

# CH402 Chemical Engineering Process Design

## Capstone Problem 3

### Blue Hydrogen Production by Autothermal Reforming

#### 1.0 Background

Hydrogen plays an important role to play in the global energy economy. While not an energy source itself, hydrogen is an energy carrier. It can also be used directly as a fuel and can be used to produce other chemicals and fuels. “Green” hydrogen is produced through electrolysis powered entirely by renewable energy. “Gray” hydrogen is produced from fossil fuels, primarily natural gas, through steam methane reforming, where the carbon dioxide from the process is not captured and is emitted into the atmosphere. “Blue” hydrogen is hydrogen produced from fossil fuels with low greenhouse gas emissions via carbon capture and sequestration. Most hydrogen production today is gray.

Your client, Company A, has recently completed an FEL-1 (Front-end loading) study of a Blue Hydrogen Auto-thermal Reforming (ATR) Project regarding the potential development and operation of a merchant “Blue Hydrogen” business in the U.S. Gulf Coast region to support the growing hydrogen demand from a variety of end user operations in the vicinity. Company A’s principals have approved advancing the Blue Hydrogen ATR Project into FEL-2 to:

- Determine a specific Blue Hydrogen process design to meet the business objectives.
- Estimate the fixed capital investment (FCI) and operational costs of the design.
- Assess the overall economics of this project.

#### 2.0 High-Level Project Objectives

The overall objectives of the project are to participate in the increasing demand for clean energy sources, such as hydrogen, and to provide shareholder value for Company A through distinctive, low emission investments. The primary objectives of the FEL-2 stage of the Project are to design and evaluate a process to produce 250 million SCFD of blue hydrogen from natural gas through an Auto-Thermal Reforming (ATR) reaction process coupled with CO<sub>2</sub> capture. The asset will be built and operated in the U.S. Gulf Coast region. Additional details about these process units can be found in the Process Design Considerations section.

The project team is charged to provide a feasible and optimized process design that meets the commercial objectives of the project under the constraints of the process chemistry and available technology. Where possible, the team should develop and evaluate options for various parts of the process such as separations, heat integration, and potential markets for which to place the hydrogen product. The project team will base their design and economic assessment on the process parameters and economic guidance provided in this problem statement document.

### 3.0 Auto-Thermal Reforming High-Level Process Overview

The main components of an Auto-Thermal Reforming process for Blue Hydrogen production are:

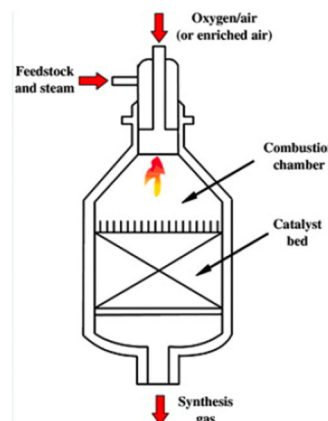
- 1) One or more Feed Pretreatment reactors to eliminate undesirable impurities in the feed,
- 2) An Auto-Thermal Reforming (ATR) reactor to heat the feed and convert methane to carbon oxides,
- 3) One or more Water Gas Shift (WGS) reactors to further convert CO to CO<sub>2</sub> and H<sub>2</sub>, and
- 4) A “back-end” separation and purification process for the product H<sub>2</sub> and byproduct gases.

The following primary reactions occur in the Blue Hydrogen process [1–3].

Reactions	Stoichiometry	$\Delta H$
<b>Pre-Reforming</b>	$C_nH_m + nH_2O \rightarrow nCO + (n + \frac{1}{2}m)H_2$	Endothermic
<b>Partial Oxidation</b>	$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$	Exothermic
<b>Autothermal Reforming</b>	$CH_4 + H_2O \rightarrow 3H_2 + CO$	Endothermic
	$CH_4 + 2H_2O \rightarrow 4H_2 + CO_2$	Endothermic
<b>WGS (Water-Gas Shift)</b>	$CO + H_2O \leftrightarrow CO_2 + H_2$	Exothermic

Since natural gas typically is not pure methane, ethane (and higher hydrocarbons) must first be converted to CO and H<sub>2</sub> before the feed enters the Auto-Thermal Reformer to avoid the creation and buildup of coke on the ATR catalyst. The Pre-Reforming reactor operations between 800-900°F and with a steam to hydrocarbon ratio of 0.8. Stoichiometric conversion of the higher hydrocarbons can be assumed, and some of the CO produced will react to CO<sub>2</sub> and H<sub>2</sub> (via the WGS reaction).

Because reforming of methane is endothermic, the ATR operates at very high temperatures (~2000°F) to drive the reforming reactions forward. To provide the heat required for main reforming reactions, some of the natural gas is first partially oxidized with pure O<sub>2</sub> prior to the Steam and Methane passing over a reforming catalyst. Oxygen feed is limiting in the methane combustion portion of the reactor at a rate sufficient to achieve the reaction temperatures required for the ATR reactions. Because the ATR also operates with a steam to hydrocarbon ratio of roughly 0.8, the hot ATR effluent can be utilized to generate the steam from Boiler Feed Water (BFW) for both injection into the process and for other heating requirements through a heat integrated process.



To increase the yield of methane to hydrogen, the ATR effluent is cooled and sent to one or two Water Gas Shift (WGS) reactors. Because the WGS reaction is exothermic, the reaction is favored at lower

temperatures and is equilibrium limited. Addition Steam is provided to achieve a Steam to Dry Gas ratio of 0.5 in the WGS section [4].

The primary product of the ATR/WGS reactions,  $H_2$ , is typically separated from other reaction byproducts (unreacted  $CH_4$ ,  $H_2O$ ,  $CO$ ,  $CO_2$ , etc.) using a Pressure Swing Absorption (PSA). The  $H_2$  must meet commercial specifications and will be delivered by pipeline to its end users. The design must include compression and cooling to achieve the required injection pressure and temperature of the  $H_2$  pipeline. Choose the most ideal order of separation to optimize your process. The remaining gas, known as the “tail gas,” is then processed to recover  $CO_2$  at 70+% recovery. The  $CO_2$  must be compressed to pipeline injection pressures and either shipped in a third-party pipeline to be permanently sequestered at an EPA-certified storage facility, or it can be sold into other markets provided it meets the industry specifications, available volumes, and profitability for such markets.

Figure 1 shows a high-level view of the ATR Blue Hydrogen Process.

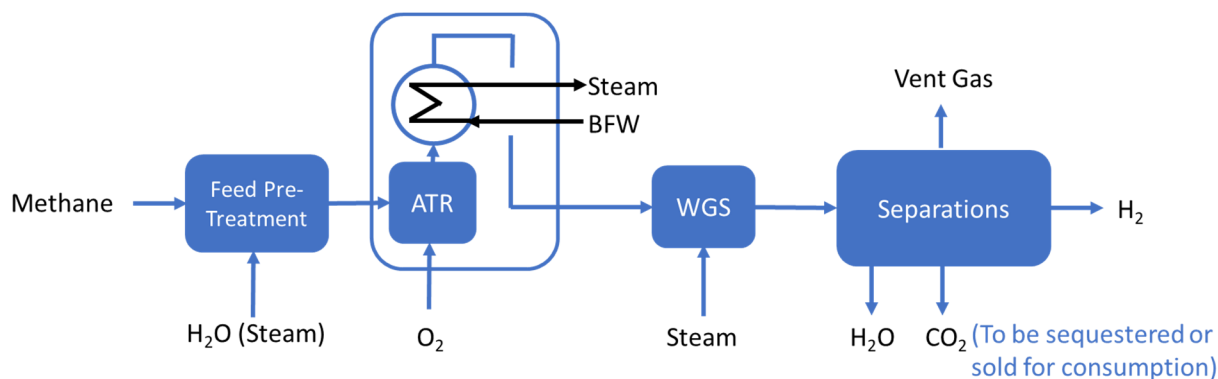


Figure 1: High-level process for Blue Hydrogen ATR production

The team should evaluate the capital and operating expenses balance between using the  $CO_2$  lean stream for recycle into the ATR process and utilizing this gas for process heat (while still meeting the greater than 70%  $CO_2$  recovery target). This may include recycling any remaining ( $CO_2$  lean) process gas that has a residual heating value and could be utilized elsewhere in the process for overall energy efficiency of the process. Note that gases combusted in other locations can negatively impact the carbon intensity (CI) of the process, and to maintain a low CI, additional carbon capture may be required (to greater than 70% overall  $CO_2$  recovery).

#### 4.0 Project Design Basis

In preparing and evaluating your design, use the following basis/assumptions:

- The Blue Hydrogen plant must produce hydrogen to meet commercial contract requirements averaging 250 MMSCFD on an annual basis.
- The natural gas feedstock is 97.0%  $CH_4$  and 3.0%  $C_2H_6$ , and it is available from the local utility at 700psi and 60°F.

- There is a trace amount of sulfur-containing species that must be removed prior to entering the ATR reactor as sulfur is a poison to the ATR Catalyst (assume 3ppm H<sub>2</sub>S in the natural gas feed). This should be handled via a ZnO unit, and does not require a detailed process model, but sizing, operating strategy (i.e., sparing), and costs need to be considered.
- Because there are other hydrocarbons in the methane feed, a pre-reformer reactor is needed to address the ethane impurity.
- The Auto-Thermal Reforming reactor includes a combustion zone followed by a fixed bed reforming zone with a Ni catalyst.
- Oxygen is supplied by a third-party gas company, assumed to be 100% pure O<sub>2</sub>, and is available at 700psi and 60°F.
- The High Temperature Shift (HTS) reactions occur in a fixed bed reactor over a Fe catalyst with an inlet temperature in the 650 - 750°F range.
- The Low Temperature Shift (LTS) reactions occur in a fixed bed reactor over Cu/Zn/Al catalyst with an inlet temperature of 360-400°F and at least 35°F over the dew point of the feed.
- All conversion processes occur adiabatically.
- The pressure drops across each reactor must be estimated.
- Hydrogen will be separated from the other process gases using Pressure Swing Absorption (PSA). In this service, PSAs can typically achieve 85% recovery to a purity of 99.9+ % H<sub>2</sub>.
- At least 70% of the CO<sub>2</sub> must be recovered either by
  - Using a vent gas scrubber and stripped from the CO<sub>2</sub> rich solution deploying a commercially available monoethanolamine (MEA) solvent,
  - Through a cryogenic separation process,
  - Via an electrochemical process that is viable and economical at this scale, or
  - Via any other viable, economical separation technology.
- CO<sub>2</sub> must meet the purity specifications of the party receiving this stream whether that be consumer of the CO<sub>2</sub> or a carbon sequestration service.
- If your process has a CO<sub>2</sub> recovery less than 95%, please add a discussion about how you would improve CO<sub>2</sub> recovery in the future.

The process design must additionally include a process to capture at least 70% of the CO<sub>2</sub> produced in the overall manufacturing of the hydrogen. CO<sub>2</sub> which is captured and sequestered is eligible for a 45Q tax credit, which was recently increased to \$85 per (metric) tonne of CO<sub>2</sub> sequestered as part of the Inflation Reduction Act signed in 2022. Alternatively, a 45V tax credit (of up to \$3/kg H<sub>2</sub>) can be applied depending upon the overall carbon intensity of the H<sub>2</sub> production process (45V tier to be calculated by the team based on the Carbon Intensity of the process). Only one of these credits (45Q or 45V) can be used [5–8].

Because of the uncertainty of handling a steam imbalance (either a need to procure steam or to sell excess steam), the process design must strive to be balanced with respect to steam production and steam consumption. Any steam needed that is in above what can be generated through heat integration of the hot flue gas must be purchased at 1.5x the heating value of the steam, and any excess steam generated can only be sold for 0.5x its heating value. Should the design include any excess power generation, any excess power generated can only be sold back to the utility at 0.9x the cost of power

purchased from the local power company. Note that efficient heat integration is required to achieve a low CI process. Refer to examples in literature of “Pinch Analysis” [9,10].

FEL-1 studies determined preliminary values of key operating conditions that FEL-2 should initially be based on. These process conditions from FEL-1 are shown in Table 1 below.

*Table 1: Key process design conditions from the FEL-1 Study*

Process Parameter	Value	Units	Comments
Sulfur Adsorbent Bed-Inlet Temperature	650	°F	
Sulfur Adsorbent Bed-Inlet Pressure	680-700	PSI	
Pre-Reformer-Inlet Temperature	800-900	°F	
Pre-Reformer-Inlet Pressure	660-680	PSI	
Pre-Reformer Steam/Hydrocarbon Ratio	0.8	Ratio	Molar basis
ATR Reactor- Inlet Temperature	1200	°F	
ATR Reactor- Outlet Temperature	1800-2000	°F	To be optimized in design
ATR- Inlet Pressure	600	PSI	
HT WGS Steam/Dry Gas Ratio	0.5	Ratio	Molar basis
High Temperature (HTS) WGS Reactor Inlet Temperature	TBD	°F	To be optimized in design
Low Temperature (LTS) WGS Reactor Inlet Temperature	TBD	°F	To be optimized in design
Shifted Gas Entering Separations Section	TBD	PSI	To be optimized in design

These conditions may be changed in FEL-2, but an economic justification for alternative operating conditions should be provided. For other process conditions not specified, use the references provided or other data from your own literature search.

Other relevant design considerations taken from the FEL-1 study are listed in Table 2 below.

*Table 2: Design Considerations from the FEL-1 Study*

Process Parameter	Value	Units	Comments
Natural Gas Supply Pressure	700	PSI	At 60 F
O <sub>2</sub> Pipeline Supply Pressure (100% O <sub>2</sub> )	700	PSI	At 60 F
Hydrogen Pipeline Injection Pressure	900	PSI	At 120F max
CO <sub>2</sub> Pipeline Injection Pressure	900	PSI	At 120F max, gas phase
Boiler Feed Water Supply Conditions	900	PSI	Supplied by vendor
	200	°F	
Cooling Water Supply Conditions	70	PSI	Supplied by vendor
	70	°F	
Cooling Water Return Conditions	40 min	PSI	Supplied by vendor
	110 max	°F	
Heat Exchanger Approach Temperature	15	°F	Minimum for design

## 5.0 Economics

### Revenues and Operating Costs

- Company A's marketing department estimates that Hydrogen produced from operations is sold at 2x its fuel value.
- CO<sub>2</sub> Sequestration Tax Credit from either:
  - The 45Q Regulation at \$85/tonne (metric ton) for any CO<sub>2</sub> captured and sequestered. CO<sub>2</sub> sold to another industry must have market price that has a documented justification
  - The 45V regulation at up to \$3/kg H<sub>2</sub> depending upon the calculated overall carbon intensity of the hydrogen production process.
- Historical power and natural gas pricing data in the U.S. Gulf Coast area should be used to forecast future prices using an appropriate commodity forecasting model based on your historical data.
  - Estimate current power costs in the U.S. Gulf Coast area from historical data.
  - Natural gas prices are to be estimated from historical data in the U.S. Gulf Coast Area.
- Boiler Feed Water (BFW) is provided by a 3rd party at \$3.50/ thousand gallons. Make-up BFW requirements (due to losses) are 1% of the total usage rate of steam used in heat exchangers; Water condensed from the steam used in the process can be resold to the BFW provider for 80% of the BFW costs.
- Cooling water (CW) is also provided by a 3rd party supplier at \$0.20/thousand gallons.
- Oxygen is priced at \$34/tonne (Metric Ton).
- The project must estimate the costs of operating staff and include a 1.5 factor for benefits.
- Other fixed operating costs (maintenance, insurance, and plant overheads, etc.) are assumed to be 4% of you ISBL FCI on an annual basis.

### Capital Costs

- Purchased/Bare Equipment Cost may be estimated using methods in your process design course, 3rd party costing tools, or by methods in other appropriate sources.
- The fully installed costs of the equipment by an appropriate method for FEL-2 should include:
  - Appropriate delivery costs if the equipment is not purchased in the U.S. Gulf Coast area.
  - A contingency of 20% for this stage of the project.
- Assume the equipment will be purchased in the first year of construction (see timeline below). Use the CECPI to adjust FCI costs to this date.
- Hydrogen will be discharged directly into a pipeline provided by a 3rd party; Include \$2m for on-site connection and metering equipment. Shipping costs for the H<sub>2</sub> are paid by the consumer
- CO<sub>2</sub> will be discharged directly into a pipeline provided by a 3rd party; include \$2m for on-site connection and metering equipment; Company A must also pay a \$0.50/thousand SCF CO<sub>2</sub> shipping tariff.
- All pumps and other rotating equipment with a Bare Equipment Costs less than \$100,000 (e.g. low-pressure blowers) must be spared for overall asset reliability.
- Land is purchased immediately before construction starts; The project must estimate the amount and cost of land needed.

- Off-sites/OSBL capital cost is assumed to be 10% of ISBL FCI for your process design.
- Working Capital is 5% of the ISBL FCI and is required immediately before start-up.

### **Economic Analysis**

- Use MACRS Depreciation for a 15-year asset class.
- Economic analysis should run through the complete depreciation of the asset.
- Use an Income Tax Rate of 35%.
- Discounted Cash Flow Analysis should assume:
  - FEL-3 starts January 1 of next year, takes one year to complete, and costs 8% of FCI,
  - Construction starts January 1 of the following year and takes 3 years,
  - Operations commence on January 1 three years after construction begins,
  - For discounting cashflows, assume  $t = 0$  to be the start date of FEL-3.
- Inflation – use 2% inflation applied to your BFW costs (and sales of your process water), cooling water, and your CO<sub>2</sub> transportation tariff (if used).
- Assume your plant operates 350 days each year on average. This accounts for periodic outages every few years for Turnaround activities and catalyst changes.
- Assume that the 45Q or 45V tax credits apply whether you sequester your CO<sub>2</sub> or sell your CO<sub>2</sub> for commercial uses/other industries.
- The Weighted Average Cost of Capital (WACC) for Company A is 15%. Company A uses the WACC as their discount rate for its analysis of large economic investments.
- The team should determine the NPV (at 15%) and IRR for the project.
- If the IRR of the project is less than 15%, then determine what the Hydrogen fuel value factor needs to be (i.e., greater than 2x) to achieve a 15% IRR.

### **6.0 Environmental, Health, and Safety:**

Environmental, Health and Safety (EHS) aspects are critical to the economic viability, sustainability and social responsibility of business and operations. These aspects must be carefully considered during design to ensure that processes minimize raw material consumption, safely contains process materials, and effectively treat potentially harmful discharges prior to release to the environment.

These aspects are especially important when processes employ or produce toxic and flammable materials. Your design will be judged in part on your recognition of the potential environment, health and safety hazards inherent in the process, as well as on the mitigation steps you incorporate to ameliorate these potential hazards.

#### **6.1 Minimizing Environmental Impacts**

- a. You should assume that regulatory authorities require the application of *Best Available Control Technology (BACT)* to minimize and treat gaseous and liquid waste generated by the process.
- b. Prepare a detailed process and instrumentation diagram (P&ID) for ATR reactor. Your P&ID should include automatic controllers for pressure, level, and composition (temperature) management as well as alarms informing operators of abnormal conditions.

- c. Size a pressure relief for this ATR reactor assuming an onsite flare is available for the safe and complete combustion of vented material.
- d. Finally, tabulate the annual failure rate of ATR reactor automatic controllers and alarmed indicators based on literature data. This analysis builds confidence that your design is reliable, minimizing off-spec production and flaring.

## 6.2 Assessing and Mitigating Potential Health Impacts

Your design must recognize the hazards associated with potential human exposure to process materials, including raw materials, intermediate and finished products, by-products and wastes. To recognize hazards, compile a table of OSHA chemical exposure limits, NFPA diamond classifications, and lethal dose (LD50) limits for process chemical constituents. This information is often found on Safety Data Sheets (SDS), formerly known as Material Safety Data Sheets MSDS. Screening for potential health risks is an important element of process safety that informs personnel protective equipment and standard operating protocols.

## 6.3 Safety – Learning from Experience

The design must consider relevant lessons learned from the industry, especially with respect to hydrocarbon explosions.

- a. Highlight the risks in hydrocarbon processing by performing a TNT equivalency calculation for the atmospheric detonation of all chemicals from the largest, by inventory. For worst-case consideration assume all gaseous contents instantly vent to atmosphere.
- b. Review and identify appropriate hydrogen safety resources from H<sub>2</sub>Tools.org
  - [Safety Planning for Hydrogen and Fuel Cell Projects \(latest version\)](#)
  - [Example Hydrogen Safety Plan - September 2020](#)
  - [Hydrogen Safety Checklist](#)
  - [Hydrogen Incident Examples, Select Summaries of Hydrogen Incidents from the H2tools.org Lessons Learned Database](#)
- c. List appropriate safety codes and standards
  - OSHA PSM (29 CFR 1910)
  - NFPA 55 for CO<sub>2</sub>, CO
  - NFPA 2 for H<sub>2</sub>, enriched Natural Gas
  - NFPA 497 for hazard area classification

## 6.4 Process Hazard Assessment

A thorough risk analysis is essential to implementation, approval, and operation of a project such as Blue Hydrogen. The risk analysis evolves in diligence and detail through each project stage and is a critical gate for advancement of the project from one stage to the next. A formal, facilitated process risk analysis draws on the collective expertise and experience of a dedicated team comprised stakeholders including designers, contractors, operators, process safety experts, and regulatory officials. A thorough risk analysis process necessarily involves a significant time and resource investment and will apply the hazard analysis methods appropriate for each project stage.

To assist the FEL-2 Safety risk analysis team, build a preliminary risk analysis table using the template provided in Appendix A. Apply the known FEL-2 process design and the safety information gathered in



sections 6.2 and 6.3. To build the risk analysis table, follow the example given in the first rows of the template to:

- Populate the 'Hazard' column with the key process hazards that can present with each major process stream, informed by SDS and safety information resources
- List up to 3 likely causes for each Hazard in the "Cause" column
- Use the Appendix A Frequency and Severity matrices to estimate how often the hazard might occur (Frequency) and how bad the hazard may impact safety (Severity).
- Propose safeguards in order of effectiveness that may eliminate or reduce risk of the hazard (preventative, mitigative)

## 7.0 Additional Information

Below are some additional details to assist with the design and economics of the project.

### Process Design Information

- Sulfur ( $\text{H}_2\text{S}$ ) Removal/ $\text{ZnO}$  Unit
  - Several papers/references [11] discuss the purpose and technologies for this purification step.
  - A detailed kinetic reactor model is not needed for this unit, but a simplified model may be useful for the Material and Energy Balance.
  - This unit must be accounted for in the PDF, in the FCI, and in the overall economics.
  - Assume the following:
    - Assume 3ppm  $\text{H}_2\text{S}$  in Natural Gas feed
    - The  $\text{ZnO}$  bed operates as follows:  $\text{ZnO} + \text{H}_2\text{S} \rightarrow \text{ZnS} + \text{H}_2\text{O}$
    - To avoid sulfur breakthrough, the bed operates up to 80% of its absorptive capacity.
    - $\text{ZnO}$  catalyst are 98%  $\text{ZnO}$  and 2%  $\text{Al}_2\text{O}_3$ , have a filing density of 81 lbs/ft<sup>3</sup>, and costs \$350/ft<sup>3</sup> of catalyst.
- For the pre-reforming reactor, assume the following:
  - Stoichiometric conversion of the ethane can be assumed, and some of the  $\text{CO}$  produced will react to  $\text{CO}_2$  and  $\text{H}_2$  (via the WGS reaction).
  - The catalyst requirement for the pre-reformer is one ft<sup>3</sup> for every 1,500 SCFH of dry gas throughput.
  - The catalyst costs \$600/ft<sup>3</sup> and has a 4-year life expectancy.
- For the ATR reactor, assume the following:
  - There are two zones in the ATR reactor: a combustion zone and a reforming zone.
    - The combustion zone operates to stoichiometric depletion of the oxygen.
    - The reforming zone operates at a 20°F approach to equilibrium.
  - The catalyst requirement for the ATR reforming zone is one ft<sup>3</sup> for every 4,500 SCFH of dry gas throughput.
  - The catalyst costs \$3,400/ft<sup>3</sup> and has a 4-year life expectancy.
  - The combustion zone is 25% of the reforming zone and does not require any catalyst.

- For the High Temperature Shift (HTS) reactor, assume the following:
  - An inlet temperature in the 650 - 750°F range.
  - The conversion operates at a 20°F approach to equilibrium.
  - The catalyst requirement for the HTS reactor is one ft<sup>3</sup> for every 4,000 SCFH of dry gas throughput.
  - The catalyst costs \$500/ft<sup>3</sup> and has a 4-year life expectancy.
- For the Low Temperature Shift (LTS) reactor, assume the following:
  - An inlet temperature of 360-400°F and at least 35°F over the dew point of the feed.
  - The conversion operates at a 20°F approach to equilibrium.
  - The catalyst requirement for the LTS reactor is one ft<sup>3</sup> for every 3,000 SCFH of dry gas throughput.
  - The catalyst costs \$600/ft<sup>3</sup> and has a 4-year life expectancy.
- If a single, Medium Temperature Shift reactor is determined to be optimal, interpolate the reactor requirements between the HTS and LTS data above.
- All conversion processes occur adiabatically. The pressure drops across each reactor must be estimated.
- For the Pressure Swing Absorption (PSA), assume an 85% recovery to a purity of 99.9+ % H<sub>2</sub>. Size the PSA unit to have the appropriate number of beds, the proper amount of absorbent, a spare bed, and the vessels and instrumentation. Use Towler and Sinnott, Chapter 10 or other appropriate resource to size and cost your PSA Equipment
- Heat Integration Considerations:
  - To reduce both capital costs for heat exchangers and operating costs (steam, fired heater fuel, cooling water costs, etc.) you must implement as much cross exchanging of process streams as possible to reduce the amount of utilities (steam or cooling water) or fuel (H<sub>2</sub>) you need to burn in a furnace. See references on Pinch Analysis [9,10].
  - Design note: To avoid generating CO<sub>2</sub> emissions, H<sub>2</sub>, produced by the process, could be used as a fuel for fired heaters/furnaces operating at temperatures around 1200 °F. If you chose to use H<sub>2</sub> for fired heaters/furnaces, the plant will need to be sized to include this fuel requirement in addition to the required H<sub>2</sub> for commercial sales (250 MMSCFD).

## Economic Information

- Catalyst Treatment/Costs:
  - In industry, catalysts are generally treated as a Capital Expense or as an Amortized Expense. This can be complicated to do in spreadsheet models.
  - For this economic model, treat your catalyst costs as an operating expense.
  - For example, if you spend \$4m for catalyst that has a 4-year life, then charge your operating costs with \$1m/year for catalyst expense. You do not need to inflate the cost of future catalyst purchase.
- MEA Usage/Costs (if you use an Amine absorber/stripper in your process to capture CO<sub>2</sub>):
  - CO<sub>2</sub> capture using monoethanolamine (MEA) in an absorber/stripper configuration is a well-established process [12,13].

- MEA is generally not consumed, but it does degrade to the point where you must effectively replace the in-process hold-up annually.
- Estimate the MEA in-process inventory using a physical basis that you define.
- Find an MEA cost on-line or from an SME.
- Do not escalate the purchase of make-up MEA over time (no inflation).
- In your report, be sure to explicitly state
  - Total FCI and the costs that make it up
  - Average annual operating income (excluding taxes and depreciation) for the first year of operation - Why? This demonstrates whether the operation makes money on a day-to-day basis (excluding the need to pay back the capital investment). This needs to be positive to even try to pay off the asset investment.

## 8.0 Notes

1. The prefix “m” means 1,000 unless it is used with currency where “\$m” is millions and “\$k” is thousands of dollars. The prefix “mm” or “MM” means 1,000,000.

## 9.0 References

- [1] Tariq R, Maqbool F, Abbas SZ. Small-scale production of hydrogen via auto-thermal reforming in an adiabatic packed bed reactor: Parametric study and reactor’s optimization through response surface methodology. *Comput Chem Eng* 2021;145:107192. <https://doi.org/10.1016/j.compchemeng.2020.107192>.
- [2] Murmura MA, Diana M, Spera R, Annesini MC. Modeling of autothermal methane steam reforming: Comparison of reactor configurations. *Chem Eng Process Process Intensif* 2016;109:125–35. <https://doi.org/10.1016/j.cep.2016.08.019>.
- [3] Halabi MH, de Croon MHJM, van der Schaaf J, Cobden PD, Schouten JC. Modeling and analysis of autothermal reforming of methane to hydrogen in a fixed bed reformer. *Chem Eng J* 2008;137:568–78. <https://doi.org/10.1016/j.cej.2007.05.019>.
- [4] Katebah M, Al-Rawashdeh M, Linke P. Analysis of hydrogen production costs in Steam-Methane Reforming considering integration with electrolysis and CO2 capture. *Clean Eng Technol* 2022;10:100552. <https://doi.org/10.1016/j.clet.2022.100552>.
- [5] 26 USC 45V: Credit for production of clean hydrogen n.d. <https://uscode.house.gov/view.xhtml?req=granuleid:USC-prelim-title26-section45V&num=0&edition=prelim>.
- [6] Fakhry R. Implementing the 45V Clean Hydrogen Tax Credits n.d. [https://www.eesi.org/files/Rachel\\_Fakhry\\_060123\\_Slides.pdf](https://www.eesi.org/files/Rachel_Fakhry_060123_Slides.pdf).
- [7] Financial Incentives for Hydrogen and Fuel Cell Projects n.d. <https://www.energy.gov/eere/fuelcells/financial-incentives-hydrogen-and-fuel-cell-projects>.
- [8] Zacarias M, McGeady C. How the 45V Tax Credit Definition Could Make or Break the Clean

Hydrogen Economy n.d. <https://www.csis.org/analysis/how-45v-tax-credit-definition-could-make-or-break-clean-hydrogen-economy>.

- [9] Linnhoff B. Pinch analysis : a state-of-the-art overview : Techno-economic analysis. Chem Eng Res \& Des 1993;71:503–22.
- [10] Pinch Technology: Basics for Beginners - Heat Transfer - Articles - Chemical Engineering n.d. <http://www.cheresources.com/content/articles/heat-transfer/pinch-technology-basics-for-beginners>.
- [11] Frilund C, Simell P, Kaisalo N, Kurkela E, Koskinen-Soivi M-L. Desulfurization of Biomass Syngas Using ZnO-Based Adsorbents: Long-Term Hydrogen Sulfide Breakthrough Experiments. Energy & Fuels 2020;34:3316–25. <https://doi.org/10.1021/acs.energyfuels.9b04276>.
- [12] Soltani SM, Fennell PS, Mac Dowell N. A parametric study of CO<sub>2</sub> capture from gas-fired power plants using monoethanolamine (MEA). Int J Greenh Gas Control 2017;63:321–8. <https://doi.org/10.1016/j.ijggc.2017.06.001>.
- [13] Jung J, Jeong YS, Lim Y, Lee CS, Han C. Advanced CO<sub>2</sub> capture process using MEA scrubbing: Configuration of a split flow and phase separation heat exchanger. Energy Procedia 2013;37:1778–84. <https://doi.org/10.1016/j.egypro.2013.06.054>.

## 10.0 Appendix A – Template for Process Safety Hazards/Risk Analysis

Hazard	Cause	Frequency (See Table A1)	Severity (See Table A2)	Safeguards (Preventative, Mitigative)
Asphyxiation from release of CO <sub>2</sub>	<ul style="list-style-type: none"> <li>• Loss of containment (mechanical leaks, carbonic acid corrosion)</li> <li>• Improper venting of waste CO<sub>2</sub></li> <li>• Inadequate dilution ventilation</li> </ul>	Remote	Catastrophic	<ul style="list-style-type: none"> <li>• Proper CO<sub>2</sub> / MEA service materials (stainless steel) in CO<sub>2</sub> removal process</li> <li>• Designed to minimize stagnant water in piping</li> <li>• dehydrate CO<sub>2</sub> stream prior to pipeline</li> <li>• Waste gas vertical stack plume away from building intakes per code</li> <li>• Fixed gas monitors for CO<sub>2</sub> with alarms and connection to control room</li> <li>• Portable detection devices for personnel working in proximity</li> <li>• Area ventilation for heavier than air gas</li> <li>• Standard operating procedures for inspection, service, maintenance, Management of Change (MoC)</li> <li>• Warning signs, training re</li> </ul>

Hazard	Cause	Frequency (See Table A1)	Severity (See Table A2)	Safeguards (Preventative, Mitigative)
				asphyxiation threat
Ignition, fire in high purity oxygen stream components	<ul style="list-style-type: none"> <li>Internal particle impact on materials exceeds ignition energy</li> <li>Residual oils from assembly or maintenance</li> <li>Loss of containment (leaks)</li> <li>Incompatible materials</li> </ul>	Remote	Serious	<ul style="list-style-type: none"> <li>Materials, components recommended, rated for O2 service</li> <li>O2 service clean per O2 industry practice</li> <li>Particle filtration at source and after particle generators</li> <li>Design for lower stream velocity</li> <li>Standard operating procedures for inspection, service, maintenance, management of change (MOC)</li> <li>Warning signs, training</li> </ul>
<i>hazard</i>	•			•
<i>hazard</i>	•			•
<i>hazard</i>	•			•
<i>hazard</i>	•			•
	•			•
	•			•
	•			•

**Table A1: Hazard Frequency** assessment

Category	Description
<b>Frequent</b>	Expected to occur more than once each year at the plant
<b>Occasional</b>	Expected to occur more than one in ten years at the plant

<b>Remote</b>	Expected to occur once in the life of the plant
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**Table A2:** Hazard **Severity** assessment

Category	Safety	Public	Environmental	Economic
<b>Negligible</b>	First Aid	Inquiry	Recordable event	\$0-\$50,000
<b>Serious</b>	Hospitalization	Injury/illness	Land or water contamination	\$50,000-\$500,000
<b>Catastrophic</b>	Death	Hospitalization	Permanent land or water contamination	>\$500000