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In This Issue

Volume 8 | Number 32

SPECIAL FEATURES

- 04** **Letter from the Editor**
Karl Kolmetz
- 05** **Assessment of Techno-Economic Feasibility of Green Hydrogen Production Using Low Carbon Electricity**
Rupsha Bhattacharyya
- 10** **Update for the South East Asia Oil and Gas Industry**
- 11** **Update for the American Oil and Gas Industry**
- 13** **The Underbelly of Ethanol Blends and Battery Electrical Vehicles**
Jayanthi Vijay Sarathy
- 16** **Managing Corrosion in Sour Water Stripping Units**
Dr. Marcio Wagner da Silva
- 22** **Rock Bottom View**
Ronald J. Cormier
- 25** **Renewable Energies as Preventive Solution of Climate Change And Clean Renewable future Energy**
Hamid Reza Seyed Jafari , Seyed Mohammad Reza Seyed Jafari
- 27** **Guidelines for Propylene Splitters**
Timothy M. Zygula, Karl Kolmetz, Johor Bahru

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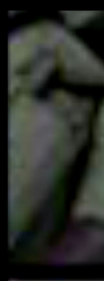
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Letter from the Editor



Things that should be obvious, but are not obvious to many people

Adapted from Charlie Munger.

1. A simple way to get what you want, it to be worthy of what you are seeking. In your life or business deliver to the world what you would buy if you were on the other side of the table. People who have this physiology win in life, not just money and honors, they win respect. The deserved respect of the people they deal with, and there is a huge pleasure in life of getting deserved respect.
2. Gaining knowledge is the key to long term success and it is an ongoing challenge. The knowledge and success you have today will not sustain you for the rest of your life. People who are successful long term are constantly gaining knowledge.
3. Learn the main concepts from many fields. Famous quote from Cicero, "A man who does not know what happened before he was born, goes through life as a child." You should know the main concepts of history, finance, philosophy, engineering, medicine, law and others or you will go through life as a child. **It is not enough that you gain knowledge to get a passing grade on a test, you need to learn them like your future success depends on them, because it does.**
4. Risk management is the minimization of things that might cause a business or endeavor to fail, and we conduct studies to determine what might cause a business or process to fail. What are some things in life that might cause you to be unsuccessful? Sometimes it is hard to find the right keys that might make you successful, but it is easy to find the things that might make one unsuccessful. So, what are things that will make you fail in life. Two easy choices are laziness and being undependable. If you are unreliable, it does not matter what other great qualities you have, you are on the road to failure. Therefore, do faithfully what you are engaged to do should be an automatic part of your conduct.
5. In your life you will have terrible problems, terrible blows of bad luck, expect them they are coming. They will be horrible and unfair. Some people recover and some people never recover. We need to adapt the attitude of Epictetus, every mischance in life is an opportunity to behave well, every mischance in life is an opportunity to learn something. Your goal is not to dwell submerged in self-pity, but to utilize the issue in a constructive fashion.
6. In life you need to hope for the best but be prepared for the worst. No one really wants to go though life anticipating trouble, but you need to anticipate trouble and be adequately prepared to perform if it happens.

We believe that IAPCE can give you the main concepts from engineering, you will need to study other fields as well. We have modules on each of the main engineering fields for you to master.

All the best in your career and life,
Karl

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Assessment of Techno-Economic Feasibility of Green Hydrogen Production Using Low Carbon Electricity

Rupsha Bhattacharyya

INTRODUCTION AND CONTEXT

Rising concerns about anthropogenic climate change, global warming, air pollution (due to use of fossil fuels) have brought to the forefront the need for a massive and economy wide energy transition. There is an urgent need for shifting to clean raw materials, feed-stock, energy forms and low or zero carbon energy vectors across all kinds of industries. One of the promising options for this transition is the use of low carbon hydrogen (often referred to as 'Green Hydrogen' in common parlance) to substitute the use of fossil fuels or their derivatives in specific hard-to-decarbonize sectors such as chemicals synthesis, ammonia and fertilizer production, metallurgical operations, heavy and long-distance transport, particularly freight transport, shipping and aviation.

To be classified as green, the hydrogen must be obtained from a carbon free feed stock like water using very low carbon energy forms such as renewable, hydro or nuclear electricity such that its green credentials can be truly maintained. But depending on the energy form used, the life cycle economics of the process can vary greatly and thus a formal decision regarding which kind of green hydrogen production would be adopted at a given location depends not only on the availability of clean electricity but also on the differing costs associated with harnessing clean electricity and converting it to hydrogen. This work looks at this issue and compares the cost of hydrogen from commercial scale water electrolyzers using different forms of clean energy.

THE PROCESS TECHNOLOGY

The process (shown in Fig. 1) for producing green hydrogen by water splitting involves electrolysis of a 30% potassium hydroxide solution in demineralised water in compact electrolyzer stacks consisting of several cells connected in series. The electrolyte mixed with hydrogen on one side and oxygen on the other are cooled and separated in dedicated gas liquid separators and are then sent through catalytic recombiners or purifiers to remove trace amounts of oxygen from hydrogen and vice versa. This ensures hydrogen purity of greater than 99.9% (v/v) at the plant battery limit. Demineralised water must be replenished from time to time to make up for its consumption during electrolysis and it is therefore a part of the operating expenses of the plant. The electrical power required for electrolysis is calculated as the product of the current required (which in turn depends on the desired hydrogen production rate, as governed by Faraday's laws of electrolysis) and the decomposition voltage required for the electrochemical reaction which is governed by the current density, pressure, and temperature at which the electrolysis is to be carried out.

MATHEMATICAL MODEL AND DATA

The techno-commercial model for calculation of levelized hydrogen production cost is described in this section. A simplified schematic of a water electrolyser plant is shown in Fig. 1.

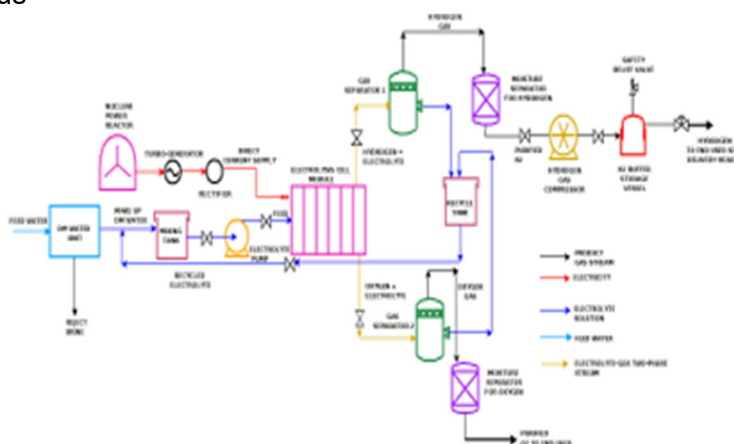


Fig. 1: Schematic diagram of water electrolyser plant coupled to a low carbon electricity source

Let the capital cost (overnight costs and interest during construction included) of the electrolyser stack and the balance of plant components (e.g., gas-liquid separators, purifiers, tanks, piping, utilities, electrical and electronic components, etc. in Fig. 1) be C_{stack} and C_{BOP} and their design life be n_{stack} and n_{BOP} years respectively. For a weighted average cost of capital or project discount rate of $d\%$ per annum, the annualized life cycle capital investment for the entire electrolyser plant is given by

$$C_{plant} = C_{stack} \times CRF_{stack} + C_{BOP} \times CRF_{BOP} \quad (1)$$

where the capital recovery factor (or CRF, calculated separately for the stack and the balance of plant due to different design lives) is expressed as [1]

$$CRF = \frac{d(1+d)^n}{(1+d)^n - 1} \quad (2)$$

For an electrolyser plant capacity of $Q \text{ Nm}^3/\text{h}$ of H_2 , the annual electricity consumption for electrolysis is expressed as

$$E = Q \times E_{sp} \times 8760 \times AF \quad (3)$$

where AF is the annual capacity factor of the electrolyser plant and the factor E_{sp} indicates specific energy consumption in water electrolysis (in $\text{kWh}/\text{Nm}^3/\text{h}$), which depends on type of electrolyser, electrolyte, electrodes, operating temperature, pressure and current density [2].

The annual electricity cost for electrolysis is

$$E_{cost} = E \times C_{elec} \quad (4)$$

where C_{elec} is the clean electricity generation tariff. Therefore, annualized green hydrogen production cost is calculated as

$$LCOH = \frac{C_{plant} + E_{cost} + OM + F}{Q \times 8760 \times AF} \quad (5)$$

where the term OM represents annual expenditure on plant maintenance and other overheads and F represents the feedstock (i.e., high purity water) cost. From reaction stoichiometry of water splitting, about 9 kg of fresh water must be electrolyzed per kg H_2 produced.

If the hydrogen plant of capacity Q_1 has a total capital cost of CC_1 in a given base year 1, the capital cost of a plant of capacity Q_2 in another year 2 is expressed by the following cost scaling law, using the reported values of Chemical Engineering Plant Cost Index (CEPCI) for the respective years [1]:

$$CC_2 = \left(\frac{CEPCI_2}{CEPCI_1} \right) \times CC_1 * \left(\frac{Q_2}{Q_1} \right)^n \quad (6)$$

The value of the exponent n in the above equation is generally taken to be between 0.7 to 0.85 since the water electrolyser technology under evaluation is mature and commercially well established [3]. In this study, 0.7 is the value used for the balance of plant components (which are conventional equipment/components) and 0.85 is used for the electrolyser stack (which is a modular component now available in standardized configurations). The base model is a 1.25 MW(e) electrolysis stack which can produce up to 250 Nm^3/h of hydrogen. Other data are provided in Table 1.

RESULTS AND ANALYSIS

The calculated costs based on data in Table 1 are shown in Table 2. A few points immediately become apparent from these results. Even though solar and wind electricity cost much less than nuclear electricity in many parts of the world, they are available for much shorter duration of time each day. This means that the electrolyser which is only drawing power from the renewable field operates for few hours per day and that too with variable capacity. Thus, the annualized cost of hydrogen goes up, simply because the electrolyser operates for low capacity or does not operate at all for a significant number of hours per day. Wind turbine capacity factors are somewhat higher on average compared to solar PV capacity factors. Nuclear power reactors on the other hand, operate at high capacities in base load mode for practically their entire lifetime, thus the electrolyser coupled to it also operates at nearly rated capacity throughout and produces hydrogen at lower cost. This analysis shows that for a nation that wishes to produce green hydrogen in bulk quantities, renewable hydrogen alone will not be the most cost-effective routes to achieving that. If nuclear or hydel power plants are available, they should also be leveraged to produce a part of the national green hydrogen demand to achieve optimal costs. In nations without nuclear power,

Table 3 shows that for renewable driven green hydrogen, operating cost contributes a far greater share to the annualized cost of the produced hydrogen whereas the reverse is true for nuclear driven hydrogen production. This follows directly from the fact that nuclear electricity is somewhat costlier than renewable electricity owing to a variety of factors (market conditions, policies, and incentives, etc) and thus it makes up a greater share of the cost of hydrogen. Even though this factor is of secondary importance in the overall decision-making process, it does indicate that reduction of electrolyser costs is the most important consideration for bringing down renewable derived hydrogen.

	Solar H ₂ cost (\$/kg)	Wind H ₂ cost (\$/kg)	Nuclear H ₂ cost (\$/kg)
Electrolyser @250 Nm ³ /h	19.39- 23.80	15.36- 17.48	9.18- 9.47
Electrolyser @500 Nm ³ /h	11.64- 14.16	9.83-11.03	7.03- 7.19
Electrolyser @1000 Nm ³ /h	7.28-8.69	6.71-7.39	5.81- 5.91

Table 2: Estimated costs of clean hydrogen production using different sources of electricity (at 8% discount rate)

	Solar H ₂ cost components	Wind H ₂ cost components	Nuclear H ₂ cost components
Stack capital cost	62.90%	48.70%	21.90%
Balance of plant capital cost	15.30%	11.90%	5.30%
Operating cost	21.80%	39.40%	72.80%

Table 3: Estimated contributions of various components to green hydrogen production cost

CONCLUDING REMARKS

There are multiple routes to obtaining green hydrogen via water electrolysis, each with its own techno-commercial implications. Even though renewable energy technologies such as solar PV modules and wind turbines have seen massive price reductions, their intermittency and diurnal and seasonal variability mean that electrolysers working solely with them as the power source will operate at low average capacity factors and thus the hydrogen price will be higher. If the same electrolyser were to be coupled with a nuclear or

hydel power plant, they will experience better utilisation factors and hence lower life cycle costs. Another factor to keep in mind is that while renewable electricity can be harnessed even in arid regions without the need for a water source nearby, renewable hydrogen production at the same sites may be severely constrained by water availability considerations. Nuclear reactors or hydel power projects are sited considering long-term water availability from nearby natural water bodies and are therefore also well suited as sites of large-scale hydrogen production. Thus, just like the electricity mix must include contributions from all low carbon electricity sources, so must the green hydrogen mix available to a nation for deep decarbonization be diversified. This is to ensure affordability and supply security of green hydrogen as well.

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Update for the South East Asia Oil and Gas Industry

May 2022

KEY DRIVING FACTORS

- Russia's invasion of Ukraine forces nations to focus more on energy security and temporarily away from energy transition.
- USA release of 1million bpd has helped moderate crude pricing.
- Petrochemical margins are squeezed as feedstock prices remain high.
- Most SEA countries begin opening-up borders with minimal requirements.

HIGHLIGHTS: SINGAPORE

Malaysia land border reopens from 1st April. The world's busiest land border that caters to more than 300,000 daily travellers.

South East Asia refiners crack spread spike due to higher fuel demand during holiday period and reopening of movement and region filling the gap created due to Russian shortfall. Profit margins for complex refineries in Singapore, the bellwether for Asian refiners, tipped over US\$20 a barrel.

Some refineries are looking at opportunistic crudes to further improve their margins in lieu of the high crude price.

Most Asian crackers plan or have cut rates to between 60-90% due to the naphtha-ethylene spread. This is especially for non-integrated sites.

Exxonmobil to supply sustainable aviation fuel (SAF) as part of Singapore pilot. The Civil Aviation Authority of Singapore (CAAS), Singapore Airlines (SIA), and Singapore-headquartered global investment company Temasek have selected ExxonMobil to supply and deliver sustainable aviation fuel (SAF) in Singapore.

MALAYSIA

Schlumberger Announces Collaboration with PETRONAS to Explore Opportunities in Sustainability, Digitalization, Internet of Things and Joint Technology Developments.

INDONESIA

Indonesia's Pertamina to double down on geothermal energy. Oil company taps binary technology as part of \$8.3bn renewables push.



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Update for the American Oil and Gas Industry

May 2022

KEY DRIVING FACTORS

Most of the supermajor Oil & Gas companies have a relationship with Russia. Russia's move into Ukraine will provide a basis for changes:

- BP will exit its shareholding Russian Company Rosneft.
- Norway's State-Controlled Oil company, Equinor will also exit their \$ 1.2 Billion stake in Russia.
- ExxonMobil has billions of dollars at stake in Sakhalin Island in Russia. Exxon has said that they will no longer invest in Russian Projects.
- Total Energies said that they will no longer finance projects in Russia.

American crude oil production was relatively flat and down 1.3 million barrels per day (mb/d) compared with its highest levels in late 2019 and early 2020. The U.S. was a petroleum net importer of 1.6 mb/d, including crude oil and refined products. By contrast, the U.S. was a petroleum net exporter of 0.8 mb/d in January 2021.

HIGHLIGHTS

U.S. Exporters of LNG are benefiting from increased sales to Europe. This is, of course, at the expense of Russia.

- Cheniere is working through France's Engie to supply more LNG from its export plant in Corpus Christi, TX. Bechtel has been given an EPC Contract for the Stage 3 expansion there.
- Europe's call for U.S. Gas is fast-tracking two or more Gulf Coast LNG Projects: New Fortress Energy off the coast of Louisiana, and a second Mexican Project Vista Pacifica LNG.

Continental Resources plans to spend \$ 250 Mil to create a Carbon Sequestration project in the Mid-western US. The CO₂ will come mostly from Ethanol Units.

ExxonMobil Baytown Texas plans a hydrogen Plant, plus a Carbon Capture and Sequestration operation. This is more than double the current capacity and will help XOM meet Greenhouse gas emissions by 2050.

Cleco announced a \$ 900 Million carbon capture project to be located NW of Alexandria, Louisiana. The project will be called the Diamond Vault.

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The Underbelly of Ethanol Blends and Battery Electrical Vehicles

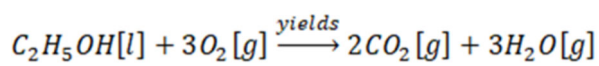
Jayanthi Vijay Sarathy

With climate change and green energy narratives becoming the central themes of the 21st century energy industry, the following article makes a quantitative argument of the aspects overlooked and why ethanol blending, and battery electrical vehicles would fail to meet green energy objectives.

DOES ETHANOL HELP REDUCE CO₂ EMISSIONS?

While shrouded in political fumes, the price of ethanol blended fuels depends on both prices of crude oil and sugar rich crops, such as sugar cane, sugar beet and corn. While the former is independent of seasonal changes, the later depends on annual weather patterns and the water table. A chief parameter that has a direct effect on fuel consumption is calorific value., i.e., how much energy is contained in a unit of fuel and how much can be extracted with technology.

Performing a mass balance between ethanol and gasoline with the assumption of complete combustion, the stoichiometry is as follows,



$$46.07 \text{ kg } C_2H_5OH \xrightarrow{\text{yields}} 88 \text{ kg } CO_2$$

$$\text{Or, } 1 \text{ kg } C_2H_5OH \xrightarrow{\text{yields}} \frac{88}{46.07} = 1.91 \text{ kg } CO_2$$

Taking ethanol density at 0.789 kg/lit,

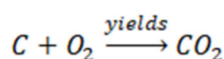
$$0.789 \frac{\text{kg}}{\text{lit}} \rightarrow \frac{1}{0.789} = 1.2674 \frac{\text{lit}}{\text{kg}}$$

i.e., 1.2674 litres of C₂H₅OH gives 1.91 kg CO₂.

Therefore relating a unit volume of ethanol to mass of CO₂,

$$1 \text{ lit } C_2H_5OH = \frac{1.91}{1.2674} = 1.507 \text{ kg } CO_2$$

Performing a similar analysis for gasoline with a typical density of 0.75 kg/lit and a carbon content of 85% in 1 litre of gasoline,



$$12 \text{ kg Carbon} \xrightarrow{\text{yields}} 44 \text{ kg } CO_2$$

$$\text{Or, } 1 \text{ kg Carbon} \xrightarrow{\text{yields}} \frac{44}{12} = 3.667 \text{ kg } CO_2$$

Therefore, with 85% carbon content,

$$0.75 \times 0.85 = 0.6375 \text{ kg } C \text{ per litre of Gasoline}$$

Or,

$$0.6375 \frac{\text{kg}}{\text{lit}} \xrightarrow{\text{yields}} 3.667 \times 0.6375 = \sim 2.33 \text{ kg } CO_2$$

From an energy perspective, LHV is a better indication of a fuel's useful heat since the combustion products are above the boiling point of water. Taking a typical lower heating value (LHV) of 26.9 MJ/kg for ethanol and 44.3 MJ/kg for gasoline, for every litre of gasoline combusted to attain the same mileage, 44.3/26.9 = 1.647 times more ethanol is needed. Therefore the CO₂ produced for the same mileage would be,

$$1.647 \times 1.507 \text{ kg } CO_2 = \sim 2.482 \text{ kg } CO_2$$

Summarizing, for a given unit of mileage, the CO₂ produced is ~2.408 kg/lit while with gasoline, the CO₂ produced is ~2.33 kg/lit.

From the analysis made, it is clear that,

1. There is no significant improvement in reducing CO₂ emissions between combusting ethanol and gasoline. Increasing the ethanol content in fuel blends would only cause motorists to purchase more fuel for the same mileage.
2. Ethanol is an agricultural product and any seasonal variations due to weather or famines and droughts, is bound to create price fluctuations to consumers.
3. Subsidies tend to encourage more sugar rich crops which can offset production of other types of crops resulting in inflationary pressures.
4. When climate change is a cause of concern, it would be imprudent to depend on the uncertainty of climate for energy security.

BATTERY OPERATED VEHICLES FOR HEAVY TRANSPORTATION

Taking a Lithium-ion battery at 100 kWh with an energy density of 0.16 kWh/kg, and an expected mileage of 30 kWh/100 miles, and gasoline with a calorific value of 44,300 kJ/kg [44,300/3600 = 12.30 kWh/kg] with average density of 0.75 kg/lit [2.84 kg/USG].

For Gasoline vehicles, the energy density in kWh/lit is,

$$\text{En. Density} = 12.3 \frac{\text{kWh}}{\text{kg}} \times 0.75 \frac{\text{kg}}{\text{lit}} = 9.23 \frac{\text{kWh}}{\text{lit}}$$

Taking an average vehicular mileage of 15 km/lit, the kWh for every km for a gasoline vehicle is,

$$\text{Gasoline Vehicle Capacity} = \frac{9.23}{15} = 1.625 \frac{\text{kWh}}{\text{km}}$$

The quantity of gasoline required to deliver 100 kWh

$$\text{Gasoline Volume} = \frac{100}{9.23} = 10.834 \text{ litres}$$

$$\text{Gasoline Weight} = 10.834 \times 0.75 = \sim 8.13 \text{ kg}$$

Comparing with battery electric vehicles (BEV), the BEV mileage would be

$$\text{BEV Mileage} = \frac{30}{100 \times 1.6} = 0.1875 \frac{\text{kWh}}{\text{km}}$$

But the battery pack weight would be,

$$\text{BEV Weight} = \frac{100 \text{ kWh} \times \text{kg}}{0.16 \text{ kWh}} = 625 \text{ kgs}$$

Therefore, to conclude,

1. To deliver 100 kWh of energy, ~8.13 kg of gasoline [with a calorific value of 44.3 kJ/kg] is required, whereas with a battery pack at 0.16 kWh/kg, the weight added to the vehicle would be 625 kgs, i.e., nearly 77 times increase in weight.
2. Considering the case of heavy transportation, with chassis weight, body components, poorly maintained roads, road traffic, weather issues such as rain & snow causing temperature variations in battery performance, an erratic power supply sources to charge and weight of passengers with goods, then both the size and weight of the vehicular battery pack weight would also increase drastically.
3. This would cause the battery pack to be unable to deliver the required power since most of the power would be wasted to overcome the weight of the goods/inventory and the vehicle itself.

From the basic mass and energy balance performed, the author would like to convey, that the shortcomings that plagues the world's

energy transition effort is in the "Energy Density" of the fuel sought after. What is required is a source of fuel which is highly energy rich for a given mass of fuel.

Nuclear, LNG, diesel, gasoline, jet fuel and natural gas offer high energy densities while renewable sources such as solar, wind are very diluted sources of energy requiring more efforts to concentrate them. Towards this, significant efforts need to be made to enhance the energy density by many folds.

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Vijay Sarathy holds a master's degree in Chemical Engineering from Birla Institute of Technology & Science (BITS), Pilani, India and is a Chartered Engineer from the Institution of Chemical Engineers, UK. His expertise over 12 years of professional experience covers Front End Engineering, Process Dynamic Simulation and Subsea/Onshore pipeline flow assurance in the Oil and Gas industry. Vijay has worked as an Upstream Process Engineer with major conglomerates of General Electric, ENI Saipem and Shell.



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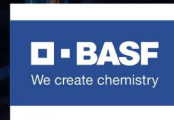
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Managing Corrosion in Sour Water Stripping Units

Dr. Marcio Wagner da Silva

INTRODUCTION AND CONTEXT

Just like any other chemical processes industry, oil refining presents a great environmental impact. Through decades engineers, scientists and researchers have dedicated efforts to minimize the environmental footprint from petroleum refining.

Some of the major impacts produced by crude oil processing are water and atmospheric emissions.

Aiming to keep under control the environmental impacts in the crude oil refining sector, some processing technologies were developed over the years and installed in the refining hardware. Nowadays, it's impossible to think in the downstream sector without the environmental processes units due to the current environmental requirements and the performance and reliability of these units are fundamental to the refiners strategy to achieve most profitable and cleaner operations. Taking into account the current regulations and the necessity to reduce the environmental impact of the crude oil refining industry, ensure the reliability and availability of environmental process units is fundamental to the players of the downstream industry and the corrosion processes are great threat to this objective. Due to his characteristics and process conditions, the corrosion phenomenon in sour water stripping

units is a special concern in the crude oil refining industry.

SOUR WATER STRIPPING TECHNOLOGIES – AN OVERVIEW

The petroleum derivatives production needs a large amount of water for cooling fluid, steam generation or to direct use in the process like in the crude oil desalting step. Water has become an increasingly scarce resource and any effort dedicated to reducing the volume applied in the process is welcome.

One of the most important environmental process units in a petroleum refinery is the so-called sour water stripping unit. Sour water is the water that had contact with the petroleum or his derivatives during some step in the process, this contact can be like rectification steam in distillations columns or in contact with hydrocarbon phases. Contaminants like NH_3 and H_2S tends to concentrate in the aqueous phase, so the sour water commonly has high concentration of these compounds.

The Sour Water Stripping Unit apply the concept of fluid rectification with steam and the partial pressure reduction to move the phase equilibrium to the vapor phase, releasing the contaminants from the liquid, like presented in Figure 1.

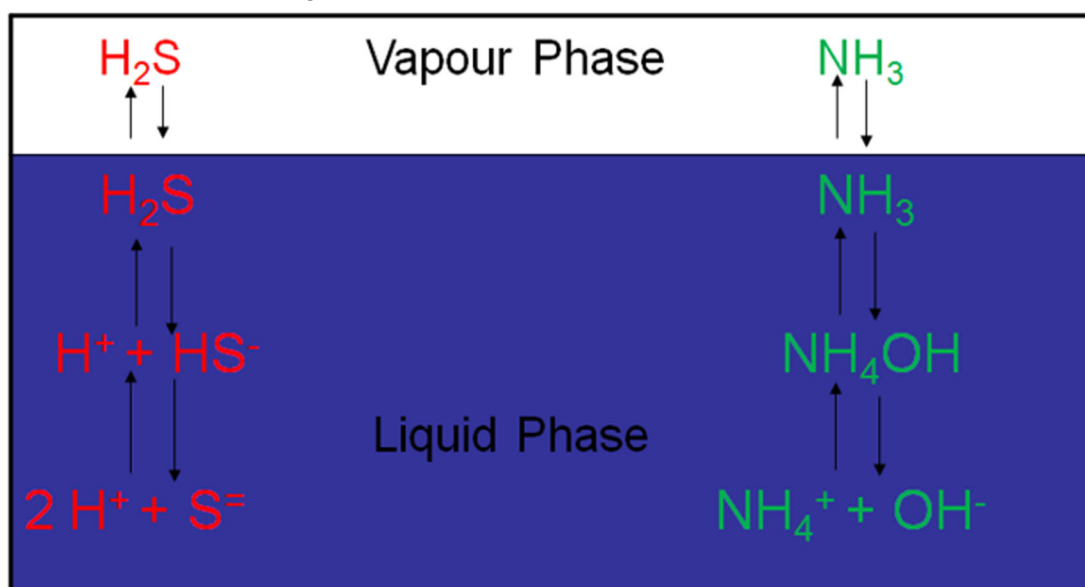


Figure 1 – Phase Equilibrium in Sour Water Stripping Process

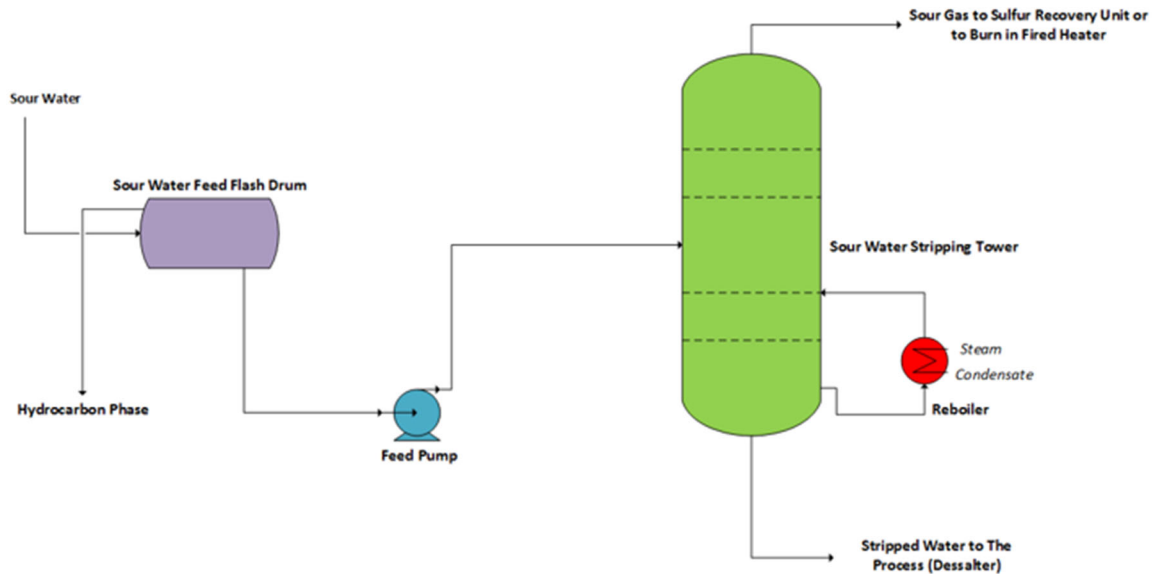


Figure 2 – Typical Arrangement for Sour Water Stripping Unit with Single Tower

Like any other process technology, the sour water stripping unit was developed and improved along the time, mainly to reduce atmospheric emissions and to raise the water reuse in the refineries.

The initial design concept for sour water stripping units had one rectifying tower, in this tower both contaminants (NH_3 and H_2S) were removed and form the stream called sour gas like described in Figure 2.

In these cases, the tower operates with relatively low pressure (about 1,0 kgf/cm²).

Initially, the designs predict to send the sour gas to burn in fired heaters, like in distillation units. Nowadays, with the environmental restrictions and the necessity to reduce SO_x and NO_x emissions the project concept was changed and the sour gases are directed to sulfur recovery units with a chamber to convert the NH_3 to N_2 , this is necessary to avoid that the NH_3 prejudice the H_2S conversion in elemental sulfur through Claus process.

The modern designs rely upon the installation of two towers, one for the H_2S removal and the second for the NH_3 removal like described in Figure 3.

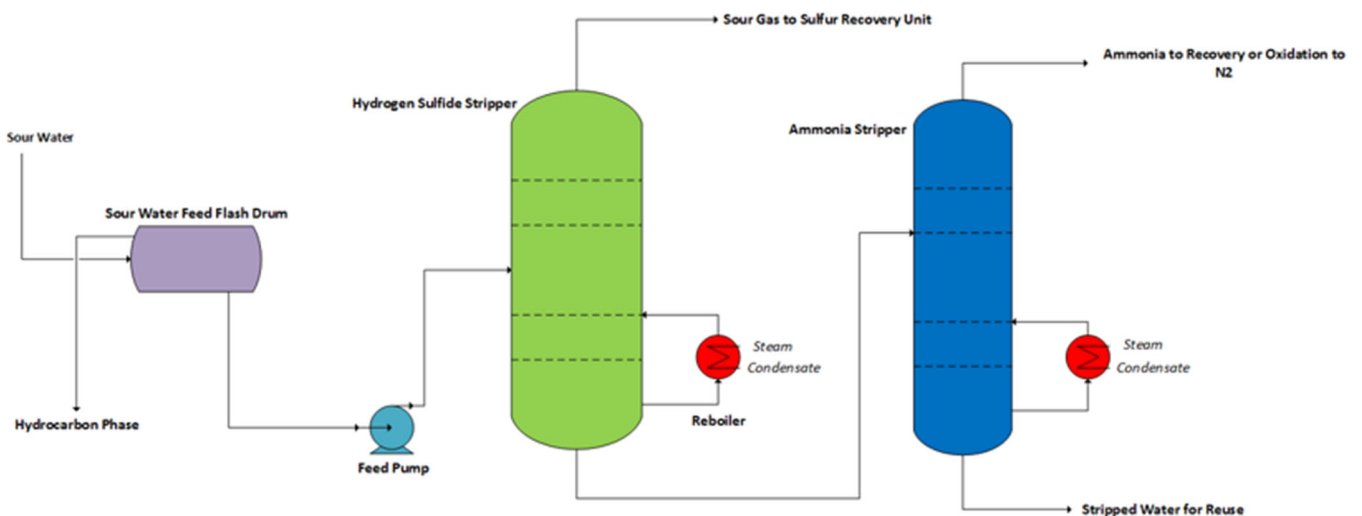


Figure 3 – Typical Arrangement for Sour Water Stripping Unit with Two Towers

For units with two towers, the H₂S rectifier operates under pressures about 5 to 11 kgf/cm², while the ammonia rectifier operates under pressures about 1 to 2 kgf/cm².

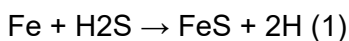
The arrangement with two towers show some advantages in relation to the project with a single tower, once that allows higher recovery of H₂S like elemental sulfur, reducing the SO_x emissions. Furthermore, the design with two towers allows recover the ammonia present in the sour water or converts this stream to N₂.

As a disadvantage in comparison with the single tower design, we can mention the higher initial investment, higher energy consumption and increased operational complexity.

CORROSION PROCESS IN SOUR WATER STRIPPING UNITS

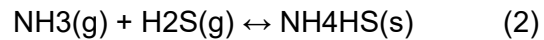
The corrosion process in sour water stripping units normally occurs through hydrogen attack, salts deposition (Ammonium Bissulfide and Ammonium Chloride), and hydrogen attack. The main regions under corrosion attack are the heating system of bottom columns and top sections, as presented in Figure 4.

The feed systems of the sour water stripping units tend to suffer corrosion by hydrogen attack. The hydrogen attack mechanism involves the contamination of steel by hydrogen (H₂) leading to the risk of failure, especially in periods of instability such as stopping and starting the process unit. Reaction 1 presents the atomic hydrogen formation in regions with presence of hydrogen sulfide (H₂S).



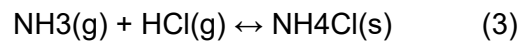
The main failure mechanisms provoked by hydrogen attack are the cracking induced by hydrogen, carbon steel embrittlement due to molecular hydrogen formation, and stress corrosion induced by hydrogen attack.

The main corrosion process observed in the sour water stripping units is related to ammonium salt deposition (bisulfide and chloride). The corrosion process resulting from the salts deposition is associated with the contaminants present in the charge currents of the sour water units. The processing of currents containing sulfur and nitrogen leads to the formation of H₂S and NH₃ in the unit's outlet currents as recycle gases, such gases can be combined to produce ammonium bisulfide (NH₄HS), according to reaction 2.



The concentration of salts formed depends on the content of contaminants in the unit's load, so sour water from units dedicated to processing bottom barrel streams tend to show more severe processes of corrosion by this mechanism.

The ammonium chloride formation process is similar, as presented in reaction 3.



The salt deposition leads basically to two corrosion processes in the units, in regions with low flow speed it's observed corrosion under salt deposits while in regions with high flow speed the most significant corrosion process is the corrosion-erosion mechanism. Due to these characteristics, the salts concentration tends to occur in the top of stripping towers, as presented in Figure 4.

The ammonium bisulfide tends to accumulate and provoke corrosion after the top condensers, while the ammonium chloride tends to suffer deposition in the top of the columns, as presented in Figure 4.

Among the actions to keep under control the corrosion process associated with ammonium salts deposition the design of the process

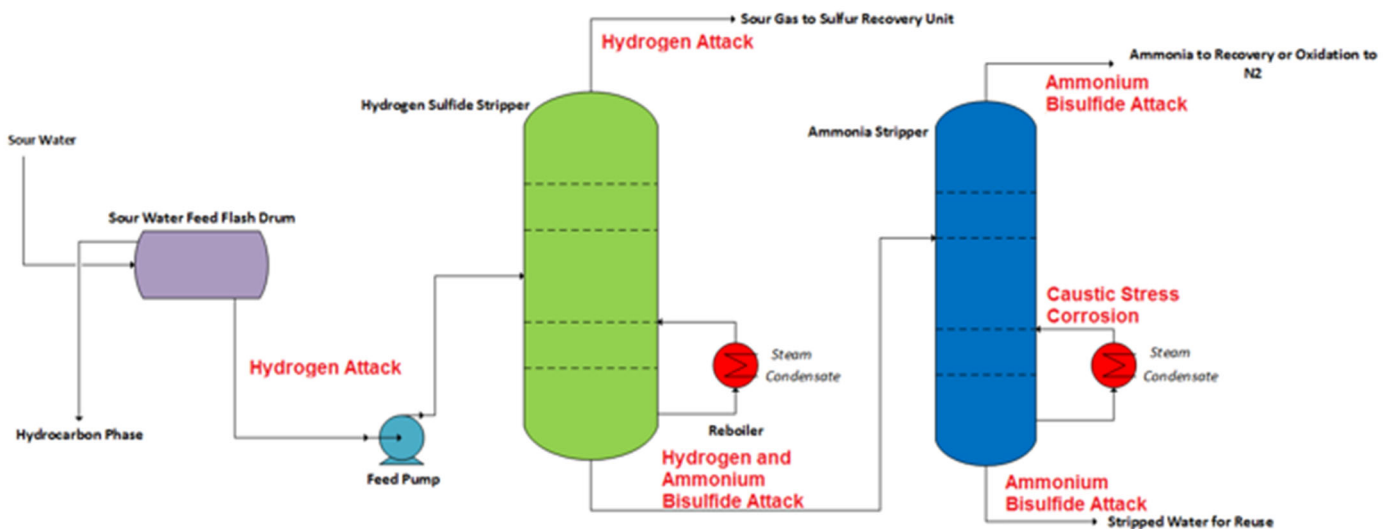


Figure 4 – Predominant Corrosion Mechanisms in Sour Water Stripping Units

units needs to consider the flow speed in the systems, a recommended range is 3,5 to 7,0 m/s. The injection of wash water is another key parameter to ensure salt removing and corrosion prevention, the water needs to be injected upstream of top condensers aiming to reduce the salt concentration.

To avoid salt deposition in sour water stripping units, especially ammonium bisulfide, its need considering the NH_4HS precipitation curve, as presented in Figure 5.

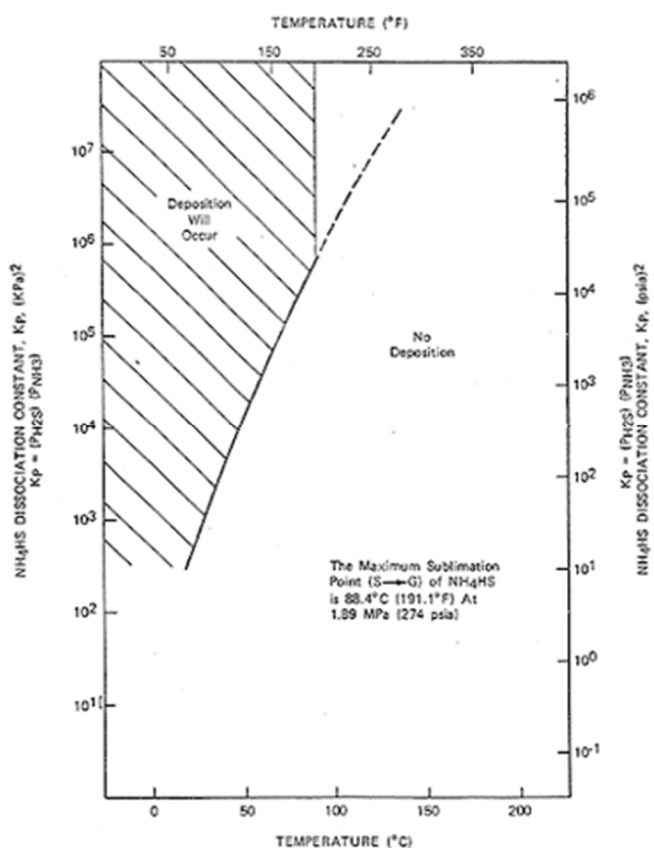


Figure 5 – NH_4HS Precipitation Curve (The Hendrix Group)

In the great part of the sour water stripping units, the deposition temperature of ammonium bisulfide is between 45 and 80 °C, and the operating temperature of the stripping sections are controlled above these values to avoid NH_4HS deposition. The corrosion under deposits tends to occur with salts concentration higher than 2,0 % in mass and with low flow speed (< 3,5 m/s), while the corrosion-erosion process tends to occur with high flow speed (> 7,0 m/s).

In the heating section of NH_3 stripping tower it's common to observe caustic stress corrosion process. This phenomenon tends to occur in welded connections that do not suffer

thermal treating to stress relief, the main susceptible materials are the carbon steel and stainless steel (300 series). Some refiners tend to inject caustic solution to achieve adequate pH to NH_3 stripping (close to 10), this practice need to take into account the risks of failures related to caustic stress corrosion and the design of the process units needs to consider the thermal treatment to stress relief of the welded connections in the bottom section of NH_3 stripping columns.

CONCLUSION

The availability of the refining hardware is a key parameter to ensure the economic sustainability of the refiners, especially those inserted in highly competitive markets, this is especially true to the environmental units like sour water stripping.

Despite the efforts over the years the crude oil refining industry still presents great environmental impact, as well as crude oil derivatives are fundamental to sustain economic development. In this sense, the availability of the environmental units is fundamental to allow sustainable operation of the refining hardware in economic and environmental point of view. The unavailability of environmental processing units, like sour water stripping, can lead to the reduction in the processing capacity of the refining hardware aiming to keep under control the atmospheric emissions and leading to great economic losses and the risk of a shortage of crude oil derivatives in the market, in extreme cases and the corrosion processes are among the main threats to the reliability and availability of the sour water stripping units. It's always important taking into account the relevance of the environmental units to the refining hardware and the optimization and maintenance priorities of these units need to be put in the same level of processing units.

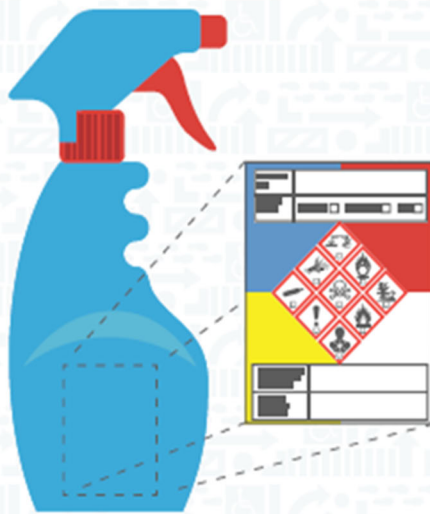
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Rock Bottom View

Ronald J. Cormier, *Engineering Practice* Contributing Author



Well, howdy again from the porch on this old central Texas ranch. Summer is around the corner, and in this part of the western world, that means heavy energy consumption for air conditioning, making for civilized living amidst brutal summer temperatures. Since we were together last time in March, the winds of global events have changed dramatically. War abroad, energy instability, inflationary aspects (read higher priced food, clothes, building products, automobiles, and economic inequality) and conflicting sovereign strategies remain high barriers to progress. Going forward, these geopolitical effects are being directly experienced in our global energy industry.

Of these implications, the Russia/Ukraine war looms largest, with huge ramifications for energy markets. Energy security has jumped to the top of the agenda. Especially, European economies (significantly supplied by Russian Btu's) want to lessen their dependence on imported fuels.

How will the Russia/Ukraine war change the geopolitical landscape? There will be seismic consequences however the war plays out. Before February 26th, geopolitics and global trade were about 'the West' versus China; now it's the West versus Russia and China. Moreover, Russia's goal to limit NATO expansion has backfired spectacularly, with Sweden and Finland now preparing to join NATO. President Emmanuel Macron's re-election in France April 23rd is another boost to solidarity both for NATO and Europe.

This intrigue is a profound shift. Europe will never rely on Russia for most anything again for the foreseeable future; energy is front and center of this loss of trust.

And what about energy policy? Policymakers must balance sustainability, security, and affordability, though supply security has jumped to the top of the agenda. European economies

want to decrease dependence on imported fuels, especially coming from their east. Countries with domestic oil and gas production will maximize the extraction of molecules. Signs of usual higher-cost extraction tiers are already underway (UK's post-invasion pivot on North Sea offshore, Canadian oil sands, and various deep-water deposits, are but a few examples).

Sustainability can also boost energy security, though it will take time. Countries will double down on low-carbon energy supply; more attention also needs to be paid to managing demand. Affordability is a societal and political challenge. Some governments have already utilized subsidies in different forms to cushion consumers from spiraling prices. But there is no easy answer to sustained high prices (no, Joe Biden doesn't not set gasoline pump prices).

Has the war breathed new life into fossil fuels? It's been a brutal reminder that the world still relies on oil, gas, and coal for 80% of its energy needs. Before the war, we expected this decade to be an "ok" decade for fossil fuel producers, perhaps a last hurrah before the world moves towards a 1.5-degree pathway. It's now shaping up to be a bonanza; the flip side of that is that high prices will accelerate the shift to low-carbon energy security.

Europe is committed to rapidly minimizing Russian imports that now meet about one-third of its demand. The proposed Nord 2 pipeline capacity increase was abruptly halted when war was on the horizon, further worsening tight supply.

There's nowhere near enough alternative gas supply available for the next four years until new volumes of LNG from the US and Qatar, increase. Meanwhile, for Europe, it's about maximizing pipeline imports from Norway,

Azerbaijan, and North Africa; outcompeting Asia for flexible LNG; and managing demand. Throughout this period, Russia has leverage and can manipulate volumes; after 2026, prices should ease. Hence natural gas prices will remain high until 2026 at least. Europe wants to rapidly minimize Russian imports that now meet about one-third of its demand. There's nowhere near enough alternative gas supply available for the next four years until new volumes are commercial.

Lastly, high oil prices could lead to some incremental upstream investment mainly on short-cycle projects (longer-term ROIs are jittery vs. still-evolving auto electrification displacement), and more supply. The crisis is super-bullish for power markets as electrification will become central to energy security of LNG from the US and Qatar become available.

End result.... Coal hasn't been this promising since the Industrial Revolution. High gas prices mean coal's outlook is brighter. But thermal coal is less influential than it was in the power system and load factors are already close to being maxed out. Gas-to-coal switching has supported carbon prices in the EU and UK *emissions trading system* (ETS) after a wobble at the start of the war. The popular view remains that carbon market development lifts the ETS price to over US\$100/t by 2030 (about twice the estimate just since 2015).

Energy shortages across the world are forcing desperate nations to abandon their climate promises in favor of keeping people warm this winter. And to be fair, everyone knew China's flimsy commitment to green energy would come second to economic growth. Overall, there's little surprise in such behavior. At the end of the day, people need energy. And they are going to get it from the cheapest source possible unless forced to do otherwise. That said, the willingness of some governments to jump directly back into coal power is still somewhat sadly amusing.

Now, it's unfair to lay all the blame at the feet of the green energy movement. The well-intentioned environmentalists aren't totally to blame here. If anything, the world's addiction to cheap natural gas is behind the domino effect of shortages we're seeing. Green energy's main role in this drama was convincing the public that renewables and natural gas were more than enough for the world.

With that in mind, we started shutting down coal plants under the assumption that things would stay the same. If the last few years

have taught us anything, it's that assumptions can be deadly. Bearish investors are patting themselves on the back over this one. While the rest of the world thought the green energy revolution was right around the corner, others saw it for what it was: wishful thinking.

American miners need to seize this opportunity. The EIA's short-term outlook tells you everything you need to know: By the end of 2021, the U.S. burned through 588 million short tons (MMst) of coal. That's a steady 10% increase over 2020. So far in 2022, forecasts are calling for yet another 10% jump in consumption, over 2021. And nearly all of it will be headed straight into power plants to stockpile against the coming winter.

But if coal is making a resurgence, the U.S. is going to get a piece of it. And domestic coal miners just caught one of the luckiest breaks they've seen in years. Despite Trump's commitment to the "China virus," it was Australia that took the blame for raising concerns about China's role in COVID-19. So, in return, China completely shut down its imports of Australian coal. Millions of tons are still taking up space in warehouses along China's coast, frozen in place until customs decide what to do with them.

The U.S. is practically drooling at the opportunity to become China's coal hookup. After reporting almost zero coal exports in 2020, the U.S. is looking at a fire sale in the making now. In 2021-2022, the top five export destinations for US coal were/are India, China, Japan, the Netherlands, and Korea.

There's no denying it now: Coal is staging a comeback, and it carries with it a stern warning about putting too many eggs in one basket. Natural gas supplies are heavily dependent on international supply chains, and renewables are limited by inconsistency. Failing to prepare a backup for when gas prices shoot up and the wind stops blowing means the world has no choice but to turn back to coal.

We will chat again in July. Until then stay cool (or warm, as the case may be).



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Renewable Energies as Preventive Solution of Climate Change And Clean Renewable future Energy

Hamid Reza Seyed Jafari , Seyed Mohammad Reza Seyed Jafari

ABSTRACT

One of the most important and most widely used environmentally friendly renewable fuels in the coming decades are green hydrogen gas & bio- methane gas in which will have a very positive effect on climate change control and reducing global warming in various industries and domestic energy consumptions in which producing of greenhouse gases (especially carbon dioxide).

KEYWORDS

Green Hydrogen Gas (H₂), Climate Change, Renewable Energy, CCS, Electrolysis of water

Today, the world's leading companies design different infrastructures for green hydrogen production and bio-methane gas. For example, some countries have selected the process of green hydrogen production by electrolysis of water, and other countries have programmed their process path by making synthetic gas from reaction of carbon dioxide and green hydrogen gas converting it to bio-methane gas.

A. Electrolysis of (sea)water to produce green Hydrogen gas

The general mechanism to produce green hydrogen from water with the help of renewable energy can be as follows:

1st step - Renewable electricity generation in one of the methods such as: Wind, Solar

2nd step - Transfer of previously generated renewable electricity to an electrolysis facility near the sea (preferably)

B. Using green hydrogen and carbon dioxide to produce syngas, renewable bio - methane and electricity

In this mechanism, the following general steps are:

1st step – Collection of CO₂ gas from the chimneys of large industries due to combustion, then storage it, known as CCS (Carbon Capture Storage).

2nd step - In the next stage, carbon dioxide gas is dissociated and reacted with green hydrogen gas to produce synthetic gas (CO) with its own processes and catalysts in the reformer furnaces.

3rd step - In the final stage, the synthesis gas is reacted again with green hydrogen gas then converted to bio-methane gas in the vicinity of the catalyst.

Renewable Reactions:

- [2H₂O (sea) +Renewable Electricity] (Electrolysis)>>O₂+Green H₂
- [2Green H₂+2CO₂ (CCS)]>>2CO+2H₂O (goes to Electrolysis)
- [2CO (SNG) +6Green H₂]>>2CH₄ (Bio-Methane or RNG) +2H₂O (goes to Electrolysis)
- Bio-Methane (goes to CHP)>>Electricity
- [Green H₂ (goes to FCHE engine produces electrons and positive Hydrogen ion H⁺) +O₂]>>Steam (emitted to atmosphere from: vehicle, airplane, train, truck, bus, marine, power plants...) +Electricity
- [Green H₂ + N₂]>>Green Ammonia(as clean fuel)
- [Green Ammonia + CO₂(CCS)]>> Green Urea

Note: SNG: syngas, CHP: combined heat & power, RNG: renewable natural Gas, CCS: carbon capture storage, Renewable electricity: solar, wind, hydro power, FCHE: fuel cell hybrid electricity

ACTUAL EVENTS

At present, various countries that have

access to sea water and have the potential to produce renewable electricity, have started planning and investing heavily in the field of green hydrogen gas production as a new source of energy or a green reactant to convert CO₂ gas captured in CCS to bio-methane.

- a. Oman (Green Hydrogen Production): A 25GW solar and wind megaproject in Oman is set to be coupled with electrolysis to generate industrial-scale quantities of green hydrogen and green ammonia. Once operational, the project will be capable of producing over 1.8 million tonnes of low-carbon green hydrogen per year in addition to 10 million tonnes of green ammonia. This could be a major development for the global hydrogen economy by presenting an opportunity to create large amounts of green hydrogen and be able to export this to lucrative international markets to enable decarbonisation. (Source: <https://www.h2-view.com>, By George Heyneson Apr 05, 2022).
- b. Japan (Green Hydrogen Production): ITOCHU announces investment in Clean H₂ Infra Fund, the World's largest Clean Hydrogen Infrastructure Investment Fund (Aiming to realize a decarbonized society through the development of a hydrogen value chain) ITOCHU Corporation (headquartered in Minato-ku, Tokyo; Keita Ishii, President & COO; hereinafter "ITOCHU") announced today that it has acquired shares of Clean H₂ Infra Fund. S.L.P. (French company) ("the Fund"), the world's first large-scale clean hydrogen infrastructure investment fund, through its wholly owned special purpose corporation ("SPC"). To coincide with this move, ITOCHU also entered into an agreement with Tokyo Century Corporation (headquartered in Chiyoda-ku, Tokyo; Makoto Nogami, President & CEO; hereinafter "Tokyo Century") regarding the third-party allotment of shares in the SPC to Tokyo Century, enabling the joint ownership of shares in the Fund (investment ratios of 50% to ITOCHU and Tokyo Century respectively following the allotment), contingent upon the approval of regulatory authorities. (Source: <https://www.itochu.co.jp/en/news/press/2022/220325.html> March 25 ,2022)
- c. EU (Green Hydrogen Production): Proposal for a Regulation of the European Parliament and of the Council on the internal markets for renewable and natural gases and for hydrogen (Source: Council of the European Union, Inter institutional File: 2021/0424, Brussels, 16 December 2021)

- d. Norway (Green Hydrogen Production): Saga Pure ASA: Acquires Hyon and partners with Nel and Norwegian Green Hydrogen. Saga Pure ASA announced the acquisition of Hyon AS (Hyon), and following a subsequent transaction, Saga Pure, Nel ASA and Norwegian Hydrogen AS will have equal shareholdings in the company. Hyon has a strong position within maritime hydrogen solutions and will mainly focus on developing green hydrogen and bunkering solutions for harbors going forward. (Saga Pure press release, June 29, 2021 - Oslo, Norway).
- e. USA(Green Hydrogen Production) : A major new green hydrogen project in the united states of USA, touted as the "world's largest" green hydrogen production and storage hub, is set to be developed in Texas with a capacity of 60GW. (LinkedIn h2-view.com, April ,2022)

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4. Saga Pure press release, Oslo-Norway, June 29, 2021

LinkedIn, h2-view.com, April 2022

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Seyed Mohammad Reza Seyed Jafari is a B.Eng. Chemical Engineer from Azad University of Iran (IAU) and post graduate student of Information Technology Management Tehran with 6 years' experience.

Guidelines for Propylene Splitters

Timothy M. Zygula, Karl Kolmetz, Johor Bahru

INTRODUCTION

The Chemical Processing Industry has been continually pushing the capacity envelope of new and existing distillation columns. While increasing the capacity of existing columns is not unusual, great care needs to be taken when a revamp is being considered. There is a fine line between success and failure when a column is designed at or near the upper end of the capacity envelope. The authors will detail the methodology used when designing a new or considering a capacity increase for an existing propylene splitter. This paper will discuss design aspects that need to be considered when designing a propylene splitter. The authors will also present a generic case study of a propylene splitter revamp. Some of the topics that will be covered by the authors are:

1. Process simulation of a splitter – proper simulation techniques
2. From the simulation to the field – tray efficiencies
3. Utilizing a process simulation to develop column hydraulics
4. The types of internals that have been used in propylene splitter columns
5. Design considerations that need to be addressed when considering a revamp.

GENERAL DESIGN OF DISTILLATION COLUMN

Separations are a major part of the chemical processing industry. It has been estimated that the capital investment in separation equipment is 40-50% of the total for a conventional fluid processing unit. In a plant one of the main unit operations is material separation. This includes distillation, storage tanks, flash drums and other equipment of this nature. Of the total energy consumption of an average plant, the separation process accounts for about 50% to 70% of the energy consumption of the plant. Within that area of the material separation, the distillation unit operation method accounts for normally greater than 80% of the energy consumed for this process.

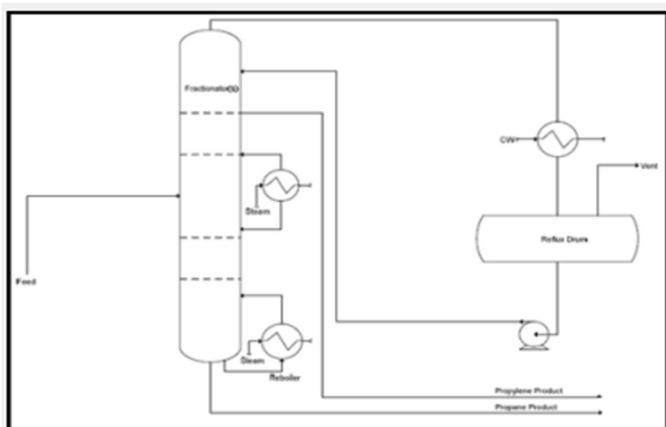
In general, initial design of a distillation tower involves specifying the separation of a feed of known composition and temperature. Constraints require a minimum acceptable purity of the overhead and the bottoms products. The desired separation can be achieved with relatively low energy requirements by using a large number of trays, thus incurring larger capital costs with the reflux ratio at its minimum value. On the other hand, by increasing the reflux ratio, the overhead composition specification can be met by a fewer number of trays but with higher energy costs.

DESIGN OF A PROPYLENE SPLITTER

Determining the design of a Propylene Splitter requires an understanding of the simulation model used to generate the internal loads and physical properties, vapor and liquid equilibrium data utilized, tray hydraulics, and how the selection of the internals will affect the actual efficiency of the installed equipment in the field. The typical design of a propylene splitter is not complex and there are two general variations in design. The first is called a high-pressure system, and the second is called a heat pumped system. A high-pressure system is designed to utilize cooling water as the source to cool the overhead vapor, and a high pressure is needed to condense the propylene vapor at ambient temperatures of about 40 degrees C. A heat pump system utilizes a compressor to reduce the tower pressure to allow the distillation column to be smaller. In most distillation application, relative volatilities can be improved by lowering the pressure. This results in lower number of stages required and reflux ratios, but at the cost of higher energy requirements of the compressor. A good rule of thumb is that if the propylene system is associated with an ethylene plant, in which there is typically an abundance of quench water that can be used to heat the C3 Splitter reboiler a non heat pump system may be the best choice. If no source of sufficient low grade heat is available for example in a refinery FCC unit or

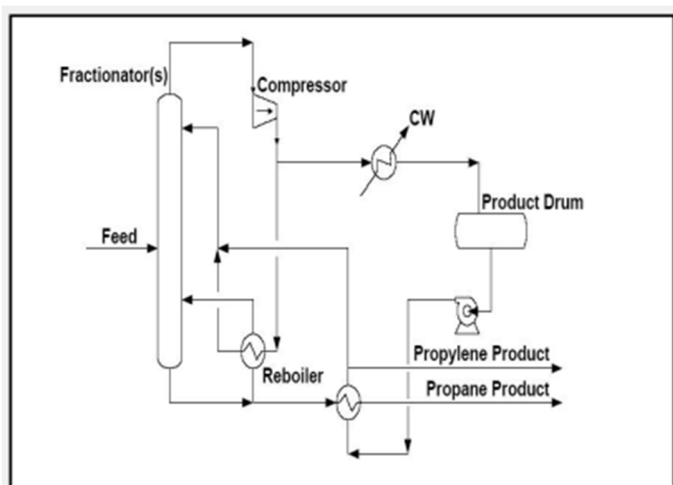
propane dehydrogenation unit, then the use of a Heat Pump is typically the economical choice.

HIGH PRESSURE SYSTEM



The first is called a high-pressure system, and the second is called a heat pumped system. A high-pressure system is designed to utilize cooling water as the source to cool the overhead vapor, and a high-pressure system is needed to condense the propylene vapor at ambient temperatures of about 40 degrees C.

HEATED PUMP SYSTEM



A heat pump system utilizes a compressor to reduce the tower pressure to allow the distillation column to be smaller. In most distillation applications relative volatilities can be improved by lowering the column pressure. This results in lower number of required theoretical stages and reflux flow. These savings are offset by the required energy cost of the compressor.

PROCESS SIMULATION OF A PROPYLENE SPLITTER – PROPER SIMULATION TECHNIQUES

Simulation of a propylene splitter seems very simple and can be done quickly by 3rd year engineering students. There are a small number of components and the equipment layout is not complex.

The challenge of a propylene splitter is that, unless you use the correct vapor and liquid equilibrium data, the simulation can have greater than 15% inaccuracies as compared to actual field data.

Physical properties are critical to the success of a simulation model and are also very important to the accuracy of the model. Poor physical property data may prevent your simulation model from converging. The most typical problem is missing parameters in the thermodynamic package utilized. This is not unusual in most commercial simulation packages.

Physical property parameters for most compounds are not known for every thermodynamic model at every pressure and temperature range. Many times this fact is overlooked when a design model is constructed. Simulation models are constructed and executed with thermodynamic parameters missing. Although the model may appear to be correct but may be incorrect because of the missing thermodynamic data. Then there is the problem that all of the thermodynamic data are present but the data are not accurate. This problem is even worse than the problem of missing data since the results from the simulation model will appear to be correct, but are totally wrong. Most simulation packages won't alert the users that there is a problem. It is the job of the user to determine if the results from a simulation model are accurate (1).

The best way to confirm if your thermodynamic data are correct is to see if you can find any laboratory data or data from literature on your system. This may not always be practical because good thermodynamic test data are hard to find.

Sometimes that data may have to be generated in a pilot plant before any design work begins.

Research the system being modeled. Published thermodynamic data on the system being modeled may exist. If data is obtained, the data must cover the same temperature and pressure range that you are designing. Next, run a simulation with the same system and see if you can match the data. Most data on propylene splitters has been compiled from years of operating experience. Many companies that license technology have done extensive testing and have developed propylene splitter data for design purposes.

The most accurate Vapor Liquid Equilibrium (VLE) data for Propylene Splitters might be Ping Robinson, but there is a huge data base

of distillation columns designed and built utilizing Soave Redlich Kwong (SRK), and many designers utilize SRK to be able to utilize the existing database for actual tray efficiency in the field. The standard SRK equation of state model handles the propane / propylene binary K values adequately over the typical operating pressure range of these towers, which is 5 bar (100 psia) to 20 bar (320 psia). The other miscellaneous lights, heavies and intermediate boilers, such as Propadiene (PD), methyl-acetylene (MA or propylene), ethane and isobutane, are also adequately modeled using the SRK equation of state. (3)

Methyl-acetylene (MA) is an intermediate boiler that is lighter than propane and heavier than propylene. Even at small ppm concentrations in the feed will, over time, result in a build up of MA in the tower. Concentrations inside C3 Splitter towers 10 to 20 trays from the bottom can be as high as 15% to 20% depending on the severity of propylene recovery required. Many propylene splitter systems have a sample point in this 10 to 20 tray range from the bottom to be able to sample the MA concentration in the column. MA, being a triple bonded hydrocarbon at elevated concentrations, above 40%, can auto decompose with potential adverse consequences. Propadiene (PD) is heavier than both propane and propylene and will never have a significant concentration in the overhead product.

Many choices are available for enthalpy models in simulation packages. SRK will do an adequate job but there may be better choices. This is important because there are always light components (i.e. methane, ethylene) that will be present in the feed and they will be close to their critical temperature. The choice of enthalpy model will help in the tower consistently achieving convergence. (3)

One other area of concern is the specific heat of liquid propylene. Some Propylene Splitters will have subcooled reflux return or a subcooled feed. The performance of a C3 Splitter tower is heavily dependent on a proper heat balance on the tower. Sub-cooled streams have to be accounted for properly. Propylene pure component liquid specific heat data is quite varied. The variation in reputable data has an error band of over 15%.

There are many choices in the vapor density, enthalpy, specific heat, viscosity, and surface tension model correlations. It is important to be able to tune your model to actual field data so that your model will reflect the real world.

High Pressure:

High-pressure distillation in a column can have challenges. There are many factors to be considered when designing at high operating pressures. (1).

At higher operating pressures the relative volatility of the system is lower which increases the separation difficulty. As a direct result of increased separation difficulty the reflux requirements for the column would increase. The column would also require more stages and increased duties for the reboiler and condenser to perform the separation. Propylene Fractionators are high liquid traffic columns that require internals that can handle high liquid traffic.

At higher operating pressures the reboiler temperature rises, thereby requiring a more expensive heating medium. If the same heating medium is used a reboiler with a larger heat transfer area would be required.

At high operating pressures the vapor density would increase and therefore lower the required vapor handling capacity. This would lead to a reduction in the diameter of the column, which would reduce the capital equipment costs.

High Pressure Distillation Tray/Column Design:

As the distillation pressure is increased, the vapor density increases. When the critical pressure is approached, the compressibility factor of a saturated vapor usually has a value less than 0.75. Thus the vapor density of the gas phase is quite high at pressures greater than 40% of critical. As the operating pressure is increased for the same Cs (Capacity Factor) value, the vapor mass flow rate will be much greater than at atmospheric operating pressure because of the high vapor density. While at the same time the liquid mass flow rate will be greater at high operating pressure than at atmospheric operating pressure. Therefore, liquid flow rates per unit of column cross-sectional area will be higher as operating pressure increases. The capacity of the fractionating device at high pressure may be dependent on its ability to handle these high liquid flow rates.

In a propylene fractionator column, the tower cross sectional area is the sum of the trays active area plus the total downcomer area. .

The amount of required active area (Vapor-Liquid Bubbling Area) is determined by vapor

flow rate. The downcomers handle a mixture of clear liquid, froth, and aerated liquid. The downcomer area required to handle the high liquid flow not only increases with the liquid flow rate, but also with the difficulty in achieving separation between the liquid and vapor phases. The volume required for the downcomer increases at a lower surface and a smaller density difference between the liquid and vapor. Because of the large downcomer area required to handle the high liquid flow rates the area may be 40% to 80% greater than the calculated tray active area for the vapor flow rates for propylene fractionator distillation. The downcomer area becomes a significant factor in the determination of the tower diameter.

Simulation Accuracy:

In order to determine the accuracy of a simulation it is always desirable to construct a McCabe-Thiele diagram from the data generated from the simulation. The data from the simulation can be easily transferred to a software package where the graph can be constructed. This graph is used more as a tool to identify possible problems that won't be discovered until the column fails. The following is a list of the areas where a McCabe-Thiele diagram can be used as a powerful analysis tool (1).
 pinched regions - Pinching is readily seen on an x-y diagram.

Mislocated feed points - the feed point should be where the q-line intersects the equilibrium curve. This is generally the rule in binary distillation. However, it is not always true in multi-component distillation. A key ratio plot is often developed in the design phase. This type of plot is far superior to an x-y diagram for identifying mislocated feeds, especially with large multicomponent systems.

Determining if the column is being over refluxed or reboiled - this can be recognized by too wide of a gap between the component balance line and the equilibrium curve throughout the column.

Identify cases where feed or intermediate heat exchangers are needed.

Most commercial simulation programs will provide the information required to generate these plots.

COLUMN SIZING

Once the internal liquid and vapor traffic is obtained from the simulation model, the diameter of the column must be obtained. Most simulation packages have tower sizing routine.

These routines are fairly easy to use and yield quick results. However, these results should be verified by calculation.

Column sizing is done on a trial and error basis. The first step is to set the design limits. The design limits are as follows:

1. Maximum Design rates - Vapor/Liquid Traffic is needed at Maximum Operating rates.
2. Design rates - Vapor/Liquid Traffic is needed at Design Operating Conditions.
3. Minimum Design rates - Vapor/Liquid Traffic is needed at Minimum Operating rates.

Sizing calculations need to be performed in areas of the column where the vapor/liquid traffic is expected to be highest and lowest for each section. For example,

- The top tray and bottom tray in the column
- The feed tray
- Any product draw-off tray or heat addition/removal tray.
- Tray where the vapor liquid loading peaks.

There are also shortcut methods to sizing a column, which involve using a flooding correlation. These methods minimize the amount of trial and error calculations. Using the method as outlined by Kister (2) the first step is to determine the C-Factor at the most heavily loaded point in the column. Using an entrainment flooding correlation like the Kister and Haas correlation the C-Factor at flood can be calculated.

$$CSB = 0.144 [d_{2H} s/rL]^{0.125} [rG rL]^{0.1} [S/hct]^{0.5} - \text{Kister and Haas (2)}$$

Next the vapor velocity at flood based on net column area minus the tray downcomer area needs to be calculated. This calculation is done for the top and bottom section of the column (2).

$u_N = CSB [(rL - rV)/rV]^{(1/2)}$ - Flooding Vapor Velocity, ft/s
 Next, the bubbling area required for the top and bottom sections of the column need to be calculated using equation 3. In new designs columns should be design for 80% flood (2).

3. $AN = CFS/[(SF)(0.8)u_N]$ - Bubbling Area Required (Column Cross Sectional Area less downcomer top area, ft²)

Next, the downcomer top area needs to be

calculated using equation 4. This calculation is done for the top and bottom section of the column (2).

$AD = GPM/VDdsg$ – Downcomer Area.

Once this has been completed the tower cross sectional area can be calculated using equation 5. The tower diameter can be calculated from the tower area. (2).

5. $AT = AN + AD$ – Tower Cross Sectional Area, ft²

The following are the definitions of the parameters used in the above equations.

CSB – C-Factor at flood, ft/s

dH – Hole Diameter, in

S – Tray Spacing, in

hct - Clear liquid height at the transition from the froth to spray regime, in of Liq

rG, rL - Vapor and Liquid Density, lb/ft³

s - Surface Tension, Dyne/cm

SF – Derating Factor or Foaming Factor

GPM – Tray Liquid Loading, GPM

VDdsg – Downcomer, GPM/ft²

AN – Tray Bubbling Area, ft²

AD – Downcomer Top Area, ft²

AT – Total Tower Cross Sectional Area, ft²

COLUMN INTERNAL DESIGN

Once the preliminary tower diameter has been set the internals can be chosen. The task of choosing the type of tower internal to use is very important.

The type of column internals used dictates a column's efficiency and capacity. All of the modeling and careful design work will mean nothing if the wrong type of column internals is chosen. For propylene fractionation trays are the only type of internal that should be considered. The types of internals that have been used in propylene splitter columns are:

- Conventional Cross Flowing Trays
- Counter Contacting Trays
- Structured Packing
- High Capacity Trays
- Multiple Downcomer Trays

CONVENTIONAL MULTIPASS TRAYS

Conventional Multipass trays are typically used when a column is initially designed. Four pass or six pass trays are usually used because of their ability to handle high liquid loads like seen in propylene fractionation. The downside

to using multipass trays is the reduction in separation efficiency that is experienced due to the reduction in active area. Great care must be taken when sizing downcomers in high-pressure distillation applications. The difference between vapor and liquid densities becomes smaller and separation of vapor from liquid in a downcomer becomes more difficult. This can result in increased aeration back-up and possible premature downcomer flooding. (2)

Multi-Downcomer Trays:

Multi-Downcomer trays are used for large liquid loads, particularly when the volumetric ratio between vapor and liquid rates is low. These situations occur in medium to high-pressure distillation, in absorption and stripping, and in direct contact heat transfer applications.

Multi-Downcomer trays can be used at close tray spacing. This will allow a reduction in both height and diameter of a new column compared to a column fitted with conventional multi-pass trays. Vessel shell costs can be significantly reduced with the use of Multi-Downcomer trays. When retrofitting an existing column with Multi-Downcomer trays, a significantly greater number can be installed, providing increased product purities and recoveries, as well as reduced reflux ratio for reduced energy consumption and/or increased column capacity.

The use of Multi-Downcomer trays has often reduced the number of columns needed in difficult separations, such as the fractionation of propylene-propane.

Tray Efficiencies: From the simulation to the field – tray efficiencies

The tray efficiencies in Propylene Splitters have been a widely discussed issue. In actual operation they have ranged from 40 percent to 100%, so it is easy to see why this is a widely discussed issue. In general if the boiling points of the overhead product (light key component) and bottoms product (heavy key component) are close, less than 5 degrees C, the actual tray efficiency in the field will be high. If the boiling points of the overhead and bottoms product are far apart, the actual tray efficiency will be low. The ratio of the boiling points is classified as the relative volatility.

For example, a Propylene Splitter has close boiling points between the overhead and bottoms product, about 7 degrees C. This requires many ideal stages for separation in a process simulation, but each stage will have high efficiency in the field. For a normal cross-

flowing tray 90% tray efficiency can be obtained. For chemical grade propylene, 95% purity, about 100 ideal trays might be required in a simulation, and 110 actual stages may be required in the field.

Tray efficiencies are generally classified as either overall efficiency (Fenske), point efficiency, or average tray efficiency (Murphree). The overall efficiency term is quite straightforward. It is the number of actual stages achieved versus the number of trays in the tower or section of the tower. Point efficiency and Murphree tray efficiency are similar. They represent the ratio of the actual compositional change and the theoretical compositional change at equilibrium. (2)

The compositional change is usually measured in the vapor phase but can be measured in the liquid phase. The difference between the point efficiency and Murphree tray efficiency calculation is the reference point. Point efficiency is measured at a specific point and the Murphree tray efficiency is measured across a complete tray. Therefore, the compositional gradients normally found on a tray will affect the Murphree tray efficiency but will not affect the point efficiency. When the liquid and vapor both have homogeneous compositions, point efficiency and Murphree tray efficiency will be equal.

In practical terms, trays with little or no liquid flow path length will essentially achieve point efficiency while trays with conventional flow path will achieve a higher Murphree tray efficiency due to the compositional gradient of the liquid flowing across the tray deck.

There are various aspects of equipment design that can affect efficiency. Any time a device can maximize the vapor/liquid contact while maximizing the compositional approach between the vapor and liquid, that device will maximize the efficiency of the tower. Conversely, any device characteristics that limit contact or compositional approach will lessen the efficiency of the tower.

Characteristics that may affect efficiency are discussed below.

Weir Height: With trays operating in the froth regime, an increase in weir height will directionally increase the efficiency. Kister has noted that the removal of even a small outlet weir can noticeably decrease the effective tray efficiency. Weir height is especially important in liquid limited systems or systems where a slow chemical reaction is taking place. (2)

Flow Path Length: Directionally, an increase in flow path length will increase efficiency. This was discussed earlier in the difference between point efficiency and Murphree tray efficiency. This holds true unless the length of the flow path creates anomalies in the tray operation such as liquid backmixing or vapor cross flow channeling. (2)

Liquid and Vapor Maldistribution: As would be expected, vapor and liquid maldistribution will cause decreases in efficiency. Generally, maldistribution problems are generated by the distribution of feeds to the columns rather than by the contacting devices themselves. When reviewing internal designs it is very important to pay attention to feed pipe designs. Good liquid distribution across the tray is essential for high efficiency. Feed pipe designs that distribute liquid at high velocities should be avoided. Vapor distribution is also an important factor to consider. Most columns use chimney trays vapor distribution devices.

Weeping and Entrainment: Weeping and entrainment will also directionally cause decreases in efficiency. When considering the effects of weeping, it is important to differentiate between inlet side weeping and outlet side weeping. With inlet side weeping, the liquid will effectively miss two tray decks and the effects can be substantial. With outlet side weeping, only a small portion of the deck is missed and the effects

Design Case: Below is a typical design case for a propylene splitter. Typically a propylene splitter would be designed with 200 theoretical stages or between 290 to 310 actual trays. The column design being presented in this paper was designed with 200 theoretical stages. Simulation models showed that 200 theoretical stages produced 99.6-mole% propylene in the overhead product stream of this column. This is based on a column feed rate of 2700 lbmol/hr and an overhead heat duty of -282 mmBTU/HR. The reboiler duty of the column design being detailed is 180 mmBTU/HR. This column has only one feed location.

The composition range of the feed stream feeding the propylene splitter column is detailed in TABLE 1.

Typical design parameters for a propylene splitter column have been compiled in TABLE 2.

DESIGN CONSIDERATIONS THAT NEED TO BE ADDRESSED WHEN CONSIDERING A GRASS ROOTS OR REVAMPED COLUMN Operating Flexibility

The column should be designed with some operating flexibility. When reviewing the required efficiency, it is usually a good idea to review the sensitivity of the product purity to losses of efficiency in the tower. One way to do this is to construct a plot of required stages versus reflux ratio. (4) Knowing the sensitivity that reflux has on product purity will allow the designer to make a decision if the available reflux is sufficient to achieve the purity goal under different operating scenarios.

Minimum Reflux or Minimum Amount of Required Internals: One design consideration is to determine the minimum reflux needed to achieve the required separation. In order to determine the amount of minimum reflux is required, one develops a reflux-stage plot and extrapolates from it. To develop this plot, simulation runs are performed at a various number of stages while keeping the material balance, product compositions, and the ratio of the feed stage to the number of stages constant. The reflux ratio is allowed to vary.

Then a plot of the number of stages versus reflux or reflux ratio is plotted. The curve is extrapolated asymptotically to an infinite number of stages to obtain the minimum reflux ratio. Once the minimum reflux has been determined then it must be decided if the design will be done at minimum reflux or with less installed internals. This is usually an economic choice. If the column is designed at minimum reflux the savings is lower required energy for the column operation. Usually the reboilers and condensers are smaller. The diameter of the column is also smaller. This choice may hinder future capacity revamps due to the size of the equipment. If the column is designed for minimum required internals required energy would be higher. The condenser and reboiler will be larger.

This option does give flexibility for future capacity upgrades. (1)

Optimization of Feed Stage: Another design consideration is to design the column at the optimum feed stage location. Once all of the simulation runs are completed two main plots can be created. One plot will be a McCabe-Thiele diagram and the other will be a concentration versus feed stage diagram. The McCabe-Thiele diagram is plotted using the mole fraction data calculated for each stage by the simulation. The equilibrium data and the

TABLE 1

Feed Stream Component	Composition Range (Mole%)
Propadiene	0.0 to 0.03
Propylene	92.0 to 96.0
Propane	4.0 to 7.0
Butanes	0.02 to 0.05
C5 non Aromatics	0.04
C6 non Aromatics	0.11
C7 non Aromatics	0.0
C8 non Aromatics	0.03
Benzene	0.0
Toluene	0.0

TABLE 2

Design Specifications	Overhead Of Column	Bottom Of Column
Tower Diameter(in)	200 to 300	200 to 300
Typical Number of Theo. Stages	133	70
Column Temperature(° F)	20(Top)	44(Bottom)
Column Pressure (PSIA)	71(Top)	86(Bottom)
Propylene Mole% Column	99.6(Top)	14(Bottom)
Propane Mole% Column	0.04(Top)	81(Bottom)
PD Mole % Column	0.002(Top)	0.01(Bottom)

operating lines are also determined from the simulation results. Determining the optimal feed stage will help to maximize efficiency of the column. (2)

In the second type plot, the key component concentration in the product streams are plotted against the feed stage numbers. The minimum in the curve will represent the optimum feed stage. One can generally assume the ratio of optimum feed stage to total number of stages is independent of the number of stages. (2) In this type of plot it is important to note that the total number of stages is kept constant. Also, if the distillate rate is increased, it is normal to move the feed stage up the column as required. (1)

CONCLUSIONS

In conclusion, it is important to note that when designing or revamping a propylene splitter great care must be taken during the design phase of the project. In order to get the maximum efficiency and capacity out of a propylene splitter one must consider the accuracy of the simulation and the thermodynamic model being used to model the column. Once the simulation has been completed great care must be taken when evaluating the sizing of new and existing equipment. Verification of the design which includes the amount of reflux required and feed location is essential to obtain maximum efficiency. All of these factors talked about in this paper are essential to obtain a good efficient design of a propylene splitter.

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