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ÚSTAV PROCESNÍHO INŽENÝRSTVÍ

Upgraded Methodology for Process Intensification in Natural Gas Dehydration

Rozšířená metodika pro intenzifikaci procesu vysoušení zemního plynu

DOCTORAL THESIS

DIZERTAČNÍ PRÁCE

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ABSTRACT

Natural gas dehydration (NGD) is essential in the processing of the associated and non-associated natural gas (NG). Its role is crucial in avoiding the hydraulic slugs, hydrate formation prevention, electrochemical corrosion control, beneficial production, and quality requirement fulfillment.

From the perspectives of capital and operational expenses (CAPEX and OPEX), energy consumption, pollution, and greenhouse gas emissions (GHG), the natural gas dehydration (NGD) has all the drivers that support approaching it via the modern process engineering concepts, such as process intensification (PI). The global requests to consider the health and environmental aspects of any development apply further pressure toward this implementation.

The literature review reveals a concern with the PI equipment, more than the PI methods and the concepts of the process itself. Furthermore, there is a limited utilization of the computer-aided simulation to serve the PI research.

This thesis presents an **upgraded methodology for PI in the NGD**. The developed method is a **systematic simulation-based one** that integrates the rated energy consumption (REC) with the dry gas water content specification. The presented method focuses on the mutual relation between the REC as a key driver and evaluation tool of the PI, and the water content specification which is the key input for NGD design. The REC is formulated of two components, process or equipment energy consumption, and enthalpy loss by venting.

A combination of two methods is used to formulate the upgraded methodology,

- a systematic method that recognizes unit intensification and extended process intensification.

- a concise framework for PI implementation in the O&G proposed originally by the author.

The established simulation-based method used a powerful process simulator to simulate an absorption-based dehydration unit as a case study for an existing gas plant. The studied dehydration unit uses triethylene glycol (TEG) as a solvent.

A sensitivity analysis of the unit independent variables' impacts on the PI approach is done. The method defines three scenarios to effectively intensify the process in the core unit (TEG), (1) the TEG circulation, (2) stripping gas flow rate, and (3) regeneration reboiler temperature. The defined scenarios can reduce the REC by 11%-18%, BTEX emissions up to 69%, $CO_{2\text{-(process)}}$ up to 37%, and TEG loss reduction of about 35%, without compromising the product specification. Due to no CAPEX impacts, these scenarios are valid for both, future design, and current TEG units in operation.

Furthermore, the proposed systematic method was also implemented for the upstream and downstream adjacent units. The output indicated the potential PI in terms of REC that could be achieved over the entire process. Moreover, the same method can be used for approaching the PI in any other process by incorporating the specific independent variables of the studied process.

KEYWORDS:

Process Intensification (PI), Oil and Gas (O&G), Natural Gas (NG), Dehydration, systematic simulation-based method, Absorption, Triethylene Glycol (TEG), Rated Energy Consumption (REC), Water Content, BTEX.

ABSTRAKT

Vysoušení zemního plynu, anglicky *natural gas dehydration*, je zásadní součástí procesu zpracování vytěženého zemního plynu. Klíčová úloha vysoušení spočívá především v podpoře množství a kvality produktu, ale také v prevenci tvorby hydrátů, koroze či nežádoucí vodní kontaminace při transportu plynu.

Tento petrochemický proces má všechny předpoklady pro aplikaci tzv. *intenzifikace procesů* (PI), moderního přístupu z oblasti procesního inženýrství. Jeho intenzifikace je žádoucí z pohledu investičních a provozních nákladů (CAPEX and OPEX), spotřeby energie i produkce nežádoucích emisí. Současné globální požadavky v oblasti ochrany životního prostředí i ochrany zdraví osob umocňují význam těchto snah. V odborné literatuře v posledních letech významně roste zájem o intenzifikovaná zařízení (*PI equipment*), méně však už o rozvoj metod intenzifikace procesů a jejich aplikaci na komplexnější procesy. Překvapivě málo se také při výzkumu PI využívá počítačové podpory ve formě simulace procesů.

Předložená práce představuje **novou systematickou metodu pro intenzifikaci procesu vysoušení zemního plynu** založenou na počítačové simulaci, která využívá dva hodnotící parametry: jmenovitou spotřebu energie (Rated energy consumption, REC) a předepsaný obsah vody v produkovaném zemním plynu. Spotřeba energie je tvořena přímou spotřebou samotného procesu a entalpickými ztrátami přes hranici procesu. Snižování jmenovité spotřeby energie je klíčovou motivací intenzifikace a současně prostředkem k jejímu hodnocení. Předepsaný obsah vody je klíčovým projekčním a provozním parametrem procesu. Navržená metoda vychází z interakce těchto dvou parametrů.

Práce staví na zodpovědném rozlišování mezi intenzifikací konkrétní procesní jednotky a intenzifikací komplexnějšího procesu, který zahrnuje více dílčích jednotek. Využívá přitom efektivní rámec pro implementaci PI v petrochemickém průmyslu, který byl definován autorem práce.

Navržená metoda byla aplikována na existující průmyslový provoz zpracovávající vytěžený zemní plyn. Pro analýzu nejčastěji využívané technologie sušení, tzv. triethylenglykolové absorpce (TEG), byl použit výkonný software pro simulaci procesů. Byla provedena citlivostní analýza jmenovité spotřeby energie (Rated energy consumption, REC) této jednotky na změnu tří vybraných proměnných – průtoku triethylenglykolu, teploty v koloně a průtoku stripovacího plynu. Ukázalo se, že tyto proměnné mají obrovský potenciál pro intenzifikaci TEG jednotky. Jejich cílenou změnou je možné snížit jmenovitou spotřebu energie o 11 až 18 %, BTEX emise až o 69 %, emise CO_2 až o 37 % a ztráty triethylenglykolu až o 35%, přičemž není negativně ovlivněna předepsaná kvalita produktu. Metoda předpokládá postupné provozní zásahy do procesu a nevyužívá technologických změn. Navržené úpravy jsou proto velmi dobře využitelné při optimalizaci provozu stávajících jednotek glykolové absorpce i návrhu nových.

Navržená systematická metoda byla použita i na další jednotky, které předchází a navazují na glykolovou absorpci. Výsledky potvrzují výše uvedený přínos a významný potenciál PI při snižování spotřeby energie, kterého může být dosaženo v rámci celého procesu vysoušení zemního plynu.

KLÍČOVÁ SLOVA

Intenzifikace procesů, petrochemický průmysl, zemní plyn, vysoušení zemního plynu, absorpce, adsorpce, glykolová absorpce, molekulární síta, jmenovitá spotřeba energie, obsah vody, BTEX.

BIBLIOGRAPHICAL CITATION OF WORK

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DECLARATION

I declare that I am the author of this doctoral thesis. It has been prepared on my own and under the guidance of my supervisor. The reported results are original research that was developed based on my knowledge gained during my Ph.D. study and consultation with experts. I have quoted all the sources including my publications. The used references are provided at the end of this thesis.

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Table of Contents

Chapter 1 Research objective and methodology

Oil and gas (O&G) continue to be the **dominant shareholder** of energy sources and raw materials for many applications, especially with the **steady increase** in the consumption of O&G to overcome the increase in demand. There is a need for powerful methodologies for process intensification (PI), which support achieving sustainable development of this sector in an efficient, profitable, and environmentally friendly manner.

The comprehensive literature review reveals

- a concern with the PI equipment, more than the PI methods and the concepts of the process itself, and
- limited utilization of the computer-aided simulation to serve the PI research.

This work aims to fill the gap and develop a clear systematic simulated-based method for process intensification in O&G that meets

- the PI drivers and process specifications, and
- easy and efficient utilization, that serves the **future design**, as well as the **current facilities** in operation.

To achieve the desired target, this work combines

- a method that recognizes unit intensification and extended process intensification, and
- a concise framework for PI implementation in the O&G proposed originally by the author. The used framework adopted the rated energy consumption (REC) as a key driver and evaluation tool of the intensified solution.

An overview of the objectives of the dissertation and applied methods is presented in Table 1.

Table 1 Overview of objectives and methods of implementation

Chapter	Partial target	Applied methodology
3	Investigating the current status of the process intensification in the petrochemical industries	• Establish a basic knowledge and understanding of fundamental concepts, which secure proper analyzes of the studied topic • Analysis of the statistics data about PI implementation and PI drivers in the petrochemical industries
4	Problem definition and goal setting	• Literature review of PI in the O&G • State of the art of PI in natural gas dehydration • Identifying research gaps • Goals setting • Solution approach
5	Theoretical base about natural gas dehydration	• Introduction to the natural gas processing and dehydration role • Definition of the fundamental concepts of dehydration and product specification • Review of the dehydration methods
6	Process intensification method	• Solution approach • Used tools (powerful process simulator)
7	Current state modeling	• Case study definition via the provided documentation and operator's available data • Establishing the simulation model of the facility using • Extract the current state results for benchmarking
7	Intensified solution modeling	• Simulation of the intensified solutions at the level of the unit, and the entire process • Extracting the results and graphical representation
7 & 8	Results analysis and evaluation	• Comparison between the intensified solution and the current state • Benefits evaluation • Conclusions

Chapter 2 Introduction

Natural gas dehydration (NGD) is essential in avoiding hydraulic problems such as slugs, hydrate formation prevention, electrochemical corrosion control, storage, transportation, and fulfillment of product specifications. Its role is integrated with the upstream phase separation process, and crucial for the performance of the downstream processes such as natural gas liquids (NGL$_s$) recovery. When the phase separation isolates the water in the liquid phase of the stream, the water vapor flows with the gas phase, and dehydration is the process responsible to catch this water. This guarantees a dew point less than the minimum operating temperature in the downstream units, especially those which work at a very low operating temperature [1]. The NG processing basic sequence is shown in Figure 1.
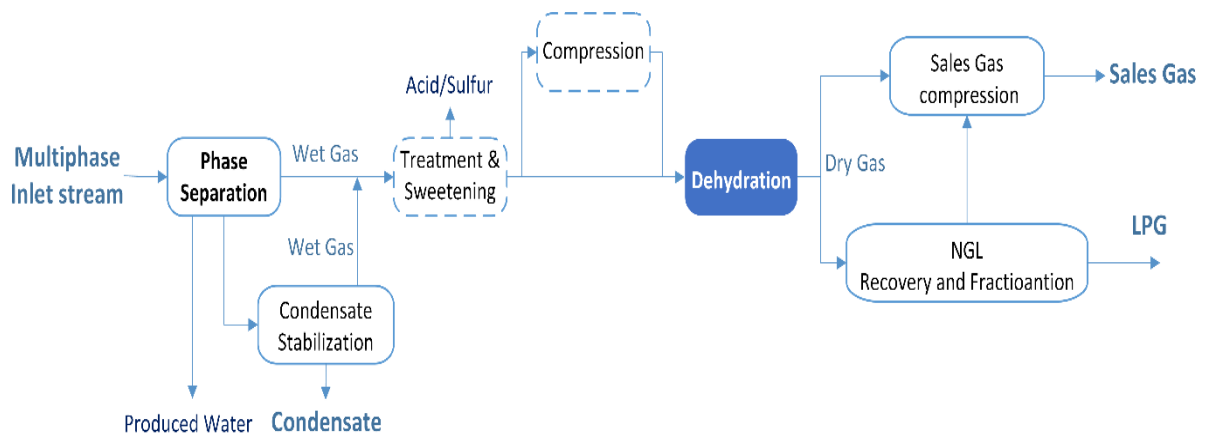
Figure 1 Basic sequence of natural gas processing

Due to its high capital and operational costs, energy consumption, greenhouse gas emissions, and some hazardous pollutants, the NGD is one of the most preferred processes for modern process engineering concepts such as **process intensification** (PI).

The O&G industry is one of the most traditional industries. It used to face multi obstacles that resist efficient development, resulting from old profitability concepts. During the last three decades, energy investment started to change driven by the modern concepts of economy, strategic management of energy sources, product quality considerations, pressing global requests regarding environmental aspects, and supportive directives. Accordingly, there was a higher interest by the stockholders in the implementation of modern process engineering concepts such as process optimization, process integration, and process intensification PI in all industrial sectors, including the O&G. These modern concepts provide useful and powerful solutions for achieving sustainable development of this industry in an efficient, profitable, and environmentally friendly manner. Moreover, they are applicable for the new projects, as well as for the development of existing facilities.

Although the PI implementation in the O&G is still limited with few technologies and inconsistent with the increasing growth of research interest after the Kyoto Protocol in 1997,

3

there were some efforts and attempts to set a road map and specific methodologies for PI implementation enhancement. Due to oil future concerns and clean energy considerations, the increase in demand on the O&G was primarily dependent on the natural gas NG since the 1980s.

The PI represents a promising solution that could serve this industry, especially since the PI drivers exist throughout the supply chain of the NG industry, from production wellheads to end consumers. This is valid for both types of NG processing facilities, the associated NG units, and the non-associated NG treatment and processing plants. The high capital and operational expenses (CAPEX and OPEX) of the NG industry are not limited to the processing facilities, as it is in the oil refining. The pipeline, transportation, and storage facilities are all of very high CAPEX and OPEX in comparison to the oil refining production and non-process facilities.

All these reasons motivated the companies to invest in the PI of the NG facilities, in addition, there were some new technologies which are developed particularly for the previously abandoned gas fields. The PI provided promising solutions for investing even in those areas with low economic paybacks.

The NGL fractionation and NGD are classified as potential candidates for any PI approach that could be applied in the NG industry. The PI in the fractionation units was studied widely in the oil refining and the results could be generalized to the NG processing units. On the other hand, further efforts have to be done concerning the PI in the NGD due to its crucial role in this industry. The development of an effective methodology for PI implementation in the NGD requires a clear understanding of the fundamentals of both PI and NG processing.

Hereinafter the main concepts of PI are summarized, and its implementation of it in the O&G generally and NGD particularly is reviewed. The NGD process itself is discussed in detail as well. This establishes a firm theoretical background for the proposed methodology.

Chapter 3 Process intensification and its drivers in the O&G

Achieving development in the traditional industries becomes an insisting need due to economic and technical reasons. The global requests to consider the energy, health, and environmental aspects of any development apply more pressure to manage this development sustainably. Oil and Gas (O&G) is one of the most preferred sectors to be considered for such sustainable development, as it has all drivers to be developed via the modern process engineering concepts, such as process optimization, process integration, and process intensification (PI), which can significantly lead this development. Hereinafter, a basic theoretical background about PI and its main drivers in the O&G are presented.

3.1 Process intensification fundamental concepts

The first use of the term "Process Intensification" (PI) in a publication is dated back to 1957 by Lebedev [2] for the intensification of the iron ore sintering process. In 1959, Planovsky [3] presented his first paper about the PI, when discussed the improvement of the flow reactor efficiency in petrochemical processes. In the next four decades, the publication about PI approached the chemical and petrochemical sector generally without concern about the O&G in particular. The term PI started to be used in the practice in the 1970s by chemical engineers [4], and the interest started arising. The first definition of PI was presented in the works of Ramshaw and Cross in the 1980s [5] [6], and it was limited to the size and cost of facilities. In 1995, the first international conference about PI in the chemical industry was held [7], followed by the works of Stankiewicz and Mouljin [7], Tsouris and Porcelli [8], the concepts of PI expanded to consider the energy, and emissions. The works of Reay and Baldea [9] [10], defined safety as an essential concept to be considered as well. The PI definition has evolved till it got to the most comprehensive one. Table 2 presents the development of the **PI definition**.

Table 2 Evolution of process intensification definition

Reference	Definition
Ramshaw,1983	"*is the devising exceedingly compact plant which reduces both the main plant item and the Installations costs*" [5]
Stankiewicz & Moulijn, 2000	"*Any chemical engineering development that leads to a substantially smaller, cleaner, and more energy-efficient technology.*" [7]
Tsouris & Porcelli, 2003	"*Technologies that replace large, expensive, energy-intensive equipment or process with ones that are smaller, less costly, more efficient, or that combine multiple operations into fewer devices (or a single apparatus).*" [8]
Reay, 2013 Baldea, 2015	"*Any chemical engineering development that leads to substantially smaller, cleaner, safer, and more end-use technology or that combine(s) multiple operations into fewer devices (or a single apparatus).*" [9] [10]

The continuous evolvement of the PI definition highlighted the need for a fundamental foundation. Van Gereven and Stankiewicz [11] proposed four **fundamental principles**, which

were updated by Arizmendi-Sanchez and Sharratt [12], and Lutze [13]. These fundamental principles could be summarized in

- Principle 1: Maximize the effectiveness of intra- and intermolecular events

- Principle 2: Give each molecule the same processing experience

- Principle 3: Optimize the driving forces at every scale and maximize the specific surface area to which these forces apply

- Principle 4: Maximize the synergistic effects

In addition to these four principles, the **domains** of PI were also defined. Spatial (structure), thermodynamic (energy), functional (synergy), and temporal (time) are the four domains for realization [14].

The Ideal PI requires realizing the four fundamental principles in one or more of the realization domains. The realization **tools** are classified into [7] [4]

- intensified equipment, such as novel apparatuses and devices, which represent the hardware structures of PI, and

- intensified technologies which represent the software part of it. Such techniques include incorporating chemical reactions into unit operations, combining two or more different processes, and using various types of energy.

In addition to the realization tools, there was a need to present the PI as a systematic method rather than a toolbox with specific equipment or technologies. Process systems engineering (PSE) for modeling and powerful systematic methods are used for the identification and synthesis of the PI processes [15], [16]. Ponce-Ortega [17] approached the PI by recognizing unit intensification and plant intensification.

Synthesis of intensified processes is classified into [18]

- knowledge or experienced-based methods (heuristic), where the PI design presented by Siirola according to task identification and integration [19] [20] [21],

- optimization-based methods, where a generated superstructure is translated into a problem with an objective function and operational constraints. The flowsheet structure and operating conditions would give then an improved design, as per the works of Grossmann, I.E. et al [22] [23] [24], and

- hybrid methods which combine the previous two methods to reduce the computational complexity of the studied design [25], [26].

Within the context of PI and synthesis methods, very helpful papers were presented by Barnicki [27], Siirola J.J. [28], Grossmann, et al [29], and Babi et al [30],

The **evaluation** of the realized PI is an essential step to be made according to specific criteria. The target of the implemented PI, completeness or ripeness of this application, and the quality and safety of the output, are fundamental in the evaluation procedure. The 2007 EU road map for PI [31], provided a good example of investigating and evaluating 72 PI technologies that cover all chemical sector industries.

Based on the presented background a **hierarchy of the PI** could be structured as per Figure 2.
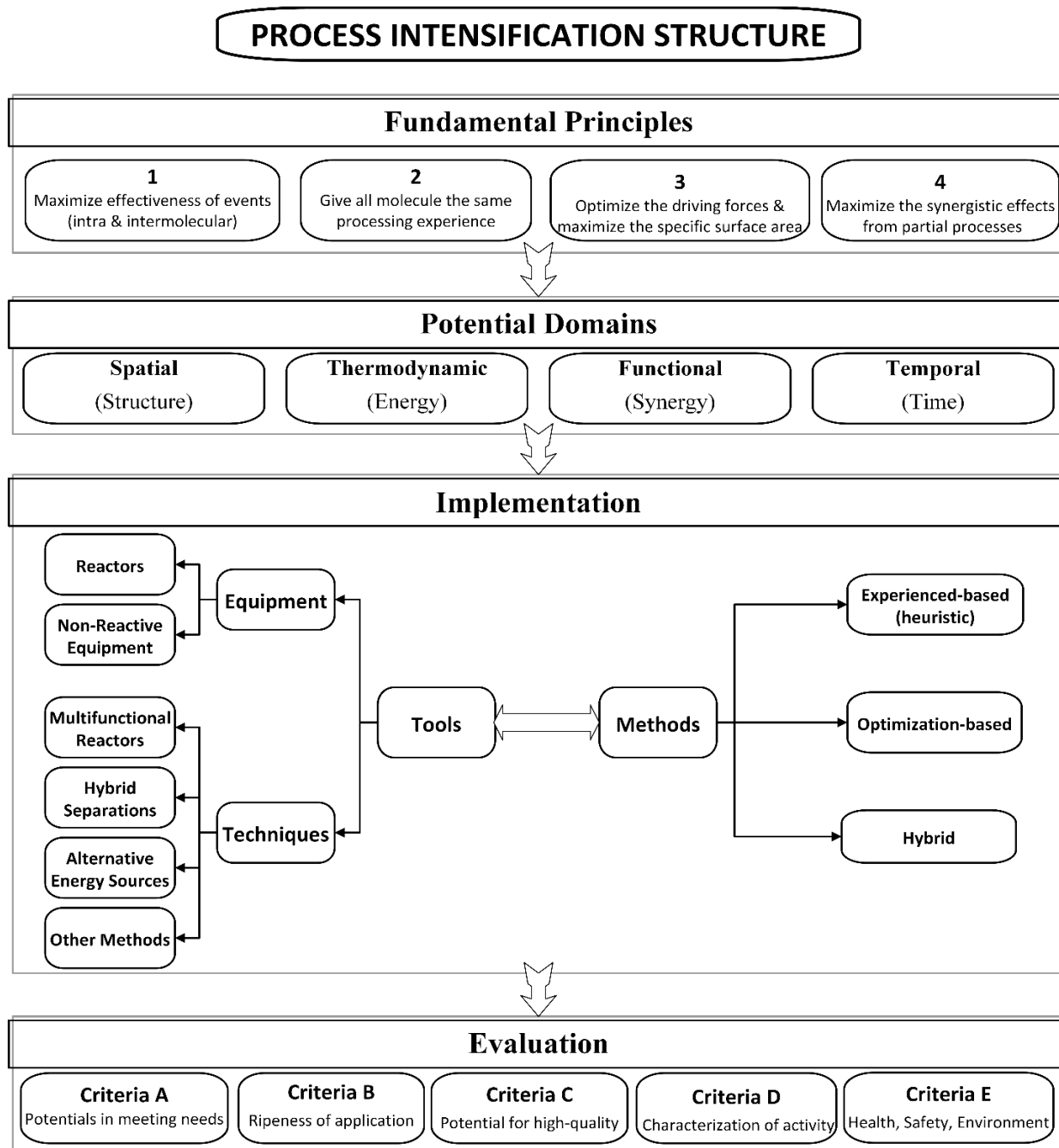

PROCESS INTENSIFICATION STRUCTURE

Fundamental Principles

1	2	3	4
Maximize effectiveness of events (intra & intermolecular)	Give all molecule the same processing experience	Optimize the driving forces & maximize the specific surface area	Maximize the synergistic effects from partial processes

Potential Domains

Spatial (Structure)	Thermodynamic (Energy)	Functional (Synergy)	Temporal (Time)

Implementation

Reactors — Non-Reactive Equipment — Equipment — Tools — Methods — Experienced-based (heuristic)

Multifunctional Reactors — Hybrid Separations — Alternative Energy Sources — Other Methods — Techniques — Optimization-based — Hybrid

Evaluation

Criteria A	Criteria B	Criteria C	Criteria D	Criteria E
Potentials in meeting needs	Ripeness of application	Potential for high-quality	Characterization of activity	Health, Safety, Environment

Figure 2 Process Intensification Structure (Extended concept from Van Gerven and Stankiewicz [9], Stankiewicz [6], European PI roadmap [36])

3.2 PI implementation drivers in the O&G

The PI definition by Stankiewicz [7] provides the best entrance to clarify the implementation drivers of PI technologies in the petrochemical industries: **energy consumption, GHG, and cost**. This subsection brings a new deeper insight into these drivers.

7

A) **Energy Consumption**

The industrial sector shares 32 % of total global energy consumption (Figure 3) [32] [33] and holds the position of the dominant energy consumer with estimations of steady growth. This is accompanied by an increase in the consumption of the O&G due to the increasing demand for energy generally, and in the industrial sector particularly. The decrease in using some other traditional sources such as coal, and the inability of the renewable sources to overcome this increase in the meantime increased the demand for the O&G, Figure 3[34]. Detailed statistics and data could be found in the work of Abdulrahman et al [35]. The projections indicate an increase in energy use of about 50 % for the next three decades. The consumption of O&G as the main feedstock for the energy market would increase consequentially. Oil would share more than 20 % of this increase, while it would be more than 40 % for natural gas [32].

ENERGY SOURCES FOR INDUSTRIAL SECTOR IN THE US
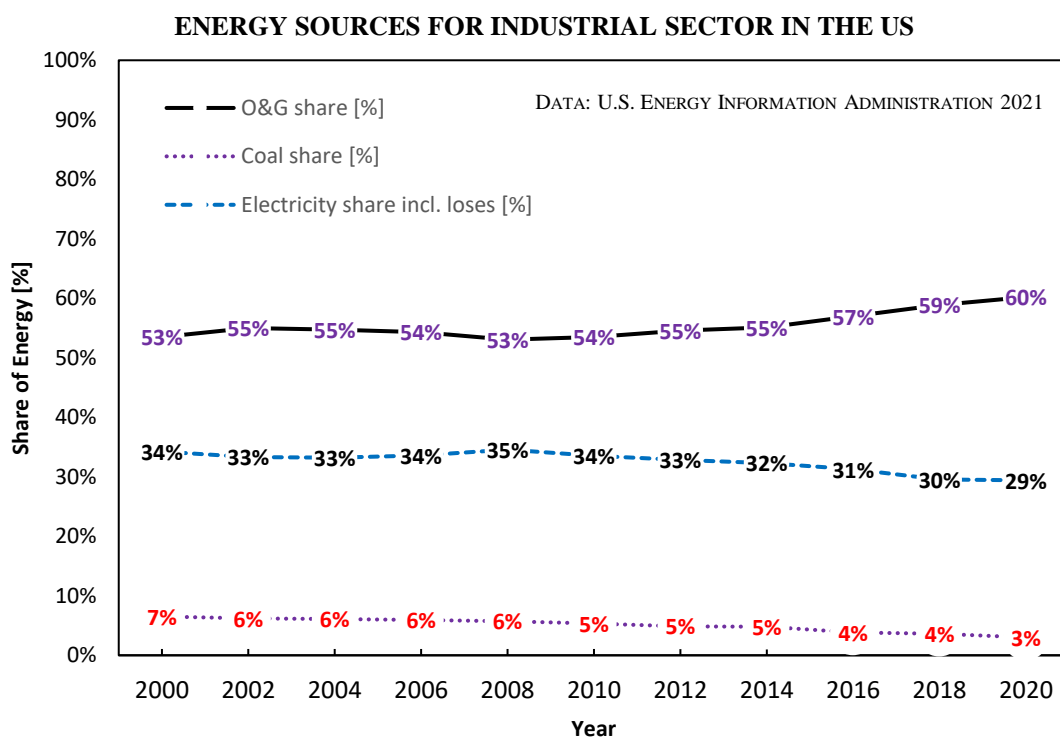
Figure 3 Energy Sources for Industrial Sector in the US [34]

While O&G is the main energy feedstock, the sector of O&G and petrochemical industries is the major consumer among all industries (30 % of industrial sector energy consumption, about 10 % of the global energy demand) [4]. Energy is needed all over the supply chain. The International Energy Agency (IEA) [36] identified oil refining as the main energy consumer. The oil itself is the main energy source for oil refining, while the natural gas is being consumed throughout the supply chain from production to distribution Figure 4.
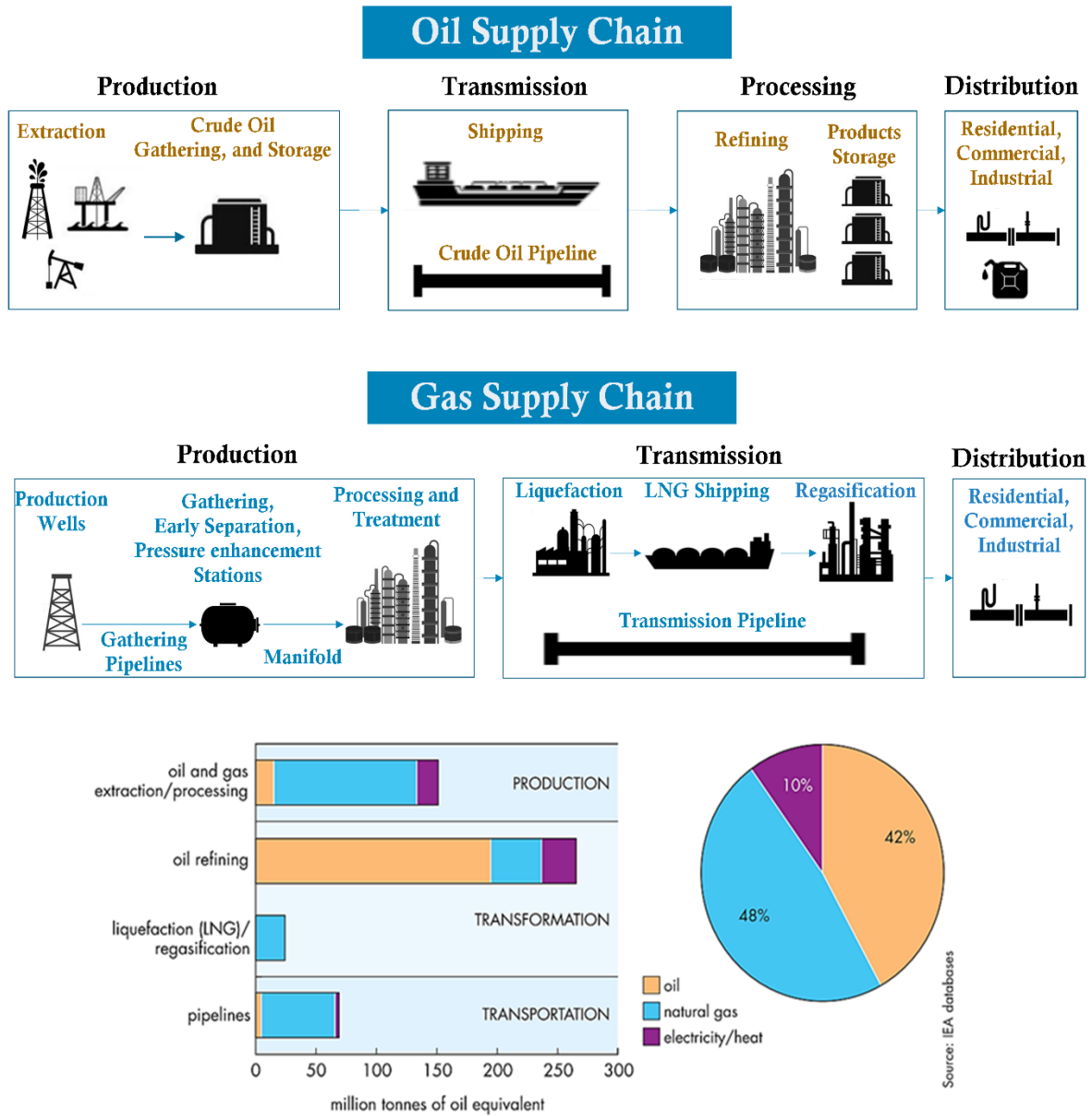
Figure 4 World energy consumption along the O&G supply chain (Data source: IEA [36]**)**

The available data from the US petroleum refining sector provides detailed data in this regard [37]. The process portion of the total energy consumption in the onsite and offsite facilities is 78.4 % of the total consumption, 23 % of this consumption is a loss. In other words, 18.1 % of the total consumption of energy in the facilities is just a loss by the process itself, 641 [TBtu] about 676×10^{15} [J], Figure 5.

TBtu = Trillion Thermal British Unit

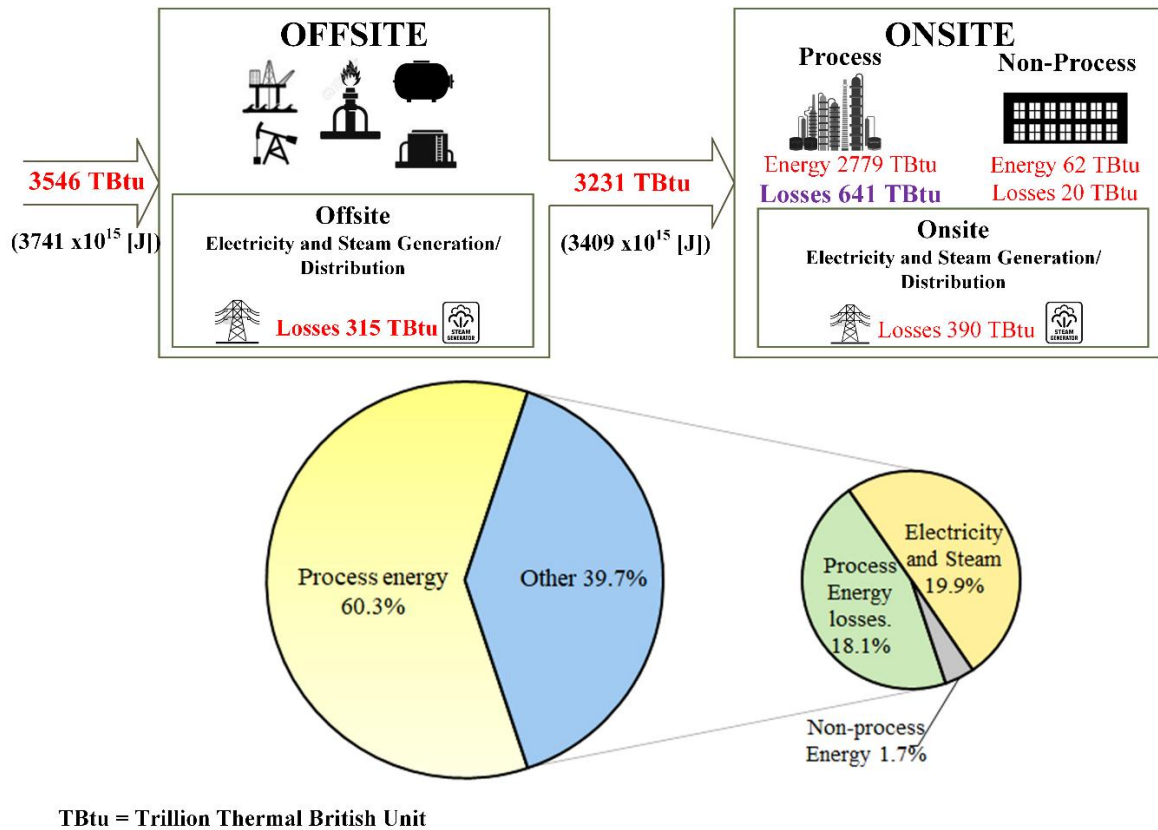**Figure 5 Total energy for the US petroleum refining sector** [35] (**Data Source: MECS** [37])

B) **Pollution and greenhouse gas emissions** (**GHG**)

Chemical and petrochemical industries (Petroleum refining, particularly) are the main producers of GHG emissions with a share of 7 % of the global CO_2 total emissions [4], [37]. Excluding the electricity production, IEA [38] and the Intergovernmental Panel on Climate Change IPCC [39] classified this sector as the 3rd in producing the CO_2 with a share of 18 % of the global emissions.

The range of pollutants produced by the O&G is much wider and more diverse than the GHG. It includes several other pollutants that cause immediate and long-impacts on health, environment, and safety, such as CO, CH_4, SO_x, NO_x, NH_3, Pb, Hg, acids, sulfides, and solid wastes [40]. Accordingly, the safety driver of the PI becomes essential.

The emissions of CO_2 are of two categories, combustion products, and process released CO_2 by venting and flaring. Combustion is the main source of CO_2 emissions in refineries with 63 % of total CO_2 emissions [41], [42]. The scenario is very similar in the natural gas (NG) industry, where the infrastructure produces 20 % of the entire global emissions of this industry [43]. In addition to CO_2, emissions include substantial quantities of CH_4, which is released by venting and fugitive leaks from equipment throughout the supply chain, from gas wells to end-users. Moreover, the released pollutants and GHG emissions from process venting and flaring are of the same considerable quantities and harmful effects, Table 3.

Table 3 Global emissions from midstream and upstream natural gas infrastructure in 2012 [43]

Industry	CO_2		CH_4	N_2O
Natural gas (midstream and upstream)		164 MMt CO_2e (combustion, flaring and removal of raw gas stream)	155 MMt CO_2e	Not estimated

C) Capital cost

To achieve cost savings in the O&G facilities, it was used to apply the two-thirds power rule to estimate the capital cost changes at various production scales [3]. Later on, and due to high error rates, capital cost statistics provided more realistic data. Anyway, increasing the facility size to save the cost is limited by technological restrictions, production, operational costs, marketing, and stability of the market [44] [4]. Nowadays, small equipment that fulfills the operational restrictions, and achieves the minimum operational costs are a need, especially for offshore and abandoned field applications.

3.3 Rated energy consumption as a key driver of PI implementation

The development of O&G as an applied field requires well-planned and applicable approaches. In addition to the research base, clear methodologies, and tools, the implementation has to have a well-defined technical framework and procedures. The measurability and feedback evaluation are essential characteristics that have to be considered. Referring to the fore clarified PI hierarchy, there is a need for defining such a technical framework. This topic was presented by the author via a specific publication in this regard "*Process intensification in the oil and gas industry: A technological framework*" [35]. The author presented a clear map for the implementation of the PI tools in the oil and gas facilities, and compared the implementation drivers, with the statistics, which led to a clear conclusion about the importance of linking the production capacity with the other drivers of the PI. Accordingly, the author suggested a concise PI framework that depends on the rated energy consumption (REC) as a key driver of the PI implementation. The physical meaning of this characteristic makes it easily meet the basic concepts of the PI, relation to the production capacity, and the measurability for evaluation needs. Such methodology is supposed to enhance the PI implementation due to its easy application by researchers, designers, and operators.

The O&G companies nowadays tend to have smaller, more efficient, synergetic, and productive facilities. These characteristics correspond to the "spatial", "thermodynamic", "functional", and "temporal" domains, respectively Figure 2. In literature, the PI implementation in terms of domains was discussed in several papers

- Law [45], Jensen [46], and Stankiewicz [47] discussed the spatial domain,

- Reay [48], Chen [49], and Khoo [40] investigated the adjective "cleaner"

- Stankiewicz [50], Thakur [51], and Reay [44] used the adjective "smaller and more efficient"

- Bernardo [52], and Kilkovsky [53] used the adjective "integrated"

Analyzing the relation between these adjectives and the PI drivers indicates a mutual and interactive relation between them. The "bigger" means more capital expenses (cost) (CAPEX) and productivity, more operating expenses (OPEX), and both means higher energy consumption. Moreover, higher energy consumption means more emissions.

The energy takes a role in between all these adjectives, but it is not able to represent all the drivers together without some adoption of the physical term. The specific energy consumption (SEC) represents the amount of energy that is needed for the unit of the product, and it is proportional to the emissions and OPEX. The CAPEX is related to the size of the equipment or facility, and together with the OPEX, has a mutual proportional relation with the processing capacity (PRC) (production per time interval). Merging the PRC with the SEC will result in the rated energy (REC) Figure 6, as per:

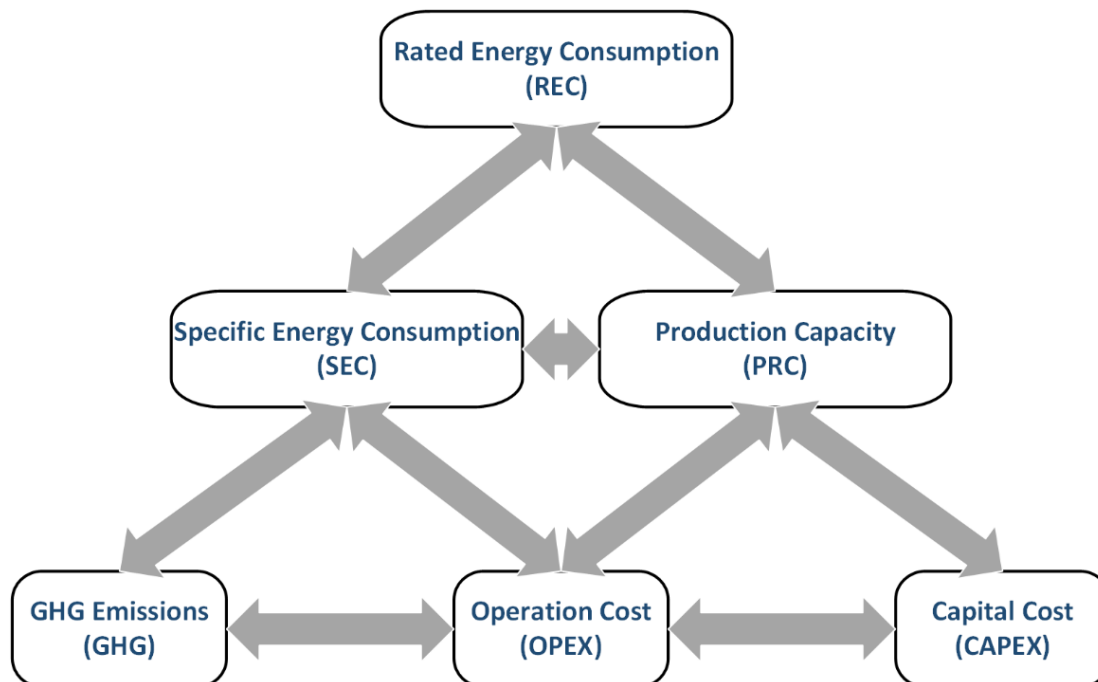Rated energy consumption (REC) = (Specific energy consumption SEC) X (the production capacity PRC)

Figure 6 PI drivers' participatory relations [35]

Within the scope of this work, the rated energy consumption is considered a key driver for PI implementation.

Chapter 4 Literature review and state of the art of process intensification in O&G and gas processing

The research interest in the PI implementation in O&G has started to show some rising only after signing the Kyoto Protocol in 1997, and more clearly after it entered into force in 2005. Synchronously, the higher interest was driven by the warmest decade ever recorded, 2000-2009. Hereinafter a literature review about the PI implementation in the O&G is provided.

After presenting the fundamental concepts of the PI, it is necessary to review the implementation in the O&G generally and NGD particularly, and figure out the gaps that have to be considered during the establishment of the proposed methodology for PI enhancement in this industry.

4.1 Literature review of process intensification in the O&G

Searching the published articles about PI in the chemical, energy, and O&G sectors, since 1994 reveals the results shown in Figure 7. (Scopus database was used)
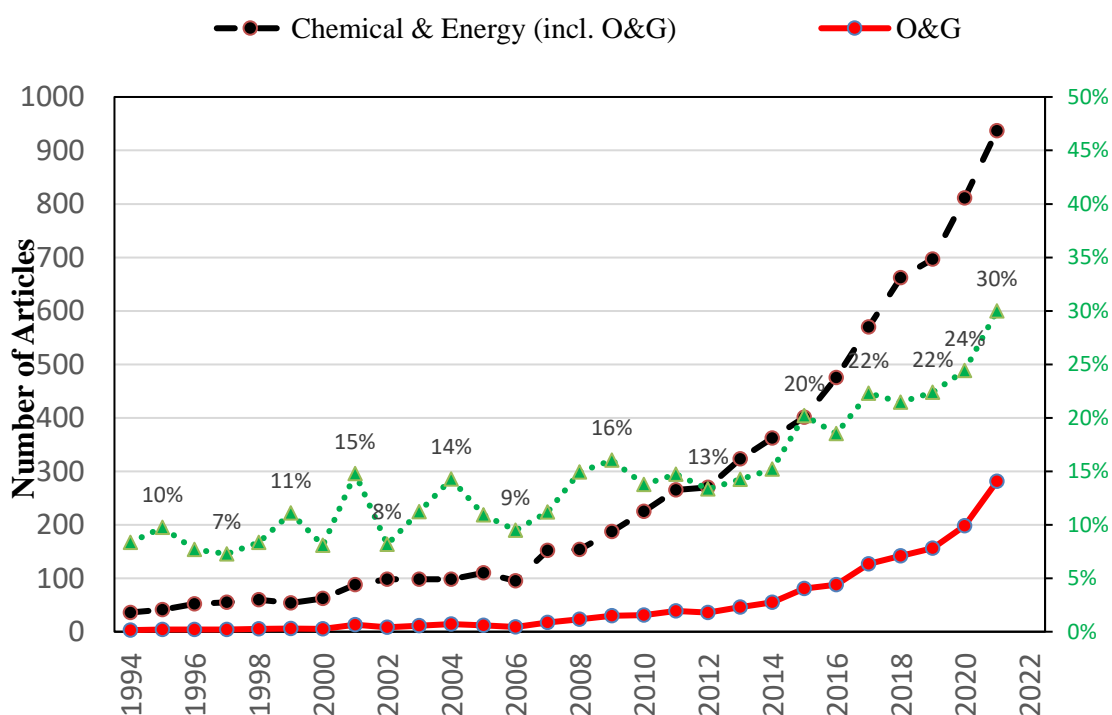
Figure 7 Number of articles about Process Intensification in the Oil and Gas (O&G) field, and in the Chemical and Energy Sectors generally (including O&G) (Data source: Scopus platform)

Analyzing the obtained results indicates

- higher research interest in PI in the O&G field since 2008, and the publications about natural gas represent 63% of the publications about PI in the O&G. O&G publications hold only 30 % of the total number of PI publications in the chemical and energy sectors.

13

- PI implementation could be classified according to the process into separation, reaction, and combined reaction-separation processes. This replaces the traditional classification of PI tools which was presented by Stankiewicz and Moulijn [7]. This was firstly done by Tian [16] and reproduced in this work to fit the O&G applications.
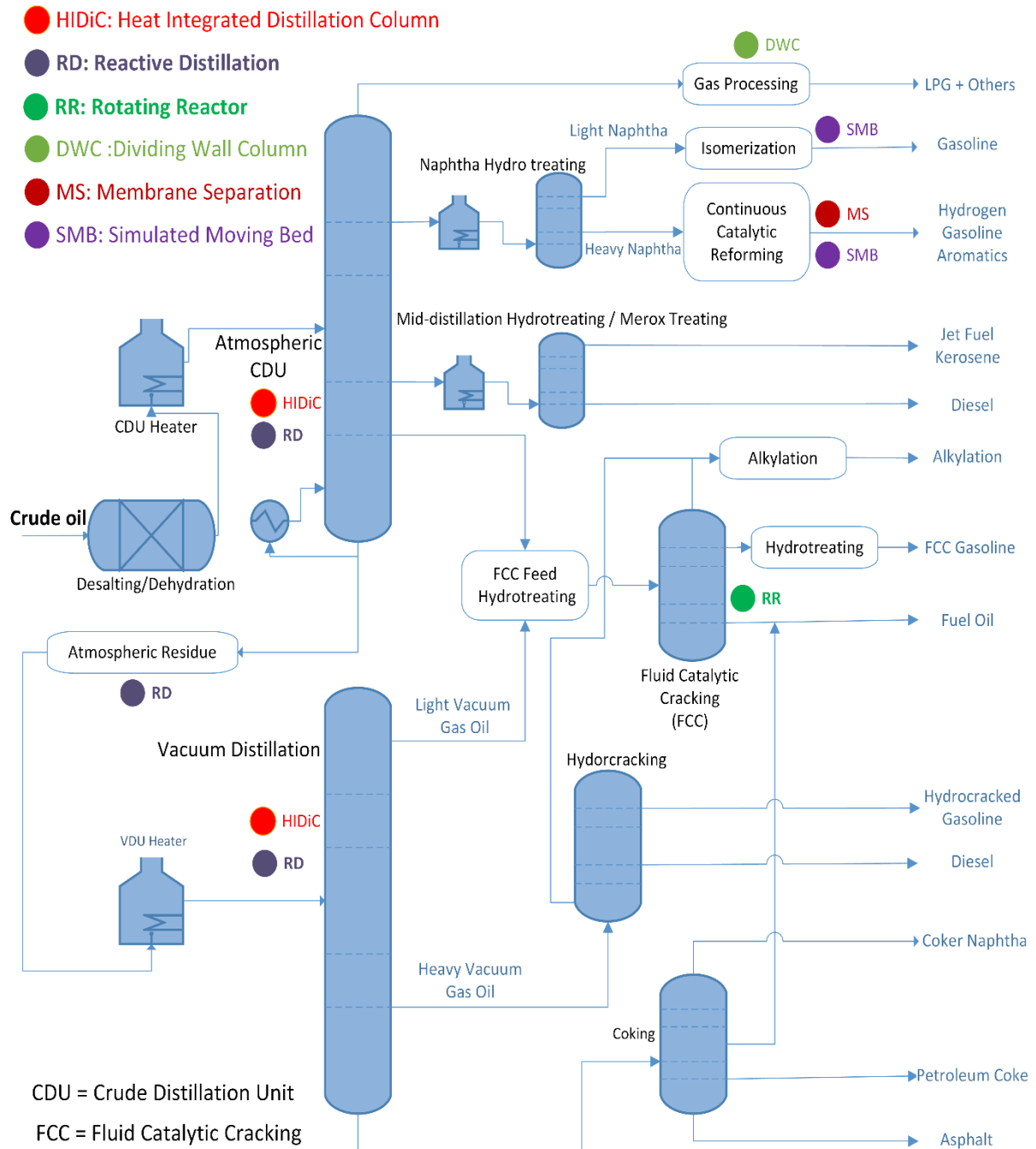
Table 4 provides a review of the category and types of the PI technologies which are applied in oil refining, and Figure 8 represents these technologies' implementation within the refinery.

Table 4 Process intensification applications in the oil refining based on the literature review [35]

Tech./ category	Implementation unit/start/rate	Savings (Capital cost; Energy)	Potentials	Constraints	Ref.
RD/RS	Started in 1951 >150	20-80%; 20-80%	Capital cost reduction via merging units. Energy savings via using the exothermic reaction heat in evaporating the liquid phase. Prevention of hot spots in catalysts due to the limited increase in temperature (boiling point). Preventing the poisoning of catalysts with impurities by locating them above the feed. Higher selectivity for the intermediate products and fewer consecutive reactions via fast removal of reactants.	Temperature reaction should be within the vapor-liquid equilibrium. Catalyst is unstable at high temperatures. Catalyst should meet long lifetime requirements due to the high cost of replacement.	[4] [16] [54] [55] [56]
DWCD/S	CDU, Started in 1949 >125	10-30%; 10-30 %	Energy-saving. Higher thermodynamic efficiency due to reduced mixing. Reduced capital cost due to the smaller size/ less number of units. Reduced operating costs. Reduced maintenance. Higher purity for all product streams. Reduced GHG emissions.	Larger equipment size (less number of equipment compared with older designs, i.e. Petlyuk column, but with larger equipment size) Lack of design validation regarding operability and controllability. Technical problems with the formation of low boiling components in the sump or high boiling components at the top affect the purity of side-draw.	[4] [16] [54] [57] [58] [59] [60]
HIDIC/ RS	Started 1951	Up to 50% in energy	Energy savings	Not found	[58]

			Zero external reflux and boil-up operation Enhanced potential of internal heat integration techniques		
RFR/R	>100	30-40%; 30-90%	Less number of equipment due to the symmetric double active design of reactor (same reaction in both directions) (exothermic) and asymmetric (reaction changes with direction) (endothermic-exothermic) Efficient heat regeneration. Process stability under parameter fluctuation. Less energy consumption. Fewer capital costs	The consequent loss of equilibrium conversion in the second half of the process cycle. Sharp increase of reaction rate with temperature which means a negative influence on the equilibrium conversion.	[54] [61] [62] [63]
MR/R	Started in 1969 Not implemented commercially	Not found	Less investment and operating costs. Higher safety level due to less residence time of the material. High potential to serve as reactors with multi-purpose functionalities, which is more economic. Improved performance and enhanced efficiency.	The incapability of handling solids. The linear increase in capital cost with the size of the reactor (not competitive for large scale applications in comparison with conventional reactors)	[64] [65]
SMB/S	Started in 1961	Not found	Less number of the unit due to continuous operation, (no regeneration mode)	Not found	[16] [66]
SMR/R	>30	Not found	Energy savings Higher yield and conversion. Higher product quality Compact size and reduced cost.	Difficulties in controlling units due to periodic operation and spatially distribution. Difficulties in scaling up.	[4] [67]
MRS/RS	Started 1983	Not found	High stability Energy savings Reduced cost compared with others (i.e. PSA)	Not found	[68] [58]
Abbreviations are listed in Nomenclature					

It can be noticed that the most implemented technologies in petroleum refining are reactive distillation (RD), dividing wall column (DWC), and reverse flow reactor (RFR) [54].

Figure 8 PI implementation in oil refining [35]

Table 5 provides a review of the category and types of the PI technologies which are applied in the NG processing, and Figure 9 represents these technologies' implementation within the gas processing plant.

It can be noticed that pressure swing adsorption (PSA) is the most hit topic for PI in Natural gas processing [69] [70].

Table 5 Process intensification applications in natural gas processing based on the literature review [35]

Tech./ category	Implementation unit/start/rate	Savings (Capital cost; Energy)	Potentials	Constraints	Ref.
PSA/S	Started in the 1960s	Low	High separation efficiency High availability >99% Easy and flexible operation (range 25%-100%) High recovery rate (up to 99%) Fully automated	High energy consumption for higher recovery in regeneration stroke	[69] [71]
TSA/S	used in gas dehydration, but not widely	Not found	High separation efficiency Easy regeneration	Difficulty in the regeneration of the adsorbent material with low thermal conductivity.	[72]
Inline/S	Used in remote fields	High; 50-80%	High separation efficiency Wide operational range Moderate to high Slug handling capacity. Reduced space. Subsea technology Saving in dehydration system weight up to 50%	Limited capacity, as this technology is not able to deal with big streams as normal separators.	[70] [73]
DWC/S	LPG recovery units	10-30%; 10-30 %	Energy-saving. Higher thermodynamic efficiency due to reduced mixing Reduced cost due to less space/ number of units. Reduced operating costs due to energy savings. Reduced maintenance cost. Higher purity for products Less GHG emissions.	Larger equipment size (less number of equipment compared with older designs, i.e. Petlyuk column, but the single equipment is larger) Lack of validated design methods considering operability and controllability.	[4] [16] [54] [57] [59] [60] [74]
MRS/RS	Started 1983	Not found	High stability Energy savings Reduced cost Beneficial in gas treatment to meet the pipeline specification.	Not found	[68] [75]
Abbreviations are listed in Nomenclature					

It can be noticed from both tables above that the more number of the used domains, the more the potential of the PI technology.
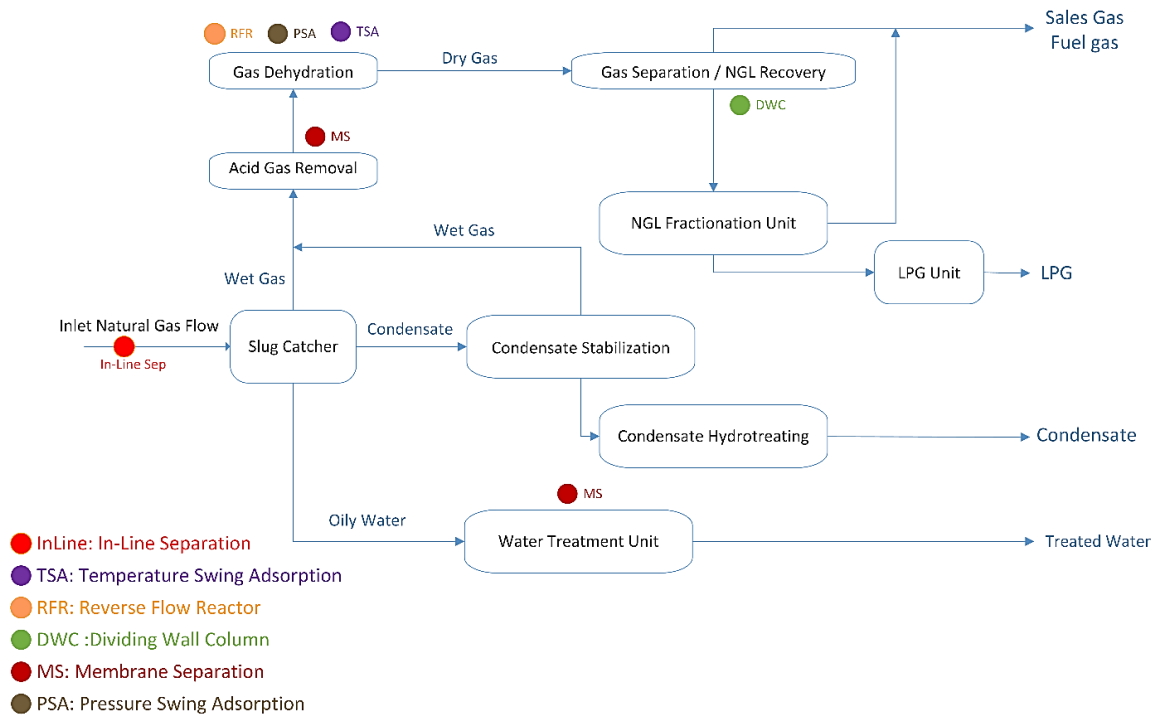
Figure 9 PI implementation in natural gas processing [35]

4.2 State of the art of process intensification in the natural gas processing and dehydration

The state of the art was studied based on the literature review of PI in the O&G, and specifically of NGD. It was realized that the implementation of PI equipment is more used than the PI systemized methods. Process Systems Engineering (PSE) is used for the modeling and simulation, and less frequently, systematic methods for the identification and synthesis of the PI processes.

Extending the above-presented review, and limiting the searching (on Scopus) to specific wording that gathers "natural gas" and "process intensification"

(TITLE-ABS-KEY("natural gas" AND "process intensification") AND (LIMIT-TO (DOCTYPE,"ar"))).

would reveal 67 articles. The majority of them were published after 2014. These 67 articles were reviewed, and the most relevant to NG processing are listed in Table 6.

The rest are more concerned about the technologies or natural gas-fueled applications, such as power generation, combustion, and hydrogen production. Several of these indirect publications discussed catalytic membrane technologies. It can be highlighted that

- most publications about the PI implementation in the NG are more concerned with the technology, **rather than the NG process itself**,

- the sweetening, NGL fractionation, and recovery process were investigated **more frequently than dehydration**, although of its high importance in terms of PI drivers,

- **very limited utilization of computer-aided tools** like CFD, or powerful process simulators such as Hysys or Promax in the PI of NG processing, and

- a very limited number of publications about **the systematic simulation-based PI methods** while the publications about the PI technologies, such as membranes and pressure swing adsorption (PSA) are dominant.

Table 6 Articles about process intensification in natural gas processing

Ref.	Topic	Method	Technique	Process	Simulation
Spoorthi, 2011 [76]	Upgrading synthetic landfill and lean natural gases by pressure swing adsorption PSA		X	Adsorption and separation of CO_2	X
Long, 2013 [77]	Heat integrated dividing wall columns for improved debutanizing and deisobutanizing		X	NGL fractionation	X
Chen, 2016 [78]	Dual-membrane natural gas pretreatment process		X	Hydrocarbons recovery	X
Duyen, 2016 [79]	PI of TEG by using sweeping gas membrane distillation		X	**Dehydration**	
Tavan, 2016 [80]	PI of amine-based gas sweetening process		X	Gas sweetening	X
Carrasco, 2017 [81]	Operability-based framework for process design and intensification of modular NG systems		X	Gas processing and power generation	X
Gong, 2017 [82]	A systematic simulation-based PI method and a novel condensation-based process design for shale gas processing and NGLs recovery	X		NGLs recovery	X
Iulianelli, 2017 [83]	Modified polymeric membrane technology for selective separation of CO_2			Gas sweetening	
Maqsood, 2017 [84]	Mixed sequential and integrated hybridization for cryogenic CO_2 capturing from NG	X		Gas sweetening	
Zhang, 2017 [85]	Hydrogen sulfide removal by catalytic oxidative absorption		X	Gas sweetening	
Ali, 2018 [86]	Optimization of cryogenic packed bed network for maximum product purity with minimum hydrocarbon losses		X	Gas sweetening	
Ghorbani, 2018 [87]	Structural, operational, and economic optimization of cryogenic natural gas plant	X		NG liquefaction and NGLs recovery	X
Miklas, 2021 [88]	PI in the TEG unit concerning rated energy consumption	X		**Dehydration**	X
Zhang, 2022 [89]	Phenomena-based approach for separation of sour gases	X		Gas sweetening	

The highlighted gaps indicate a clear need for approaching the PI in the NGD process via a systematic simulation-based method.

Hereinafter, the work is focused on the NGD particularly, and approaching the PI in this process via a systematic simulation-based method, that combines the **rated energy key driver** with a **powerful process simulator** to analyze and evaluate the potential PI and its fulfillment of the **process key specification**.

Before proceeding with the modeling and simulation of the studied case, a theoretical background about the NGD is presented.

Chapter 5 The natural gas dehydration process

Approaching the optimal design of the dehydration process requires a proper understanding of the multicomponent nature of the NG raw stream. The three-phase flow of NG, hydrocarbon condensate, and water is the typical mixture in the pipeline connecting the wellheads to the early gathering and processing facilities, the so-called trunkline. This trunkline and the entire field gathering network have the highest cost among pipelines within this industry, as they should withstand the stressful dynamics and operation of the multiphase flow, as well as the aggressive nature of this flow in the presence of water with acidic and sour impurities. Due to cost, integrity, operational, and efficiency issues removing this water and these acidic impurities are the top priority. This is done by phase separation, treatment, and dehydration.

The later stages of processing are to condition the stream to meet the desired final products. Cryogenic processes such as LNG recovery and fractionation impose extending the dehydration process scope to be deep dehydration. The basic sequence of NG processing is shown in Figure 1. Phase separation and dehydration are essential processes at the inlet of the NG processing facility, while the other two units, treatment, and compression, may present or not due to the composition and parameters of the raw natural gas. However, these four field modules should be as close as possible to the production field. **Water, sulfur, and acid gases have to be eliminated from the stream** before proceeding with processing, regardless if this processing of the NG would take a place at the same production location, or a remote one. Moreover, the gas properties downstream of these modules allow the implementation of piping of less specification than those the upstream, which is vital from the economics perspective, especially if the processing would be far from the production field.

Phase separation is responsible to isolate the liquids of the mixture stream, but the NGD objective is the elimination of the associated water vapor from the separated gas phase. This will achieve a dew point less than the minimum operating temperature in any part of the system, downstream of the separation unit. Consequently, this avoids the condensation of water, which

- eliminates the potential hydraulic problems such as slugs which could be destructive for receiving facilities, and other types of flow that enhance erosion rates,
- prevents gas hydrates formation, which could plug the piping system and result in severe damage to the asset, and high risks to the operator, as a mechanical hazard, in addition to high toxicity in case of H_2S content,
- reduces electrochemical corrosion rates and fines, especially with the natural gas of sour and acidic nature,
- guarantees beneficial transportation capacity with higher caloric value, and
- provides a wider operation margin at the low temperature and high-pressure values, which is mandatory in the case of cryogenic processes.

Before designing a natural gas dehydration process, the vapor-liquid equilibrium, and the water-hydrocarbon phase behavior should be well investigated. Proper understanding and

analyses of both aspects would lead to a proper method selection and a more efficient process. In this regard, the reader is referred to the Campell [90]. It is important to understand the thermodynamic behavior of the mixture. The water content varies inversely with the pressure of the sweet gas, as the pressure increases the water content decreases, and directly with the temperature, as the temperature increases the water content increases as well. This behavior is more complicated in the presence of acid gases. Therefore, the essential inputs for the design of a dehydration process are to define the nature of the gas which is being dehydrated, if it is sour or sweet, and to estimate or calculate the water content. Then, a designer can proceed to the selection of a dehydration method or constitute a combination of methods for natural gas dehydration.

5.1 Water content prediction

Although this topic is out of the scope of this paper, presenting a few fundamental aspects is needed. A detailed discussion about water content estimation is available in several works of Bahadori [91] [92] [93], and Carroll [94]. Estimating or defining the water content could be done via direct metering devices such as Impedance Sensors and hygrometers, or indirect methods such as correlations, equations of state (EOS), and charts.

Most of the applicable methods are based on empirical data, considering that the water content is highly affected by the acidic gas content. If the gas is sweet, numerous satisfactory methods could be used for prediction, most of these methods are based on the McKetta-Wehe chart (1958) [95], but if the gas is acidic, it is essential to define the acidic composition ratio of the gas stream.

A precise strategy for estimating the sour gas/sweet gas water content ratio was presented by Baghban [96]. Most of the existing methods for predicting the gas content are not satisfactory when the acidic composition exceeds 30%, and could lead to serious errors when exceeds 50%, particularly at higher pressure. In this case, the method by Bahadori [97] is very helpful, as well as the chart by Wichert [98], which is based on a wide set of empirical data, and provides an appropriate estimation even with high acid gas concentrations. Caroll [94] provided a very helpful summary of the models and corrections for water content prediction. Most of these models are based on the calculation and correction of the water vapor pressure to the total pressure. For this purpose, the correlation proposed by Saul and Wagner [99] is recommended.

The equation of state methods (EOS) are being used for water content calculation as well, taking into consideration that EOS methods face the same problem with the gas of acidic nature. EOS methods overestimate the content of water for the high acid gas concentration and pressures, in comparison with the empirical data [100].

5.2 Pipeline and process specification

The desired pipeline specification is affected mainly by the ambient conditions. As much as the ambient temperature is lower, the allowable water content is more restricted according to the applicable specification. For example the US pipeline specification defines the allowable water content with 7 [lb/MMscf] (112 [mg/m^3]) in the southern states, while this number is decreased to 4 [lb/MMscf] (64 [mg/m^3]) for the northern states pipelines, and <2 [lb/MMscf]

(32 [mg/m^3]) for Alaskan pipeline [1]. The water content specification could be expressed in terms of dew point at a certain pressure, for example, the dew point in the middle east is < 0 [^{o}C] at 69 [bar(a)], considering that the dew point is affected with the content of sulfur.

Knowing the nature of the gas, sweet or sour, pressure, and dew point, the water content of the wet gas (initial stage) could be defined by one of the previously mentioned prediction methods. As the required water content or dew point of the product is known, the entire water removal [%] could be found easily:

$$Water\ removal\ [\%] = \frac{Initial\ water\ content - final\ water\ content}{Initial\ water\ content}$$

The initial water content, and consequently the water removal, is highly affected by the final water content or dew point. Increasing a dew point of a few degrees celsius would result in a considerable increase in the water removal percentage.

The above-mentioned dew points or content values are valid for natural gas transportation via pipeline, and for some processes within gas processing such as compression and condensate stabilization, but cryogenic processes such as NGL recovery and fractionation are much more demanding, and the allowable water content has to be decreased to the order of few [ppm], which meets a dew point in the range of -50 [^{o}C] to -150 [^{o}C], depending on the gas composition, pressure, and cryogenic process. Such a process requires what is known as deep dehydration.

5.3 Dehydration methods

As soon as the water content is estimated or determined, the designer can proceed to select a proper dehydration method or combination of several methods in series to achieve the desired specification. Figure 10 presents the categories and sub-technologies of natural gas dehydration methods.
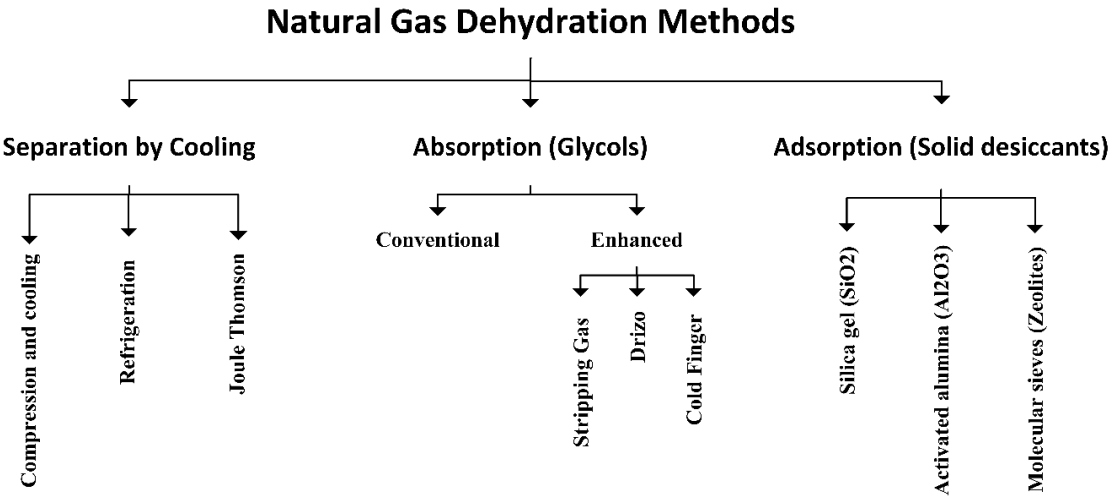

Figure 10 Natural gas dehydration methods

The natural gas dehydration process is selected based on the water dew point requirements. Helpful illustrations which define the ranges of applicability of each dehydration method could be found in the work of Netušil and Ditl [101], and Kemper [102], Figure 11.
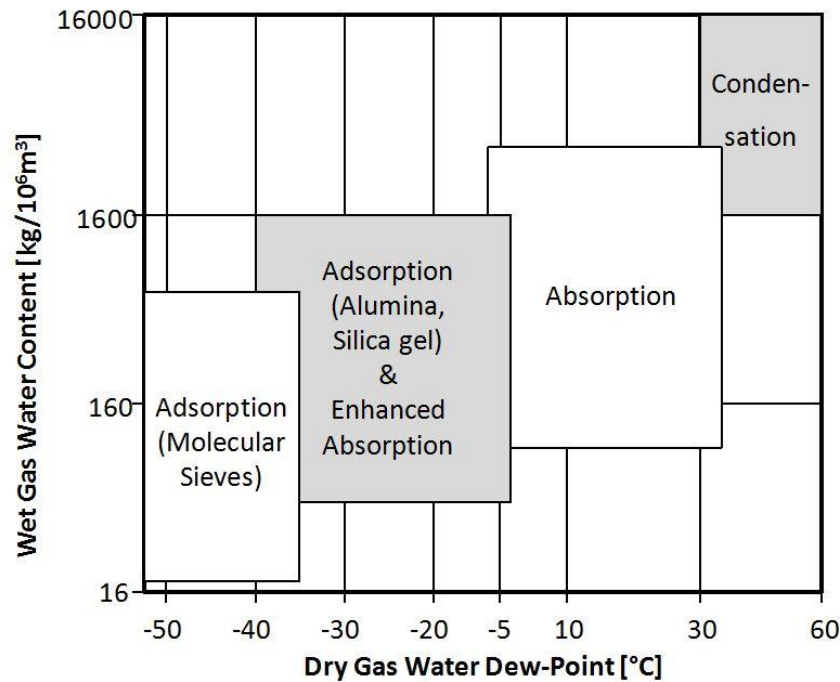

Figure 11 Natural gas dehydration methods due to water content [101]

In industry, there are three conventional methods for natural gas dehydration, direct cooling (refrigeration), absorption, and adsorption. Depending on the desired specification, a method or a combination of methods could be used.

- **Separation by cooling:** This method is very helpful in dehydrating the high-pressure NG streams with high water vapor content. It is based on condensing the water which will decrease the saturated vapor content of NG. Heat exchangers and Joul-Thomson could be used. The main problem of this method is the hydrate formation, which is usually prevented by injecting a polar solvent, such as monoethylen glycol (MEG) or methanol, as a hydrate inhibitor before cooling. The injected solvent selection depends on the low-temperature properties. This method can reduce the water content down to 1 [lb/MMscf] (16 [mg/m^3]) [94], but it is still insufficient for cryogenic processes, in addition to the high energy consumption and operational problems.

  The most applicable cooling method is the refrigeration method using Propane as a refrigerant. Figure 12 shows a simplified flow diagram of this method, which is composed of two working circuits, a refrigerant loop, and a glycol loop. To prevent ice or hydrate formation due to low temperature, as well as to catch the water, a glycol (mostly EG) is sprayed upstream of both, the inlet gas pre-cooler, and chiller. The rich inlet gas is being cooled preliminarily into the inlet gas precooler and continues to the chiller where it gives its heat content to the refrigerant (Propane) and evaporates it.

The rich gas leaves the chiller at a low temperature (-10 [^{o}C] to -40 [^{o}C]) to the low-temperature separator (LTS). At the LTS the mixture is separated into very cold lean gas, condensate, and rich glycol/water mixture. The cold lean gas continues to the inlet gas pre-cooler, where it absorbs some heat from the rich inlet gas causing the rich inlet gas to be pre-cooled as previously mentioned. The rich glycol water mixture continues to the regeneration unit where the mixture will be heated to a temperature higher than the boiling point of water and the water evaporates from the mixture while the lean liquid glycol is circulated to be sprayed again with the rich inlet gas upstream of the pre-cooler and chiller. The other loop of this process is a typical refrigeration loop, which consists of an evaporator (chiller), compressor, and condenser.
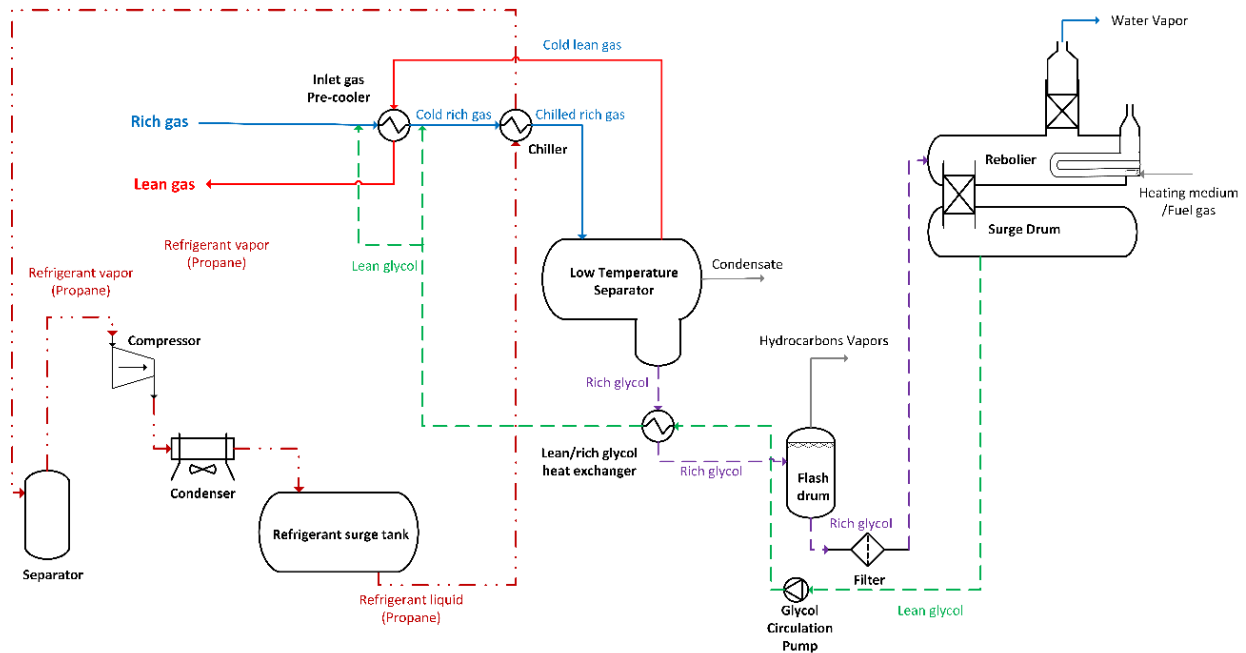

Figure 12 Natural gas dehydration by refrigeration

- **Absorption**: **It is the most common method of dehydration**, especially with using glycol compounds as absorbents. This method has higher efficiency than direct cooling in the mid. range of dew point and water content. Monoethylene glycol (MEG) (So-called EG), Diethylene glycol (DEG), and Triethylene glycol (TEG) are the main glycol solvents for use in this method, especially TEG, which is very efficient due to higher boiling point, and fewer emissions losses. Tetraethylene glycol (T$_4$EG) has less loss at high contact temperatures but it is more expensive than TEG. Figure 13 shows a typical scheme of the TEG unit. Each TEG unit consists mainly of an absorption section and a regeneration section. The lean TEG is getting in contact with the wet gas at the contactor, and the output of the contactor is a dry gas and rich TEG. The water would be separated from the rich TEG in the regeneration unit by evaporation.
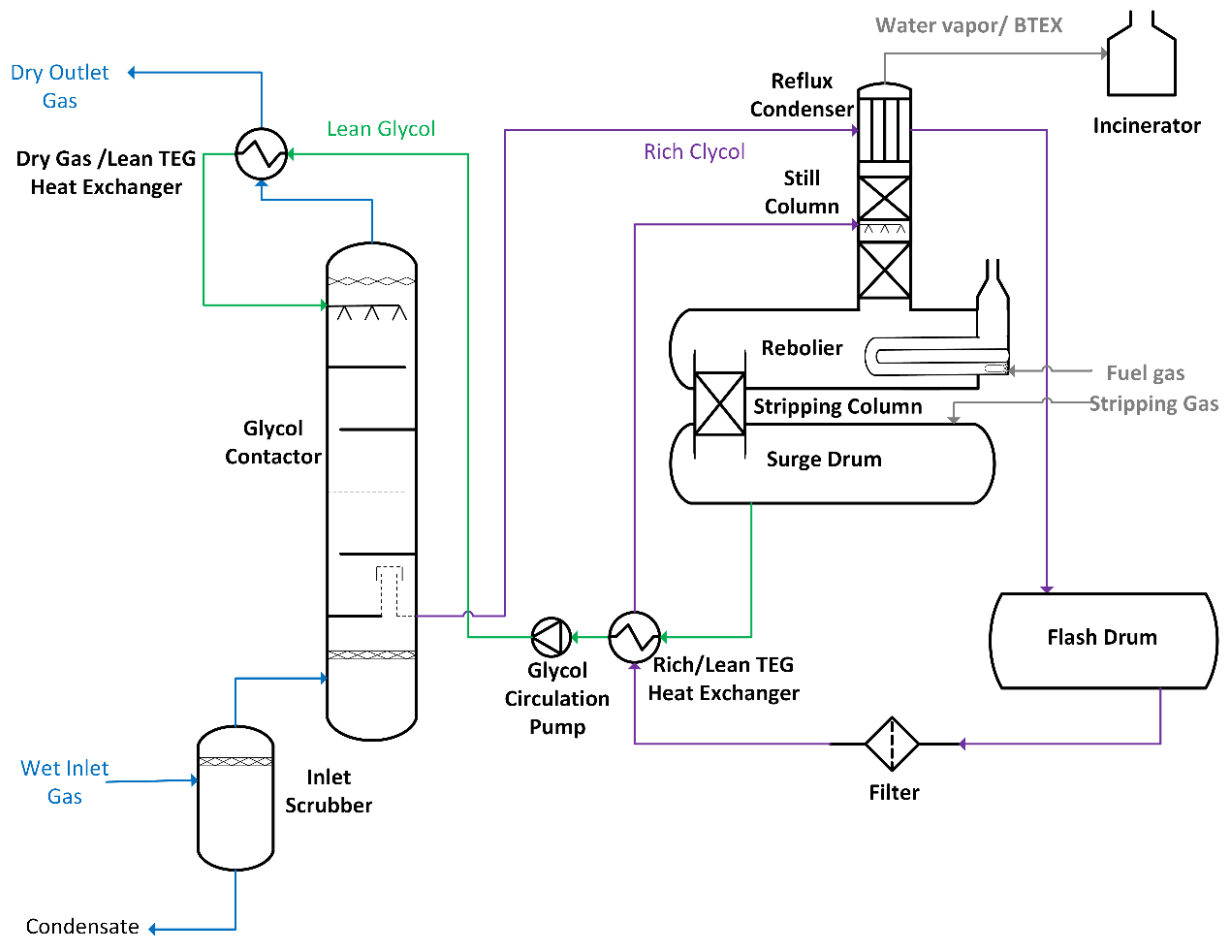
Figure 13 Typical TEG unit

The regeneration part consists mainly of the flush drum, still column, reboiler, stripping column, surge drum, filtration, and circulation pumps, in addition to some material and heat recovery units. The main consumption of the regeneration is an energy consumption at the reboiler, which could be a hot oil heat exchanger, or fired heater, in addition to material consumption, such as TEG makeup, and stripping gas in some of the enhanced configurations of TEG units.

BTEX (benzene, toluene, ethylbenzene, and xylene) is still one of the main problems of glycol dehydration units. Although TEG doesn't absorb hydrocarbons, it will remove a significant amount, up to 20% [1] of these heavy hydrocarbons that would be evaporated with the water at the regeneration. These VOCs (volatile organic compounds) must be incinerated, which will solve the harmful emissions problem, partially, but it is still a loss of energy via flaring, as well as another type of emissions.

There are several configurations of TEG units, conventional and enhanced, the main differences between these configurations are related to the regeneration process. In stripping gas, the separated water vapor is vented with the other gases to the flaring. The Drizo configuration utilizes a recoverable solvent as the stripping medium, and the Coldfinger configuration employs a cooling coil in the vapor space of the surge drum, which causes condensation of water, and this TEG rich condensed water will be further

processed to separate the TEG [103] [1]. Kong [104] presented a review of the enhanced glycol technologies that have been developed to overcome the limitation in glycol purity.

Using the absorption method can reduce the water content significantly, to be about 10 [pmmv], which means dew point temperature down to -30 [^{o}C]. This water content reduction might be insufficient for downstream cryogenic processes such as LPG recovery and fractionation, especially for the low-temperature separation process (LTS). For such a process lower dew point temperature of -30 [^{o}C] to -60 [^{o}C] has to be achieved, and further deep dehydration via adsorption becomes a must.

The combination of the low-temperature cooling (especially JT) and absorption (TEG) methods is conventional, and it was used for comparison with more modern technologies, such as supersonic separation (Twister® [105]), and a subsea gas dehydration method [106].

- **Adsorption**: The absorption dehydration might be insufficient in the case of downstream cryogenic processes. For some processes such as LNG recovery, a very low dew point temperature of about -60 to -110 [^{o}C] has to be achieved, so further dehydration via adsorption is done. The molecules of water vapor (adsorbate) adhere in an extremely thin film to the surfaces of solid bodies (adsorbent) with which they are in contact. Air drying and natural gas dehydration are the most common purification applications of adsorption. Where the deep dehydration upstream of cryogenic processes becomes a must, adsorption is the most efficient method for meeting the required inlet gas specification. There are two types of adsorption, chemisorption where a strong chemical surface bonding occurs between the adsorbent and adsorbate, and physisorption where the bonding is the week van der Waals forces. While the chemisorption forms a single layer of adsorbate on the adsorbent, the physisorption (Physical adsorption) could take a place on the adsorbent surface as a single layer (unimolecular), or more (multimolecular). The number of layers depends on the adsorbent material and surrounding conditions.

The equilibrium between adsorbent and adsorbate could be represented by the adsorption isotherm, isobar, or isostere. The adsorption isotherm represents the relation between the adsorbed amount against the varied pressure when the temperature is kept constant. There are several mathematical models for isotherm representation Figure 14, the most common are those of Langmuir for the monolayer adsorption and dual layers adsorption, where the dual-site Langmuir model is going to be used in the case study. A detailed explanation of adsorption fundamentals and isotherms extraction could be found in Brunauer [107], Ruthven [108], and Suzuki [109].
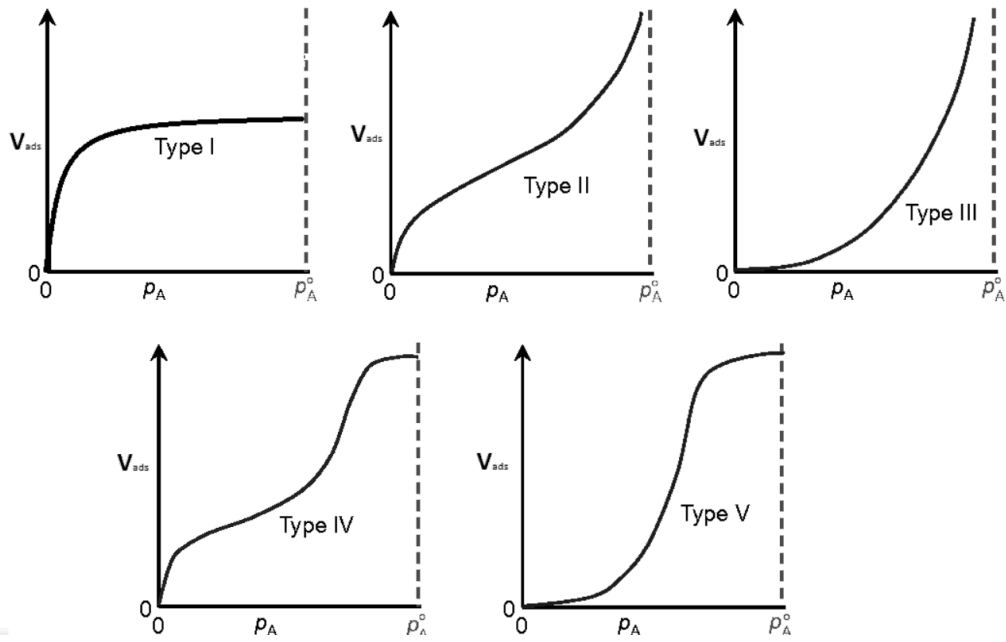
Figure 14 Physical adsorption isotherms

Due to its activated nature, the rate of the chemisorption increases exponentially with the temperature. On the contrary, and due to the exothermic nature of the physisorption, the amount adsorbed at equilibrium must always decrease with increasing temperature according to the principle of Le Chatelier Figure 15.
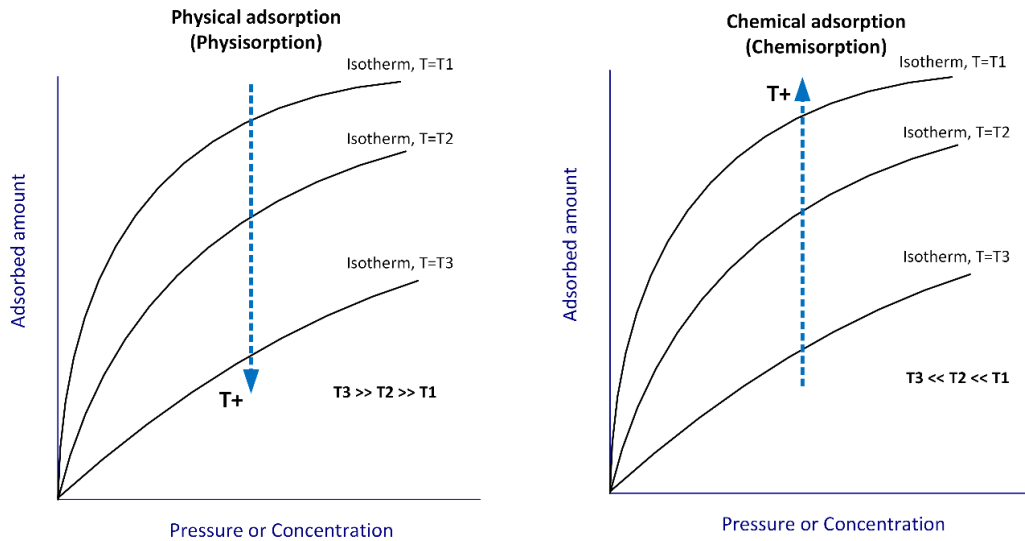

Figure 15 Adsorption isotherms with temperature change

The commonly used adsorbent in natural gas dehydration is the solid one, so-called desiccants, and the most commercially of them are the zeolites. Bahraminia [110]

presented a review of the adsorption dehydration process, where the most important solid desiccants and their improvements have been reviewed. The adsorption of water on zeolites is a process of a physical nature. This material has an extensively wide surface to its volume, approximately 750 [cm^2/cm^3]. The adsorption doesn't take a place only on the surface of the particle but within its microporous structure of it as well.

In principle, the adsorption dehydration system consists of, at least, two adsorption vessels, so-called molecular sieve contractors, in addition to a gas regeneration heating unit, Figure 16. The adsorbers are packed with desiccants, and work in a batch-type process within a continuously operating facility. One adsorber is within the active phase (adsorption phase), while the other would be in the regeneration phase. The operation is regulated via an automated control system that switches over from one adsorber to the other and controls the regeneration process of the saturated adsorber. The length of each cycle is decided according to the saturation when the leading edge of the mass transfer zone (MTZ) reaches the end of the bed, illustrated by breakthrough curves.
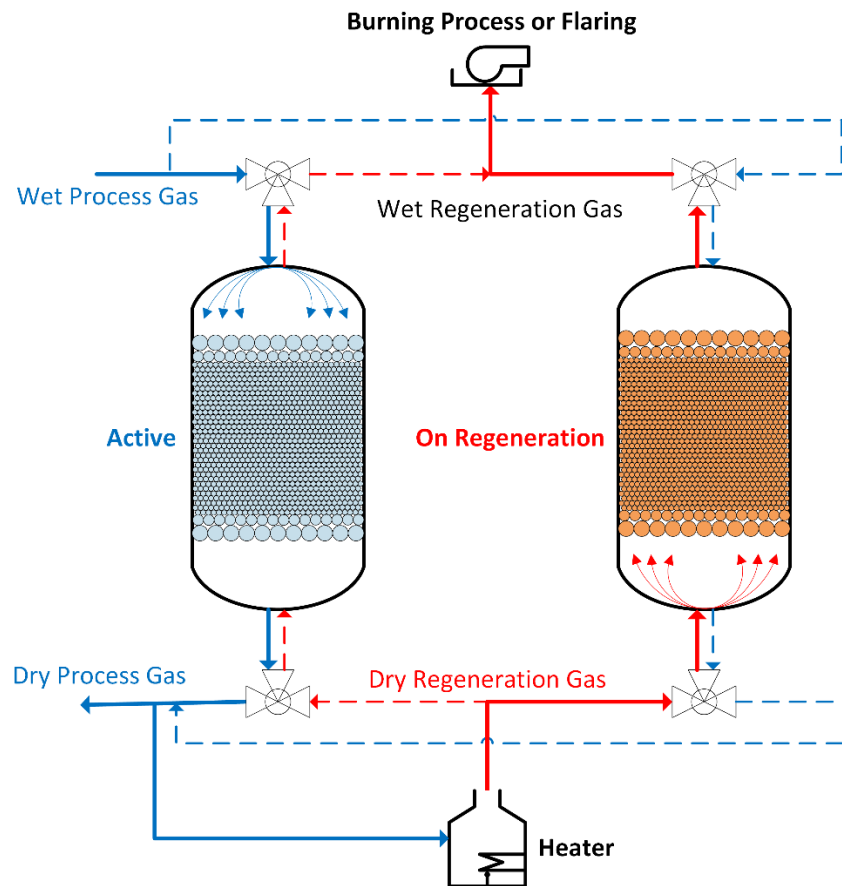
Figure 16 Adsorption unit with regeneration

The regeneration of the saturated contactor is a process of multi-stages. Its principle is the desorption of the adsorbate by breaking its weak surface bond with the adsorbent. This could be achieved by reducing the pressure, increasing the temperature, or both [109].

The first method, depressurization, is the principle of pressure swing adsorption (PSA), which is the most common in adsorbent regeneration [69] [70]. Depressurizing method is limited by the minimum pressure that we can achieve, which is not necessarily the atmospheric pressure. Also, it is affected by the strength of the bond between the adsorbate and adsorbent, which is not easy to break by depressurizing the system only. When the depressurizing is not applicable or insufficient, the second method is applied, which is heating. Through this method, a deep regeneration or drying of the adsorbent could be achieved, Figure 17. A combination of both methods could be the most common, especially in the natural gas dehydration units, where most of the adsorbed water is released by depressurizing, while heating is responsible for releasing the residual water in addition to the minor amount of hydrocarbons, that were also adsorbed during the active phase.
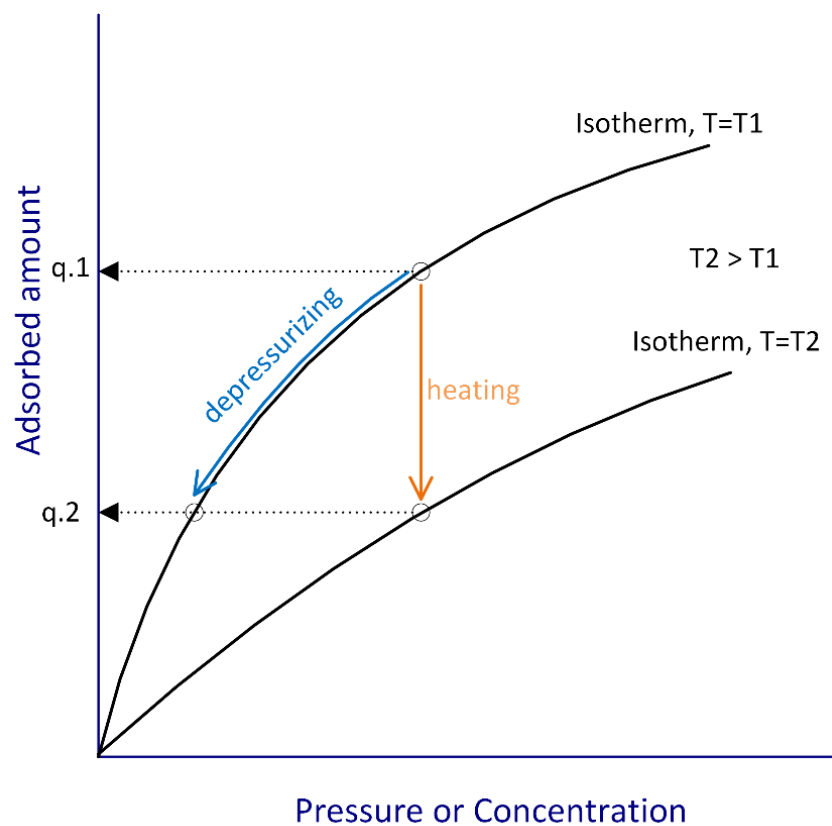
Figure 17 Desorption methods

The consumption of the regeneration phase is an energy consumption at the heating unit, which is mostly a hot oil heat exchanger, in addition to material consumption, which is the regeneration gas that is being extracted from the dry gas stream downstream of the active adsorber. While the regeneration gas could be reused as a fuel gas, the energy consumption is considered the only loss.

Within this thesis, the Triethylene glycol (TEG) absorption unit was selected as a case study, to validate the methodology.

Chapter 6 Modeling approach

This work aims to develop a clear **systematic simulated-based method** that meets

- the PI drivers and process specifications, and

- easy and efficient utilization, that serves the **future design**, as well as the **current facilities** in operation.

Investigating the **mutual relation between** the **REC** and the **water content** specification represents the cornerstone of the proposed method.

- REC is a key driver and evaluation tool of PI, and

- the water content specification is a key input for NGD design.

The $REC_{Dehyd.}$ is used as an evaluation tool of the intensified solution that has to fulfill the product specification (water content). The $REC_{Dehyd.}$ consists of the equipment or process energy consumption ($REC_{equipment}$) and enthalpy loss ($Enthalpy_{vent}$).

$$REC_{Dehyd} = REC_{equipment} + Enthalpy_{vent}$$

The announced aim is approached by using a combination of two methods,

- a systematic method that recognizes unit intensification and extended process intensification, similar method was firstly suggested by Ponce-Ortega [17], and

- a concise framework for PI implementation in the O&G proposed originally by the author [35].

The established simulation-based method uses the powerful process simulator **ASPEN HYSYS**.

As a case study, the **absorption-based dehydration** by TEG in an existing gas plant is used.

The case study's primary data is prepared from the available documentation from the operator. A current state model is established, and the reference results are calculated.

In the second stage, the intensified solution of the NGD is structured as follows

- the PI of the core unit (core node (TEG unit)), with its independent variables,

- the PI of the extended unit including the integrated upstream and downstream.

Hereinafter, the case study of the existing gas processing facility is presented, simulated, and the results are discussed.

Chapter 7 Case Study

The studied facility is an existing gas processing plant of two parts,

- a gas gathering and initial processing station denoted hereinafter as (GGS), and

- a gas treatment plant, denoted as (GTP)

7.1 Process overview: Natural gas dehydration unit

A simplified block diagram of the entire chain from wellheads to the final product, throughout the GGS, trunkline, and GTP, is shown in Figure 18.
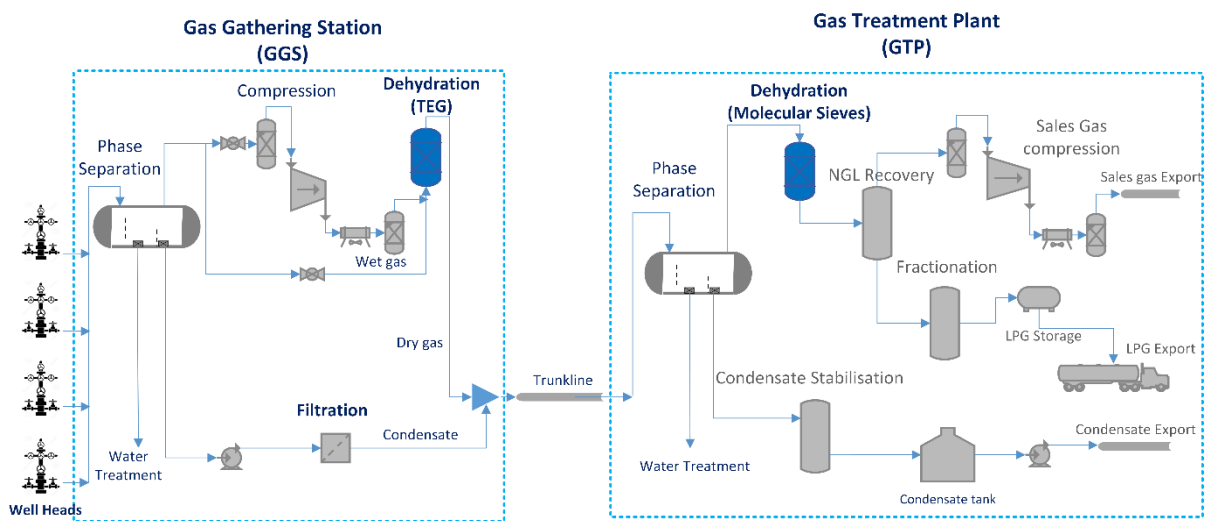

Figure 18 Gas processing facilities block diagram

The stream from WHs is transported to the GGS inlet manifold through flowlines. The 3-phase inlet stream consists of gas, condensate, and water. The liquids will be separated from the stream at the inlet 3-phase separator. Separated oily water will be treated at the oily water treatment unit. Condensate will be filtered of solids, then of water using a filter coalescer. The wet gas will proceed to the TEG dehydration unit. The carried-over liquids within the wet gas stream could be separated and recovered by 2-phase inlet separators of units, while the water vapor could be eliminated only by dehydration. Though, the major portion of water vapor will be absorbed in the first dehydration unit, the TEG unit.

The field reserve profiles indicate that the well-pressure will be high during the initial phase of production, then expected to decrease substantially. Hence the production operation mode is termed high-pressure operation (HP-Phase) and low-pressure operation (LP-Phase). The GGS is designed to work over both operational modes, though, it is equipped with a compression unit that would be in operation in the LP phase, to raise the pressure of the stream after separation to the operational pressure.

The filtered condensate and dehydrated gas are mixed at the outlet of the GGS and transported through a trunkline of a 77 [km] length to the GTP, where another round of separation and dehydration takes a place. The dehydration unit at the GTP contains Molecular Sieves to

33

remove the moisture by adsorption, which achieves a water content of 0.1 [ppmv], the required specification for the downstream cryogenic process, NGL recovery, and fractionation.

The recovered NGL products will be processed at the fractionation unit to produce the LPG. While the dry sales gas will be compressed and exported. The condensate liquids which are separated at the GTP inlet separator so-called slug catcher, and the 2-phase inlet separators of units, would be gathered and processed during the stabilization process. The oily water will be treated at water treatment.

In comparison to the basic sequence of NG processing in Figure 1, this studied facility has no treatment and sweetening unit, as the raw gas is of low content of the acidic and sour gases. Also, the compression at the GGS unit is currently bypassed due to high pressure at the inlet manifold. A preamble of the studied facilities is presented in Table 7.

Table 7 Primary data about the case study (Gas processing plant)

| Plant	Gas Gathering and Treatment Plant
Production Capacity	Gas: 80 [MMSCFD]= 2.265×10^6 [Sm3/d], Condensate: 22.5 [m^3/d], about 141.53 [bbl/d]
Production phase	High-pressure stream at the inlet manifold
Fluids	Gas, condensate, water
Acid/Sour gas	No
Products	GGS outlet: Dehydrated gas and condensate mixture GTP: Sales Gas, LPG, Condensate
Production facilities	7 Wellheads, flowlines, gas gathering station (GGS), trunkline, and gas treatment plant (GTP).
Dry gas specification	GGS outlet: Dew point: < -5 [^{o}C] ~ 35 [ppmv] GTP Fractionation inlet: <-60 [^{o}C] ~ 0.1 [ppmv]

A) Dehydration unit at GGS (TEG Unit)

The wet gas at an approximate pressure and temperature of 87 [barg] and 40 [°C] enters the integrated scrubber of the gas dehydration contactor (105-C-001), where any entrained free liquid is removed before the gas flows on into the dehydration section of the contactor, Figure 19. The recovered liquids in the inlet scrubber are drained to Flare Drum in case of high pressure operating case, or GGS inlet separator in case of low pressure operating case Figure 18.

The TEG contactor has a structured packing design where the mass transfer takes a place. The lean TEG enters the contactor from the top with an approximate temperature of 50 [°C]. It gets equally distributed over the whole section of the column above the packed bed, then flows downward through the bed, absorbs the water of the upward gas flow, and leaves the bottom

of the contactor as a rich TEG. The lean TEG inlet temperature is maintained at 5-10 [°C] above the wet gas inlet temperature to avoid hydrocarbon condensation that causes foaming in the contactor. The TEG contactor is equipped with a high-efficiency demister that catches the entrained glycol droplets from the dehydrated gas stream before it leaves the top of the contactor. The dehydrated gas from the top of the contactor enters the Dry Gas/Lean TEG Heat exchanger (105-E-004), where the lean TEG is cooled to maintain the above-mentioned contact temperature difference. The dehydrated gas continues to be mixed with the dehydrated condensate, downstream of the condensate filter coalescer, then it is exported to the GTP through the trunkline, while the rich TEG is sent to the TEG regeneration unit.
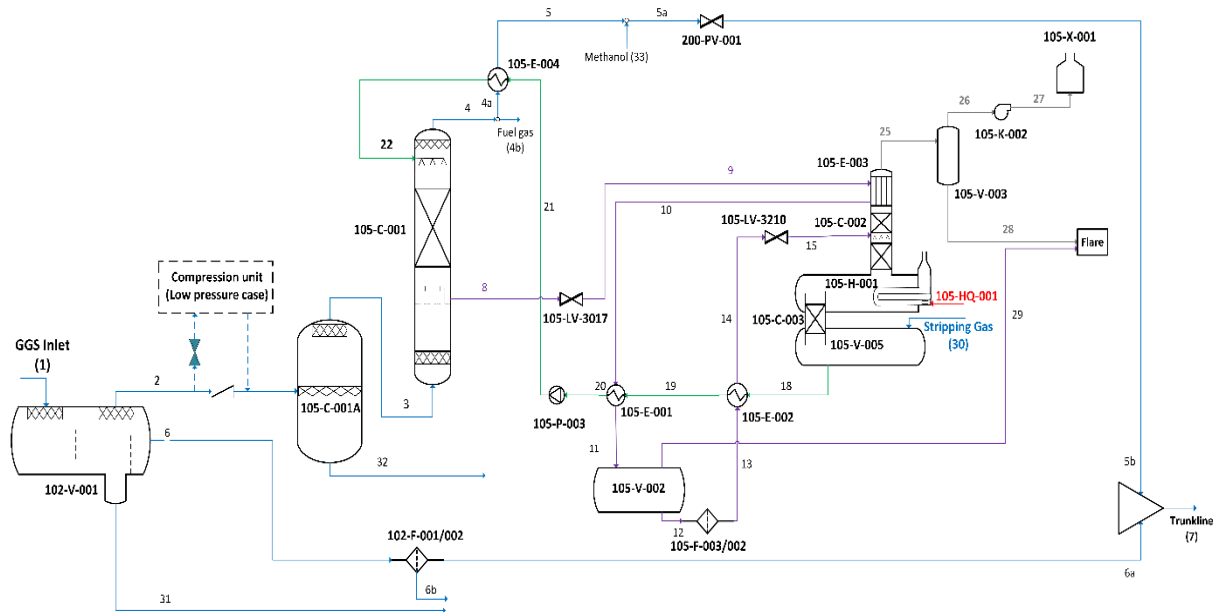
Figure 19 Gas Dehydration- TEG Unit

The rich TEG, with some dissolved light hydrocarbons such as Methane and Ethane, leaves the bottom of the contactor (105-C-001) and the pressure will be reduced at the 105-LV-3017 to the app. 5-6 [barg], and flows through the reflux condenser (105-E-003), which is installed on the top of the still column (105-C-002) and intended to cool the vapors that leave the still column to the incinerator unit (140-X-001). This will reduce the TEG losses with the water vapor exhaust. The preheated rich TEG continues to the cold lean/rich glycol heat exchanger (105-E-001), where heat will be recovered from the lean TEG stream, this will help flash off the dissolved light hydrocarbons when the rich TEG gets into the flush drum (105-V-002), rather than to have them with the still column vapor exhaust later. In this facility, these light hydrocarbons are sent off to the flaring. The flush drum is equipped with a weir, to catch the liquid hydrocarbons, which are operation problematic if exist with the rich TEG, and drain them into the closed drain.

The rich TEG at the outlet of the flush drum is filtered and flows through the hot lean/rich glycol heat exchanger (105-E-002), where another portion of the heat is recovered from the hot lean TEG, and the rich TEG is preheated. This multistage heat recovery helps to reduce the reboiler duty (105-H-001). The rich TEG enters the still column (105-C-002) and flows

downward through the packing to the direct-fire heated reboiler. The rich TEG is boiled at about 200-204 [°C], and the water vapor flows up through the packing of the still column to the reflux condenser (105-E-003) where it will cool down. The entrained TEG should not travel beyond the reflux condenser, as mentioned above, while the water vapor with the vapors of BTEX go out the top to the incinerator. To increase the purity of the TEG, the regeneration process is enhanced by a stripping column (105-C-003). When the lean TEG flows over the weir of the reboiler, it flows down the packing of the stripping column and gets in contact with the upward dry natural gas flow, so-called the stripping gas. To meet the equilibrium, the dry NG flow will drive out the traces of the water that could still exist within the lean TEG. The hot lean TEG will leave the bottom of the TEG surge drum (105-V-005) and goes through the heat exchangers (105-E-001,002 and 004) while the TEG circulation pump (105-P-003) increases the lean TEG pressure to a value higher than the contactor pressure.

The efficiency of the absorption section is measured by the percentage of the water that is being removed from the wet gas, as a rule of thumb, it is 90-95%, while the efficiency of the regeneration section is measured with the achieved purity of the TEG. To define the required TEG purity for a specific dew point, the equilibrium water dew point chart is used [111]. The chart gives the equilibrium water dew point for different TEG purity over a range of contactor temperatures.

Taking into consideration the approach temperature, which is a reserved margin below the equilibrium dewpoint, the TEG purity could be defined at the contactor temperature.

In the studied case, an equilibrium dew point of -5 [°C] in addition to a reserved safety margin (approach temperature) of -5 [°C] which is in total -10 [°C] could be obtained with a TEG of 99.0 [wt.%] at contactor temperature of 40[°C]. Using the stripping gas will enhance the purity.

In addition to the minimum purity degree of the TEG, and the temperature difference between the lean TEG and the NG inlet temp at the contactor, the reboiler temperature is another operational limitation that has to be considered very carefully. The degradation temperature of the TEG is 206 [°C], so the reboiler temperature should not get to this temperature. Moreover, a margin has to be kept due to the hot surface considerations of the heating element. In other words, the lower the temperature that meets the required TEG purity degree, the better lifetime and fewer losses of the TEG at the reboiler.

B) Dehydration unit at GTP (Molecular Sieves)

The mixture of gas and condensate is received at the GTP slug catcher, which takes the role of a 3-phase separator, the gas is separated and sent into an inlet gas knockout drum (202-V-001), then through a filter coalescer (205-F-001) to remove the hydrocarbon liquids and ensure a good molecular sieve life Figure 20. The gas then is heated 1-2 [°C] by the Inlet Gas Superheater (205-E-001) and fed to the adsorption beds so-called molecular sieve driers (205-V-001) from the top. The molecular Sieve system is a batch-type process that consists of three adsorption beds, two are in operation simultaneously, while the third is on regeneration. The used adsorbent is the 3A Zeolit. It preferentially adsorbs water due to its small molecule size

and its polar nature. The 3A zeolite has a pore diameter of 3-angstrom, which minimizes the adsorption of larger hydrocarbons, H_2S, CO_2, and other stream components. The deeply dehydrated gas leaves the bottom of the adsorption bed with an approximate water content of 0.1 [ppmv] and is filtered at the dehydrated gas filter (205-F-002) to guarantee no descent reaches the NGL recovery.
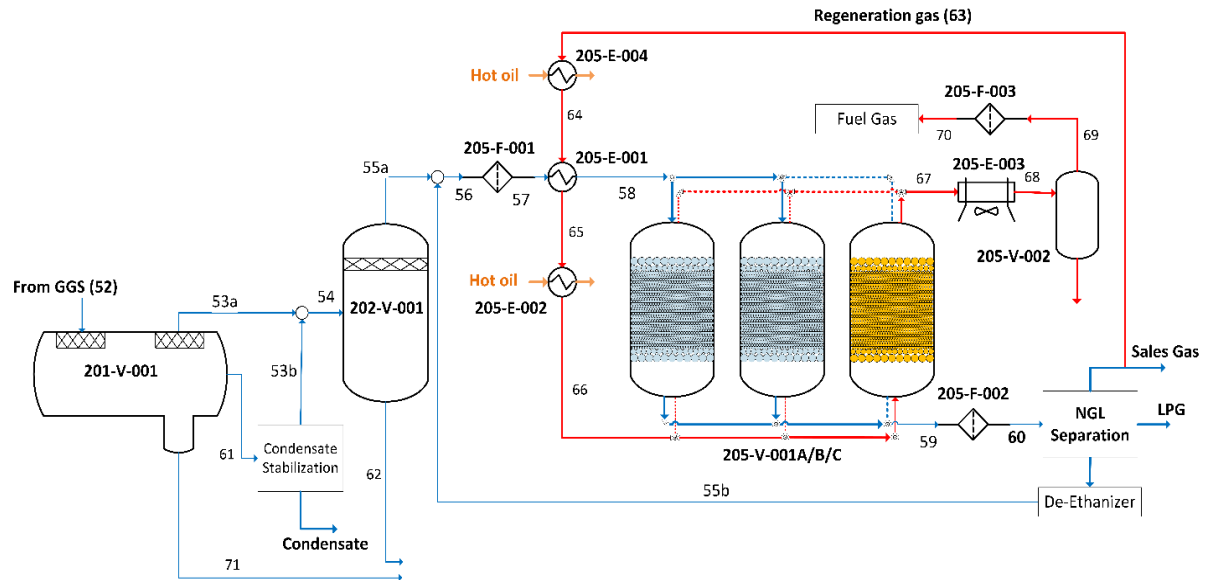
Figure 20 Gas Dehydration- Molecular Sieve unit

The saturated adsorption bed is regenerated with hot dry gas at a temperature of 240°C flowing from the bottom to the top, counterflow of process gas. The dry sales gas is heated at the regeneration gas superheater (205-E-004) and regeneration gas heater (205-E-002). When leaves the adsorption bed, the regeneration gas is cooled by the cooler (205-E-003), and the condensed water is removed at the KO drum (205-V-002). The gas then is filtered at (205-F-003) and used as a fuel gas at the GTP.

The regeneration of the saturated adsorption bed consists of the following sequence (Figure 21)

1. depressurizing from the operating pressure, approx. 61-62 [barg], to 29 [barg] over a period of 30 [min] (1[bar/min]),

2. isobar purging for 30 [min],

3. heating with regeneration gas of two stages to avoid thermal shock, with 120 [°C] reg. gas flow for 30 [min], and with 240 [°C] reg. gas flow for 5 [hours]. The bed is considered regenerated when the reg. gas outlet temp. exceeds 210 [°C]. The reg. gas is heated by the heater (205-E-002),

4. cooling by a reg. gas with a temperature of approx. 30 [°C], depends on the superheater,

5. pressurizing with a reg. gas to the same initial operating pressure, and

6. holding the adsorption bed pressurized on standby till put back in operation.
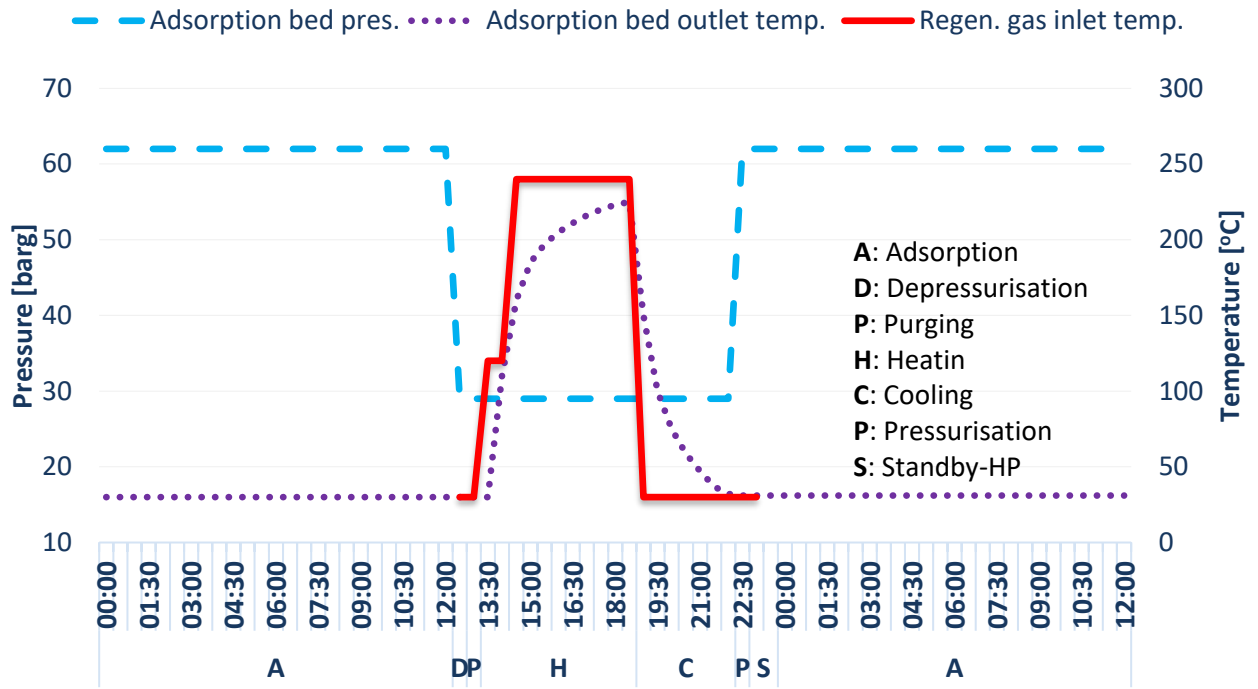
Figure 21 Adsorption/Regeneration cycle of an adsorption bed

(Figure 22) shows a complete cycle of an adsorption unit of 3 beds. It can be noticed that at any moment two beds are in the active phase while one is in regeneration.

Adsorption system cycle (36 hours)	1/3 Cycle (12 hours)	2/3 Cycle (12 hours)	3/3 Cycle (12 hours)
Bed_A	Adsorption	Regeneration	Adsorption
Bed_B	Regeneration	Adsorption	Adsorption
Bed_C	Adsorption	Adsorption	Regeneration

Figure 22 Operating sequence of the Molecular Sieves beds over a complete cycle

The available document and data of the studied facilities include:

- Heat and material balance

- Equipment datasheets

- Process and instrumentation diagrams (P&IDs)

- Operation manuals

As indicated above, the facility includes a dehydration process of two stages, 1st stage is absorption-based dehydration by TEG at the GGS, and adsorption-based dehydration (deep dehydration) with molecular sieves at the GTP.

Hereinafter, the work is **focused on the absorption-based dehydration unit**, as a defined case study to validate the PI approach.

The PI of the studied unit is approached in terms of operational and asset modifications. The simulation is done via the process simulator **ASPEN HYSYS** as follows

- the simulation of the current status is presented, and the current case calculations are done as a reference, then

- the intensified solution is structured and simulated as follows:

- Intensification of the core unit (TEG unit), with its independent variables,

- Intensification of the adjacent upstream and downstream units, intensifies the wider or entire process.

7.2 Simulation input

Table 8 provides the overall parameters of the inlet stream at the GGS and Figures 23 and 24 their composition.

Table 8 GGS Inlet overall phase parameters, characteristics, and composition

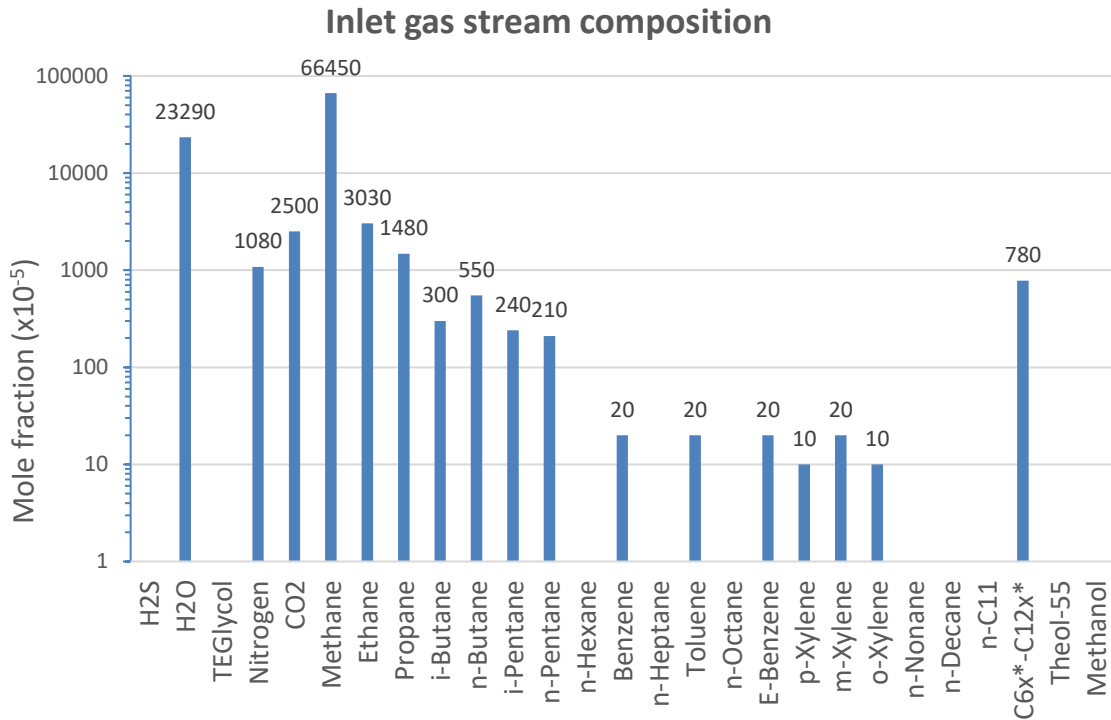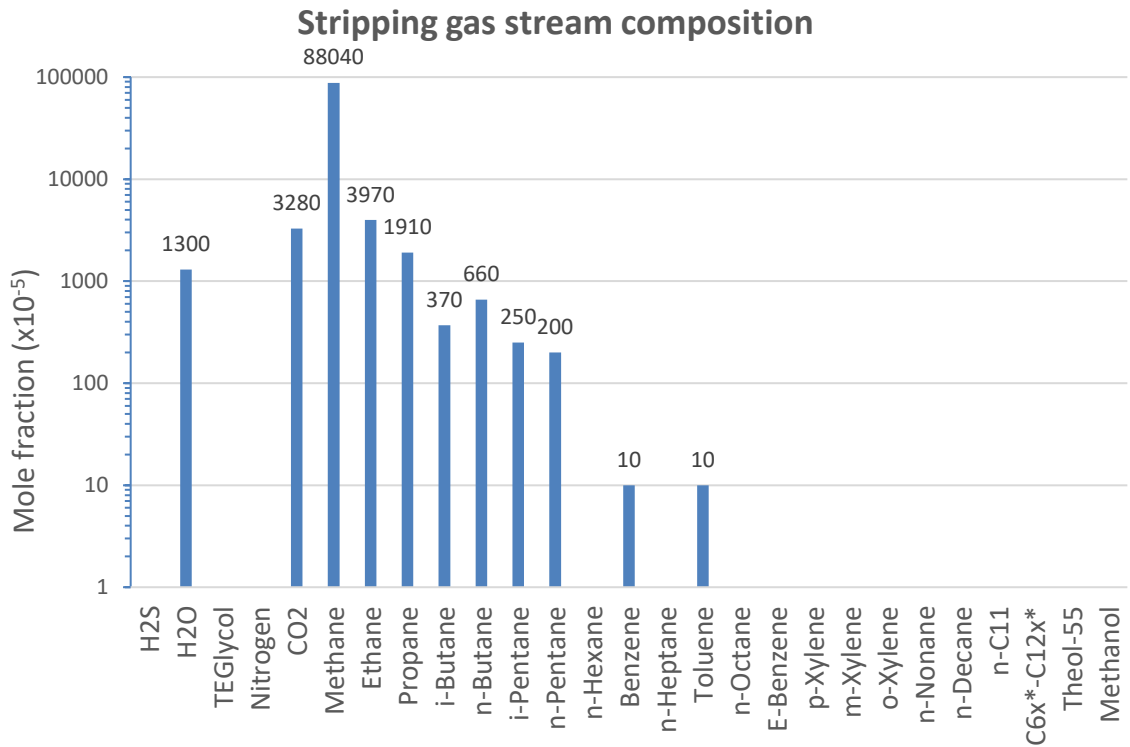|	unit	Fluid from the production manifold (to production separator) High-pressure lean summer (Case HPLS)	Stripping Gas
Vapor fraction	-	0.76	0.9996
Temperature	[ºC]	41.14	15.00
Pressure	[barg]	87.65	0.330
Mass flow	[kg/h]	130820	150.0
Molar flow	[kgmole/h]	6667.2	7.990
Enthalpy (heat flow)	[kJ/h]	-8.9707E+08	-8.980E+04
Composition		Figure 23	Figure 24

Figure 23 Inlet gas stream composition

Figure 24 Stripping gas stream composition

Table 9 provides the list of equipment with the primary input data for simulation, considering that the applicable fluid package for simulation is the glycol package of the TEG unit, and Peng Robinson for the gas stream and GTP units.

Table 9 List of equipment and primary data and inputs for design

Equipment	Code	Simulation input
GGS Production separator	102-V-001	3- phase separator ΔP: 60 [kPa] Heavy liquid in light liquid 4.4 [kgmole/h]
Condensate solid Filter / Filter coalescer	102-F-001/002	ΔP: 100 [kPa]
GGS TEG Dehydration column integrated scrubber	105-C-001A	ΔP: 50 [kPa]
GGS TEG Dehydration column	105-C-001	Packed type (No. of theoretical stages: 3) Packed layers: 1 Packed layer D/H: 1.4/4.62 [m] ΔP: 20 [kPa] (bottom-to-top) Operating pressure: 85 [barg] Packing material: Random 38 [mm] SS 316 L_Pall Ring
GGS dry gas/lean TEG heat exchanger	105-E-004	Shell&Tube type TEMA type: AEL (Counterflow) ΔP: (shell side, lean TEG): 6.3 [kPa] ΔP: (tube side, dry gas): 4.8 [kPa] Inlet/outlet temp. (shell side): 65.7/50.00 [°C] Inlet/outlet temp. (tube side): 42/42.9 [°C]
GGS rich TEG pressure control valve	105-LV-3017	P_{out}: 4.97 [barg] ΔP: 8019 [kPa]
GGS TEG Reflux condenser (Overhead vapor/Rich TEG)	105-E-003	Tube-Shell type TEMA type: BEM (Co-current flow) ΔP: (shell side, rich TEG): <1 [kPa] ΔP: (tube side, gas): <1 [kPa] Inlet/outlet temp. (shell side): 43.3/46 [°C] Reflux condenser temp. (tube side, overhead gas): 85 [°C]
GGS cold-rich/lean TEG heat exchanger	105-E-001	Shell-plate type ΔP: (shell side, rich TEG): 1 [kPa] ΔP: (plate side, lean TEG) : 7 [kPa] Inlet/outlet temp. (shell side): 46/62 [°C] Inlet/outlet temp. (plate side): 82.6/64.6 [°C]
GGS TEG flash drum	105-V-002	Horizontal ΔP: 40 [kPa]
GGS rich TEG cartridge filter/carbon filter	105-F-003/002	ΔP: 100 [kPa]
GGS Hot-lean/rich TEG heat exchanger	105-E-002	Plate-Shell type ΔP: (shell side, rich TEG) : 10 [kPa]

		ΔP: (plate side, lean TEG) : 1 [kPa]
		Inlet/outlet temp. (shell side): 62/150 [^{o}C]
		Inlet/outlet temp. (plate side): 175/82.6 [^{o}C]
GGS rich TEG pressure control valve	105-LV-3210	P_{out}: 2.4 [barg]
		ΔP: 50 [kPa]
GGS TEG Still column	105-C-002	Packed type (No. of theoretical stages: 3)
		Packed layers: 2 (rich TEG inlet in-between)
		Upper packed layer D/H: 0.496/1.00 [m]
		Lower packed layer D/H: 0.496/1.80 [m]
		Operating pressure: 0.1 [barg]
		ΔP: 2 [kPa] (bottom-to-top)
		Reflux condenser temp.: 85 [^{o}C]
		Reboiler temp.: 200 [^{o}C]
		Packing material: Random 38 [mm] SS 316 L_Pall Ring
GGS TEG Re-boiler	105-H-001	Direct fired
		Reboiler temp.: 200 [^{o}C]
GGS TEG Stripping column	105-C-003	Packed type (No. of theoretical stages: 4)
		Packed layers: 1
		Packed layer D/H: 0.387/1.90 [m]
		Operating pressure: 0.2 [barg]
		ΔP: 21 [kPa] (bottom-to-top)
		Stripping gas inlet: at bottom
		Packing material: Random 38 [mm] SS 316 L_Pall Ring
GGS TEG Surge drum	105-V-005	Horizontal
GGS TEG circulation pump	105-P-003	Reciprocating
		ΔP: 87.2 [bar]
VENT GAS KO DRUM	105-V-003	2-phase separator
		Vertical

7.3 Current state simulation results

The steady-state model of the facility is simulated via ASPEN HYSYS, Figure 19 and Figure 25 with the heat and material data of the selected streams for analysis in Table 10.
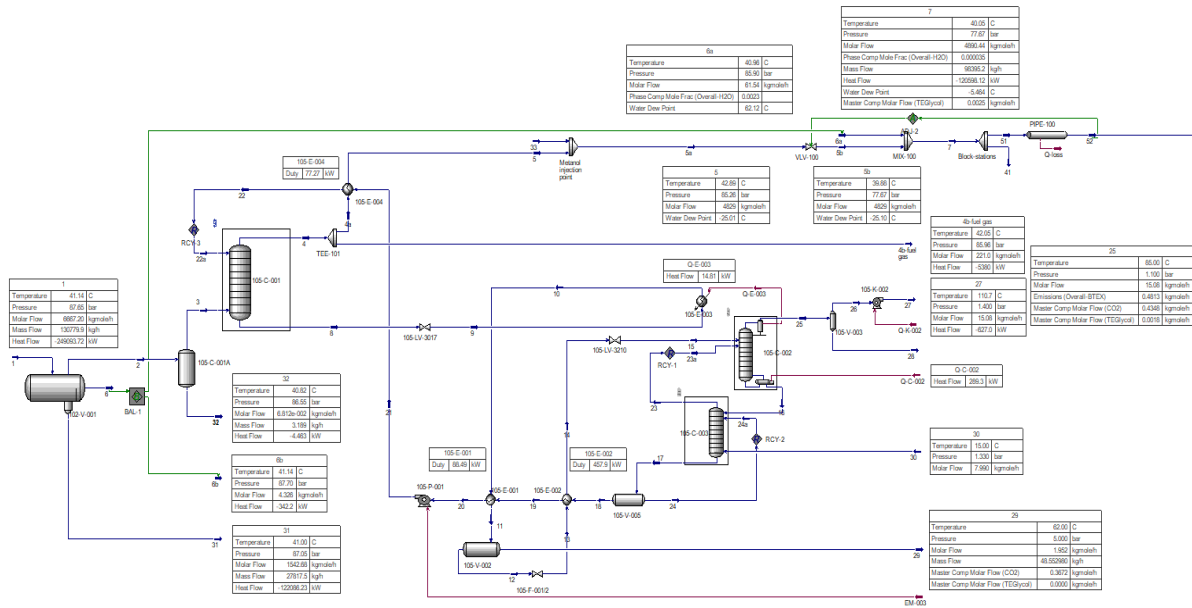
Figure 25 Case study simulated model

Table 10 Heat and material balance of the current state inlet-outlet streams

	1	4	6b	32	30	25	29	7
Main Stream Workbook and Rated Energy Consumption								
	Production separator gas **inlet**	TEG contactor dry gas **outlet**	Condensate Filter Coalescer water **outlet**	TEG contactor separator water **outlet**	Stripping gas **inlet**	TEG regeneration reflux cond. vapor **outlet**	TEG flush drum vapor **outlet**	Trunkline mixed gas-condensate (GGS **outlet**)
Temp.[C]	41.14	42.05	41.14	40.82	15	85	62	40.05
Press.[barg]	86.65	85.96	86.7	85.55	0.33	0.1	4	76.67
Molar Flow [kgmole/h]	6667.2	5050	4.33	0.068	7.99	15.08	1.95	4890.44
Mass Flow [kg/h]	130780	27818	78	3	150	322	49	98395
Liquid Volume Flow [m3/h]	319.97	-	0.079	0.004	0.445	0.629	0.11	279.26
Heat Flow [kW]	-249094	-122933	-342.2	-4.5	-199.3	-631.5	-71.8	-120598
Dew Point [ºC]	-	**-25.00**	-	-	-	-	-	**-5.46**
Reboiler energy consumption 105-HQ-001								**289.35 [kW]**
NGD vent and drain enthalpy (Heat flow loss) (streams 25+29)								**703.3 [kW]**
Other appliances' electrical consumption (motors) (considered fixed)								**23 [kW]**
Total Rated Energy Consumption REC$_{Dehyd.}$								**1015 [kW]**
Comp. Mole Fraction and Emissions								
H2O	0.23286971	**7.5E-6**	0.99789979	0.532687	0.0130387	0.428773411	0.005079883	**35.4E-06**
TEG	-	5.03E-07	-	-	-	119.3 E-06	4.24E-06	5.03E-07
CO$_2$ (Process)	0.02499675	0.0737626	0.00070007	0.0103396	0.0327974	0.028828383	0.188127235	0.032298227
BTEX	-	-	0.00E+00	2.81E-02	-	3.19E-02	2.05E-03	-
Total TEG loss (25+29+7)								0.0043 [kgmole/h]
Total CO$_2$ (Process) (25+29)								0.8017 [kgmole/h]
Total BTEX emissions (25) (BTEX in 29 is negligible)								0.4813 [kgmole/h]

7.4 Intensified solution simulation and results

As previously highlighted, the intensified solution is structured and simulated as follows:

- Intensification of the core unit (TEG unit), with its independent variables,

- Intensification of the adjacent upstream and downstream units, intensifies the wider or entire process.

7.4.1. Intensification of the core unit (TEG unit)

The intensification approach of the TEG unit is established on varying **four independent** parameters and defining the consequent impacts on the TEG unit outlet in terms of **product specification, energy consumption, and emissions**.

The independent variables with the ranges are listed in Table 11. While varying a single independent parameter, the other parameters are held at original operating values, which are the actual values in the existing facility.

Table 11 Studied cases of TEG unit operation

| Case study	Operational parameter	Fixed (Actual operating parameters of the unit)	Investigated range	unit
Case 1	TEG circulation rate	39.83	5-40	[kgmole/hr]
Case 2	Stripping gas flow	7.99	2-16	[kgmole/hr]
Case 3	Stripping gas temperature	15	15-200	[^{o}C]
Case 4	Regeneration reboiler temp.	200	160-204	[^{o}C]

Case 1 - TEG circulation rate

Figure 26 presents the impact of TEG flow rate on the gas specification, lean TEG characteristics, and reboiler power consumption. It can be noticed that the increase of the reboiler power consumption with the TEG flow rate increase is almost linear. The decrease in the water content of the dry gas outlet is almost negligible with a TEG circulation rate > 25 [kgmole/h]. The highest purity degree, 99.62 % is achieved at the TEG flow rate of 15-20 [kgmole/h]. The increase of the TEG circulation beyond 25 [kgmole/h] does not provide better

dehydration, while the purity degree of the TEG drops, even though, the gas outlet spec is still met due to the higher TEG flow rate.
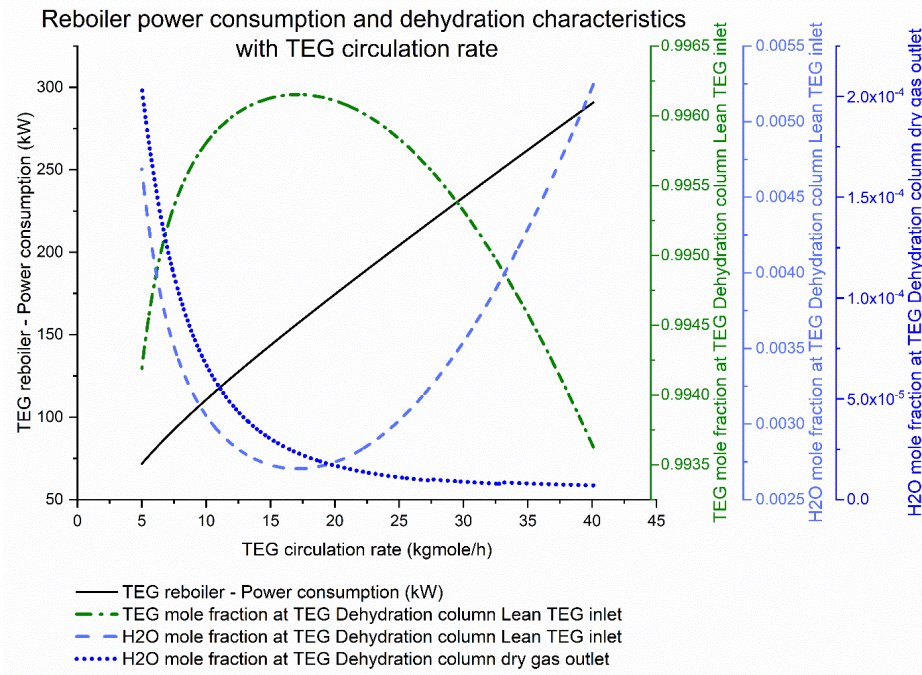
Figure 26 Reboiler power consumption and dehydration characteristics with TEG circulation rate

Investigating the impacts of increasing the TEG circulation rate on the TEG unit emissions, Figure 27, shows a steady increase of the reflux condenser outlet over all the range. While the water component of this exhaust is almost steady beyond the TEG circulation rate of 25 [kgmole/h], the increase in the reflux condenser emissions is attributed to the increase in BTEX emissions as well as the increase in TEG losses. On the other hand, the HC emissions of the TEG flash drum show a steady increase over the entire range of TEG circulation rates.

45
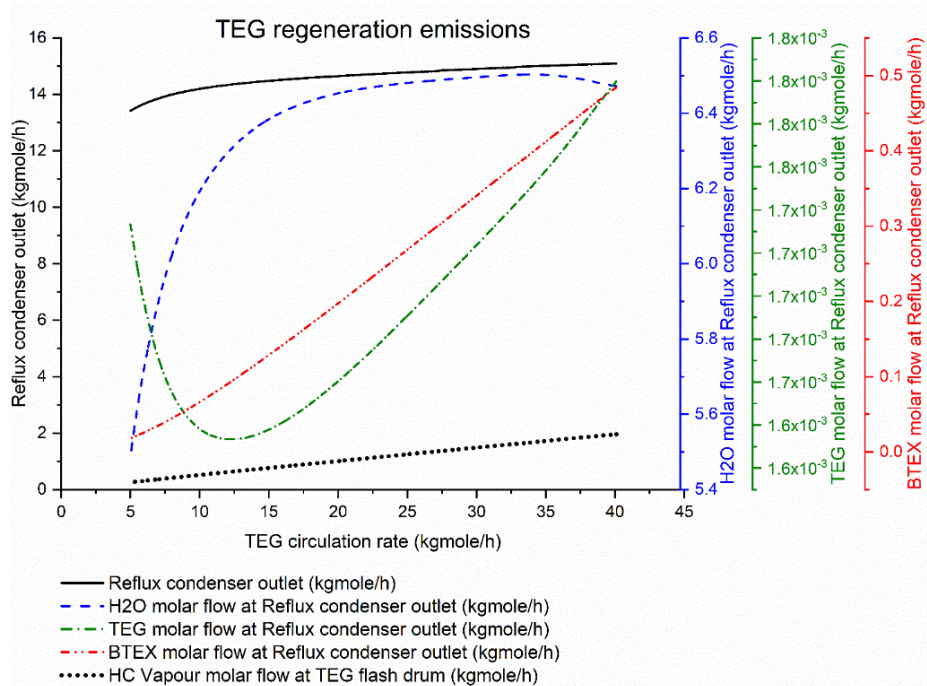
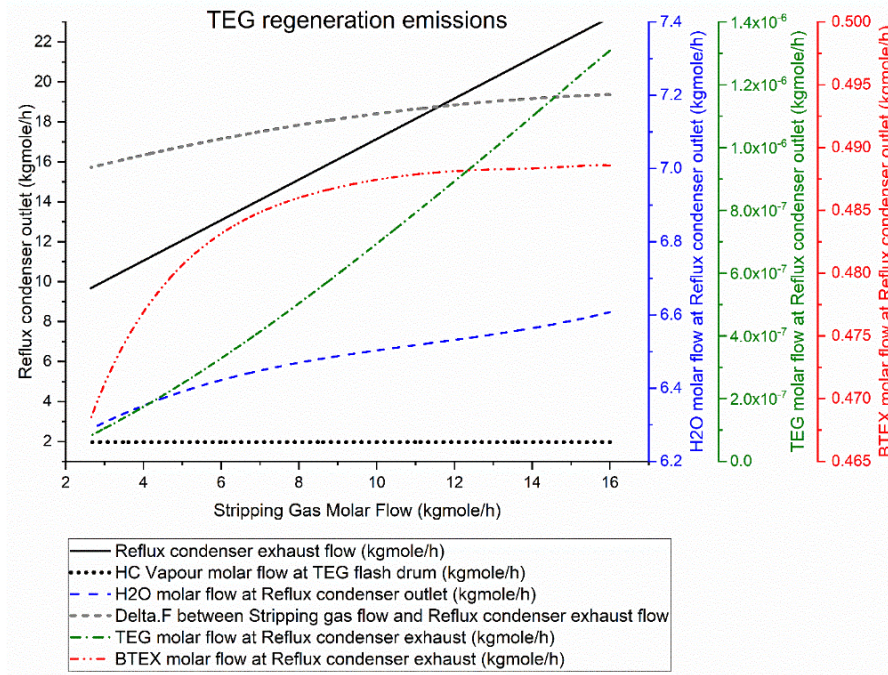**Figure 27 TEG regeneration emissions with TEG circulation rate**

Case 2 - Stripping gas flow rate

The stripping gas flow rate impacts are investigated in the range of 2-16 [kgmole/h]. Figure 28 shows a steady REC with the stripping gas flow rate increase to 10 [kgmole/h], and a very minor decrease with a flow rate increase higher than 10 [kgmole/h]. The decrease of the water content of the dry gas outlet is almost negligible with a stripping gas flow rate higher than 10 [kgmole/h], which achieves a purity degree of 99.6 %.
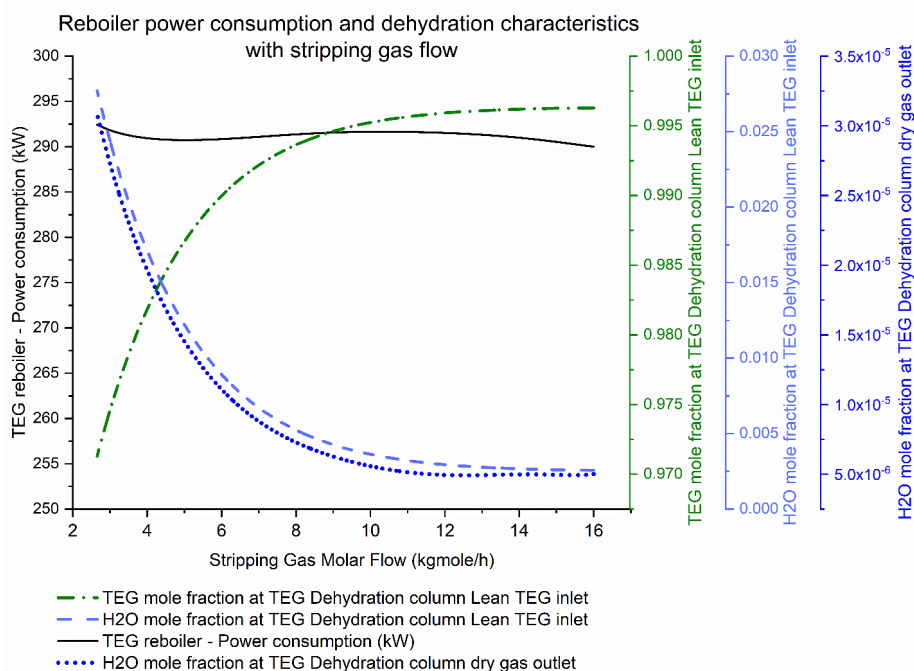
Figure 28 Reboiler power consumption and dehydration characteristics with stripping gas molar flow rate

Increasing the stripping gas flow rate shows a steady increase of the reflux condenser exhaust over all the range. This is logical, as the stripping gas is a part of the reflux condenser outlet. A better characteristic to judge the impact is to analyze the flow difference (ΔF) between the reflux condenser exhaust and stripping gas flow rate. As shown in Figure 29, this (ΔF) tends to be horizontal after a stripping gas flow rate of 10 [kgmole/h], this means that the stripping gas flow rate is almost the dominant component of the reflux condenser exhaust.

The BTEX emissions approach the highest value with a stripping gas flow rate of 10 [kgmole/h], in other words, this stream of stripping gas helps to get rid of all the BTEX content which are burned at the incinerator. The TEG losses increased considerably. The emissions of the HC transferred with the TEG from the TEG contactor are not affected by stripping gas, as they are released from a different section of the regeneration process.

Figure 29 TEG regeneration emissions with stripping gas molar flow rate

Case 3 - Stripping gas temperature

Heating the stripping gas in the TEG reboiler, before injecting it into the surge drum, increases the reboiler power consumption with the amount of energy used for this heating, Figure 30. The improvement in the TEG purity upon increasing stripping gas temperature is found to be marginal (99.361 mole% at 15°C to 99.375 mole% at 200 °C which corresponds to a 0.01% net increase). Hence, no sensible improvement is found in the dehydrated gas water content. The simulation results show no impact of the stripping gas temperature on the TEG regeneration emissions.
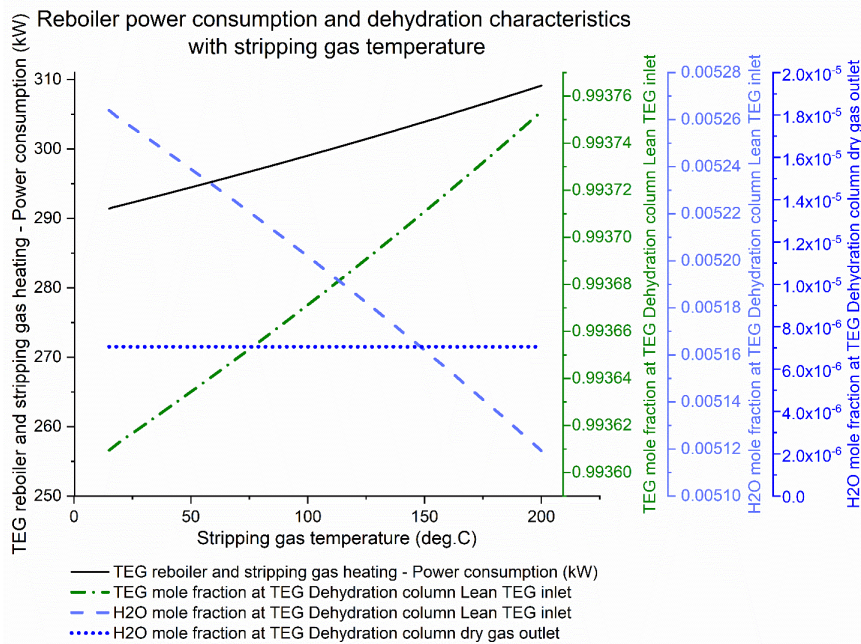
Figure 30 Reboiler power consumption and dehydration characteristics with stripping gas temperature

47

Case 4 - TEG regeneration reboiler temperature

The regeneration reboiler temp is varied within the range 160-204 [^oC], Figure 31. Reboiler power consumption increases linearly with temperature increase, while the water content in the dry gas stream approaches a minimum H_2O mole fraction of 7×10^{-6}. Although the curve is not level out but increasing the temperature beyond 204 [^oC], has a negligible impact on water content on one hand. On the other hand, it is not feasible due to the TEG degradation temperature restriction of 206 [^oC], considering the heating surface temperature safe margin.
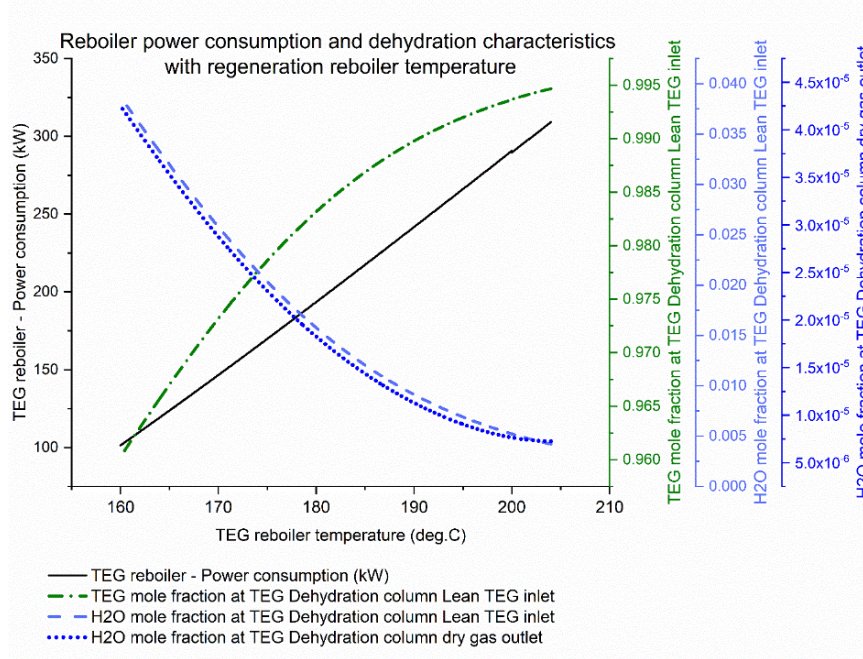
**Figure 31 REC and dehydration characteristics with reboiler temperature**

Increasing the reboiler temperature over the investigated range shows a minor increase in the reflux condenser exhaust. The BTEX emissions reach the highest value at a reboiler temperature of 195 [^oC]. The H_2O vapors molar flow increases as well over the entire range. The HC emissions at the TEG flash drum are not affected Figure 32.
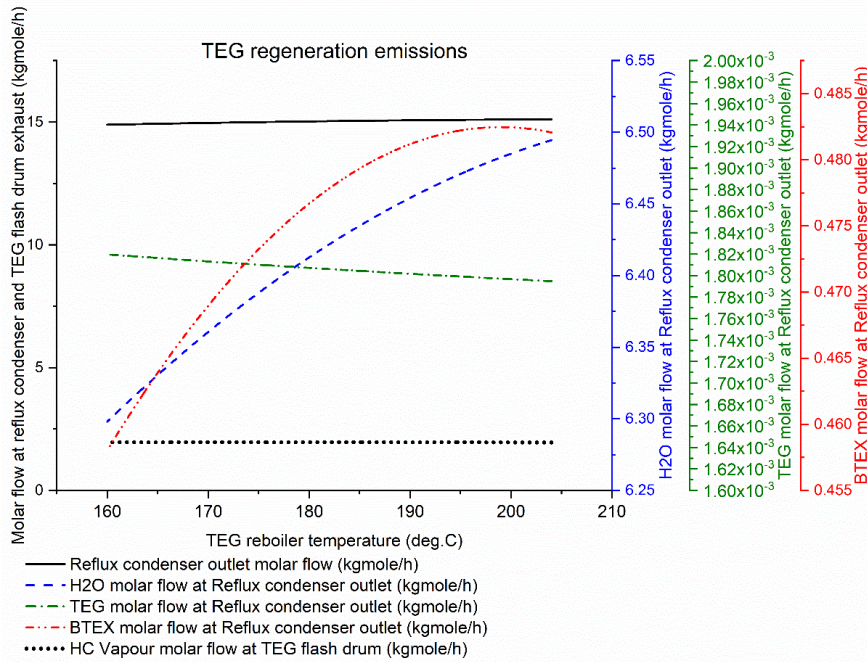
Figure 32 TEG regeneration emissions with reboiler temperature

In a summary, the maximum dryness that could be achieved at the TEG unit with the studied cases is limited to 5.0-7.0 x10^{-6}, Figure 33. It can be noticed clearly that the water content curves are mostly leveled out in the upper rate of the investigated ranges. This differs from one parameter to another. In terms of reboiler power consumption, the stripping gas molar flow increase tends to be inefficient >8 [kgmole/hr], and entirely ineffective >11 [kgmole/hr]. The stripping gas temperature increase is entirely inefficient due to no effect on dehydration over the entire range while consuming energy for heating. The TEG circulation rate is partially inefficient >30 [kgmole/hr], and ineffective >40 [kgmole/hr]. The reboiler temperature is partially inefficient at >200 [ºC], and technically limited to 204 [ºC]. The equivalent reboiler power consumption at a specific water content could be found by defining the working point on the curves of the investigated case or method. To define the required power for further dehydration using the same method, the working point slides on the curves, and the new equivalent power consumption could be defined. The same could be used for a different method, and results could be compared to define the efficient method.

49
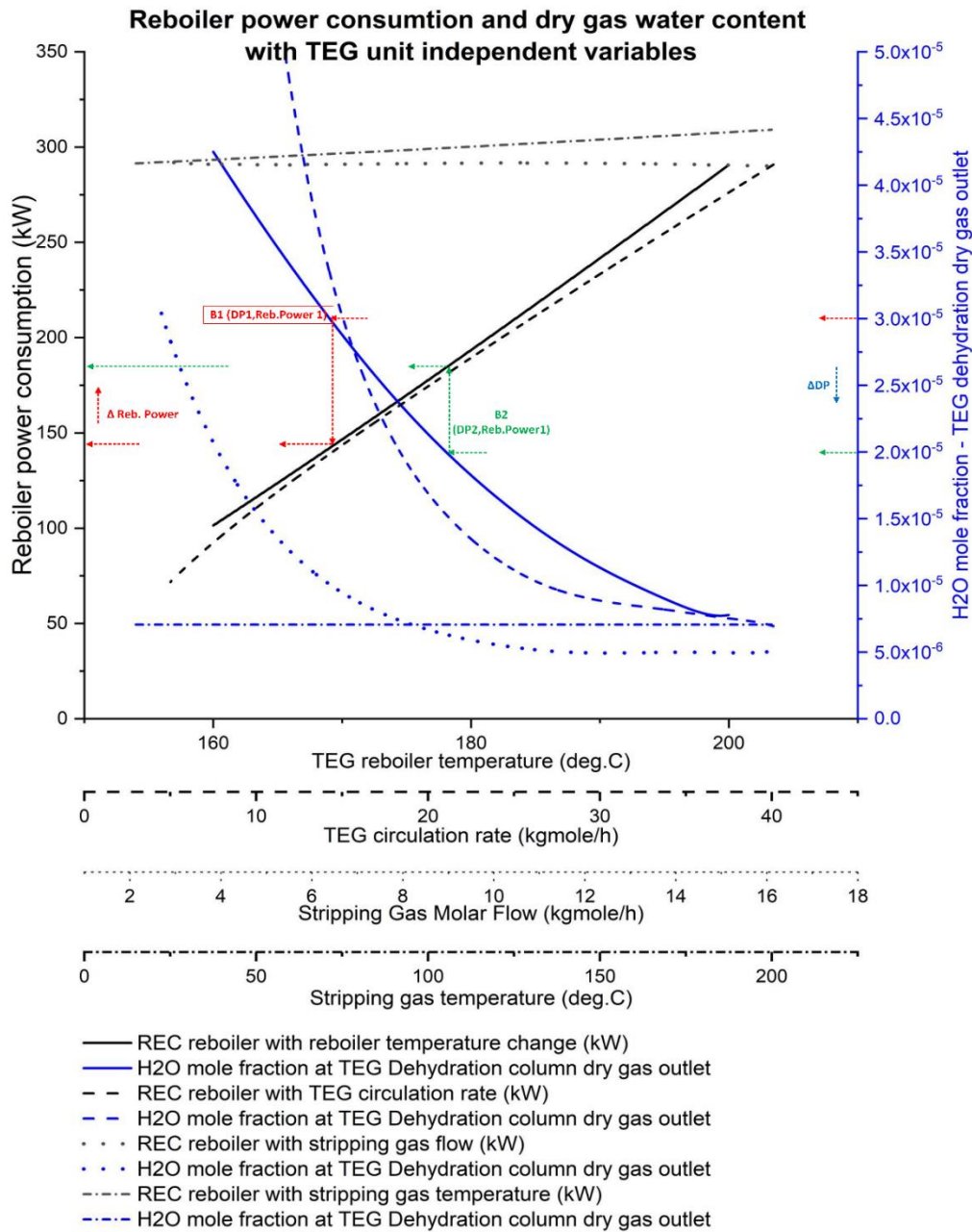
Figure 33 Reboiler power consumption and dry gas water content with TEG unit independent variables

TEG unit losses primarily consist of the condenser exhaust, and TEG flash drum released vapors. The greatest heat losses are accompanied by the stripping gas flow, Figure 34. The enthalpy changes at the exhaust due to dehydration degree, for a specific method, could be defined easily.
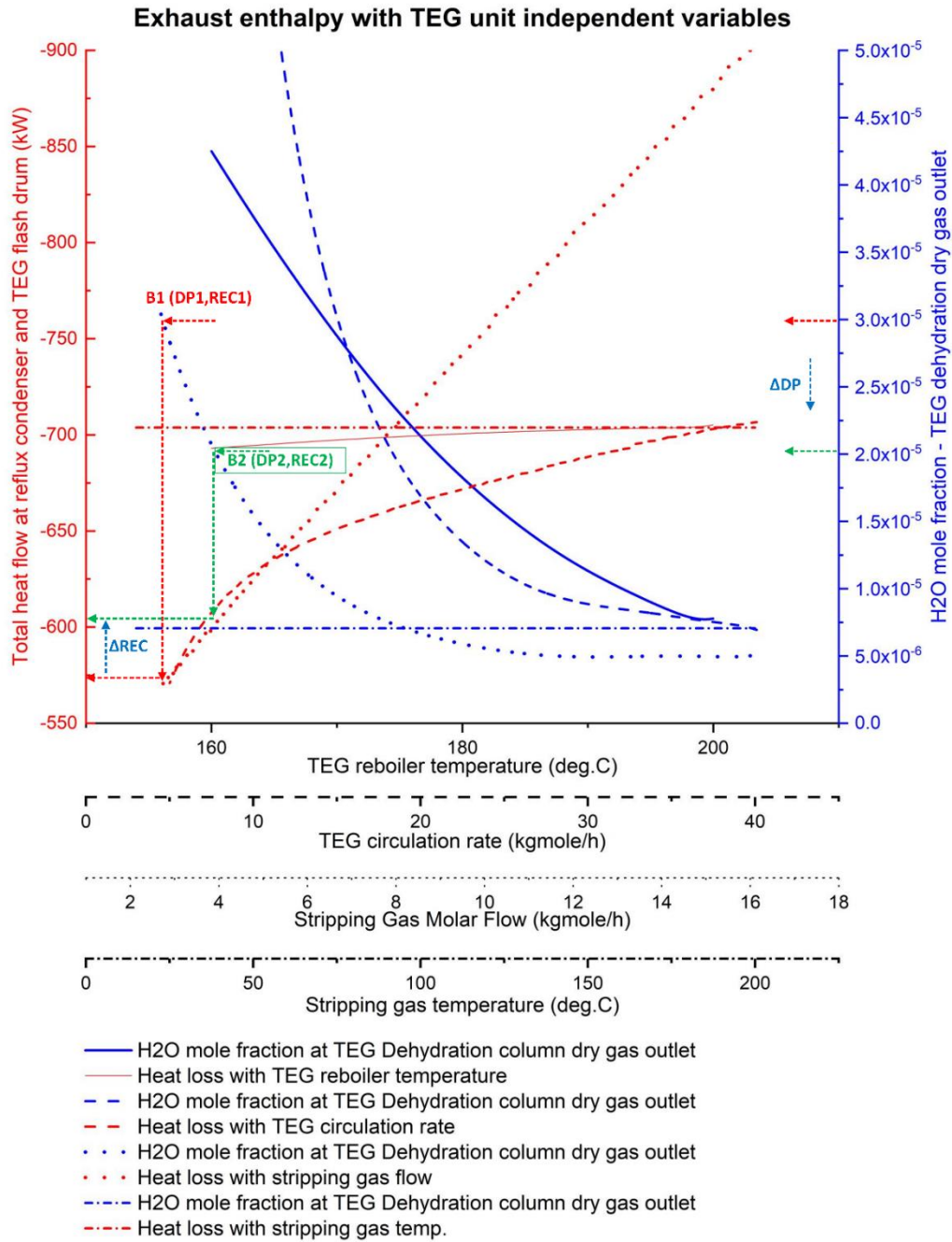
Figure 34 Exhaust enthalpy with TEG unit independent variables

- Referring to the current status simulation results in Table 10, it can be noticed that the gas is dehydrated **beyond** the dry gas specification (**dew point (DP) < -5 [°C] ~ 35E-6 [ppmv]**). Stream 4 at the outlet of the TEG contactor has a **DP – 25 [°C] ~ 7.5E-6 [ppmv]**.

The extensive dehydration results in extra energy consumption and emissions.

- The obtained graphs in Figure 33 and Figure 34, help define the proper operating points to meet the required specification with some safety margin, save the extra OPEX, and other consequences resulting from the overdesign in an operational method.

51

- Based on the obtained simulation results, it was highlighted the effects of three independent variables on the NGD process in the TEG unit, **the TEG circulation, stripping gas flow rate, and regeneration reboiler temperature**. These three scenarios represent **three routes to shift** the current operating point **DP$_0$** (-25 [^{o}C]~ 7.5 ~ 7.5E-6 [ppmv]) to the optimized one **DP$_x$** (-10 [^{o}C] ~ 25E-6 [ppmv]), where x=1,2,3 according to the used variable (1: TEG circulation, 2: stripping gas flow rate, and 3: regeneration reboiler temperature). Taking into consideration that this new point keeps a safety margin of (-5 [^{o}C]) below the dry gas specification.

- The shifting process from the current to the new operating point is represented for the case of reboiler temperature (x=3) in Figure 35, and Figure 36, which are extracted from Figure 33 and Figure 34.
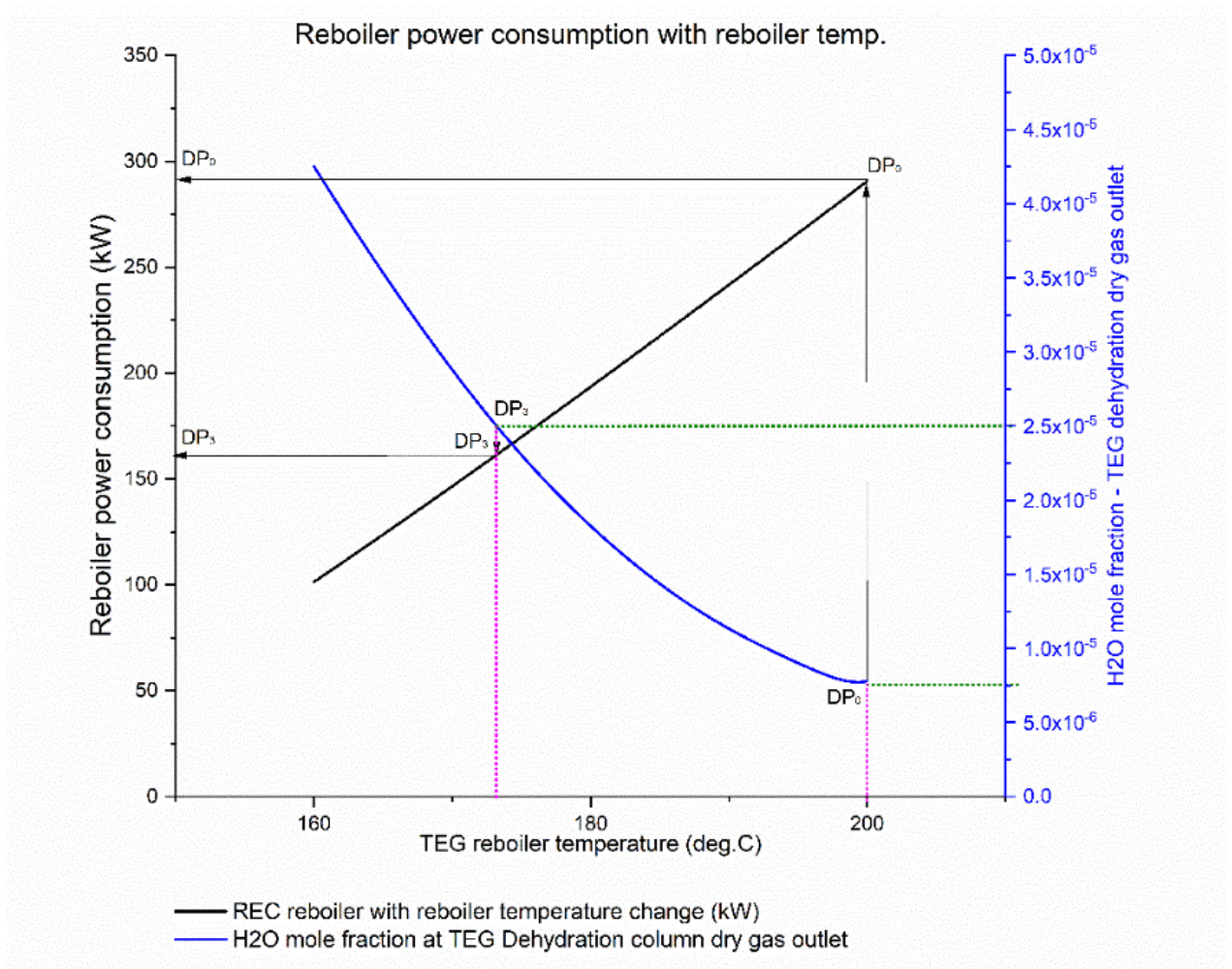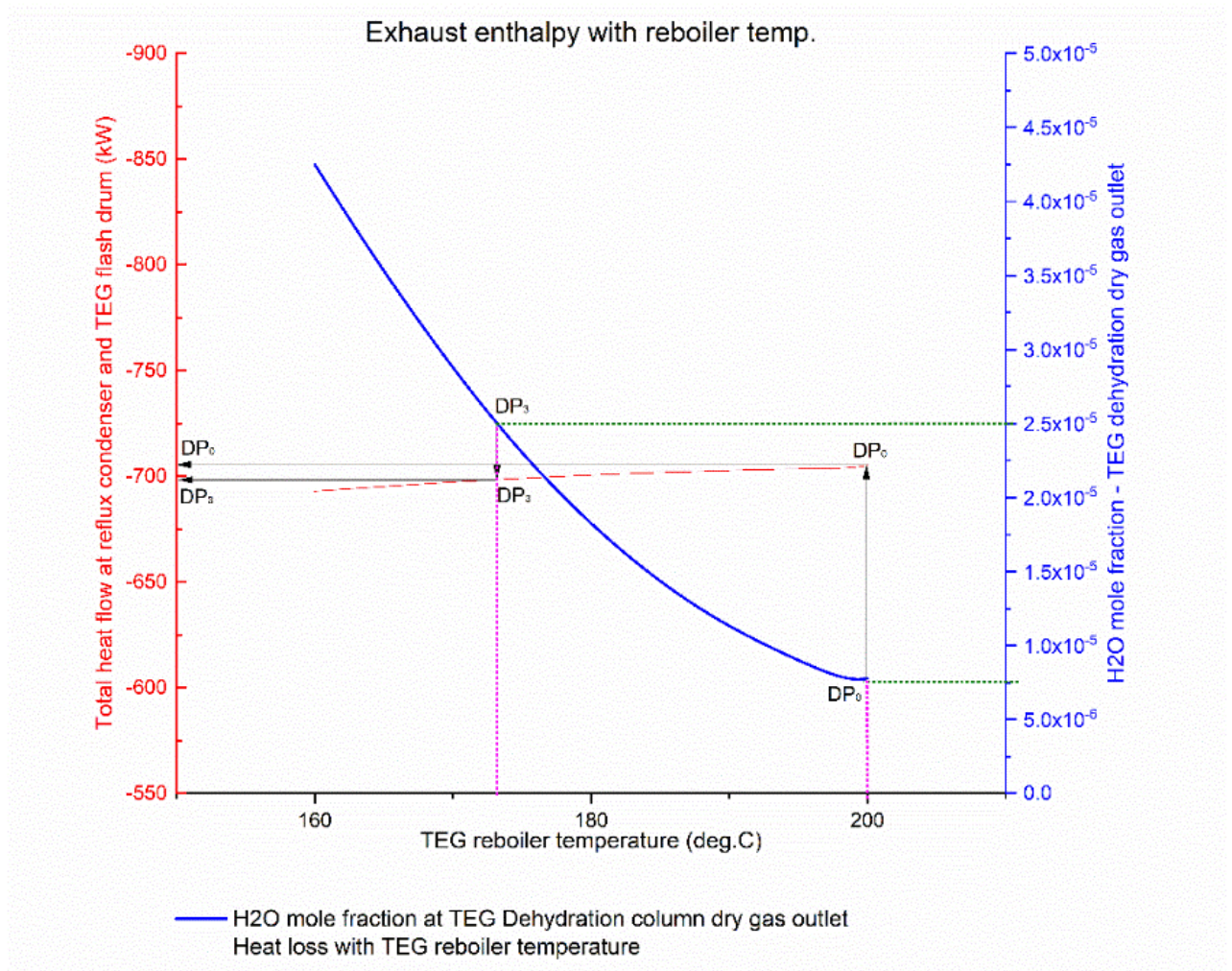

Figure 35 Reboiler power consumption with reboiler temperature change

Figure 36 Exhaust enthalpy with reboiler temperature change

The same procedure could be repeated for the other two scenarios (x=1: TEG circulation, x=2: stripping gas flow rate).

- The results of the intensified solution on the TEG unit are represented in Table 12.

Table 12 The intensified solutions impact the REC and emissions

Reference current state operating point **DP$_0$** TEG circulation =39.83 [kgmole/h] Stripping gas flow = 7.99 [kgmole/h] Reboiler temp.= 200 [ºC]			Scenario 1 TEG circulation @DP$_1$=16.46 [kgmole/h]		Scenario 2 Stripping gas flow @DP$_2$=3.3 [kgmole/h]		Scenario 3 Reboiler temp. @DP$_3$=172.8 [ºC]	
			DP$_1$	Δ_{0-1} %	**DP$_2$**	Δ_{0-2} %	**DP$_3$**	Δ_{0-3} %
Dew point (DP)	[ºC]	-25	-10	-	-10	-	-10	-
Water content	[ppmv]	7.5	25	-	25	-	25	-
Reboiler power consump.	[kW]	289	153	-136 47%	292	+3 1%	161	-128 44%
Enthalpy loss via venting	[kW]	703	652	-56 8%	586	-122 17%	699	-9 1%
Total REC*	[kW]	1015	828	-187 18%	901	-114 11%	883	-132 13%
BTEX (stream 25) **	[kgmole/h]	0.4813	0.1485	-0.3330 69%	0.4730	-0.0080 2%	0.4713	-0.0100 2%
CO$_2$ (stream 25 +29)	[kgmole/h]	0.8017	0.5073	-0.2944 37%	0.6517	-0.1653 21%	0.8053	+0.0036 >1%
TEG loss (stream 25 +29+7)	[kgmole/h]	0.0043	0.0040	-0.0003 7%	0.0029	-0.0014 35%	0.0042	-0.0001 2%

*: a fixed load of small motors = 23 kW
**: BTEX are calculated at the top of the reflux condenser outlet (stream 25) only. The BTEX emissions from the flash drum (stream 29) are negligible

The defined scenarios can reduce the REC by 11 %-18 %, BTEX emissions up to 69%, CO$_2$-(process) up to 37 %, and TEG loss reduction of about 35 %, without compromising the product specification, and due to no CAPEX impacts, these scenarios are valid for both, future design, and current TEG units in operation.

7.4.2. TEG unit upstream block intensification (Pre-cooling)

The intensified solution upstream of the TEG unit includes the integration of a pre-cooling unit which enhances the separation by direct cooling, the first method of dehydration.

There are three common types of cooling units, air coolers, water coolers, and propane chillers. From an operation point of view, the cooling unit type and design depend on the

location of the facility, compatibility with the annual ambient conditions, and the source of the cooling medium in case of water cooling. So, the air cooling systems are not feasible for a hot climate, considering that the selection and design inputs are considered in the worst case, which is the summertime, and the approximate approach temperature for design is 10 [ºC]. The cooling water systems are limited by the source of water, otherwise, a cooling system for the water cycle is needed, which means higher cost. The last method is the propane chilling process, which is feasible for all circumstances, but it has to be analyzed in terms of CAPEX and OPEX before deciding to integrate it with the TEG.

Regardless of the pre-cooling method type, the other limit for consideration is the wet gas temperature, for efficient operation of the TEG unit, which is 25-40 [ºC]. So cooling below 25 [ºC] affects the TEG efficiency.

Considering the most feasible solution, the propane chilling process, Figure 37, the cooling cycle consists mainly of an evaporator, a compressor, a condenser, an expander, and an after-cooler separator to separate the condensed water. The main energy consumer in this cycle is the compressor. The integrated scrubber of the TEG contactor itself is used as an after-cooler separator.

The suggested solution was simulated via the same proposed simulated-based method of REC that was used with the TEG, **and the only change is the process independent variable**, which is **Δt** in this case.
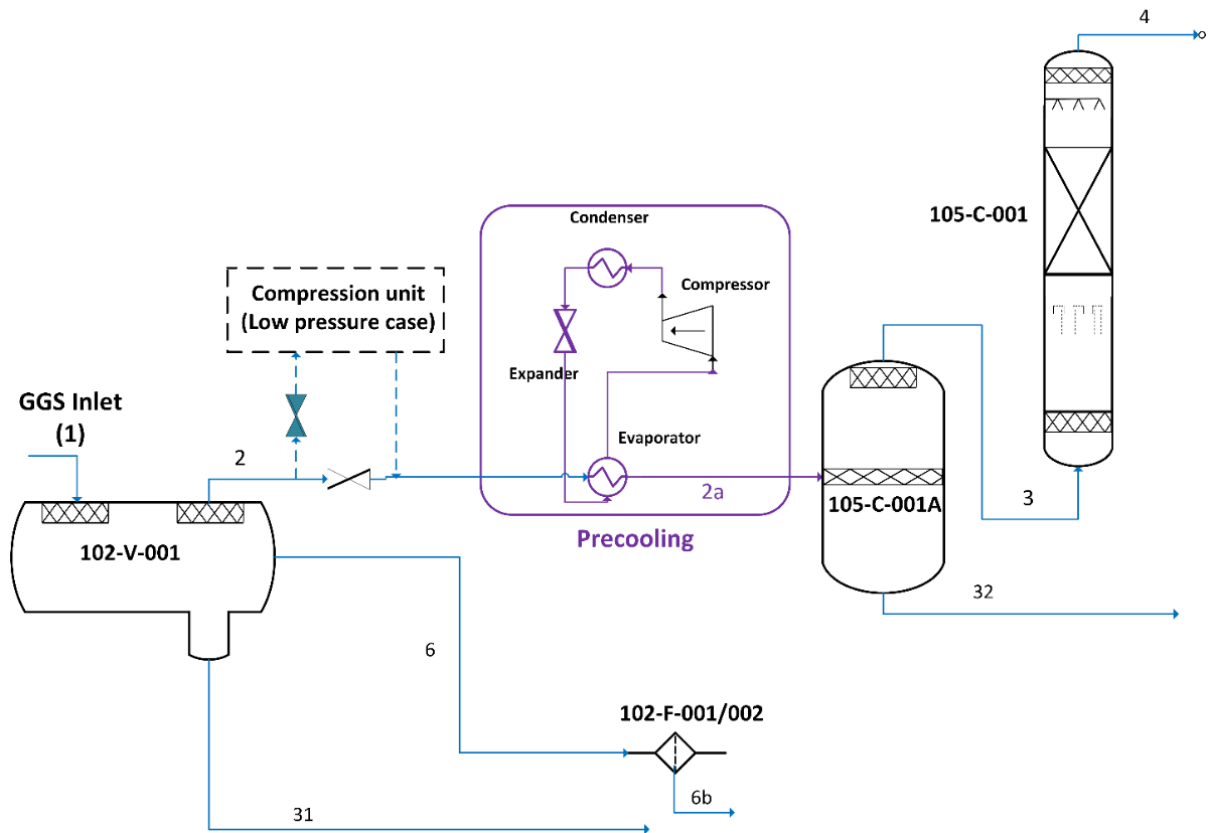

Figure 37 TEG unit pre-cooling solution

Figure 38 shows that the precooling helps separate the water over the precooling temperature range (41 to -15 [ºC]) and the water content mole fraction in the dry gas is reduced to a limit

value of approx. 3.7 x10^{-5}, but we are limited with the 25[^{o}C]. Otherwise, another solvent has to be used.

The REC including the power consumption of the compressor and the enthalpy loss at the separator drain increases considerably as far as we proceed with precooling. At the beginning of the range, the water mole fraction in the separator drain increases up to the pre-cooling temperature of 39.7 [^{o}C], then starts to decrease. This is attributed to the condensation of the HC and could be recognized with the increase of the heat flow at precooling separator liquid outlet.
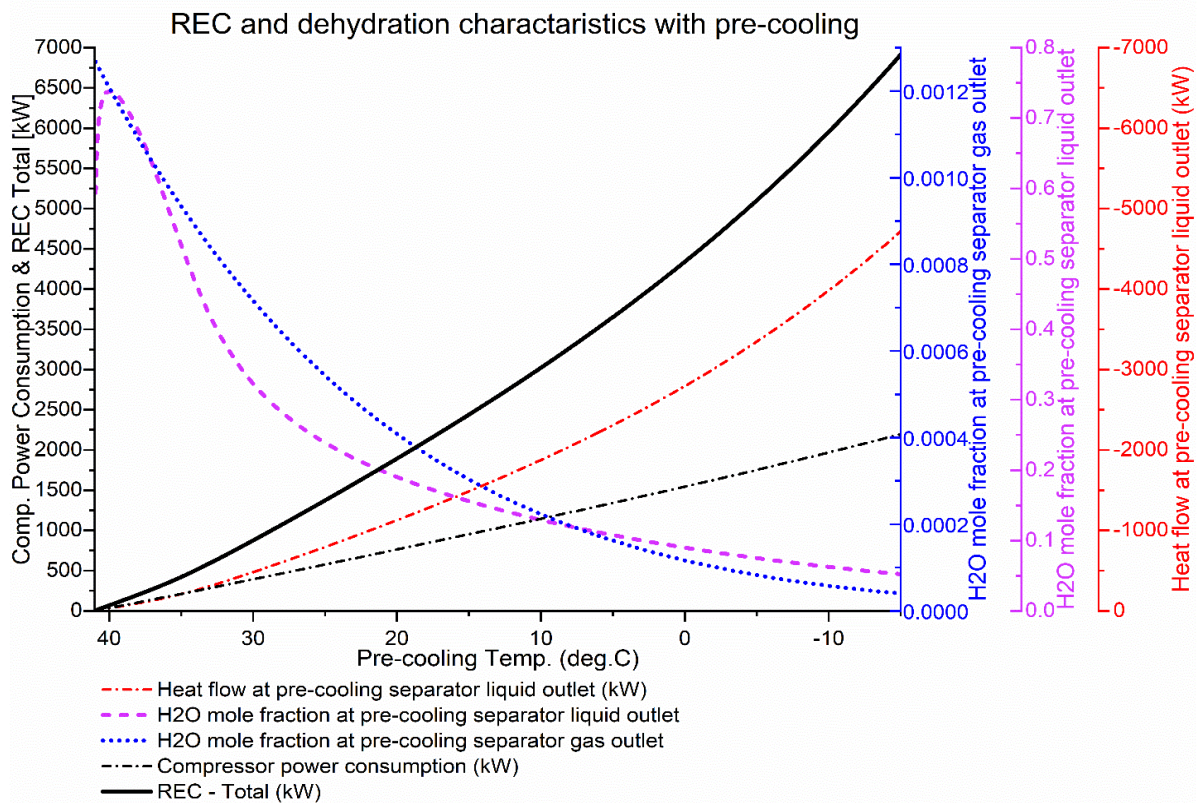
Figure 38 REC and dehydration characteristics with TEG precooling solution

7.4.3. TEG unit downstream block intensification (Mixing point)

The downstream block of the TEG is a simple mixer (Tee connection) that mixes the streams of dried gas and dried condensate to export them via one pipeline. The mixed condensate – gas pipeline itself is a common PI solution in the field of transportation. No intensification can be done at this mixer, but the simulation results of the current state drive further discussion about the input and output characteristics of the mixer. It was noticed that the dry gas downstream of the TEG unit has a dew point of -25.1 [^{o}C], while the dew point downstream of the mixer is only -5.5 [^{o}C]. The operator input data of the condensate stream is actual and obtained from the operator, and there was no simulation of the condensate processing itself, Figure 39.

The high content of water of 0.0023 [mole. fraction] within the condensate stream is clear, while this stream is supposed to be dry. This indicates a clear carry underwater with condensate. This water could be

A) free water with condensate, which means very fine droplets of water with the condensate, or

B) dissolved saturated water with the condensate,

C) or both.

The filter coalescer can remove the free water only, but it is not able to separate the dissolved water. To separate this water from condensate, a l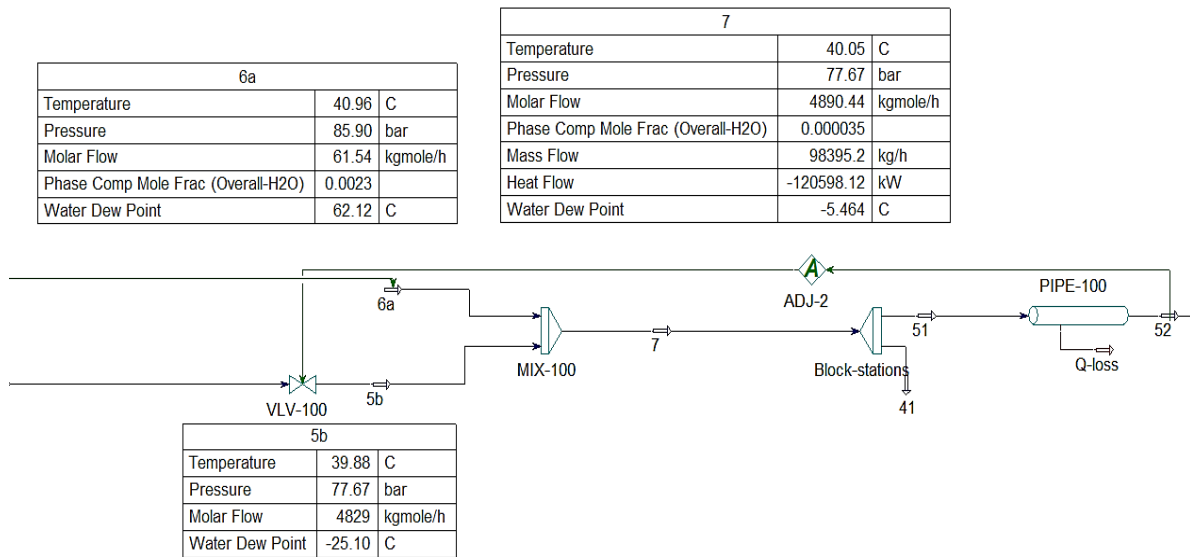iquid dehydrator is needed (adsorption catalyst), which is of considerable high CAPEX and OPEX due to the regeneration process requirement.

In such a scenario the best is to keep the TEG unit with extended dehydration scope beyond the required dry gas specification (dew point), and the extra dried gas will be sufficient to overcome the compromising of the specification, which is caused by the water in the condensate. The other solution is to optimize the design of the separator internally to achieve better settling for the water and increase the efficiency of the separator.

The water in condensate problem is beyond the scope of this paper, and it is a complicated standalone topic that has to be solved separately.

7		
Temperature	40.05	C
Pressure	77.67	bar
Molar Flow	4890.44	kgmole/h
Phase Comp Mole Frac (Overall-H2O)	0.000035	
Mass Flow	98395.2	kg/h
Heat Flow	-120598.12	kW
Water Dew Point	-5.464	C

6a		
Temperature	40.96	C
Pressure	85.90	bar
Molar Flow	61.54	kgmole/h
Phase Comp Mole Frac (Overall-H2O)	0.0023	
Water Dew Point	62.12	C

5b		
Temperature	39.88	C
Pressure	77.67	bar
Molar Flow	4829	kgmole/h
Water Dew Point	-25.10	C

Figure 39 Dry gas and condensate mixer

To investigate this finding, further data about the composition of streams upstream and downstream of the condensate filtration were obtained. The mole fraction of water within the stream from the separator to the condensate filtration is 0.0630, and downstream the condensate filter coalescer is 0.0023. This means a filter efficiency of 96.3%, which is a high-quality filtration. So the solution could be one of the following:

- Design modifications on the separator unit to reduce the transferred water with the condensate stream.

- Replacing the condensate filter element with higher quality membrane filters.

- Implementation of some demulsification methods to reduce the water content within the condensate stream.

Chapter 8 Conclusions

The developed systematic method for process intensification of the natural gas dehydration indicated PI potentials on two levels:

- the core unit or core node (TEG unit), and

- the complex NGD unit with the integrated upstream and downstream.

The PI in the core unit (TEG unit) is achievable via three scenarios, each of them deals with one independent variable in the TEG unit, (1) the TEG circulation, (2) stripping gas flow rate, and (3) regeneration reboiler temperature. These three scenarios represent three routes to shift the current operating point to the optimized one.

This upgraded methodology has the following unique characteristics:

- It is a systematic simulated-based, that can be implemented easily and efficiently.

- The proposed scenarios are applicable for future design, as well as the existing units in operation.

- In the case of implementing this method for achieving the PI in the existing TEG units, no CAPEX have to be considered.

- This method provides a guide to the operator to approach the PI in an existing dehydration unit, by applying operational procedures.

- It is ready to extend and define the optimum balance between the three scenarios.

- The proposed method with the same approach applies to the extended unit, including the upstream and downstream adjacent units.

- The results of the extended unit case study indicate the benefits that could be obtained or losses that could happen, by implementing any dehydration solution upstream or downstream of the core unit.

- The implementation of the proposed method can be extended beyond the TEG unit and its adjacent upstream and downstream units. The same method applies to the entire process of dehydration, from separation to the LPG fractionation. The only needed input is the independent variable for each studied unit.

It is recommended to extend this work by studying the adsorption unit as well, and the independent variables, in this case, would be the water content on the inlet of the unit, and the stream temperature, pressure, and flow rate. Also, the condensate filtration and dehydration loop should be further investigated as it is an essential connection to the node of mixing, and affects the entire dehydration process in terms of PI drivers.

As could be noticed, our newly developed method can be considered as a general upgraded PI methodology useful for various types of intensification in the O&G. It has a wide range of applicability, as it is related to the REC that exists in any part of the process, and all the processes. So the same method can be used for approaching the PI in any other process by incorporating the specific independent variables of the studied process.

Nomenclature

Abbreviation	Definition
bbl/d	Barrels per Day
BTEX	Benzene, toluene, ethylbenzene, and xylene
Btu/bbl	British Thermal Unit per Barrel
Btu/MMSCF	British Thermal Unit/Million std. Cubic Feet
Btu/Sm^3	British Thermal Unit per std. Cubic Meter
CAPEX	Capital Expenses
CFD	Computational Fluid Dynamics
CH$_4$	Methane
CNG	Compressed Natural Gas
CO	Carbon Monoxide
CO$_2$	Carbon Dioxid
DEG	Diethylene glycol
DOE	Department of Energy
DP	Dew Point
DWC	Dividing Wall Column
EG	Ethylene glycol
EOS	Equation of state
EPA	Environmental Protection Agency
EU	European Union
GAS	Gas Antisolvent Precipitation
GGS	Gas gathering station
GHG	Greenhouse Gas Emissions

GLMR	Gas-Liquid Microreactors
GLSMR	Gas-Liquid-Solid Microreactors
GTL	Gas to Liquids
GTP	Gas treatment plant
HAPs	Hazardous air pollutants
Hazop	Hazard & Operability Analysis
HC	Hydrocarbons
Hg	Mercury
HIDIC	Internally Heat Integrated Distillation Columns
HIGEE	High-Gravity Technology
HNS	Hybrid Nonreactive Separation
HNS	Hybrid Nonreactive Separation
HSE	Health, Safety, and Environment
IEA	International Energy Agency
Inline	Inline Separation and Dehydration
IPCC	Intergovernmental Panel on Climate Change
ISI	Inherent Safety Index
kt/y	Kilotonnes per year
LNG	liquefied natural gas
LPG	Liquefied Petroleum Gas
MBR	Membrane Bio-Reactors
MECS	Manufacturing Energy Consumption Survey
MEG	Monoethylen glycol
MFBR	Membrane Fluidized Bed Reactors

MM	Million
MMSCFD	Million std. Cubic Feet
MMt CO2e	Million Ton Carbon Dioxide equivalent
MNS	Membrane-Assisted Nonreactive Separation
MR	Microreactors
MRD	Membrane Reactive Distillation
MRS	Membrane-Assisted Reactive Separation
MS	Membrane-Assisted Separation
Mt/y	Milliontonnes per year
Mtoe	Million Tonnes Of Oil Equivalent
N_2	Nitrogen
NG	Natural gas
NGD	Natural gas dehydration
NGL	Natural gas liquid
NH_3	Ammonia
NO_x	Nitrogen oxides
O&G	Oil and Gas
OFR	Oscillatory Flow Reactors
OPEX	Operating Expenses
Pb	Lead
PGSS	Particles From Gas Saturated Solutions
PI	Process Intensification
PIIS	Prototype Index for Inherent Safety
PO	Periodic Operations

PPC	Pulsed Packed Columns
ppm	Part per million
PRC	Processing Capacity
PSA	Pressure Swing Adsorption
PSAR	Pressure Swing Adsorptive Reactor
PSE	Process Systems Engineering
QRA	Quantitative Risk Analysis
R	PI Reaction Technology
RA	Reactive Adsorption
RD	Reactive Distillation
REC	Rated Energy Consumption
$REC_{Dehyd.}$	Dehydration Rated Energy Consumption
RESS	Rapid Expansion Of Supercritical Solution
RFR	Reverse Flow Reactor
RO	Reverse Osmosis
RPB	Rotating Packed-Bed
RPC	Reciprocating Plate Columns
RS	PI Reactive Separation Technology
S	PI Separation Technology
SAS	Supercritical Anti-Solvent
SCR	Structured Catalytic Reactors
scRPE	Supercritical Reverse-Phase Evaporation Method
SEC	Specific Energy Consumption
SFEE	Supercritical Fluid Extraction Of Emulsions

SIL	Safety Integrity Level
SMB	Periodic Operations -Simulated Moving Bed
SMBR	Simulated Moving Bed Reactor
SMR	Static Mixer Reactor
SO_2	Sulfur Dioxide
SO_x	Sulfur Oxides
T_4EG	Tetraethylene glycol
TBtu	Trillion British Thermal Unit
TEG	Triethylene glycol
$TCO_2 e$	Ton Carbon Dioxide equivalent
TSA	Temperature Swing Adsorption
US	United States
VBCG	Variable-Based Causal Graph
VOCs	Volatile Organic Compounds

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Highlights of Professional and Academic Activity by Author

The author is an MSc. Mechanical Power Engineer with technical experience of 13 years in O&G and Power plants. Experienced in execution, construction, commissioning, and operation of energy projects and facilities, including a very wide range of static and rotating equipment.

Work History

05. 2018 – UP TO DATE	**DOOSAN-SKODA POWER, PLZEŇ, CZECH REP. (Steam Turbine)** **Technical Project Manager**
08. 2015 – 10.2016	**AL-BAATH UNIVERSITY, HOMS, SYRIA (Education)** **Assistant Lecturer of Thermodynamics & Energy Technologies**
10. 2010 – 08. 2015	**SUNCOR ENERGY- EBCO GAS TREATMENT PLANT, SYR(O&G)** **Mechanical Maintenance Supervisor/Engineer**
02.2009 – 10.2010	**SUNCOR ENERGY (PETRO CANADA), SYRIA (O&G)** **Mech. Construction, Installation & Commissioning Engineer**
04.2006 – 02.2009	**CONOCOPHILLIPS DEZ GAS PLANT (CSDGL), SYR. (O&G)** MECHANICAL MAINTENANCE ENGINEER
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Education

09.2017 – Up to date	Faculty of Mechanical Engineering - Brno University of Technology "VUT", Czech Rep. **Ph.D. student in Design and Process Engineering** ▪ Simulation of heat and mass transport in porous media ▪ Process Intensification in the Oil and Gas industry
10.2010 – 03.2013	Faculty of Mechanical and Electrical Engineering - Al-Baath University, Homs, Syria **Master of Science (MSc.) in Mech. Power Eng. – Fluid Dynamics Recognized by Czech Technical University ČVUT - Prague** ▪ Study of acidic multiphase flows characteristics in raw gas transportation pipelines
09.2006 – 08.2007	Faculty of Mechanical and Electrical Engineering - Al-Baath University, Homs, Syria **Diploma in Renewable Energies** ▪ Applications of Hydrogen Fuel Cell Technology.
10.2000 – 08.2005	Faculty of Mechanical and Electrical Engineering - Al-Baath University, Homs, Syria **BSc. Degree in Mechanical Power Engineering**

Publications

The author performed a review of the design considerations for internal electrochemical corrosion caused by the presence of acid gases in gas pipelines and the effect of various factors including the flow pattern.

> **Abdulrahman, I.**, 2013. Characteristics of Flows & Reliability of Gas Pipelines with Flows of an Acidic Nature. **JOURNAL OF AL-BAATH UNIVERSITY,** Paper ref. 3789-24/10/2013. ISSN-1022-467X
> https://drive.google.com/drive/folders/17-brno04Sr_Ib538Ef2T1ObW4OPE-bgV (File 7 - pages 255-284)

The author investigated the irregularity of electrochemical corrosion of the internal surface of the gas/condensate pipeline, and the effect of flow pattern on this phenomenon.

> **Abdulrahman, I.**, 2013. Analysis of Irregular Internal Electrochemical Corrosion in Multiphase Pipelines. Journal of Al-Baath University, Paper ref. 3789-24/10/2013. ISSN-1022-467X
> https://drive.google.com/drive/folders/17-brno04Sr_Ib538Ef2T1ObW4OPE-bgV (File 10 - pages 175-206)

The authors presented an approach for enhancing the PI implementation in the O&G via a novel technical manner. The proposed framework adopts a process-based PI tools classification and suggests the rated energy as a PI unified driver of implementation. Conditioning the PI concepts via adopting a unified driver as an assessment parameter will facilitate the design and evaluation which would lead to an implementation enhancement.

> **Abdulrahman, I.,** Máša, V., Teng, S.Y., Process intensification in the oil and gas industry: A technological framework, Chem. Eng. Process. - Process Intensif. 159 (2021). https://doi.org/10.1016/j.cep.2020.108208.

The authors presented a case study on absorption dehydration using triethylene glycol (TEG). A process model was built and used to evaluate REC. The outcome of this work represents an important foundation for future PI within a more complex dehydration process considering also cooling and condensation upstream of the TEG unit and adsorption as a subsequent step.

> Miklas, V., **Abdulrahman, I.**, Máša, V., Process Intensification of Triethylene Glycol Dehydration Units with Respect to Rated Energy Consumption, Emissions and Product Quality, Chem. Eng. Trans. 88 (2021) 925-930 SE-Research Articles. https://doi.org/10.3303/CET2188154.

The authors presented a systematic simulation-based method that integrates the rated energy consumption (REC) with the dry gas water content specification. An existing gas plant is simulated via a powerful process simulator as a case study. The method defines three scenarios to effectively intensify the process.

> **Abdulrahman, I.,** Máša, V., Miklas, V., 2022, A systematic method for process intensification of the natural gas dehydration with respect to the rated energy consumption, Chem. Eng. Process. Process Intensification, ISSN 0255-2701. (Under Review)