

Hydrogen Network Optimization Through Hydrogen Recovery from Refinery Off Gases

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Abstract

Hydrogen plays a crucial role in refineries primarily for two main purposes: hydroprocessing and hydrocracking. It is used in hydroprocessing reactions, such as hydrotreating, to remove impurities like sulfur and nitrogen from crude oil fractions. Additionally, hydrogen is involved in the hydrocracking process, where it saturates the heavy cracked hydrocarbon feed stocks, producing more valuable products and improving their environmental and performance characteristics. This makes hydrogen an essential component in refining operations, contributing to the production of cleaner and higher-quality petroleum products. The hydrogen demand in oil refineries is increasing due to environmental legislation, which is pushing the market towards products with lower sulfur content. In this study, we will explain the recovery of hydrogen from internal refinery sources such as Refinery Off-Gas "ROG" as a cost-effective solution to meet the rising hydrogen demand. This approach also serves as a valuable means of recovering light hydrocarbons. The simulation was conducted using computer programming Aspen HYSYS® v11.A hydrogen recovery of 13,050 Nm³/hr was achieved and recycled back into the hydrogen network, also a reduction of carbon dioxide emissions by 450.69 Nm³/hr was achieved. This method offers valuable guidelines for managing hydrogen and light hydrocarbons in refineries.

Keywords: *Hydrogen recovery; Hydrogen network; Refinery off-gases; LPG recovery.*

1. Introduction

The refinery segment is considered the second-largest hydrogen consumer, coming after the petrochemical industries segment, with 30% of the market share [1]. Recently, refineries are facing a misbalance between the hydrogen availability and demand [2]. This misbalance is caused by several reasons. Firstly, the increase in the environmental regulations towards sulfur, nitrogen and metal impurities removal. Secondly, there are strict regulations on the aromatic content in gasoline product. These regulations limit the catalytic reforming hydrogen production, which occurs as a byproduct due to the conversion of paraffins and naphthenes to aromatics [3]. Lastly, the fact that crude oils are getting heavier and containing higher sulfur content. Thus, it is a challenging issue to meet the excess in the hydrogen demand [3].

In refineries, the cost of hydrogen production has become the second largest cost coming after the crude oil cost. The recovery of hydrogen from refinery off-gases, which are rich in hydrogen, should be taken into consideration for the hydrogen deficiency problem. These off-gases were previously sent to the sweetening process for sour gas removal before being sent to the fuel gas system, where their calorific value is the only aspect being utilized [4].

The removal of acid gases from refinery off-gases is vital to prevent corrosion and catalyst poisoning [2]. Methods include chemical absorption (using solvents like MEA, DEA, and TEA), physical absorption (employing solvents like Selexol and Sulfinol), and mixed absorption (a combination of chemical and physical solvents) [5].

Methods of hydrogen separation are crucial for various applications, and several technologies are employed to achieve this:

- Polymeric membrane process:** Utilizes a polymer membrane to separate hydrogen based on varying permeabilities within a gas stream. Membranes, often with lifetimes of 4-5 years, can be enhanced for higher hydrogen recovery by increasing surface area or differential pressure. Sensitivity to liquid hydrocarbons necessitates heating the feed above its dew point temperature [4,6-7].
- Gas absorption process:** Relies on solvent absorption of light hydrocarbons from the feed gas stream, commonly using gasoline as a solvent. Gas absorption occurs at high pressure and low temperature, with solvent regeneration through pressure reduction across stages, maintaining nearly the same pressure as the feed [4,6].
- Cryogenic process:** Involves refrigeration to extremely low temperatures and high pressure, leading to partial condensation of impurities, followed by distillation for separation. Cryogenic distillation is costly due to the need for very low temperatures, around -250°C, for carbon dioxide liquefaction and separation from hydrogen [9-10].
- Pressure swing adsorption (PSA):** Employs solid bed adsorbents with high selectivity for impurities, making it the most commonly used and studied technique for hydrogen purification [10-11]. PSA operates as a batch process, cycling between high pressure for adsorption and low pressure for desorption from adsorption beds, with different adsorbents layered based on feed impurities [4,11]. PSA relies on the adsorbent's strong affinity for certain impurities over hydrogen [4].

The utilization of PSA processes to produce pure hydrogen from gas mixtures containing 60-90 mol% hydrogen has emerged as a noteworthy technology in refineries. These PSA processes are engineered to generate a dry-rich product stream at the feed gas pressure, comprising 98-99.99 mol% hydrogen, while achieving a hydrogen recovery rate ranging from 70-90% [12-13].

Table 1. Difficulty of impurities removal per adsorbent in H₂ PSA systems(modified after [11,13-15]).

	Silica gel	Activated carbon	Zeolites" molecular sieve"
H ₂ O	Very easy	Very difficult	Very easy
C ₂ H ₄	Easy	Very easy	Very easy
C ₄ H ₁₀	Easy	Very easy	Very easy
C ₃ H ₈	Easy	Easy	Very easy
C ₂ H ₆	Moderate	Easy	Very easy
CO ₂	Moderate	Easy	Very easy
CH ₄	Moderate	Easy	Moderate
CO	Moderate	Moderate	Easy
N ₂	Difficult	Difficult	Moderate
Ar	Difficult	Difficult	Difficult
H ₂	Very difficult	Very difficult	Very difficult

For units designed to treat synthesis gases from steam reformers, the first layer at the feed entrance usually consists of activated carbon and acts as a guard bed to absorb CO₂, CH₄ and C₂H₆, while the second bed, typically made of zeolite or molecular sieve, deals with CO [11].

The length ratio of different adsorbents to be used in the PSA adsorption beds is adjusted to make sure that CO₂ doesn't break through the activated carbon layer at the adsorption step. This is crucial to prevent CO₂ from contacting the zeolite "molecular sieve" layer as CO₂ is adsorbed strongly to the zeolite layer and can't be desorbed easily. This may lead to the reduction in the adsorption capacity of the zeolite layer for other impurities [11].

Similarly, C₃H₈ and C₄H₁₀ are strongly adsorbed on activated carbon. So, silica gel is primarily used, as C₃H₈ and C₄H₁₀ are absorbed and desorbed easier in case of silikagel [11].

The PSA units that are designed to remove impurities from refinery off gases, each bed contains silica gel at the feed entrance for the removal of C_3H_8 and activated carbon near the bed outlet for the removal of CH_4 and C_2H_6 [11].

A study was conducted on four different off-gases sources to recover hydrogen from at constant purity. The study employed various purification methods including counter-current absorption, pressure swing adsorption at high pressure (PSA-HP), pressure swing adsorption at low pressure (PSA-LP), and membrane adsorption. The study also involved a comparison to determine the most economical purification method across four distinct sources [8].

Another study focused on the economics of hydrogen recovery from refinery off-gas in Tehran. This research applied three different methods: pressure swing adsorption, gas absorption, and membrane separation technologies. The comparative analysis on an economic basis revealed that the PSA process incurred lower costs compared to the other two methods [4].

Some other studies investigated an integrated hydrogen production process based on coupled feeds of refinery off- gas and natural gas for steam reforming, utilizing the Aspen Plus simulator. Their work involved multi-objective optimization employing a non-dominated sorting genetic algorithm-II (NSGA-II) [3].

Most of the previously mentioned studies focus on the choice of the hydrogen purification method, and found that the PSA is often emerging as the preferred choice in refineries. However, these studies typically did not emphasize the required hydrogen purity for catalysts used in refineries, despite the potential for some impurities to permanently poison catalysts. Also, none of these studies calculated the amount of carbon dioxide reduced and the amount of LPG recovered due to hydrogen recovery from refinery off-gases (ROG).

In this paper, a refinery located in Egypt was used as a source of refinery off-gases to recover hydrogen from, with a focus on utilizing HYSYS simulation software to leverage actual data and minimize operating and fixed costs by using the refinery potential.

2. Case study

Industrial hydrogen production is primarily achieved through steam methane reforming, accounting for around 48%, while oil/naphtha steam reforming processes contribute approximately 30%. Coal gasification makes up about 18%, and electrolysis contributes about 2% of global hydrogen production [16]. Hydrogen finds numerous applications in oil refineries, including hydro-cracking processes used for transforming low-grade heavy hydrocarbons into high-grade products. Additionally, hydrogen plays a crucial role in the hydro- treating process, removing impurities like sulfur and nitrogen. Also in isomerization process, hydrogen is used in reducing the benzene content and increase the research octane number "RON", as benzene is a known carcinogenic substance [11,17].

Figure 1 provides a simple view of the hydrogen network within a refinery located in Egypt. It is evident from this illustration that there are two primary hydrogen sources, namely the steam methane reforming unit (SMR) and the platforming unit (CCR). The SMR has a capacity to produce $99,000 \text{ Nm}^3/\text{hr}$. However, the CCR plant plays a crucial role in the hydrogen network by contributing $26,400 \text{ Nm}^3/\text{hr}$ with 92 % hydrogen purity from the platforming compressor second stage, thus reducing the dependence on the SMR plant.

The SMR effluent stream has a hydrogen purity of 74.77%, it passes through PSA1 to increase the hydrogen purity to 99.9%. Also, the platforming effluent gases has hydrogen purity of 92%, it passes through PSA2 to increase the hydrogen purity to 96%. Some of the platforming hydrogen is fed to the N.H.T and the D.H.T at 44 barg, after being compressed and recontacted with naphtha in order to increase the hydrogen purity to 96%. PSA1 and PSA2 products are mixed together compressed to 45 barg for the isomerization unit, and to 160 barg to the hydrocracker unit as shown in Figure 1.

It is also evident that a significant amount of hydrogen is being wasted to the fuel gas system from mentioned units. Some of these off-gases, such as streams number (1), (2), (3), and (4), are rich in hydrogen content. These streams provide an opportunity for hydrogen recovery, which could subsequently lead to a reduction in the required S.M.R capacity.

Table 2. Streams composition and condition.

	(1)NHT-off-gas	(2)PENEX-off-gas	(3)HCK-recycle- gas-bleeding-to-FLH	(4)HCK-High-P-Sep-off-gas	Natural gas to SMR
Vapor fraction	1	1	1	1	1
Temperature (°C)	26	40	76	55	33
Pressure (barg)	10.2	10.2	160	29	3.4
Molar flow (Nm ³ /hr)	2700	3200	5000	9000	-
	Composition, % mol				
CH ₄	3.1	1.5	4.6	8.1	92.3
C ₂ H ₆	4.6	1.7	1.8	4.3	3.9
C ₃ H ₈	4.4	3.3	1.1	2.7	0.6
i-C ₄ H ₁₀	6.5	8.9	0.4	0.8	0
n-C ₄ H ₁₀	12	17.4	0.4	0.7	0
i-C ₅ H ₁₂	0.2	0	0.1	0.3	0
n-C ₅ H ₁₂	0.1	0	0.1	0.2	0
C ₆₊	0	0	0.1	0.3	0
H ₂	66.9	67.2	89	73.2	0
N ₂	0	0	0	0	0.6
H ₂ S	2.2	0	2.3	9.4	0
CO	300 vppm	<0.2vppm	0	0	0
CO ₂	1700 vppm	269 vppm	0	0	2.6
HCl	0	0	<0.2 vppm	0	0
NH ₃	0	0	8 vppm	50 vppm	0

Table 2 outlines the conditions and specifications of natural gas fed to the SMR and various off-gas streams rich in hydrogen. For instance, stream number 3, directed from the hydrocracker unit to the FLH, exhibits a flow rate of 5000 Nm³/hr. It is noteworthy that this stream is a bleed from the recycled gas to the reactor. This practice is implemented to uphold the hydrogen purity required for the hydrocracker catalyst. Additionally, the table reveals that there are three streams: 2700 Nm³/hr from the naphtha hydro-treating unit (stream number 1), 3200 Nm³/hr from the isomerization unit (stream number 2), and 9000 Nm³/hr from the hydrocracker unit high pressure separator (stream number 4). These streams are hydrogen-rich and possess minimal impurities. After undergoing desulfurization, they are directed to the fuel gas system to utilize their calorific value as shown in Figure 1. It is essential to acknowledge these four hydrogen-rich streams as a valuable source of hydrogen. So, a modification has been proposed on the previously mentioned case, such that streams (1-4) are treated from sour gases, LPG is extracted and hydrogen is purified and recycled back to the network.

3. Methods

The proposed process was simulated using Aspen HYSYS® v11. Two fluid packages were employed. The Peng- Robinson equation of state was applied to compute the thermodynamic properties of the refinery off-gas streams and the refrigerant. The second fluid package utilized is the Acid Gas-Chemical Solvent package, which was employed for sour gas treatment.

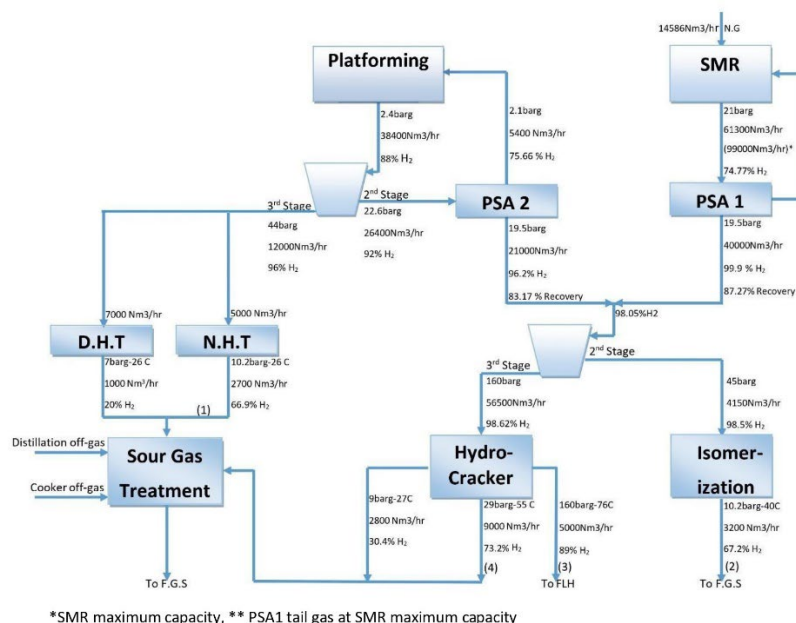


Figure 1. The current hydrogen network to be studied

3.1. Process description

As depicted in Figure 2, the off gases from the naphtha hydro-treating (NHT) unit (stream number 1) and the isomerization unit (stream number 2) are initially pressurized from 10.2 barg to 23.55 barg using compressor K-01. Subsequently, they undergo cooling to around 51°C in Ex-01. The off gas from the (HCK) bleed (stream number 3) operates at high pressure, allowing us to take advantage of its energy by directing it to turbine-01. This turbine generates sufficient energy to power compressor K-01. Due to the pressure reduction of stream 3, its temperature decreases. It then undergoes heat exchange in Ex-01 before being mixed with the other three streams.

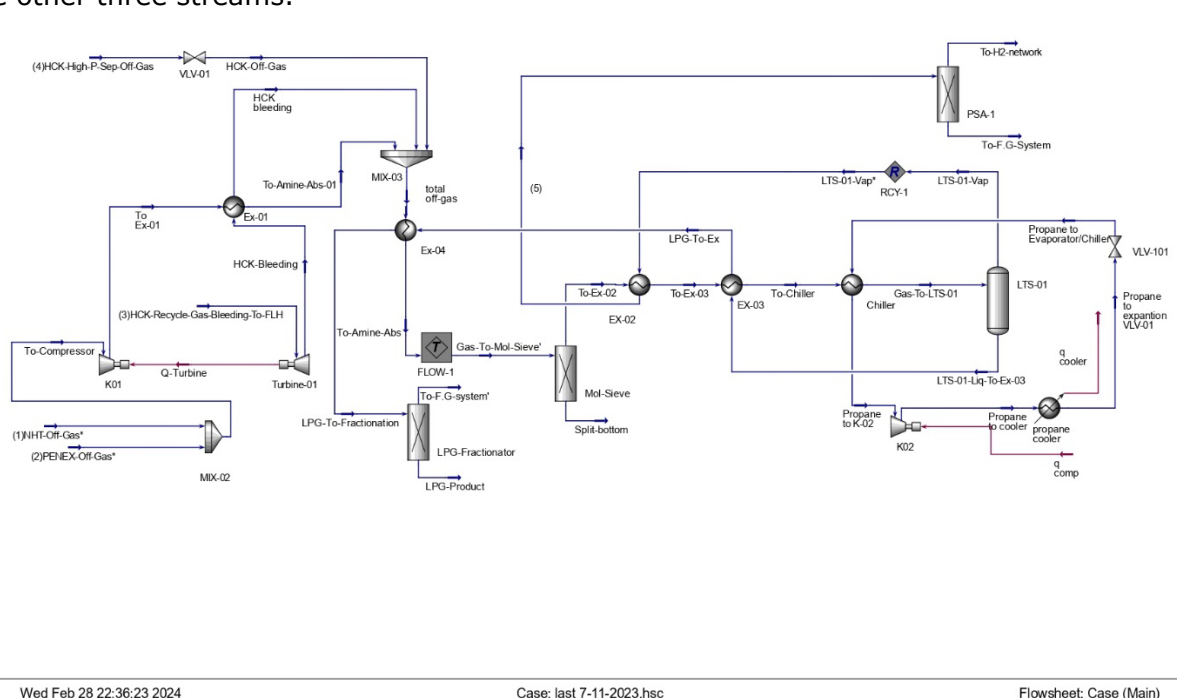


Figure 2. The HYSYS flow sheet.

These four streams are combined into a single stream, referred to as "total off gases" and are directed to Ex-04. This step not only imparts heat to the product LPG but also cools the stream to 42°C. Subsequently, the total off gases is routed to the de-sulfurization section. Here, the H₂S content is reduced from 5.13 volume % to 0 volume % through counter-current contact with a 37% weight basis methyl di-ethanol amine solution. This solution is already employed in the sour gas treatment section, as indicated in Figure 1.

The off gases are then directed to the water removal section. In this section, a molecular sieve bed acts as an adsorbent to decrease the water content from 0.7 weight % to 0 weight %. This water content arises from the contact between gas and amine solution and could potentially lead to hydrate formation in the refrigeration section. Following these steps, the off gases are progressively cooled in Ex-02, Ex-03, and a Chiller, ultimately reaching a temperature of -35°C. The chiller functions as the evaporator of the propane refrigeration loop. The cooled stream is then directed to the low temperature separator (LTS) to enhance hydrogen purity and to separate condensed LPG.

The recovered LPG is first heated in Ex02 to approximately 1°C. Subsequently, it undergoes further heating in Ex04 to about 30°C before being sent to the LPG fractionation tower for vapor pressure adjustment. The vapors from the LTS are directed to Ex02 for heating and to attain the required feed temperature for PSA1. Finally, these LTS gases, stream number 5, are directed with the effluent from the SMR unit. The combined stream is then directed to PSA1, as illustrated in Figure 3.

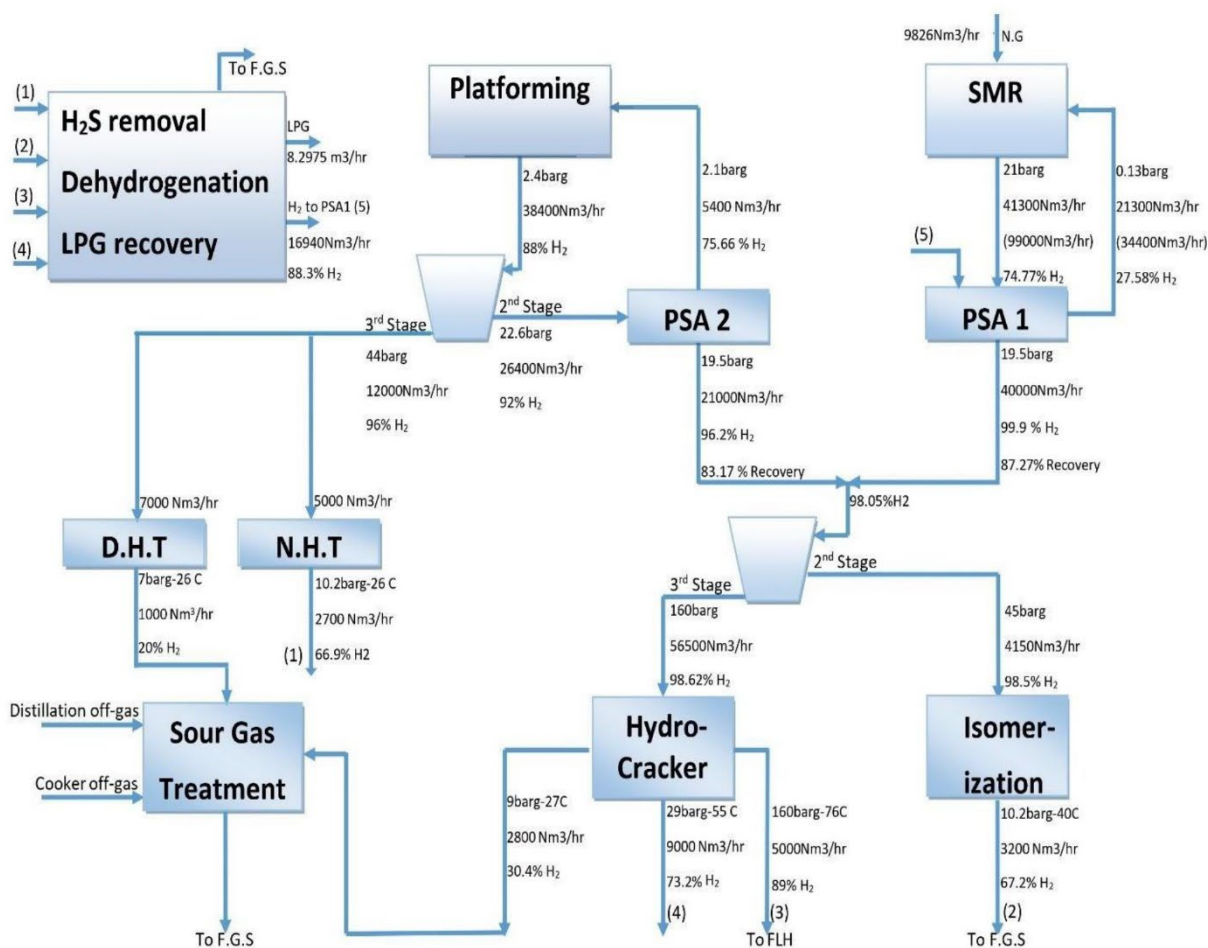


Figure 3. The hydrogen network after the proposed modification

Due to the process described above, approximately 16,940 Nm³/hr of hydrogen rich gas is successfully recovered, with a hydrogen purity of 88.3%. This recovered stream, stream number 5, is sent to PSA1 where a 13,050 Nm³/hr of hydrogen with 99.9% purity can be obtained. This highly pure recovered hydrogen is then integrated into the hydrogen network, thereby diminishing the demand on SMR production capacity. Additionally, a total of 7.35 m³/hr of stabilized LPG is successfully reclaimed as illustrated in Table 3.

Table 3. Products conditions and composition.

	Unit	Hydrogen to PSA1 (5)	LPG-Product	To-H2-network
Vapor fraction		1.00	0.00	1.00
Temperature	C	40.00	16.28	49.70
Pressure	barg	21.00	18.50	19.50
Molar flow	Nm ³ /h(gas)	16940.50	1656.77	13054.86
Mass flow	kg/h	3703.77	4166.51	1174.20
Liquid volume flow	m ³ /h	25.42	7.35	16.81
Heat Flow	kJ/h	-7289118.93	-10882980.00	410414.82
Hydrogen	mole %	88.30	0.00	100.00
Ammonia	mole %	0.00	0.00	0.00
H ₂ S	mole %	0.00	0.00	0.00
CO	mole %	0.02	0.00	0.00
CO ₂	mole %	0.02	0.00	0.00
Methane	mole %	6.28	0.00	0.00
Ethane	mole %	3.06	1.81	0.00
Propane	mole %	1.21	16.39	0.00
i-Butane	mole %	0.52	25.72	0.00
n-Butane	mole %	0.58	50.51	0.00
i-Pentane	mole %	0.01	2.13	0.00
n-Pentane	mole frac	0.00	1.52	0.00
n-Hexane	mole frac	0.00	1.92	0.00
MDEA	mole frac	0.00	0.00	0.00
H ₂ O	mole frac	0.00	0.00	0.00

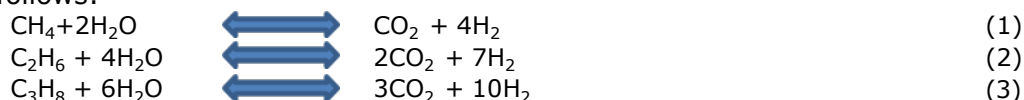
For context, the steam methane reforming process originally consumes 14,586 Nm³/hr of natural gas as per design, aiming to generate 40,000 Nm³/hr of hydrogen with a purity of 99.9%, as depicted in Figure 1. Consequently, the 13,050 Nm³/hr of recovered hydrogen at a purity of 99.9% is equivalent to utilizing 4760 Nm³/hr of natural gas as feed for the SMR unit. This equivalency translates to a remarkable 32.6% reduction in the SMR production capacity.

4. CO₂ reduction calculation

The amount of CO₂ reduced as a result of this study can be calculated from the difference between the CO₂ that was produced during the production of the 13,050 Nm³/hr of hydrogen and the CO₂ that will be produced from the natural gas that substitutes the calorific value of 13,050 Nm³/hr of hydrogen

4.1. Steam methane reformer reactions

The composition of the natural gas used in the refinery is shown in Table 2. The average molecular weight of this natural gas is 17.557, and its calorific value equals 34.29 MJ/Nm³. The reaction of natural gas with steam in the steam-methane reforming can be shown as follows:



From equation 1, it is obvious that 1 mole of methane gives 4 moles of hydrogen and 1 mole of carbon dioxide, but the conversion of this reaction is about 76%, as per design, so 1

mole of methane gives 3.04 moles of hydrogen and 0.76 mole of carbon dioxide. Similarly, for equation 2, 1 mole of ethane gives 7 moles of hydrogen and 2 moles of carbon dioxide, and the conversion of this reaction is about 100 %. For equation 3, 1 mole of propane gives 10 moles of hydrogen and 3 moles of carbon dioxide, and the conversion of this reaction is about 100 %. And according to Avogadro's law which states that the number of moles of a gas is directly proportional to the gas volume at constant conditions of temperature and pressure. Therefore, we can say that 1 Nm³ of methane gives 3.04 Nm³ of hydrogen and 0.76 Nm³ of carbon dioxide.

4.2. Calculation of the amount of natural gas substituting the recovered hydrogen

The total amount of hydrogen recovered and recycled back to the network is about 13,050Nm³/hr. Before recovery some of this hydrogen quantity was directed to the flaring system, and the rest of the hydrogen quantity was directed to the fuel gas system. The hydrogen amount that was sent to the fuel gas system from streams 1,2 & 4 is calculated from Table 2 as follows;

$$QH_2 \text{ (Nm}^3\text{/hr)} = \sum [Q_i \text{ (Nm}^3\text{/hr)} * Y_i(\%)] \quad (4)$$

where i is stream number (1,2 and 3); QH₂ total amount of hydrogen in streams 1,2 and 3; Q_i- stream I flow rate; Y_i- hydrogen mole fraction of stream i.

Therefore, QH₂=2700Nm³/hr*0.669+3200Nm³/hr*0.672+9000Nm³/hr*0.732=10,545 Nm³/hr

The hydrogen amount that was sent to the flaring system from stream 3 is calculated in the same manner as follows; 5000Nm³/hr*0.89=4450 Nm³/hr.

Since the calorific values of hydrogen and natural gas respectively are 10.8 MJ/Nm³ and 34.29 MJ/Nm³. The calorific value of the hydrogen sent to the fuel gas system can be calculated as follows;

$$CTH_2 \text{ (MJ/hr)} = QH_2 \text{ (Nm}^3\text{/hr)} * CH_2 \text{ (MJ/Nm}^3\text{)} \quad (5)$$

where CTH₂ : The total calorific value from H₂; QH₂: total amount of hydrogen in streams 1,2 and 3. CH₂ : hydrogen calorific value.

Therefore, CTH₂= 10,545 Nm³/hr * 10.8 MJ/Nm³=113,886 MJ/hr.

This quantity of energy can be substituted with natural gas of the following amount;

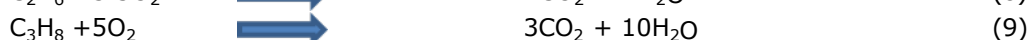
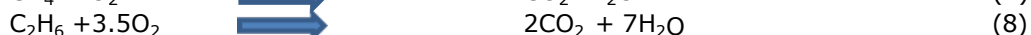
$$QNG \text{ (Nm}^3\text{/hr)} = [(CTH_2 \text{ (MJ/hr)}) / (CNG \text{ (MJ/Nm}^3\text{)})] \quad (6)$$

where QNG : amount of natural gas required; CNG: natural gas calorific value.

Therefore, QNG= (113,886Mj/hr) / (34.29Mj/Nm³) = 3,321.26 Nm³/hr.

4.3. CO₂ calculation from natural gas combustion

The complete combustion reactions of natural gas are shown as follows;



The amount of carbon dioxide produced during the complete combustion of the 3,321.26 Nm³/hr of the substituting natural gas can be calculated as follows;

$$QCO_2 \text{ (Nm}^3\text{/hr)} = (Q \text{ N.G (Nm}^3\text{/hr)}) (\sum X_i (\%) * N_i) \quad (10)$$

where QCO₂ total amount of CO₂ produced; i different components of the natural gas; X_i mole fraction of components i in the natural gas as shown in Table 2; N_i number of moles of CO₂ produced from combustion of component i as illustrated in section 5.1.1.

Therefore, QCO₂=(3,321.26 Nm³/hr)*(0.923*1+0.039*2+0.006*3+0.026)=3,470.72 Nm³/hr

4.4. CO₂ calculation from the production of 13,050 Nm³/hr of hydrogen with 99.9%purity

As mentioned previously, the amount of natural gas required to produce the saved 13,050 Nm³/hr of hydrogen is 4762 Nm³/hr. The amount of CO₂, produced at the SMR when 4762 Nm³/hr of natural gas is used, can be calculated as follows;

$$Q'CO_2 \text{ (Nm}^3\text{/hr)} = (Q \text{ N.G (Nm}^3\text{/hr)}) (\sum (X_i (\%) * N_i * Z_i)) \quad (11)$$

where $Q'CO_2$: total amount of CO_2 produced; i : different components of the natural gas; X_i : mole fraction of components i in the natural gas as shown in Table 2; N_i : number of moles of CO_2 produced from combustion of component i ; Z_i : conversion % of component i .
Therefore, $Q'CO_2 = (4762 \text{ Nm}^3/\text{hr}) (0.923 \cdot 1 \cdot 0.76 + 0.039 \cdot 2 + 0.006 \cdot 3 + 0.026) = 3,921.41 \text{ Nm}^3/\text{hr}$.
The amount of CO_2 reduced as a result of this study can be calculated as follows;
 $CO_2 \text{ reduced (Nm}^3/\text{hr)} = Q'CO_2 (\text{Nm}^3/\text{hr}) - QCO_2 (\text{Nm}^3/\text{hr}) = 3,921.41 - 3,470.72 = 450.69 \text{ Nm}^3/\text{hr}$.

4.5. Economic analysis

Aspen Process Economic Analyzer (APEA) V.14 is used to calculate the economics of this study. The results are shown in Table 4.

Table 4. APEA cost calculations.

Total project capital cost	18 202 705	USD
Total operating cost	2 473 375	USD/Year
Total raw materials cost	0	USD/Year
Total utilities cost	447,662	USD/Year
Total product sales	21 727 724	USD/Year
Desired rate of return	20	Percent/Year
P.o. period	2.91	Year

5. Conclusion

The research has demonstrated the importance of the recovery of hydrogen from refinery off gases. Our findings not only contribute to the optimization of the hydrogen network in refineries, but also offer a pathway towards reduced environmental impact, and improved economic performance for the refining industry. The most important achievement of this study is the recovery of a $13,050 \text{ Nm}^3/\text{hr}$ of hydrogen. This study shows the efficiency and feasibility of our recovery method. Equally significant is the remarkable 32.6% reduction in steam methane reforming (SMR) capacity. This reduction in SMR capacity signifies a critical shift in the resource utilization model, allowing for enhanced resource allocation and energy efficiency.

In addition to the substantial hydrogen recovery, the research has delivered a noteworthy recovery of $7.35 \text{ m}^3/\text{hr}$ of liquefied petroleum gas (LPG). This co-recovery of LPG reinforces the numerous benefits of our methods, extending beyond hydrogen to other valuable byproducts. Perhaps one of the most important outcomes of this study is the reduction in carbon dioxide emissions by an impressive $450.69 \text{ Nm}^3/\text{hr}$. This reduction aligns perfectly with the global imperative to mitigate carbon emissions and underscores the environmental stewardship inherent in our approach. From an economic standpoint, this study demonstrates a highly favorable payback period of just 2.9 years for this project. This quick return on investment underscores the economic viability of implementing our innovative hydrogen recovery methods within the refining industry. As we look forward, it is our sincere hope that these results will serve as a catalyst for industry-wide adoption of more sustainable and efficient hydrogen recovery practices. This study not only aligns with the demands of an increasingly eco-conscious world but also holds the promise of economic prosperity for refineries.

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