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OFFSHORE H_2 PRODUCTION VIA SMR WITH CO_2 CAPTURE AND STORAGE

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Abstract. *Despite being widely used in the fuel and fertilizer production and as raw material in several chemical processes, only now hydrogen has been gaining importance. The most used process for hydrogen production, steam methane reforming (SMR), has significant CO_2 emissions (9.5kgCO_2/kgH_2) in spite of the absence of carbon in its molecule. Processes that make use of renewable energy combined with hydrolysis of water are still excessively expensive for large scale production and suffer with the intermittency related to most of these energy sources. This work proposes offshore production of hydrogen via SMR with reduced CO_2 emissions through carbon capture. The production of hydrogen on oil and gas platforms, or on near barges, aims to make use of the natural gas produced, sea water for the production of steam and as cooling fluid and the oil and gas wells for CO_2 storage. These synergies are evaluated in this paper, from an energetic and environmental perspective, in order to find a technically and economically viable solution to the growing demand for hydrogen.*

Keywords: *hydrogen production, carbon capture, desalinization, offshore, AspenPlus*

1. INTRODUCTION

Hydrogen is an important energy carrier (Nanakı, Koroneos, & Xydıs, 2017) which has been considered as an alternative to reduce CO_2 emissions in various sectors of society (Bockris, 1991). With regard to the production of H_2 offshore, there is a great incentive in the coastal area, due to the intense activity of refining and chemical processes that use hydrogen based on fossil fuels in coastal industrial zones around the world such as the North Sea in Europe, Gulf Coast in North America and Southeast China (IEA, 2019). H_2 production plants can supply ships and trucks serving ports and feeding other nearby industrial facilities (IRENA, 2019).

The hydrogen production route is a determining factor for its environmental performance (Nanakı et al., 2017). This fuel can be produced by renewable energy sources, nuclear energy, natural gas and other oil derivates (ERIA, 2019). The renewable energy sources that have been used are: wind power, solar energy, hydroelectric power, and biomass (Christopher & Dimitrios, 2012). Most of the found search related to offshore H_2 production are via electrolysis supported by renewable energy sources. Serna & Tadeo (2014) evaluates the design of an offshore electrolysis plant powered by wave energy. Hogerwaard et al. (2019), defends solar energy to generate electricity in a membrane reactor and enable water electrolysis. Gondal (2019), argues that offshore wind energy has been standing out in relation to onshore due to the greater energy production capacity, in addition to the availability of space for the installation of wind farms, which must be installed in regions further from the coast. D'Amore-Domenech & Leo (2019), analyzes different methods of water electrolysis for H_2 production based on existing environmental impacts and energy consumption, it was concluded that large-scale production by this method is not profitable, as its price is not competitive compared to other methods of production. The use of renewable energies such as wind and solar, through water electrolysis, has a high cost, 3.0-7.5 USD/kgH_2 (IEA, 2019), due to underutilization of the plants which rely on intermittent energy sources and the energy consumption is 1.65-1.29kWe/kgH_2 (Suleman, Dincer, & Agelin-Chaab, 2016). The discussion about converting offshore energy resources to H_2 is not a recent one, but there are still limited studies on this approach.

Hydrogen transport, distribution and storage are the main challenges for the H_2 technology integration in the global energy saving system, the H_2 can be transported by gas pipelines in gas form or in liquid form via ships (United States Department of Energy, 2020). This gas is being used today, mainly in the production of fertilizers, oil refining, power generation and heating. In the future, it is estimated an increase in the use of hydrogen as fuel in cars, trucks, ships and airplanes (IEA, 2019). Currently, 120 million tons of H_2 are produced per year (IEA, 2019) about 95% of all this hydrogen is generated from natural gas, via steam methane reforming (SMR) (Bakey, 2015). This method is responsible for considerable carbon dioxide emissions, having an emission factor of 9.5 $kgCO_2/kgH_2$ (IRENA, 2019). When the steam reforming method is associated with a carbon capture and storage system, this factor can be reduced to 1.8 $kgCO_2/kgH_2$, reducing the impact on Earth temperature and water and soil contamination (Nanakı et al., 2017), being the production cost of these plants of 1.7-2.8 USD/kgH_2 (IEA, 2019), that is, a value significantly lower than that of H_2 from electrolysis.

The purpose of this work is to evaluate the offshore H_2 production in large scale by the SMR method from the energy and environmental point of view (CO_2 emission) considering the capture and storage of CO_2. The hydrogen offshore production aims to obtain synergies related to the availability of water, natural gas and the proximity to the carbon storage site. The synergies can be explained as follows: In the offshore process, the natural gas consumed is easily accessible, the captured CO_2 is stored in the mature well and water is available in abundance for the reform reaction and as cooling fluid. On the other hand, for onshore H_2 production, natural gas must be transported to the coast, water has a higher cost and the captured CO_2 must be sent back to the offshore platform. From the parameters presented by (Silva, 2013) for an offshore-onshore transmission pipeline, and the results of this work, it is possible to estimate the amount of energy needed for the displacement of natural gas, CO_2 and H_2 in 1253km, consisting of 10 compression stations. For this estimation must be take a count the pressure variation along the pipeline, on this way, the variation is 73-97bar for natural gas (J. A. M. Silva, 2013), 48-69bar for H_2 (Witkowski, Rusin, Majkut, & Stolecka, 2017) and 100-150bar for CO_2 (Peletiri, Rahmanian, & Mujtaba, 2018). For gas transport, in addition to using the natural gas pipeline separately, there is a potential possibility of transporting hydrogen in a mixture with natural gas (United States Department of Energy, 2020). In this context, further research should be carried out to assess which is the best transport alternative for this gas.

2. PROCESSES DESCRIPTION AND METHODOLOGY

Figure 1 presents the integration between the plants involved in this work. The hydrogen plant uses natural gas from an offshore well and seawater, after desalinization, to produce a stream rich in H_2 e generates an undesirable stream of acid water. The carbon capture plant using chemical absorption receives a stream rich in H_2 and the solvent composed of recycled amine. In the end of the carbon capture process, the stream rich in H_2 leaves the absorption column and goes to the purification process and can reach up to 99.999% (mol) of purity, while the stream rich in CO_2 leaves the regeneration column and is directed to the compression plant to be stored. The desalinization plant receives seawater and produces desalinated water and brine. The plants can use the seawater as cooling fluid (Cuchivague, 2015) reinforcing the advantage of offshore installation, and are powered by heat and electricity generated at the utility plant associated reformer furnace of hydrogen plant. This equipment is feeded by natural gas, air and gaseous residue from purification process generating an exhausted gas stream post-combustion. The reformer furnace was simulated separately: the reform was simulated by a Gibbs reactor at hydrogen plant and the combustion by a stoichiometric reactor at utilities plant.

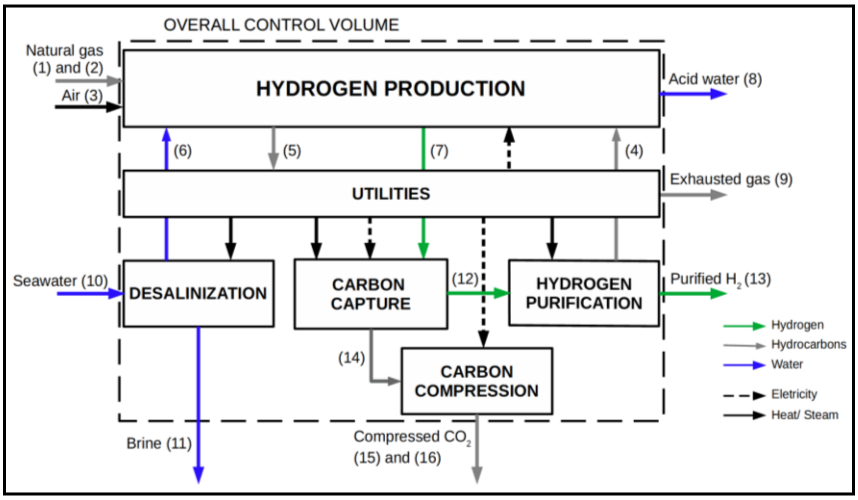

Figure 1 – Flowchart of integrated subsystems

Chemical processes were simulated in Aspen Plus™ and the results were verified using data from the literature. Mass and energy balance are used in all subsystems as well as in the global system which are modeled on a permanent regime.

2.1 Hydrogen production

The H_2 production via steam methane reform involves the natural gas pre-treatment, pre-reform, reform and the stage known as water gas shift. The thermodynamic model used to model this plant was Peng Robinson, due to the hydrocarbons reactions (Lira, Lima, & Lucena, 2018) and the simulated process is presented in Fig. 2.

The natural gas pre-treatment removes liquid components and sulfur (Molburg & Doctor, 2003). Desulfurization is necessary to avoid sulfur adsorption on the surface of the reformer's catalyst, which may cause reduction in catalytic activity and a consequent reduction in the total efficiency of the plant. In this step the sulfur is hydrogenated, forming H_2S, and it is adsorbed on a bed of zinc oxide forming zinc sulfide, which is removed as a solid residue (Jakobsen & Åtland, 2016), in this work it is assumed that natural gas has neither water nor sulfur in its composition, therefore the pretreatment stage is not simulated.

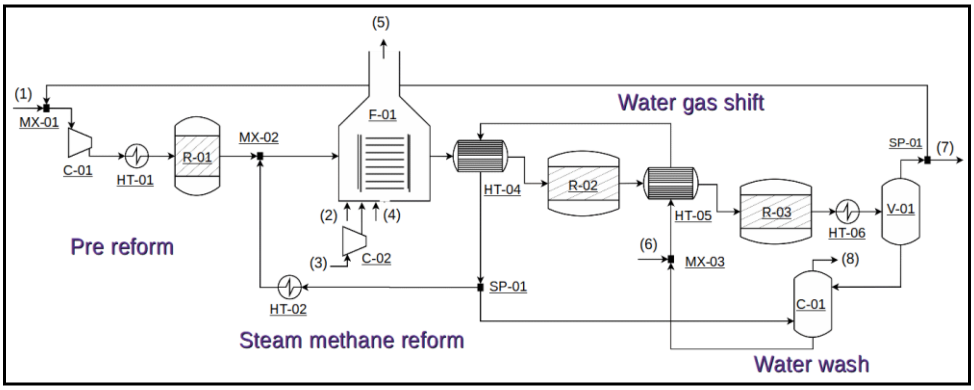

Figure 2 - Hydrogen production

In the pre-reform the heavy hydrocarbons are converted into methane using a recycling stream of H_2 (0.8% of stream) (Molburg & Doctor, 2003), was simulated in Gibbs reactor at 343ºC and 26bar. The reform concentrates the highest thermal demand of the plant due to the endothermic reactions being carried out in the reformer at 650-850^{O}C (Cruz, 2010; Wu & Kuo, 2015) and it consists of several internal tubes filled with a nickel-based catalyst (Jakobsen & Åtland, 2016). In this reformer, the methane is converted into hydrogen as Eq. (1) (Amran, Ahmad, & Othman, 2017; Al Khusaibi & Rao, 2016; Lira et al., 2018) with the steam to natural gas proportion 3:1 and for to simulate this reaction was used the Gibbs reactor. The reform is supported by a reformer furnace that provides heat for the reform reaction and heats process currents through the heating sections (F. E. Da Cruz, 2010).

$$CH_4 + H_2O \rightarrow CO + 3H_2 \qquad \Delta h_r = + 251 \text{ MJ/ kmolCH}_4 \qquad (1)$$

After the reform, the gas stream is direct to the water gas shift stage, characterized by two exothermic conversion reactors, normally using iron-based or copper-based catalyst in order to guarantee 90% of carbon monoxide conversion into hydrogen, Eq. (2). The first reactor operates at 350ºC and the second at 204ºC (Molburg & Doctor, 2003) both in the stoichiometric reactor.

$$CO + H_2O \leftrightarrow CO_2 + H_2 \qquad \Delta h_r = - 41.2 \text{ MJ/ kmolCO} \qquad (2)$$

After the water gas shift stage, the H_2 rich product is taken to a flash type separator, where the liquid and gaseous phases are separated (Jakobsen & Åtland, 2016) at 37.77ºC and 5.6bar. The liquid phase has hydrocarbons traces, acid water, and is washed by a steam stream for process reuse (F. Cruz & De Oliveira, 2008; Molburg & Doctor, 2003) by an Radfrac column at 1bar. The final stream of hydrogen production plant is sent to the carbon capture plant and, subsequently, to the pressure swing adsorption purification process.

2.2 Desalinization

The main commercial desalinization technologies are thermal distillation and membrane separation. Membrane separation methods (reverse osmosis) use pressure gradients across a semipermeable membrane to filter out unwanted minerals and salt dissolved in seawater. In thermal distillation, changes in sea water pressure and temperature cause fresh water to evaporate and condense as distillate (Hogerwaard, Dincer, Naterer, & Patterson, 2019). Multi-effect distillation and multi-stage flash distillation are of better quality and require less post-treatment for demineralization (Meier, 2014). Post-treatment includes chemical treatment to bind the remaining ions and other solids dissolved in desalinated water to prevent scale and the use of chemicals is undesirable on an offshore platform (Meier, 2014).

In view of the large-scale production of H_2, associated with the production of steam, the desalinization of seawater by multiple effect distillation is the alternative with the greatest synergy. This method is characterized by the water and salts separation by evaporators in series, in which the heat necessary for the evaporation of the first effect is transferred to evaporate the fluid of the next effect. The ELECNRTL thermodynamic model and eight evaporators configuration was used to model the desalinization plant, as show in Fig. 3. The first evaporator is at 57.7ºC and 0.17 bar and the last at 40ºC and 0.07 bar.

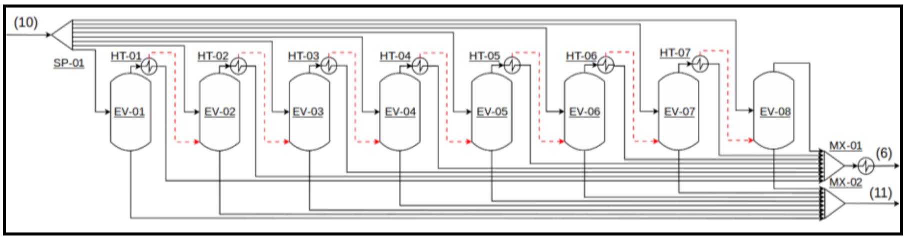

Figure 3 – Desalinization

2.3 Carbon Capture

The most known technologies for CO_2 capture are: absorption, adsorption and membrane permeation. They are already applied in large scale and are commercially mature (Queiroz, Araújo, Carvalho, Paula, & Musse, 2017). The most used technology for removing CO_2 from acid gases is absorption with amine. Although chemical absorption presents economic and environmental disadvantages in relation to solvent loss, high energy demand and equipment corrosion, this technology has an excellent performance in terms of solubility and selectivity (Barbosa, 2018), furthermore availability of energy is usually not a constrain in gas and oil offshore platforms (R. de P. Silva, 2018). In the CO_2 absorption process, ethanolamines solutions are generally used. Among the most used ethanolamines are monoethanolamine (MEA), diethanolamine (DEA) and methyl diethanolamine (MDEA), which are primary, secondary and tertiary amines, respectively (Marin, Ferreira, Queiroz, & Fortini, 2016). In Brazilian refineries, almost all gas treatment units from fluidized catalytic cracking and hydrotreating plants operating with amines use DEA. In Hydrogen Generation Units (UGH), MEA are used to purify the synthesis gas with chemical absorption by amines instead of physical adsorption in order to obtain greater purity (Martins, 2011). The carbon capture by absorption used in this work, as shown in Fig. 4, consists of an absorption column followed by the regeneration column.

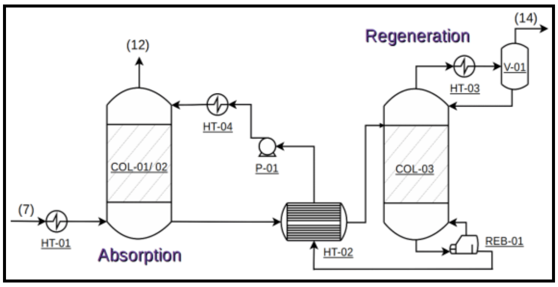

Figure 4 – Carbon capture

In the absorption column, the gas stream enters from the bottom and the solvent, monoethanolamine (MEA), from the top, forming a counter current flow in which CO_2 is captured by the solvent leaving through bottom, and the H_2 rich gas leaves through the top (Martins, 2011). Usually the solvent flow is between 10 to 15 times the gas flow (R. de P. Silva, 2018), in this work is considered 10 times, at 52ºC and 26bar, with 20% (wt) MEA. The top stream rich in H_2 is directed to the PSA process to guarantee the purity of the final product and the bottom stream rich in captured CO_2 by solvent is directed for the regeneration column. The regeneration performance is very sensitive in relation to the reboiler temperature, which cannot exceed 130°C to avoid degradation of solvent (Barbosa, 2018). After the regenerating, a stream rich in free captured CO_2 and a recovered solvent stream, which is recirculated back to absorption column, are generated. Following capture, the CO_2 rich stream is sent to the compression train to achieve the pressure required for injection. The ENRTL-RK thermodynamic model is adopted for modeling, due to eletrolytes of solvent and gases presence (Cuchivague, 2015). Due to an offshore scenario and the height restrictions of the absorption column, two absorption columns with 10 plates at 26 bar each and a regeneration column with 10 plates at 3 bar were considered.

2.4 Hydrogen purification

Pressure swing adsorption (PSA) is an industrial unit for the separation and purification of gas mixtures that operates from the capacity of adsorption of solids and selective separation of gases. In this technique, the separation of a specific gas occurs with adsorption at a higher pressure and desorption at a lower pressure (Asgari, Anisi, Mohammadi, & Sadighi, 2014), from a dynamic that operates cyclically, in which each bed goes through the same sequence of steps (Nikolic, Giovanoglou, & Georgiadis, 2006). Currently, this technology is used in petrochemical industries to produce pure H_2 from a mixture containing between 60 and 90% (mol) of H_2 (Asgari et al., 2014 ;Nikolic et al., 2006). At the end of the process, a product rich in H_2 in the pressure of the feed gas is guaranteed with a purity of 98-99.999% (mol) with recovery of 70-90% H_2 from the initial gas stream (Nikolic, Giovanoglou, & Georgiadis, 2006). The PSA plant was not simulated, so the process is presented by a fluxogram as Fig. 5 and according to Villaça, the specific thermal consumption of this process is 2.6kWt(medium pressure steam)/kgH_2.

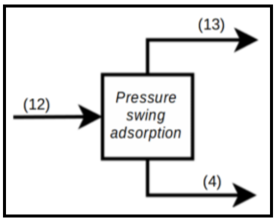

Figure 5 – Hydrogen purification

2.5 Carbon compression

The injection of CO_2 in an offshore well can, sometimes, help in oil recovery, technique known as advanced oil recovery (EOR). CO_2 injection reduces the viscosity and density of the oil, due to the miscibility between them, improving fluidity and increasing its elevation (R. de P. Silva, 2018). This technology is intended to increase productivity and useful life of the well, recover additional oil and mitigate climate change by reducing CO_2 emissions (Eide et al., 2019). According to (ARAÚJO et al., 2016), the compression pressure must be 30,000kPa, whereas used 45,000kPa (F. C. N. Silva, Freire, Orrego, & Oliveira, 2020). Compression is characterized by a compressor followed by a heat exchanger and separating vessel. Thus, it is possible to avoid the temperature rise due to the pressure increase and remove residual water from the stream. The ENRTL-RK thermodynamic model was used to obtain the thermodynamic properties during compression. Two parallel compression trains with three compressors in series were used, as is presented in Fig. 6. Polytropic efficiency of 75% and intercooling at 40 ºC were assumed.

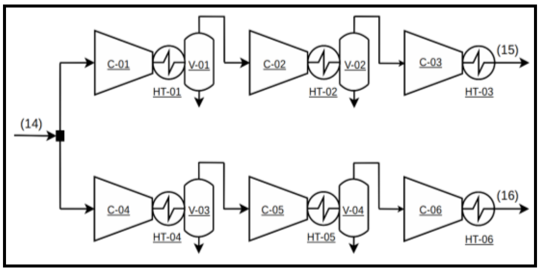

Figure 6 - Carbon compression

2.6 Utilities

In conventional offshore oil platforms, the combined production of heat and electricity currently depends on gas turbine systems that operate with less efficiency and greater environmental impact compared to modern onshore cogeneration plants (FCN Silva, Freire, Orrego, & Oliveira, 2020). The energy demand of H_2 production plants, carbon capture, CO_2 compression, H_2 purification and desalinization is met by H_2 reformer furnace. This equipment can provides heat for the reform reaction and heat H_2 plant streams through the heating sections (F. E. Da Cruz, 2010).

According to Fig. 7, there are two heating sections on the reformer furnace, the preheating section, for stream before pre reform reactor (HT-01), and a section for generating superheated steam for SMR feed (HT-02) of H_2 plant. To supply the thermal demand of the other processes, a heat recovery is used to generate steam from the use of the exhausted gas stream from the reformer furnace. Electric demand is supplied by an exhaust gas expander before being directed to the heat recovery. The combustion zone of the reformer furnace is fed by natural gas, air and also by the gas residue from the PSA plant (F. E. Da Cruz, 2010).

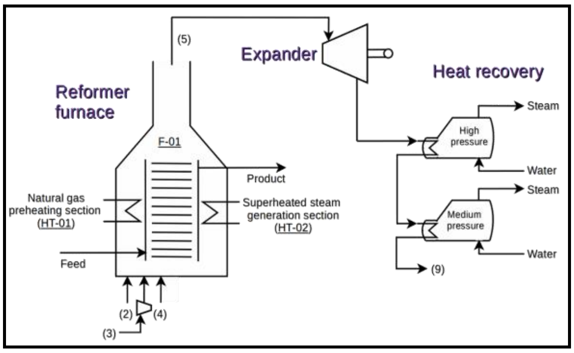

Figure 7 – Utilities

3. RESULTS

The results obtained for H_2 plant (Tab. 1), carbon capture (Tab. 2), desalinization (Tab. 3) and CO_2 compression (Tab. 4), were verified using previous works.

Table 1 – Specific consumption for the hydrogen production plant

	Al Khusaibi & Rao (2016)	Jakobsen & Åtland, (2016)	Molburg & Doctor (2003)	Boyano et al. (2011)	Simulated
$kgNG/kgH_2$	0.32	0.27	0.38	0.38	0.34
kgH_2O/kgH_2	-	0.81	1.09	1.21	1.03
$kgH_2O/kgNG$	-	3.00	2.90	3.20	3.00
kWt/kgH_2	6.37	11.36	-	10.37	9.31
kWe/kgH_2	-	-	-	-	0.02

The differences regarding H_2 production are mainly due to the following points: Al Khusaibi & Rao (2016) do not take into account the thermal energy required by the pretreatment reactor; Jakobsen & Åtland (2016) simulated the stage of pre-reform, it increases hydrogen production; in the plant studied by Boyano et al. (2011), the reform temperature (700ºC) is smaller than the simulated in this work (850ºC). The temperature variation of the reforming furnace and the H_2 concentration at the end of the process were evaluated to determine the ideal temperature for this work.

Table 2 – Specific consumption for the carbon capture plant

	Martins (2011)	Silva (2018)	Simulated
$kgSolvent/kgCO_2in$	24.20	15.00	13.50
$kgSolvent/kgCO_2out$	27.50	21.07	19.98
$kWt^*/kgCO_2out$	1.19	0.56	1.19

The differences in relation to the carbon capture plant are: Silva (2018) simulates the CO_2 capture from a gas stream of 5,78%CO_2 (mol) with solvent 50% (wt) MDEA and PZ, while Martins (2011) used 20% (wt) MEA to capture carbon from an H_2 rich gas stream with 24.48%CO_2 (mol). In order to obtain a concentration greater than 95% CO_2 (mol) of the final stream, a high capture of CO_2 from the gas stream that enters the absorption column and less thermal demand in the reboiler, parameters such as pressure in the columns and distillate ratio and rate of reflux in the regeneration column are evaluated.

Table 3 – Specific consumption/ production for the desalinization plant

	Hamad et al. (2018)	Guimard et al. (2019)	Dong et al. (2019)	Elsayed et al. (2018)	Simulated
kgSeawater/kgH_2O	2.62	-	2.45	2.94	2.52
kgBrine/kgH_2O	1.62	2.02	1.45	1.95	1.52
kgSalt/kgBrine	0.058	-	-	0.07	0.065
kWt*/kgH_2O	0.07	0.07	-	-	0.10

The main differences found in relation to desalinization plant are: Hamad et al. (2018) and Guimard et al. (2019) simulated the plant with a steam thermo-compressor and Guimard et al. (2019) simulate 12 evaporators while the plant modeled in this work uses 8 evaporators.

Table 4 – Specific consumption for the compression

	Cuchivague (2015)	Rodrigues (2013)	Silva (2018)	Simulated
kWe/kgCO_2	0.09	0.18	0.12	0.09

In the work of Rodrigues (2013), the compression pressure is 568,600kPa using three compression stages. Cuchivague (2015), simulates two compression trains with four stages to reach a final pressure of 25,000kPa and Silva (2018) assesses six compression stages to 30,000kPa, while in this work is used two parallel compression trains with three compressors in series to 45,000kPa.

4.2. The Process

The processes were simulated in the Aspen Plus separately. In Tab. 5 is presented the results of the main currents highlighted in the methodology.

Table 5 - Results of main currents

	Flow rate (kg/h)	Temperature (ºC)	Pressure (bar)	Composition (mol)
1	6,000.00	21.11	20.00	0.95CH_4
2	6,500.00	21.11	20.00	0.95CH_4
3	130,000.00	25.00	4.00	0.79N_2; 0.21O_2
4	547.39	21.7	1.00	0.91H_2; 0.04N_2; 0.03H_2O; 0.02CO
5	137,047.39	700.00	5	0.70N_2; 0.15H_2O; 0.079CO_2; 0.028O_2; 0.024H_2
6	13,097.30	40.00	2.00	1.00H_2O
7	18,511.80	37.77	5.86	0.76H_2; 0.19CO_2; 0.03CH_4
8	745.24	101.52	1.00	0.99H_2O; 0.01CO_2
9	137,047.39	123.10	1.6	0.70N_2; 0.15H_2O; 0.079CO_2; 0.028O_2; 0.024H_2
10	33,000.00	37.00	1.00	0.98H_2O; 0.01Na^+; 0.01Cl^-
11	19,902.7	40.07	0.07	0.96H2O; 0.02Na+; 0.02Cl-
12	2,794.15	25.00	10.00	0.99H_2; 0.004N_2; 0.003H_2O
13	2,246.76	25.00	1.00	0.9999H_2
14	11,067.00	20.00	15.00	0.95CO_2; 0.05CH_4
15	5,922.58	172.09	450.00	0.95CO_2; 0.05CH_4
16	5,922.58	172.09	450.00	0.95CO_2; 0.05CH_4

4.3 Electrical and thermal demand

The thermal and electricity demand of the subsystems are listed in Tab. 6 and 7, respectively.

Table 6 – Thermal demand of subsystems

Subsystem	kWt
Hydrogen production	HT-01=824.02; HT–02=3,332.47; F–01=26,793.60
Carbon capture	HT–01=315.03; REB–01=22,317.00
Hydrogen purification	6,000.00
Carbon compression	-
Desalinization	EV-01=1,400.00

Table 7 – Electric demand of subsystems

Subsystem	kWe
Hydrogen production	C–01=662.25; C–02=7,300.81
Carbon capture	P–01 = 223.32
Hydrogen purification	-
Carbon compression	C–01 à C–06=1,050.00
Desalinization	-

For cooling subsystems, seawater can be used as a cooling fluid. As already discussed, the H_2 plant has a thermal integration capable of generating heating/steam and electricity as was shown in Fig. 7. Based on the thermal load required for heating process streams of the H_2 plant and the need to generate medium and high pressure steam, the capacity of the reformer furnace was dimensioned at 1400ºC and 5 bar. Table 8 lists the utilities necessary for the thermal demand of the processes and the temperature variation of each utility stream.

Table 8 - Utilities necessary

Subsystem	Equipment	Utility	T_{in}/T_{out} (ºC)
Hydrogen production	HT-01	Heated gas	82.32/343.01
	HT-02	Heated gas	100.00/ 343.00
	F-01	Heated gas	387.13/ 855.88
Carbon capture	HT-01	Medium steam	200.00/120.00
	REB-01	High steam	350.00/128.00
Hydrogen purification	PSA Heating	Medium steam	200.00/120.00
Desalinization	EV-01	Medium steam	200.00/120.00

4. CONCLUSION

From the results obtained, it is possible to verify that the energy consumption for the production of H_2 via SMR is 13.92kWt/kgH_2 and 0.35kWe/kgH_2. In comparison to electricity consumption, it was found that consumption by SMR is, on average, four times less than consumption by water electrolysis (1.65-1.29kWe/kgH_2). Due to the high electrical consumption by electrolysis and the high cost of installing technologies for the supply of electricity by renewable sources, the SMR method is still a viable technology for the production of H_2, in addition to being an already consolidated process.

From the data obtained by simulations, is possible to estimation the energy amount needed for the displacement of natural gas, CO_2 and H_2, for the pressure variation, 73-97bar gas natural, 100-150bar CO_2 and 48-69bar H_2, in the natural gas pipeline discussed in introduction. The energy consumption for the transportation of H_2 is greater than that of natural gas and CO_2, 4.74kJ/kmkgH_2, 0.38kJ/kmkgCH_4 and 0.12kJ/kmkgCO_2 respectively. The difference between these values can be explained by the fact that the H_2 compression work is greater to achieve a certain mass flow rate at different gas volumes. Therefore, the compressors of the natural gas infrastructure are not sufficient to transport H_2 in the same proportions and the installation of transport lines exclusive for H_2 is not yet economically viable (Gondal, 2019). In this context, a potential alternative is to transport hydrogen in a mixture with natural gas (United States Department of Energy, 2020), where the pipeline supports between gas mixtures between 15-30% by volume of H_2 without significant effects on gas flow and compression (Gondal, 2019; United States Department of Energy, 2020).

The alternatives for the transport and subsequent storage of H_2 must still be better evaluated to guarantee an economically and environmentally viable process.

Undesired products of the simulated process are acid water, exhausted gas and brine. Due to the water washing step in the H_2 plant, the undesired acidic water flow is much lower than the seawater flow used for desalinization and the CO_2 concentration is negligible. In this case, this water can be disposed of in seawater without causing environmental damage. The exhausted gases must be treated to capture CO_2 to be injected into the well or to be sold or used as raw material for other products. The brine, on the other hand, can be sent back to the sea due to the high dilution power of the oceans or injected into deep wells, viable alternatives considering an offshore scenario, or used in the production of domestic salt.

For future works there are possibilities to do the dimensioning of equipment for adaptation on an offshore platform, explore other destinations for CO_2 such as methanol production and evaluate the feasibility of replacing natural gas with H_2 as fuel in the reformer furnace.

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