidly,[1] however, energy production from these sources cannot be easily controlled by manipulation of the rate of input of raw materials, unlike energy production from fossil fuels, natural gas, and hydropower plants. For wind and solar energy, excess from times of high energy production must be accessible during times of low energy production, so it is necessary to store energy in some way. Battery technology is a popular topic today, and the technology is advancing; however, battery storage would be more feasible for short-term storage, whereas a combined cycle hydrogen gas turbine is feasible regardless of length of storage.[2] The COVID-19 pandemic brought uncertainty to all parts of life in 2020, including energy demand, which brought attention to a need for “a reliable, affordable, and secure electricity supply that is able to accommodate sudden changes in behavior and economic activity while continuing to support vital health and information services.”[1] Storage of energy during times of excess production for use when energy demand exceeds wind and solar energy production provides a reliable source of energy during times of uncertainty and change.

DESIGN

In a combined cycle hydrogen-fired turbine power plant, such as the one modeled in Figure 1, hydrogen is combusted with air at high pressure. The reaction produces large amounts of thermal energy. The combustion product is sent through a gas turbine, which converts thermal energy to mechanical energy to produce electricity. The turbine effluent still has a significant amount of thermal energy. To extract the remaining thermal energy and increase the plant's efficiency, the gas turbine effluent is sent through a series of heat exchangers, which causes water to turn to superheated steam at elevated pressures. The steam is sent through a series of turbines, which converts its thermal energy to mechanical energy to produce additional electricity. Currently, no operational plants operate using this technology; however natural gas power plants operate in a very similar manner.[2] The objective of this specific design is to mimic existing natural gas power plants, so that, in theory, a natural gas power plant could easily transition to hydrogen as a fuel source when the fuel contract for natural gas expires. The design for combined cycle hydrogen turbine power plant in Figure 1 and simulated in Aspen HYSYS in Figure 2, was developed based on the assumption that the plant would supplement wind and solar energy sources by producing electricity when demand is higher than those renewable sources are producing. The design was not intended to be a primary power source. A design like this would best be paired with a hydrogen producing process. One example of this may be electrolysis driven by solar power when grid energy requirements are being met with excess energy left over. Another promising place for a hydrogen-fired turbine is after at the end of a nuclear cycle as high temperature requires less energy input to undergo electrolysis. In addition to this nuclear energy has the capability of directly producing hydrogen.[3]

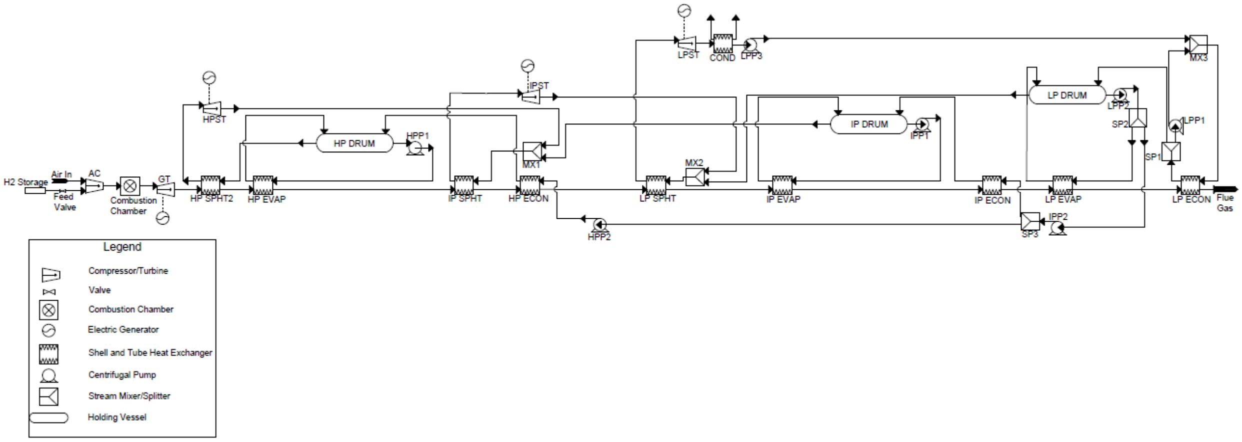

Figure 1. The basic design of the power generation plant is represented in the above P&ID. Hydrogen fuel and air are mixed and combusted within the combustion chamber, and the resultant exhaust gases are then utilized in a gas turbine for the generation of power. The residual heat from the gases exiting this turbine is then used gradually through a series of heat exchangers to support three separate steam turbine sections operating at different pressures. This design makes use of both the hot gas itself and its residual heat to create energy through several turbine cycles.

Storage

Numerous methods have been proposed for storing hydrogen, and some of these are in use currently in moderate-to-large scale applications. These methods include storage in large above ground tanks in either gaseous or liquid form, storage underground in pipeline sections, and storage in large underground salt caverns in the case of very large scale storage.[4] Within the constraints of this project, storage in underground pipelines seemed to be the optimal choice, due to onsite storage requirements and necessary storage space. Underground pipes allow for higher pressures, and therefore higher storage densities, than aboveground tanks.[4] Literature indicates that storage pressures near 10 MPa and pipe diameters up to 1.4 meters allow for storage of approximately 12 metric tons of hydrogen per kilometer of pipeline.[4] While this is much greater than aboveground tanks, which retain a sizeable footprint and only allow storages in the hundreds of kilograms (due to lower allowable storage pressures), this still yields a relatively high amount of required piping in light of the plants hydrogen flow rate.

Given that the plant design necessitates a flow rate of roughly 200 kg of hydrogen per hour, significant amounts of storage are required for self-sufficient operation over a feasible time span. Assuming 12 hours of continuous operation every day, a week's worth of run time utilizes nearly 17,000 kg of hydrogen, or 17 metric tons. If hydrogen is delivered on a weekly basis, and the desired storage quantity is extended to run the plant for 9 days to include a time buffer, the required hydrogen storage onsite is increased to roughly 22 metric tons. This equates to just under 2 km of buried storage piping, which must be constructed of a material that can resist hydrogen interaction and embrittlement - likely 316 Stainless Steel or similar.[5] The projected installed cost of this piping, using a Lang Factor of 5.5 and a Towler and Sinott cost correlation (See Appendix D, Equation D.15), comes to $138M. This is an extremely high cost for onsite storage, especially considering the relatively short 9-day operation capability given by this solution. As such, onsite generation through electrolysis or integration with a hydrogen-producing process would likely be a desirable alternative to physical storage.

Safety Considerations

Like any flammable chemical used in industry, hydrogen must be handled carefully. Compared to other fuels used for energy production, such as natural gas, hydrogen gas has a wide flammability range (4-80 vol% H_2 in air) and is easily ignited at ambient conditions, even when great care is taken to eliminate sources of ignition.[6] Since large quantities of hydrogen would be stored on-site, the plant engineers would be responsible for writing safe procedures for equipment maintenance and refilling the hydrogen storage vessels. Additionally, the hydrogen gas is stored at 10 MPa (100 times atmospheric pressure), so the storage system and pipelines from storage to the combustion chamber must be rated for very high pressures and have systems in place for emergency situations. Note in Figures 1 and 2 that the hydrogen feed to the combustion chamber is fitted with a directional valve, to ensure that the contents of the combustion chamber will not have access to backflow to the hydrogen storage under any circumstances. Additionally, this valve is fail-closed, so the flow of hydrogen is stopped when safety systems must engage. The combustion chamber and every major unit operation must be fitted with pressure relief valves for safety purposes. Finally, there must be a system in place for the steam cycle in the event that the combustion chamber does not provide heat to the steam cycle, since standard pumps are designed for liquid only and standard turbines can only process vapors.

SIMULATION

The process was modeled using Aspen HYSYS simulation software, and the overall simulation is depicted in Figure 2.

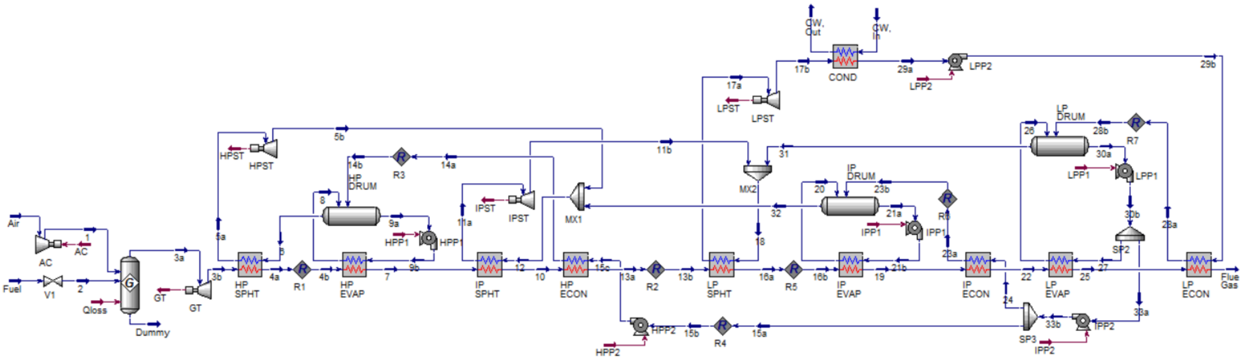

Figure 2. Overall flowsheet for a simulation of the process model in Aspen HYSYS. Material stream properties and compositions are presented in Appendix B in Tables B1 and B2, respectively. Energy flow is presented in Appendix B in Table B3. Duty and log mean temperature differences for all heat exchangers are collected in Appendix B in Table B4.

The model assumes a basis of 100 kmol/h hydrogen gas and 240 kmol/h dry air. The combustion reaction was simulated in a Gibbs reactor, which predicts a product that contains 0.72% nitrogen oxides (NO_x), which exceeds the maximum allowable NO_x emissions standard set by the Environmental Protection Agency (EPA) for natural gas power plants of 42 ppm per hour.[7] However, the Gibbs reactor is only a model of the combustion chamber, which assumes no pre-dilution of hydrogen gas with air or fuel injection strategies, which can be employed to reduce the formation of NO_x. Additionally, most NO_x are water-soluble and could be captured through sequestration with water, and the reactor produces no nitrogen dioxide, which is the compound that is measured as a surrogate for the family of NO_x compounds as part of the National Ambient Air Quality Standards (NAAQS).[8]

At a raw material flow rate of 100 kmol/h, the simulated plant produces 3.3MW of power, but the production of power scales proportionately to the flow rate of hydrogen (i.e. 10 kmol/h produces 0.33 MW and 1000 kmol/h produces 33 MW) when the flow rates of air to the reactor and water in the steam cycle are also scaled to the flow rate of hydrogen. Since this technology is designed to be supplementary to renewable energy sources, such as wind and solar, the current capacity is reasonable for calculating estimates for the size and cost of equipment.

In the current configuration, which is based on a simulation of a combined cycle natural gas-fired turbine power plant,[9] the hydrogen-fired turbine power plant in Figure 2 achieves 41.37% efficiency. See Appendix C for the efficiency sample calculation. Accounting for the energy used in the plant for operation, the plant has a net efficiency of 30.72%. Energy losses occur at the reactor, since it was not reasonable to make the assumption of a perfectly insulated combustion chamber and in each turbine as mechanical losses; each turbine has an isentropic efficiency of 90%. Current natural gas-fired turbine power plants operate at an average of 33 to 35% and have a wide range of capacities based on the power demands of the region for which the plant provides energy.[10] Current fuel cells operate at about 60% efficiency and the "stackability" of fuel cells allows for a very wide range of power generation.[10]

ECONOMICS

Economic analysis was carried out per unit operation, using several cost correlation methods.[11,12,13] The specifics of the costing correlations carried out can be seen in Appendix D, and the overall results of these calculations separated by equipment type can be seen in Appendix E. Equipment was grouped for costing as follows: heat exchangers, pumps, pressure vessels, turbines, the combustion chamber, storage piping, operator costs, and raw material costs. While the different kinds of equipment utilized various costing correlation methods, they are seen as equally valid once converted to present dollars. The heat exchangers and pressure vessels were sized and subsequently priced based primarily on specific flow parameters and residence times, while other equipment such as the gas and steam turbines utilized a single key metric such as power output.

The total installed equipment cost of the plant came to roughly $22.2M, along with $7.69M in yearly raw material cost and $720k in operator salaries. The raw materials cost was priced for purchasing all the raw hydrogen required for the plant, receiving no hydrogen from electrolysis on site. Yearly plant costing was calculated with constant operation, this is likely unrealistic as this plant would be used to supplement the power grid during energy shortages.

SUMMARY AND SUGGESTIONS

Renewable energy sources, especially solar and wind technologies, will grow rapidly in the next ten years, but the power production from these sources is not user-controlled, unlike fuel-burning power plants. This means that power supply from wind and solar sources, and consumer demand are rarely equal. During times of excess production, energy can be used to produce hydrogen from water through electrolysis. When demand exceeds the supply from wind and solar, hydrogen can be combusted in a combined cycle hydrogen-fired turbine power plant to produce additional power. The team investigated a design that was loosely based on an existing natural gas-fired turbine power plant, and simulated the process in AspenHYSYS. The design achieved a maximum output of 33MW per 100kmol/h of hydrogen and an efficiency of 41%. A plant of this size would have a capital investment of $22.2 million, plus $8.41 million per year in operating costs, although these costs could potentially be reduced through future research and investigation, including the following suggestions.

The low-pressure economizer has a large cool water flow rate and a small log mean temperature difference, so the area required for heat exchange is large, making this heat exchanger expensive. Additionally, use of the low-pressure economizer requires a large cooling water flow rate in the condenser. If the condensate was only cooled to a saturated liquid, then pumped to the low-pressure phase separator drum, it would save on capital and operating costs. The low-pressure economizer helps to cool the combustion products but, the heat exchanger cost is very high to achieve a small temperature difference. The reaction product has a dew point of about 72^{o}C at atmospheric pressure due to its high water content, so it would be useful to research cooling this stream below its dew point before release to the atmosphere to avoid the possibility of heat pollution.

For this hydrogen-fired power plant design to be more profitable, hydrogen should be supplied on site from a different process, such as excess power from a renewable energy source being used for electrolysis. A second option may consist of a nuclear plant feeding the hydrogen it produces to a hydrogen-fired turbine. High temperature electrolysis (600^{o}C) of process steam would require just two-thirds the amount of power as required for low temperature electrolysis, reducing the cost of electrolysis and raw materials for the turbine design.[3] The operating costs of the plant could be reduced by up to $4.4 million per year by eliminating or reducing the cost of hydrogen for the plant. Finally a paired model would reduce operating time reducing other raw material costs such as cooling water.

REFERENCES

[1] IEA (2020), *World Energy Outlook 2020*, IEA, Paris https://www.iea.org/reports/world-energy-outlook-2020

[2] Patel, S. by P. O. W. E. R. High-Volume Hydrogen Gas Turbines Take Shape. https://www.powermag.com/high-volume-hydrogen-gas-turbines-take-shape/ (accessed Apr 18, 2021).

[3] World Nuclear Association (2021), *Hydrogen Production and Uses,* WNA. https://www.world-nuclear.org/information-library/energy-and-the-environment/hydrogen-production-and-uses.aspx. (Accessed Apr. 18, 2021)

[4] J. Andersson, S. Gronkvist. Large-scale storage of hydrogen. International Journal of Hydrogen Energy 44 (2019) 11901-11919. https://doi.org/10.1016/j.ijhydene.2019.03.063

[5] C. San Marchi, B.P. Somerday. *Technical Reference for Hydrogen Compatibility of Materials;* SAND2012-7321; Sandia National Laboratories: Livermore, CA, 2012. https://h2tools.org/tech-ref/technical-reference-for-hydrogen-compatibility-of-materials

[6] Leachman, J. So just how dangerous is hydrogen fuel? https://hydrogen.wsu.edu/2017/03/17/so-just-how-dangerous-is-hydrogen-fuel/ (accessed May 15, 2021).

[7] Texas Administrative Code. Rule §117.305. https://texreg.sos.state.tx.us/public/readtac$ext.TacPage?sl=R&app=9&p_dir=&p_rloc=&p_tloc=&p_ploc=&pg=1&p_tac=&ti=30&pt=1&ch=117&rl=305. (Accessed Apr. 17, 2021).

[8] Clean Air Technology Center. Technical Bulletin: Nitrogen Oxides (NOx), Why and How They Are Controlled, 1999.

[9] Z. Liu, I.A. Karimi. Simulation of a combined cycle gas turbine power plant in Aspen HYSYS. Energy Procedia 158 (2019) 3620-3625. https://doi.org/10.1016/j.egypro.2019.01.901

[10] U.S. Department of Energy. *Energy Efficiency & Renewable Energy*, 2015.

[11] J. Couper et. al. *Chemical Process Equipment - Selection and Design;* Gulf Professional Publishing, 2009.

[12] G. Towler, R. Sinott. *Chemical Engineering Design;* Elsevier Inc., 2010.

[13] *Peters & Timmerhaus Costing Website.* http://www.mhhe.com/engcs/chemical/peters/data/ce.html (Accessed 2021-05-18).

[14] Robinson, W. R. Chapter 5: Thermochemistry. In *Chemistry*; Rice University: Houston, TX, 2016.

APPENDICES

Appendix A: Assumptions

1. A 100 kmol/h hydrogen gas feed is assumed. Scaling all flow rates by a factor X also scales the energies of the system by that factor X, so the plant can be scaled to whatever feed is appropriate for the maximum needs of the plant, depending on location.
2. Air is fed to the system in slight excess, to make sure that hydrogen is the limiting reagent.
3. In the Aspen HYSYS simulation, a Gibbs reactor is used to simulate the reactions that occur in the combustion chamber, including the formation of NO_x as a secondary source of oxygen consumption.
4. The average natural gas plant has an efficiency of 33 to 35% at full loading.[10]
5. Current fuel cells have an efficiency of approximately 60% and a wide range of capacities due to their "stackability".[10]
6. Heat transfer coefficient U is assumed to be 28 [W/m^2 C] in low temperature vapor-vapor processes and 50 [W/m^2 C] high temperature vapor-vapor processes. U in the evaporative condenser is set to 1560 [W/m^2 C].
7. Tube count was assumed at 20 in shell-and-tube heat exchangers.
8. Market research done to select 3 pumps within spec.
9. A weld efficiency of 1 assumed for separator vessel manufacture.
10. Sizing the pressure vessels assumed the fraction of liquid in the tank is 0.5.
11. When calculating installation costs a Lang factor of 5.5 was used.

Appendix B: Aspen HYSYS Simulation Tables

Table B1. Material stream properties for all material streams presented in Figure 2.

Stream	**Air**	**Fuel**	**1**	**2**	**Dummy**	**3a**	**3b**	**4a**
Vapor Fraction	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	1.0000	1.0000
Temperature [ºC]	20.00	20.00	435.8	21.28	2229	2229	1598	1259
Pressure [kPa]	101.0	5000	2000	2000	1900	1900	281.0	261.0
Molar Flow [kmol/h]	240.0	100.0	240.0	100.0	0.0000	289.9	289.9	289.9
Mass Flow [kg/h]	6924	201.6	6924	201.6	0.0000	7126	7126	7126
Stream	**4b**	**5a**	**5b**	**6**	**7**	**8**	**9a**	**9b**
Vapor Fraction	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000
Temperature [ºC]	1259	1000	834.0	196.6	864.0	196.6	196.6	196.6
Pressure [kPa]	261.0	1405	625.0	1425	241.0	1425	1425	1475
Molar Flow [kmol/h]	289.9	122.2	122.2	122.2	289.9	122.2	122.2	122.2
Mass Flow [kg/h]	7126	2201	1902	2201	7126	2201	2201	2201
Stream	**10**	**11a**	**11b**	**12**	**13a**	**13b**	**14a**	**14b**
Vapor Fraction	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000
Temperature [ºC]	790.0	690.0	528.0	592.2	719.8	719.8	195.0	195.0
Pressure [kPa]	221.0	605.0	225.0	625.0	201.0	201.0	1425	1425
Molar Flow [kmol/h]	289.9	197.2	197.2	197.2	289.9	289.9	122.2	122.2
Mass Flow [kg/h]	7126	3552	3552	3552	7126	7126	2201	2201
Stream	**15a**	**15b**	**15c**	**16a**	**16b**	**17a**	**17b**	**18**
Vapor Fraction	0.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature [ºC]	124.0	124.0	124.1	716.7	716.6	450.0	365.8	446.7
Pressure [kPa]	700.0	700.0	1500	181.0	181.0	205.0	110.0	225.0
Molar Flow [kmol/h]	122.2	122.2	122.2	289.9	289.9	250.0	250.0	250.0
Mass Flow [kg/h]	2201	2201	2201	7126	7126	4504	4504	4504

Stream	**19**	**20**	**21a**	**21b**	**22**	**23a**	**23b**	**24**
Vapor Fraction	1.0000	1.0000	0.0000	0.0000	1.0000	0.0000	0.0000	0.0000
Temperature [ºC]	428.5	188.0	160.7	160.7	412.6	149.0	149.0	124.0
Pressure [kPa]	161.0	625.0	625.0	675.0	141.0	625.0	625.0	700.0
Molar Flow [kmol/h]	289.9	75.00	75.00	75.00	289.9	75.00	75.00	75.00
Mass Flow [kg/h]	7126	1351	1351	1351	7126	1351	1351	1351
Stream	**25**	**26**	**27**	**28a**	**28b**	**29a**	**29b**	**30a**
Vapor Fraction	1.0000	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Temperature [ºC]	161.0	124.0	124.0	110.0	110.0	70.00	70.01	124.0
Pressure [kPa]	121.0	225.0	275.0	225.0	225.0	110.0	300.0	225.0
Molar Flow [kmol/h]	289.9	60.00	60.00	250.0	250.0	250.0	250.0	257.2
Mass Flow [kg/h]	7126	1081	1081	4504	4504	4504	4504	4633
Stream	**30b**	**31**	**32**	**33a**	**33b**	**Flue Gas**	**CW, in**	**CW, out**
Vapor Fraction	0.0000	1.0000	1.0000	0.0000	0.0000	1.0000	0.0000	0.0000
Temperature [ºC]	124.0	124.0	160.7	124.0	124.0	72.79	20.00	75.00
Pressure [kPa]	275.0	225.0	625.0	275.0	700.0	101.0	101.3	101.3
Molar Flow [kmol/h]	257.2	52.81	75.00	60.00	197.2	289.9	3004	3004
Mass Flow [kg/h]	4633	951.4	1351	1081	3552	7126	54120	54120

Table B2. Material stream compositions for all material streams presented in Figure 2.

Stream	**Air**	**Fuel**	**1**	**2**	**Dummy**	**3a**	**3b**	**4a**	**4b**	**5a**
H_2	0.0000	1.0000	0.0000	1.0000	0.0239	0.0239	0.0239	0.0239	0.0239	0.0000
H_2O	0.0000	0.0000	0.0000	0.0000	0.3210	0.3210	0.3210	0.3210	0.3210	1.0000
O_2	0.2100	0.0000	0.2100	0.0000	0.0005	0.0005	0.0005	0.0005	0.0005	0.0000
N_2	0.7900	0.0000	0.7900	0.0000	0.6473	0.6473	0.6473	0.6473	0.6473	0.0000
NO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2O	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NO	0.0000	0.0000	0.0000	0.0000	0.0011	0.0011	0.0011	0.0011	0.0011	0.0000
N_2O_4	0.0000	0.0000	0.0000	0.0000	0.0061	0.0061	0.0061	0.0061	0.0061	0.0000

Stream	**5b**	**6**	**7**	**8**	**9a**	**9b**	**10**	**11a**	**11b**	**12**
H_2	0.0000	0.0000	0.0239	0.0000	0.0000	0.0000	0.0239	0.0000	0.0000	0.0000
H_2O	1.0000	1.0000	0.3210	1.0000	1.0000	1.0000	0.3210	1.0000	1.0000	1.0000
O_2	0.0000	0.0000	0.0005	0.0000	0.0000	0.0000	0.0005	0.0000	0.0000	0.0000
N_2	0.0000	0.0000	0.6473	0.0000	0.0000	0.0000	0.6473	0.0000	0.0000	0.0000
NO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2O	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NO	0.0000	0.0000	0.0011	0.0000	0.0000	0.0000	0.0011	0.0000	0.0000	0.0000
N_2O_4	0.0000	0.0000	0.0061	0.0000	0.0000	0.0000	0.0061	0.0000	0.0000	0.0000
Stream	**13a**	**13b**	**14a**	**14b**	**15a**	**15b**	**15c**	**16a**	**16b**	**17a**
H_2	0.0239	0.0239	0.0000	0.0000	0.0000	0.0000	0.0000	0.0239	0.0239	0.0000
H_2O	0.3210	0.3210	1.0000	1.0000	1.0000	1.0000	1.0000	0.3210	0.3210	1.0000
O_2	0.0005	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000	0.0005	0.0005	0.0000
N_2	0.6473	0.6473	0.0000	0.0000	0.0000	0.0000	0.0000	0.6473	0.6473	0.0000
NO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2O	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NO	0.0011	0.0011	0.0000	0.0000	0.0000	0.0000	0.0000	0.0011	0.0011	0.0000
N_2O_4	0.0061	0.0061	0.0000	0.0000	0.0000	0.0000	0.0000	0.0061	0.0061	0.0000
Stream	**17b**	**18**	**19**	**20**	**21a**	**21b**	**22**	**23a**	**23b**	**24**
H_2	0.0000	0.0000	0.0239	0.0000	0.0000	0.0000	0.0239	0.0000	0.0000	0.0000
H_2O	1.0000	1.0000	0.3210	1.0000	1.0000	1.0000	0.3210	1.0000	1.0000	1.0000
O_2	0.0000	0.0000	0.0005	0.0000	0.0000	0.0000	0.0005	0.0000	0.0000	0.0000
N_2	0.0000	0.0000	0.6473	0.0000	0.0000	0.0000	0.6473	0.0000	0.0000	0.0000
NO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2O	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NO	0.0000	0.0000	0.0011	0.0000	0.0000	0.0000	0.0011	0.0000	0.0000	0.0000
N_2O_4	0.0000	0.0000	0.0061	0.0000	0.0000	0.0000	0.0061	0.0000	0.0000	0.0000

Stream	25	26	27	28a	28b	29a	29b	30a	30b	31
H_2	0.0239	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.3210	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
O_2	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.6473	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2O	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NO	0.0011	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2O_4	0.0061	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Stream	**32**	**33a**	**33b**	**Flue Gas**	**CW, in**	**CW, out**				
H_2	0.0000	0.0000	0.0000	0.0239	0.0000	0.0000				
H_2O	1.0000	1.0000	1.0000	0.3210	1.0000	1.0000				
O_2	0.0000	0.0000	0.0000	0.0005	0.0000	0.0000				
N_2	0.0000	0.0000	0.0000	0.6473	0.0000	0.0000				
NO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000				
N_2O	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000				
NO	0.0000	0.0000	0.0000	0.0011	0.0000	0.0000				
N_2O_4	0.0000	0.0000	0.0000	0.0061	0.0000	0.0000				

Table B3. Magnitude of energy flow for all energy streams presented in Figure 2.

Stream	AC	Qloss	GT	HPST	IPST	LPST
Heat Flow [kJ/h]	3,040,000	-5,000	8,906,000	872,700	1,268,000	785,000
Stream	**LPP1**	**LPP2**	**IPP1**	**IPP2**	**HPP1**	**HPP2**
Heat Flow [kJ/h]	277.6	977.7	84.00	1809	142.6	2110

Table B4. Duty and log mean temperature difference for all heat exchangers presented in Figure 2.

Heat Exchanger Name	LP ECON	LP EVAP	LP SPHT	COND
Duty [kJ/h]	806,600	2,387,000	32,890	13,220.000
Log Mean Temperature Difference [ºC]	16.66	116.2	269.9	62.10

Heat Exchanger Name	**IP ECON**	**IP EVAP**	**IP SPHT**	
Duty [kJ/h]	15,550	2,907,000	782,400	
Log Mean Temperature Difference [ºC]	284.0	393.7	185.6	
Heat Exchanger Name	**HP ECON**	**HP EVAP**	**HP SPHT**	
Duty [kJ/h]	733,500	4,342,000	3,974,000	
Log Mean Temperature Difference [ºC]	595.0	851.0	798.5	

Appendix C: Efficiency Calculations

From the literature,[14] it is known that the heat of combustion of hydrogen gas (H_2) is:

$$q_{Theor.} = 286{,}000 \text{ kJ/kmol}$$

The amount of energy produced by the complete combustion of the hydrogen fed to the reactor is,

$$Q_{Theor.} = F_{Fuel} q_{Theor.} \tag{C.1}$$

$$Q_{Theor.} = (100\frac{kmol}{h})(286,000\frac{kJ}{kmol}) = 28,600,000\frac{kJ}{h}$$

The total energy produced by the four turbines is,

$$Q_{Total} = Q_{GT} + Q_{HPST} + Q_{IPST} + Q_{LPST} \tag{C.2}$$

$$Q_{Total} = 8,906,000\frac{kJ}{h} + 872,700\frac{kJ}{h} + 1,286,000\frac{kJ}{h} + 785,000\frac{kJ}{h} = 11,800,000\frac{kJ}{h}$$

The efficiency of the plant's energy production is therefore,

$$Eff. = \frac{Q_{Total}}{Q_{Theor.}} \times 100\% \tag{C.3}$$

$$Eff. = \frac{11{,}800{,}000\ kJ/h}{28{,}600{,}000\ kJ/h} \times 100\% = 41.37\%$$

However, operation of the plant requires electricity to produce electricity. This is factored into the plant's efficiency by calculating the net energy produced,

$$Q_{Used} = Q_{AC} + Q_{LPP1} + Q_{LPP2} + Q_{IPP1} + Q_{IPP2} + Q_{HPP1} + Q_{HPP2} \tag{C.4}$$

$$= 3,040,000\frac{kJ}{h} + 277.6\frac{kJ}{h} + 977.7\frac{kJ}{h} + 84.00\frac{kJ}{h} + 1809\frac{kJ}{h} + 142.6\frac{kJ}{h} + 2110\frac{kJ}{h} = 3,050,000$$

$$Q_{Net} = Q_{Total} - Q_{Used} \tag{C.5}$$

$$Q_{Net} = 11,800,000\frac{kJ}{h} - 3,050,000\frac{kJ}{h} = 8,790,000\frac{kJ}{h}$$

Therefore the efficiency of the plant when considering net energy produced is,

$$Eff_{Net} = \frac{Q_{Net}}{Q_{Theor.}} \times 100\% \tag{C.6}$$

$$Eff_{Net} = \frac{8.790.000\ kJ/h}{28,600,000\ kJ/h} \times 100\% = 30.72\%$$

Appendix D: Unit Operation Sizing and Costing

Lang Factor Adjustment to calculate cost with installation,

$$C_{total} = C_{capital}LF \tag{D.1}$$

Heat Exchanger Area Calculation for Shell and Tube (sample IP SPHT),

$$A_{HX} = \frac{Q}{U\Delta T_{LM}} \tag{D.2}$$

$$A_{HX} = \frac{782400\ J/s}{(50W/m^2C)(185.6^oC)} = 23.4\ m^2$$

Pricing is done utilizing the following correlations,

$$C = 1.218 f_d f_m f_p C_b \tag{D.3}$$

Where,

$$C_b = e^{(8.821-0.30863(ln\,A_{HX}) + 0.0681(ln(A_{HX}^2)))} \tag{D.4}$$

And,

$$f_d = e^{(-0.9816+0.0830(ln(A_{HX})))} \tag{D.5}$$

$$f_p = 0.7771 + 0.04981(ln(A_{HX})) \tag{D.6}$$

$$f_m = g_1 + g_2(ln(A_{HX})) \tag{D.7}$$

For stainless steel 316 utilized in the heat exchangers and condenser, g_1=0.8603 and g_2=0.23296.

Pressure Vessel Sizing (LP Drum),

$$d^2L = \frac{t_{sv}Q_{sv}+t_{sl}Q_{sl}}{1.4F_l} \tag{D.8}$$

Where,

t_{sv} is the residence time of saturated vapor; [d]

t_{sl} is the residence time of saturated liquid; [d]

Q_{sv} is the flow rate of saturated vapor; [B/d]

Q_{sl} is the flow rate of saturated liquid; [B/d]

F_l is the fraction of the tank filled with liquid; 0.5

$$d^2L = \frac{(130800\ B/d+723.7\ B/d)}{1.4(0.5)} = 187891\ in^2\ ft = 37\ m^3$$

Assuming a 1:1 inner length to inner diameter sizing, L_i=3.3 m and d_i=3.3 m

The thickness is calculated by,

$$T = \frac{(P_{design}S_{max})}{(2P_{max}W_{eff})-(1.2P_{design})} \tag{D.9}$$

$$T = \frac{(35737\ Pa)(172.3\ m)}{((2)(17300000\ Pa)(1.0))-((1.2)(35737Pa))} = 0.18\ m$$

Where,

P_{design} is the design pressure; Pa

P_{max} is the maximum allowable pressure; Pa

S_{max} is the maximum allowable stress; m

W_{eff} is the weld efficiency; 1.0

Pressure Vessel Costing,

$$C_e = a + bS^n \tag{D.10}$$

$$C_e = (-\ 2500) + (200)(39457^{0.6}) = \$111,973$$

Where,

C_e is the expected cost in 2010 dollars

a, b, and n are tabulated constants for horizontal pressure vessels for carbon steel

S is the mass of carbon steel required for the vessel

Extrapolation of Costs using CEPCI Index Values

$$C_P = C_O\left(\frac{CIV_O}{CIV_P}\right) \tag{D.11}$$

$$C_P = \$111,973\left(\frac{600.6}{532.4}\right) = \$126,317$$

Where,

C_P is the present cost in current USD

C_O is the original cost given by the correlation

CIV_O is the CEPCI Index Value for the year the correlation was produced

CIV_P is the current CEPCI Index Value

Operator Cost Calculation 3 operators 4 shifts,

$$C_{Op} = n_{Op} SW \tag{D.12}$$

$$C_{Op} = (3)(4)(\$60,000) = \$72,000/y$$

Where,

C_{Op} is the yearly operator costs

n_{Op} is the operators required per shift

S is the number of shifts per day

W is the yearly wage of an operator

Storage Piping Steel Mass Determination,

**Note: Piping thickness determined using the method shown in Eq. D-9. Assumed values for P_{design}, W_{eff}, and S_{max} taken from source 4 and ASME pressure vessel code.*

$$V_{steel} = \left(\pi \cdot \frac{(ID + 2t)^2}{4} \cdot L\right) - \left(\pi \cdot \frac{ID^2}{4} \cdot L\right) \tag{D.13}$$

$$= \left(\pi \cdot \frac{(1.4m + 2(0.059m))^2}{4} \cdot 2000m\right) - \left(\pi \cdot \frac{1.4m^2}{4} \cdot 2000m\right) = 538\, m^3$$

$$m_{steel} = V_{steel} \cdot \rho_{steel} \tag{D.14}$$

$$= 538\, m^3 \cdot 8000 \frac{kg}{m^3} = 4.3M\ kg\ Stainless\ Steel$$

Where,

V_{steel} is the volume of steel needed for piping

ID is the internal diameter of the piping

t is the required thickness of the pipe walls

L is the necessary length of the pipe

m_{steel} is the mass of steel needed

ρ_{steel} is the approximate density of steel

Storage Piping Cost Determination,

$$C_e = a + bS^n \tag{D.15}$$

$$C_e = -15,000 + (560 \cdot 114,000kg^{0.6}) = \$591,000$$

Where,

C_e is the expected cost in 2010 dollars

a, b, and n are tabulated constants for horizontal pressure vessels for carbon steel

S_{max} is the maximum mass of stainless steel allowed by the correlation

Thus, for each 114,000 kg "unit" of stainless steel (304SS is used in the correlation, costs are assumed similar to 316SS within a reasonable margin), the cost will be $591k in 2010 dollars. The cost must be extrapolated to current dollars (see D.11) and total cost must be determined from the total amount of steel required, as noted by D.14.

$$C_{total} = C_{unit} \cdot \frac{m_{total}}{m_{unit}} \tag{D.16}$$

$$= \$666,500 \cdot \frac{4{,}300{,}000kg}{114{,}000kg} = \$25.1M$$

APPENDIX E: Economic Tables

Table E.1: Capital Cost Totals

Costing Totals	
HX	$1,127,233.24
CS Pressure Vessels	$2,887,788.43
Operators (High Val.)	$720,000
Turbines	$10,125,883.70
Combustion Chamber	$7,199,865.57

Pumps	$167,750.00
Total Installed Costs:	**$22,228,520.94**

Table E.2: Raw Material Costs (Calculated here for offsite hydrogen purchasing)

Raw Material Costing	kg/h	kg/y	**Yearly Cost**
Hydrogen	201.6	1766016	$4,415,040.00
	m^3/h	m^3/yr	**Yearly Cost**
Water	55.88	489508.8	$3,279,708.96
		Total	**$7,694,748.96**

Table E.3: Operator wages calculation based on unit operations and shifts.

Operator Cost	# Unit	Operator Per Shift	Operator #	Yearly Wage per shift
Pump	3	0	0	$60,000
HX	10	0.1	1	Shifts
Reactor	1	0.5	0.5	4
Turb	4	0.1	0.4	
Air Comp	1	0.1	0.1	
		Total Opps	2	3
		Total Cost	**$480,000**	**$720,000**