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# Thermodynamic analysis of an upstream petroleum plant operated on a mature field

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**Abstract**

Oil and gas processing on offshore platforms operates under changing boundary conditions over a field lifespan, as the hydrocarbon production declines and the water extraction increases. In this paper, the processing plant of the Draugen platform is evaluated by performing an energy and exergy analysis. This facility exploits an end-life oilfield and runs at conditions deviating significantly from its optimal operating specifications. Two different operating modes were assessed, and process models were developed using the simulation tools Aspen Plus® and Aspen HYSYS®, based on measured and reconciliated process data. The total energy demand is moderately sensitive to daily and monthly variations: it ranges between 22 and 30 MW, of which 18–26 MW and about 3–4 MW are in electrical and thermal energy forms. The greatest exergy destruction takes place in the gas treatment (51%), recompression (12%) and production manifold (10%) modules. The separation work performed on this platform is greater than in similar facilities because of higher propane and water fractions of the well-streams. These findings emphasise the differences between *peak* and *end-life* productions: they suggest (i) to set focus on processes including gas expansion and compression, (ii) to investigate possibilities for an improved energy integration, and (iii) to consider and evaluate alternative system designs.

*Keywords:* Energy, Exergy, Efficiency, Petroleum, Oil and gas platform, End-life

## 1. Introduction

Reservoir fluids from oil and gas fields are complex multiphase mixtures, containing a large range of chemical compounds, from light to heavy hydrocarbons, contaminants such as carbon dioxide, and subsurface water. The hydrocarbon fraction must be separated and purified to be further transported, while the impurities must be removed and the water phase cleaned. The processing plant of an offshore platform should meet these criteria, and the design phase should consider parameters such as the reservoir fluid composition, flow rate, pressure and temperature [1–4]. The on-site processes are generally designed for near-peak hydrocarbon production and have a lower performance at the end life of an oilfield, when the oil and gas production decreases and the water production rises. The specific environmental impact and energy intensity of the processing facility will increase, because large amounts of power are consumed to enhance oil recovery by water injection, gas injection and/or gas lift [5–7].

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Nowadays, the performance of an offshore platform is measured by indicators related to the energy demand and environmental impact of the processing plant. Svalheim and King [8] mentioned: (i) the *energy efficiency* of the platform, defined as the ratio of the energy exported to the shore to the energy entering the processing plant, (ii) the *energy intensity*, defined as the ratio of the energy used on-site to the energy exported onshore, (iii) the *specific power consumption* and (iv) the *specific $CO_2$-emissions*. The authors emphasised the limitations of these indicators: a comparison of different plants with these metrics may be misleading, since different oilfields present different characteristics [9]. These metrics are based on conventional energy analyses, which yield information on the energy inputs needed to produce a given product [10]. As stated by the 1st Law of Thermodynamics, energy is conserved and cannot be destroyed. An energy analysis indicates therefore changes from one form of energy to another and allows the tracing of energy flows throughout a given system.

Unlike energy, exergy is destroyed via conversion technologies and losses in real processes because of entropy generation. An exergy accounting reveals the locations and extents of the thermodynamic irreversibilities of the system under study [11,12], and the amount of exergy destroyed throughout successive processes accounts for the additional fuel use because of system imperfections [13–16]. Rivero [17] argued that the application of an exergy analysis to petrochemical systems would provide valuable insights. Voldsund et al. [18] suggested the use of exergy analysis as a tool for performance benchmarking and evaluation of offshore platforms.

A few studies on energy, exergy and offshore processes exist. Oliveira and Van Hombeeck [19] analysed a Brazilian plant: the gas compression and the separation processes were the main energy users and exergy consumers, and both systems were associated with significant exergy destruction. Voldsund et al. [18,20,21] studied a North Sea oil platform. They showed that the greatest thermodynamic irreversibilities were associated with processes where large changes in pressure took place. Nguyen et al. [22] conducted an analysis of generic North Sea oil platforms. In *most cases*, the gas compression step was the most power-consuming and exergy-destroying process, but one particular case showed different characteristics. Their work indicated that the thermodynamic performance of an oil and gas platform was optimal with low reservoir fluid contents of gas and water.

The literature appears to contain little on the application of energy *and* exergy analyses to offshore *processing plants*, and none with a special focus on their thermodynamic performances when the oilfield is mature. The goal of this study is to help close these gaps. This work aims at (i) quantifying, in terms of energy and exergy, the several transformations taking place within the processing plant of a specific oil and gas platform, (ii) comparing two types of production days, and (iii) investigating the effects of end-life boundary conditions. It is part of a larger research project dealing with the optimisation of electrical energy production and consumption on offshore platforms and builds on earlier work conducted by the same authors [22,23].

The facility investigated in this work is similar to other plants in the North Sea [7,24,25], with two main differences: (i) the oilfield is characterised by a high propane content of the reservoir fluid and a small gas-to-oil ratio, and (ii) oil is not exported continuously via pipelines but in batch operation with shuttle tankers and intermediate storage in on-site tanks. The oil recovery rate is expected to reach 65–75%, which is much higher than the typical rate of 45–50% for Norwegian fields. This has encouraged an extended exploitation of this petroleum field, despite the high water cut of the feed. However, the processing plant is still run within its design range, although far from its optimum conditions. This is the first study on the thermodynamic performance of this particular plant, and on the performance assessment of offshore facilities operated during an extended production life period.

The present work was structured as follows:

- development and calibration of the processing plant model based on measured process data;
- evaluation of the material, energy and exergy flows throughout the processing system;
- analysis of the energy and exergy consumption patterns and of the plant inefficiencies.

The processing plant system and model are described with the methods of analysis in Section 2. Section

3 reports the results obtained, which are discussed and critically reviewed in Section 4. Concluding remarks are outlined in Section 5.

## 2. Methods

### *2.1. Case study*

#### *2.1.1. System overview*

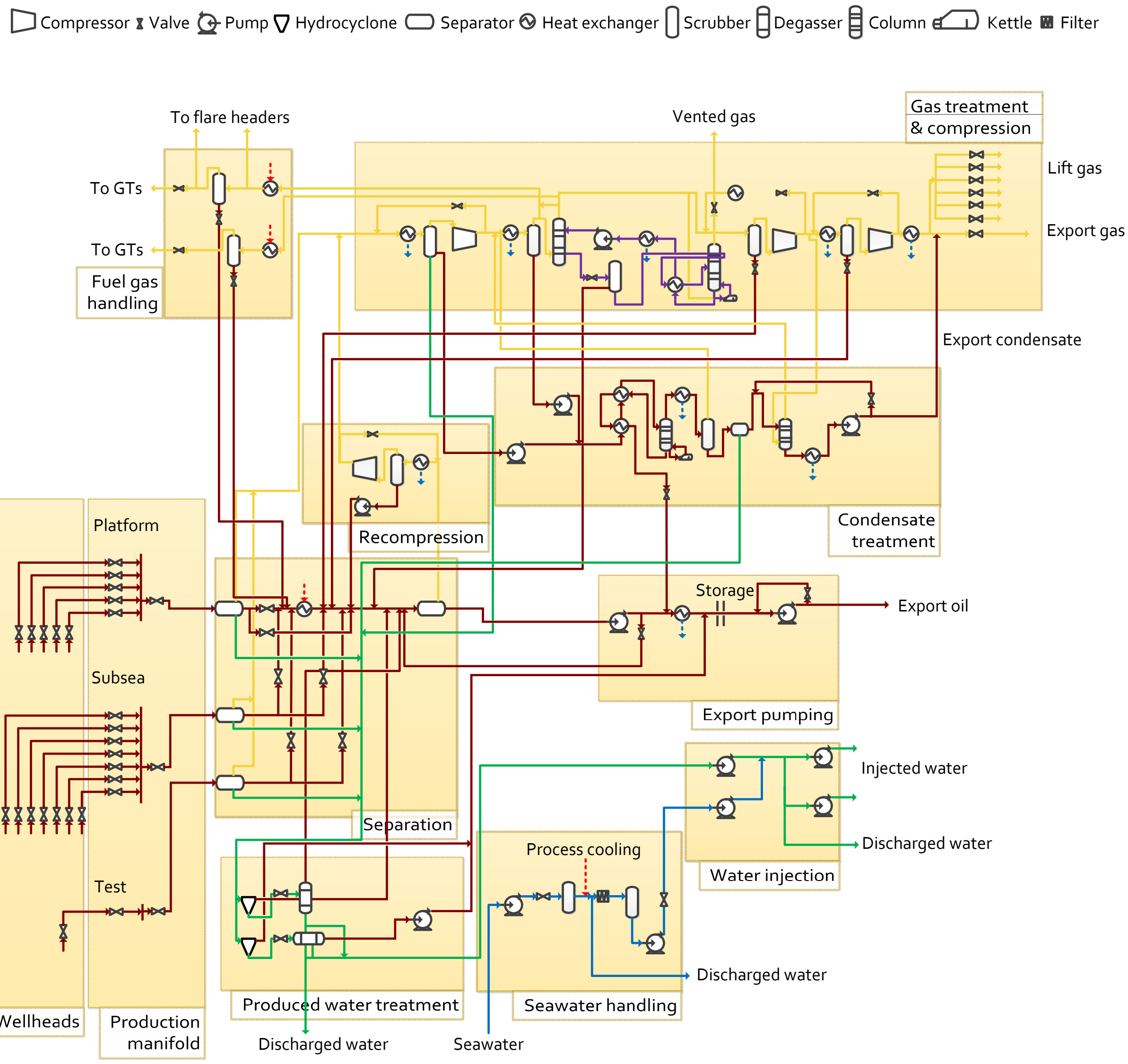


Figure 1: Process flow diagram of the processing plant of the Norwegian Sea offshore platform investigated in this work, based on data available at [26–31] and input by Norske Shell A/S. Reservoir fluid, oil and condensate streams are indicated in brown, gas streams in orange, produced water in green, seawater in blue and glycol in indigo. For ease of reading, only the most important recycling loops are drawn. Control valves, connections to pilot flares, and storage units are not presented. Addition of chemicals such as biocides and methanol is not indicated. The hydrocyclones of the produced water handling system are merged into two, the scrubbers of the fuel gas treatment into one, and the oil and condensate pumps are shown as a single pump per two pumps run in parallel.

The oil field investigated in this study (Figure 1) is located in the Norwegian Continental Shelf region: the construction of the platform was finalised in 1993 and the oil production started the same year. This facility is characterised by the seven oil tanks located at the base of the structure and operates on two different deposits. The aim has been to maximise the oil production: associated gas was used for petroleum lift, and gas export only started in 2000 [31]. Water injection started in 1994 to sustain a high reservoir pressure: seawater was filtered, treated by addition of chemicals and injected into the reservoir at high pressure. Water production started in 1998 and has drastically increased, reaching a water cut (water fraction in the reservoir fluid) above 90% in the last years. The oil production peak was reached between 2000 and 2002 [28,32,33].

At present, stabilised oil is stored in the tanks in the base of the facility, and exported to the shore from once every single week to once every month via a floating load buoy. Produced gas is (i) mainly used for gas lift, i.e. is injected into the oil wells to ease the reservoir fluid lift and maximise oil production, or (ii) is transported to the coast via the Åsgard pipeline network, along with condensate. Gas injection was considered, and dedicated wells were drilled. There is a foreseen fuel gas deficiency at Draugen, and the diesel system is likely to be used at times in continuous operation, while it was designed as a back-up solution [32,33].

*2.1.2. System description*

The structural design and general building blocks of oil and gas processing plants stay similar across platforms [7,24,25], but differences in the detailed design exist from one processing plant to another, depending on the reservoir characteristics (e.g. temperature and pressure), reservoir fluid properties (e.g. chemical composition and thermophysical properties), technical requirements (e.g. need for dehydration and compression) and operating strategies (e.g. gas export and water injection). The field produces oil from two reservoirs located at the same depth, which is extracted via seven platform- and six subsea-wells connected to the platform. The initial hydrostatic pressure and reservoir temperature were about 165 bar and 71 °C [29].

The field and platform differ from other oil and gas fields and facilities: (i) the water cut exceeds 75% (mass), illustrating the mature properties of this oilfield; (ii) gas condensate is treated in a dedicated process section, where there are a fractionation and a dehydration column; (iii) the processing plant is run in different operating modes: oil may be loaded for export to the shore, and seawater and produced water may be injected into the reservoir. This study focuses on two operating modes: in the first one, oil is sent for storage, while in the second one, oil is loaded and exported. Both cases consider injection of seawater in the reservoir.

**Production manifold**

The well-streams are gathered and transported to the main production facility via a network of pipelines and manifolds. They are mixed and de-pressurised by chokes, which are set to control both flows and pressures. A fraction of the well streams, usually from a dedicated well, is processed through a test manifold and a test separator, to allow further flow analysis [1]. The other fraction is sent to the normal production manifolds and is normally processed in two three-phase separators run in parallel.

Four platform wells are connected to the same 3-phase separator at the 1st stage, while the seven subsea wells are connected to another 3-phase separator, as well at the 1st stage. There are two other platform production wells that can be connected to the test manifold and separator, or to any of the two regular 3-phase separators. Some wells only extract gas, some extract mainly oil and water, and a few are not operated at the moment. The water cut is slightly higher for platform wells, resulting in a greater water production from the 1st stage separator connected to the platform wells. The pressures at the inlet of the production manifold range between 8 and 11 bar for the platform wells and between 38 and 46 bar for the subsea wells. The average temperatures are about 63 °C, with a deviation of ± 5 °C.

**Separation**

Oil, gas and water are separated by gravity in two stages, operated at two different pressure and temperature levels [1]. The 1st stage consists of two three-phase separators run in parallel and at similar operating conditions. The temperatures and pressures are close to the outlet conditions of the production manifolds dedicated to the platform wells. The 2nd stage consists of a two-phase separator operated at about 65–75 °C and 1.6–1.8 bar [26]. The pressure is decreased by throttling valves, the oil streams from the three 1st stage separators are mixed, and their temperature is increased in a crude heater from the 1st to the 2nd stage, easing recovery of light hydrocarbons [26].

**Oil export**

The crude oil leaving the separation system enters a storage and pumping section. It is first mixed with condensate recycled from other sections of the processing plant. It is then cooled to a temperature close to 30 °C by seawater and placed in tanks located at the bottom of the sea. Stabilised oil is pumped later for export onshore with shuttle tankers, and additional power is required during the loading periods [32].

**Gas recompression and treatment**

The gas recovered from the 2nd separation stage is sent to the recompression system, where it is cooled to about 30 °C, sent to a scrubber, where condensate and water droplets are removed, and re-compressed to the pressure of the 1st separation stage. It is then mixed with the gas recovered from the 1st separation stage and enters the gas treatment and compression train. This system is divided in three stages operated at 19–23, 57–60 and 179–189 bar [26]. Each stage includes a cooler operated with seawater, a scrubber to separate liquid droplets from the gaseous phase, and a compressor, as in the re-compression process. The 2nd compression stage also includes a dehydration process to prevent corrosion issues and hydrate formation in the gas pipelines.

In a packed contactor, wet gas flows counter-currently to liquid and dry TEG (triethylene glycol) that captures water by physical absorption. The glycol-water mixture is then depressurised, flashed and heated before entering a desorption column where water and glycol are separated. Liquid TEG exits at the bottom of the desorber while water vapour exits at the top, along with hydrocarbon impurities. Dry gas from the 2nd stage of the compression train is introduced in the desorber to increase the glycol purity to about 99.5% on a mass basis. The temperature of the bottom stage is controlled by a reboiler and is about 204.5 °C, which is the highest temperature level of the processing plant. The temperature of the top stage is controlled by a reflux condenser and is about 98.5 °C to prevent excessive glycol losses with the vented gases [1,34]. Regenerated glycol is pumped to the pressure of the 2nd compression stage and is recycled to the absorber. Most high-pressure dry gas is used for gas lift, while the remaining is sent onshore via pipelines.

**Fuel gas handling**

A certain fraction of the wet and dry gases from the 2nd separation stage is processed through the fuel gas system: it is heated in electric heaters, scrubbed in three parallel trains and is combusted in five gas turbines installed on-site for power generation. Three are used to generate the power required by the gas compressors and other process utilities, two are dedicated to the water injection pumps, and one is at present in standby for backup.

**Condensate treatment**

Condensate recovered from the 1st and 2nd stages of the gas compression system is not sent to the crude separators, as done on most offshore platforms, but is handled in a separate process to achieve the desired separation between light and heavy hydrocarbons. A condensate treatment section was integrated on this facility because the feed contains a high fraction of propane. This process avoids recycling of propane and butane between the separation and gas treatment sections, and reduces significantly the power demand of the recompression train. The recovered condensate is heated by integration with other process streams and sent to a fractionation column. At the difference of conventional scrubbers, which operate without any heat addition, this condensate scrubber is equipped with a reboiler. The heat input is regulated to control the temperature at the 15th stage of the column, which is about 158 °C [26].

Liquid hydrocarbons exiting this column are mixed with the crude oil entering the cooler of the oil pumping section, while gaseous ones are processed further in the condensate treatment system. They are cooled and dried in a condensate dehydrator, using stripping gas from the 3rd stage of the gas compression train as a drying agent. The dry condensate is then cooled, pumped and finally mixed with the gas for export. Wet gas from the condensate dehydrator is reprocessed through the gas treatment system at the inlet of the 2nd stage [26].

**Produced water treatment, water handling and seawater injection**

Produced water from the oil and gas processing sections enters a wastewater handling train. Suspended particulates and traces of dissolved hydrocarbons are removed by hydrocyclones operated in parallel, and entrained gases are removed in degassers. Most cleaned produced water is discharged into the sea [27,28,30]. Seawater is supplied to the system, pumped and processed in the cooling water system. A large fraction is filtered and used for further injection into the oil reservoir.

**Flaring and venting**

The introduction of an offshore $CO_2$ tax by the Norwegian Authorities in 1991 [35] and stricter environmental regulations under the Petroleum Act [36–38] have encouraged efforts to reduce flaring practices [9,25]. This tax was increased from 210 to 410 NOK/ton $CO_2$ (28 to €55) in 2013 [39]. However, for safety reasons, the possibility of releasing gas by flaring to the atmosphere in emergency and shut-down situations is essential. The 1st stage separators, as well as the gas treatment and fuel gas processes, are connected to high pressure flares. The 2nd stage separator and the produced water handling system are connected to low pressure ones.

### *2.2. Modelling and simulation*

#### *2.2.1. Fluid modelling*

Reservoir fluids are complex multiphase mixtures containing a large variety of chemical components: their composition and properties differ significantly from one reservoir to another. The reservoir fluid processed on this platform is a light volatile crude oil [40]. The term *volatile oil* implies that the hydrocarbon compounds are mostly present in the liquid phase, but that large quantities of light hydrocarbons may evolve out of the liquid phase to the gaseous one when the reservoir pressure declines. The chemical composition of this reservoir fluid, excluding the subsurface water, is remarkable. The propane fraction is as high as 9% on a molar basis, and the content of medium-weight hydrocarbons such as butanes and pentane is significantly higher than in most conventional volatile crude oils [26].

In this work, the chemical compositions and flow rates of the feed streams, i.e. at the *inlet* of the processing plant, were deduced and adjusted from the crude oil, fuel gas and water compositions and flow rates, measured at the *outlets*. This backward approach was applied in the work of Voldsund et al. [18] and was suggested by the platform engineers. A direct and forward approach may be inappropriate, as there is a lack of knowledge on the exact composition of each well-stream. On the opposite, the application of a backward approach is eased by the measurements of, among others, the oil, gas and water compositions at the fuel gas and export systems.

Complete compositional analyses of crude oils are rarely conducted, and crude oils are often characterised by their bulk and distillation properties [2,4,40]. *Bulk properties* refer to properties measured when analysing the complete crude, such as density and viscosity. On the opposite, *distillation properties* refer to properties measured when analysing individually smaller fractions of the crude mixture. Thermophysical properties such as heating value and thermal conductivity may be estimated by empirical correlations. Crude oils are therefore modelled as mixtures of known and unknown, named *hypothetical*, or *pseudo-*, components [1]. Hydrocarbons forming the heavy fractions are lumped into hypothetical components, and each pseudo-component represents a certain number of real chemical compounds.

In this study, crude oil was modelled as a blend of 83 chemical compounds, including 29 hypothetical components. The properties of the pseudo-components were calculated by using the analyses and assays from 2002 [41,42]. The lumping of the real chemical compounds into pseudo-components results in additional uncertainties in the calculations of thermodynamic (e.g. enthalpy and entropy) and physical (e.g. molecular weight and liquid density) properties [43,44].

The following bulk properties were considered: an API (American Petroleum Institute) gravity of 39.9, a specific gravity of 0.826, a density of 825.5 kg/m$^3$ and a light ends fraction of 27.2% in volume. The API gravity varied between 39 and 41° this last decade. Thermophysical properties of the whole crude oil (e.g. density and kinematic viscosity) were similar from one assay to another with a deviation of ± 2%, and the kinematic viscosity at 20 °C is around 4 centistokes. Sulphur and nitrogen concentrations varied within a range of ± 5%, and vanadium and nickel contents by ± 0.03 ppm. The yield of light hydrocarbons, with a carbon number smaller than 4, slightly decreased, while the yield of heavy fractions increased.

*2.2.2. System modelling*

Data used for the model calibration and validation were based on values received from the process database. These were *not* the values directly received from the sensors, but values that were received after post-processing between the sensor and the database. They were given on a rate of 1/s and had an accuracy of up to 15 digits. The evaluation of the data quality showed that:

- when several sensors are placed at the same location (e.g. venting and flaring systems), a single averaged-value was stored, and no information on the averaging algorithm was available;
- some values were kept as constant values inside the database, as long as the standard deviation of the new measurement did not exceed a certain threshold limit;
- it was not possible to identify if the updated values were measured and registered at the same point in time, and this generated an additional uncertainty.

These constraints were considered in the modelling and simulation phases, when comparing the available measurements with the simulated values. The chemical compositions of the well streams were adjusted in the simulation to match the surveillance overviews provided by the operators, and measurement uncertainties were analysed, following the method proposed in Voldsund et al. [18]. Flow rates of produced oil, gas and water for each separator and well are measured and allocated, as well as of the gas, produced water and oil outflows, including the flow rates of lift gas per well. Information on the anti-surge recycling rates were provided by the operators. Cooling water flow rates were not measured, but were allocated based on the cooler data-sheets. Finally, an overview of the main energy users at a given point of time was presented.

Measurements on the well outflows, i.e. on the water, gas and oil flows for each well, are less accurate than the measurements on the platform outflows, because of the multiphase properties of the well-streams and the changes in the field conditions. Sources of uncertainties may be (i) the sensor's precision and calibration, and (ii) the operation instabilities.

The measurement uncertainties of each sensor were unknown: they were fixed to 10% for the inflows, as suggested in several technical documents, and to the limit values set by the Norwegian Petroleum Directorate for the outflows. The uncertainty levels at 95% confidence stated by the Norwegian Petroleum Directorate [45–47] are 0.30% for oil, 1.8% for fuel gas, 1.0% for sales gas and 5% for flared gas. Values for lift gas and for vented gas were not found and were assumed to be 1.8% and 5%, as for fuel and flared gas.

Data was reconciliated [48] by respecting conservation balances on the mass and standard volumes:

$$f(x^*) = \min_{x_i} \sum_{i=1}^{n} \left( \frac{x_i^* - x_i}{\delta_i} \right)^2 \tag{1}$$

where, given $n$ measurements, $x_i^*$ is the reconciled value of the variable of interest, $x_i$ the corresponding measured value and $\delta_i$ the measurement uncertainty, chosen as the limits set by the authorities at a 95% confidence interval. The term $\left( \frac{x_i^* - x_i}{\delta_i} \right)^2$ is called the penalty of measurement. The objective of the data reconciliation process is to minimise the overall correction $f(x_i^*)$, which is defined as the sum of the measurement penalties. Redundant measurements that can be used for data reconciliation were only available for the flowrates of the input and output streams. However, other measured variables, such as temperatures and pressures, are used in this study to calculate the energy and exergy flows.

The sample standard deviations [49] over the day of study $\sigma_x$, meaning that there is a 95% probability that a measurement taken in this day falls in the range $\bar{x} \pm 2\ \sigma_x$, are calculated as follows:

$$\sigma_x = \sqrt{\frac{1}{n-1}\sum_n (x_i - \bar{x})^2} \tag{2}$$

The sample standard deviation gives information on the variations of the flow and field conditions over a production day but does *not* reflect systematic errors caused by calibration imperfections. The deviations between the model results on the main process variables (standard volume flows, power and heating demands, fuel gas compositions) were smaller than 3%.

*2.2.3. System simulation*

The processing plant was simulated using Aspen Plus® [50] and Aspen HYSYS® [51]. The Peng-Robinson EOS (equation of state) [52], the Non-Random Two Liquid model [53] and the Schwartzentruber-Renon EOS [54,55] were used for calculating physical and thermodynamic properties. The first property package is recommended for oil and gas applications (e.g. gas processing), the second for electrolyte simulations (e.g. seawater treatment), and the last one for highly non-ideal liquid systems (e.g. glycol dehydration). Uncertainties associated with the equation of states are not considered in this work: the Redlich-Kwong equation of state with Soave modifications [56] was tested in parallel for the oil and gas processing and did not return significant differences, as suggested in several works [57,58].

Process data were measured and available from 2000 to 2013, showing that operating conditions and process variables change considerably from year to year because of variations in the well-fluid composition and flow rates. However, variations on an hourly, daily or weekly basis were not significant, with the exceptions of urgency or shut-down situations, as well as cases where load set-points were changed by the operators. Normal production days can be grouped into two different categories. The first one is called *low energy use* production days: oil, gas and water are processed and treated on-site, and oil is stored in the storage tanks located at the bottom of the plant. The second one is denoted *high energy use production days*: oil, gas and water are processed onsite, and oil is pumped from the storage tanks to the floating loading buoys, which results in a greater power consumption. The second operation mode is the least frequent. A representative day is studied for both cases, but this work focuses mainly on the first type of production day, as the oil export system is run separately.

*2.3. Thermodynamic analysis*

*2.3.1. Energy balance and exergy accounting*

*Energy* may be transformed from one form to another and transferred between systems, but can neither be created nor destroyed. *Exergy* is defined as the *maximum theoretical useful work as the system is brought into complete thermodynamic equilibrium with the thermodynamic environment while the system interacts with it only* [11,59,60], and, at the opposite of energy, is not conserved in real processes. Some is destroyed due to internal irreversibilities [11], and the exergy rate balance can be written as:

$$\begin{aligned}\dot{E}_d &= \sum \dot{E}_{\text{in}} - \sum \dot{E}_{\text{out}} \\ &= \sum_k \left(1 - \frac{T_0}{T_k}\right)\dot{Q}_k - \dot{W} + \sum \dot{m}_{\text{in}} e_{\text{in}} - \sum \dot{m}_{\text{out}} e_{\text{out}}\end{aligned} \tag{3}$$

where $\dot{E}_d$, $\dot{E}_{\text{in}}$ and $\dot{E}_{\text{out}}$ are the destroyed, inflowing and outflowing exergy rates. $e$ is the specific exergy of a stream of matter, $T_0$ and $T_k$ are the environmental and instantaneous temperatures. Alternatively, the exergy destruction rate can also be calculated from the Gouy-Stodola theorem [12]. The exergy destruction is related to the irreversibilities taking place within the system, whereas the exergy losses, written $\dot{E}_l$ on a time rate basis [11], refer to the exergy discharged to the environment without any practical use (e.g. exergy lost with cooling water) [16,61].

In the absence of nuclear, magnetic and electrical interactions, the exergy associated with a stream of matter is a function of its physical $e^{\text{ph}}$, chemical $e^{\text{ch}}$, kinetic $e^{\text{kn}}$ and potential $e^{\text{pt}}$ components [11], and is expressed as:

$$e = e^{\mathrm{ph}} + e^{\mathrm{ch}} + e^{\mathrm{kn}} + e^{\mathrm{pt}} \tag{4}$$

Physical exergy accounts for temperature and pressure differences with the environment and can be derived as:

$$\begin{aligned} e^{\mathrm{ph}} =& (h - h_0) - T_0(s - s_0) \\ =& \underbrace{h - h(T_0, p) - T_0\left(s - s(T_0, p)\right)}_{I} + \\ & \underbrace{\left(h(T_0, p) - h_0\right) - T_0\left(s(T_0, p) - s_0\right)}_{II} \end{aligned} \tag{5}$$

where $s$ is the specific entropy of a stream of matter per unit-of-mass. *Term I* and *Term II* refer to the thermal $e^{\mathrm{t}}$ and mechanical $e^{\mathrm{m}}$ components of the physical exergy [16,62]. In this work, the ambient conditions are taken to be 8 °C, as measurements conducted by the operators indicated that the air temperature oscillated between 7 and 9 °C, and 1 atmosphere.

Chemical exergy accounts for deviations in chemical composition from reference substances present in the environment. In this work, chemical exergy is calculated based on the concept of *standard chemical exergy*, using the reference environment defined by Szargut [63–65]. The specific chemical exergy of hypothetical components $e^{\mathrm{ch}}_{\mathrm{hyp}}$ is computed with the heuristic correlations of Rivero [66]. The specific chemical exergy of a given mixture $e^{\mathrm{ch}}_{\mathrm{mix}}$ is calculated as [67]:

$$\begin{aligned} e^{\mathrm{ch}}_{\mathrm{mix}} =& \underbrace{\sum_i x_i e^{\mathrm{ch}}_{i,\mathrm{mix}}}_{I} \\ =& \underbrace{\sum_i x_i e^{\mathrm{ch}}_{i,0}}_{II} + \\ & \underbrace{\left(\sum_i x_i \left(h_{i,\mathrm{mix}} - h_{i,0}\right)\right) - T_0 \left(\sum_i x_i \left(s_{i,\mathrm{mix}} - s_{i,0}\right)\right)}_{III} \end{aligned} \tag{6}$$

where the mass fraction, the chemical compound and the mixture are denoted by $x$, $i$ and mix. The specific exergy of a given chemical compound is written $e^{\mathrm{ch}}_{i,\mathrm{mix}}$ when it is in the mixture and $\bar{e}^{\mathrm{ch}}_{i,0}$ when it is in a pure component state. Term $I$ illustrates the chemical exergy of each individual chemical compound in the mixture, Term $II$ the chemical exergy of these compounds in an unmixed form, and Term $III$ the reduction in chemical exergy due to mixing effects. Term $II$ is called the *pure-component* term in the rest of this study, while Term $III$ is called the *compositional exergy* in the work of Rivero [66]. The *compositional exergy* is negative, as it is defined as the reduction of chemical exergy caused by mixing, and is always smaller than the pure-component exergy. Kinetic and potential effects are neglected.

### *2.3.2. Thermodynamic performance parameters*

Oil and gas companies developed their own indicators to assess the energy performance of oil and gas processing and offshore platforms. Others have been discussed in the open literature [9], such as:

- the energy efficiency $\eta$, defined as the ratio of the energy of the oil and gas leaving the plant, including the lift gas, to the energy entering the system with crude oil.
- the energy intensity $\iota_{\hat{h}}$, defined as the ratio of the energy contained in the fuel gas used for power generation to the energy exported to the shore.

- the energy waste $\omega_{\hat{h}}$, defined as the ratio of the energy of the flared and vented gases to the energy exported to the shore.
- the specific power consumption $w$, defined as the power consumed per unit of oil equivalent exported onshore, on a standard volume basis.

The subscript $\hat{h}$ for the second and third indicators denotes that they are defined on an energy basis. The three first indicators may be calculated on a daily basis and compared to benchmark values given by the platform operators, which are 97%, 2% and 0.15% for the present platform. They are generally calculated on a chemical energy basis, meaning that the thermal energy of the gas and oil streams are neglected. These indicators illustrate the use of the energy resources on-site, as well as the possible leaks (e.g. flaring) during the system operations. They allow therefore for a comparison of some aspects of the performance of oil and gas platforms.

Similar indicators to the energy intensity and waste parameters may be developed on an exergy basis: they are written in the rest of the study with $\hat{x}$ as a subscript to avoid confusion. Besides them, the values of the exergy destruction and loss rates provide information on the system inefficiencies, and other performance parameters related to these variables were developed [11,14–16]. They can illustrate the possibilities for improvement and indicate the components and sub-systems on which attention should be focused:

- the exergy destruction ratio $y_d^*$, defined as the ratio of the exergy destruction rate $\dot{E}_{d,k}$ within a specific process component $k$ to the exergy destruction rate within the whole system $\dot{E}_d$:

$$y_d^* = \frac{\dot{E}_{d,k}}{\dot{E}_d} \tag{7}$$

- the exergetic efficiency $\varepsilon$ of a given component or sub-system $k$, which is defined as the ratio of the product exergy to the fuel exergy. The product exergy $\dot{E}_{p,k}$ represents the desired effect of a given thermodynamic transformation or process, while the fuel exergy $\dot{E}_{f,k}$ represents the resources expended in this component/sub-system to generate this desired result [16,68].

$$\varepsilon_k = \frac{\dot{E}_{p,k}}{\dot{E}_{f,k}} = 1 - \frac{\dot{E}_{d,k} + \dot{E}_{l,k}}{\dot{E}_{f,k}} \tag{8}$$

  The fuel and product exergies are not necessarily equal to the exergy flows entering $\dot{E}_{\mathrm{in},k}$ and leaving $\dot{E}_{\mathrm{out},k}$.

The exergy destruction ratios give information on the share of exergy destruction within the plant, and are therefore very useful for locating and evaluating the main sources of thermodynamic inefficiencies. The exergy efficiency indicators can be used to compare the performance of similar components in the same system. It should be noted that the inefficiencies within a given component can be caused by the inefficiencies of the components present in the remaining system (e.g. aftercooler after a compressor).

## 3. Results

### *3.1. Energy analysis*

#### *3.1.1. Energy flows*

About 3870 ± 100 MW of energy enters the oil and gas plant, of which $\simeq$ 98.5% enters with the well-streams, $\simeq$ 0.5% with power and less than 0.3% with heat carried with the triethylene glycol. Most inflowing energy is exported to the shore along with oil, gas and condensate (76% ± 0.4), while a small fraction is used as lift gas (18% ± 0.2), fuel gas (2.3% ± 0.1), and injection seawater (0.4% ± 0.1). Only a small fraction of the total energy input is lost to the environment, along with wastewater (2.0% ± 0.1), cooling water (0.5% ± 0.1), injection water (0.4% ± 0.1), flared and vented gases ($\simeq$ 0.1%).

*3.1.2. Energy transformations*

Energy flowing through the oil and gas plant is dominated by the chemical energy of the material streams. Electrical energy is converted in the recompression, compression and pumping sections, with negligible heat losses, while energy is partly dissipated along with cooling water. The total power consumption varies between 18 and 26 MW, as the oil loading system is not run continuously. For any production day, the gas treatment ranks as the most power-consuming sub-system, representing 44 to 60% of the total power demand of the processing plant. The major power consumers of this system are the compressors, using about 13.3 MW of power. The total heating demand is about 4.2 MW, of which 90% is provided by the thermal oil loop and and the remaining part by electric heating (Figure 2). The latter increases slightly when oil is exported, since more fuel gas needs to be heated. The discharge of thermal energy with cooling water amounts to about 18 MW. However, it should be noted that the well-streams enter the platform at a high temperature and are consequently cooled, which explains these large energy losses.

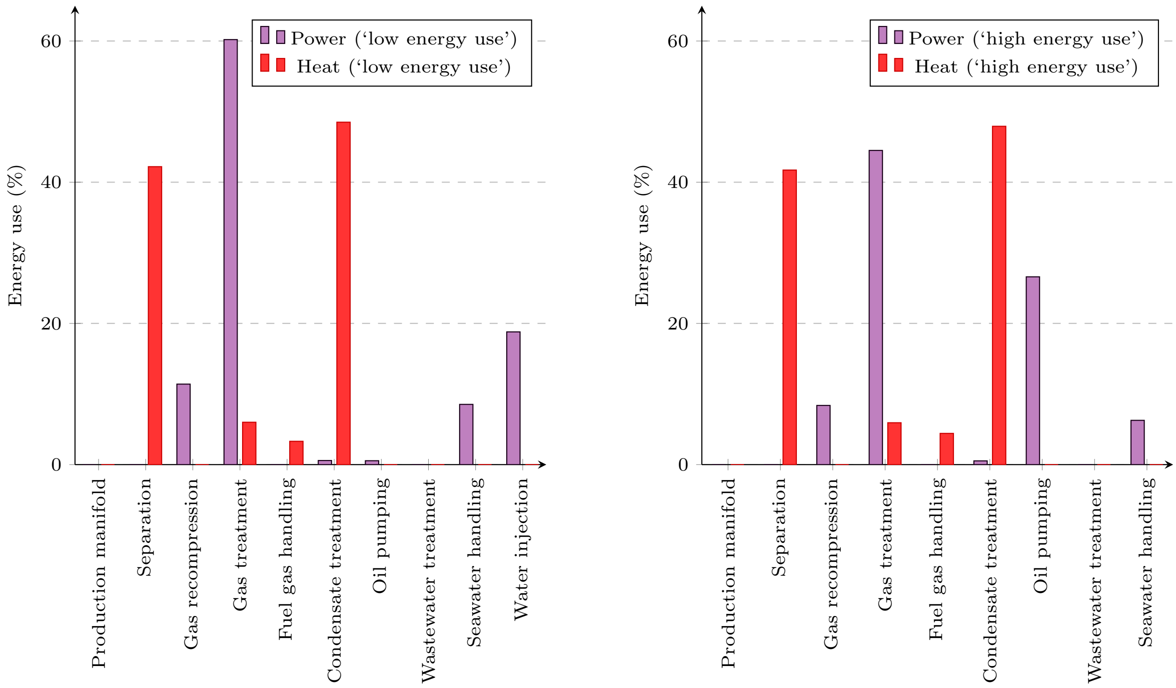


Figure 2: Energy demand of the processing plant for a 'low' and 'high energy use' production day, expressed in relative values.

*3.1.3. Energy-based indicators*

The energy efficiency $\eta$, the energy intensity $\iota_{\hat{h}}$ and the energy waste $\omega_{\hat{h}}$ amount to 93.4% $\pm$ 0.25, 2.3% $\pm$ 0.05 and 0.13% $\pm$ 0.03 for a *low energy use* production day, and about 93.2%, 2.8% and 0.18% for a *high energy use* one. In all cases, the energy efficiency of the processing plant is lower than the benchmark value, while the energy intensity and waste are higher. The changes of the energy contained in the lift gas are small, as the volume of gas used for lift varies only slightly, while the amount of fuel gas changes from day to day, since the water injection turbines are not operated at a constant load. These values, as returned by the model, are similar with a deviation of $\pm$ 1% point to the values retrieved by the operators. The largest difference is found with the values of the energy intensity factor, and this may be caused by (i) a difference in the literal expression of this indicator, (ii) discrepancies between measurements and estimations or (iii) deviations between the measured composition at a certain time point, and the actual one. The specific power consumption $w$ of the processing plant varies between 75 kWh/Sm$^3_{\mathrm{o.e}}$ and 100 kWh/Sm$^3_{\mathrm{o.e}}$ of exported oil and gas for *low* and *high* energy use production days. The specific power consumptions of the gas recompression and treatment systems amount to about 1050 kJ/Sm$^3$ and 650 kJ/Sm$^3$ of processed gas.

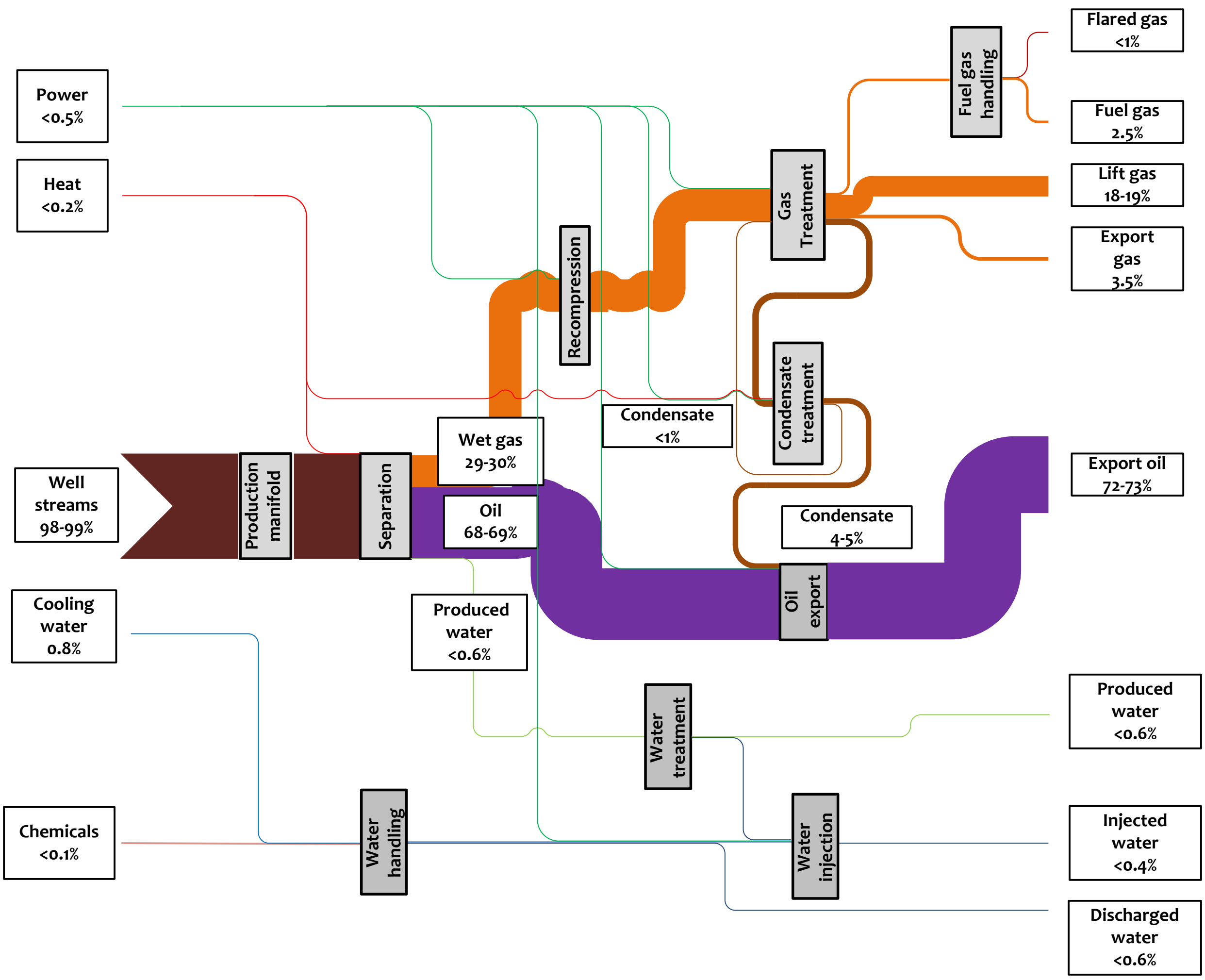


Figure 3: Grassmann diagram of the processing plant of the Norwegian Sea offshore platform investigated in this work.

This significant difference may result, among other factors, from the larger fraction of gas that is recirculated in the recompression section to prevent surge. For the condensate treatment process, this indicator has a value of about 60 kJ/Sm$^3$ of condensate.

*3.2. Exergy analysis*

*3.2.1. Exergy flows*

The exergetic accounting shows that exergy enters the processing plant along with material flows (about 3850 MW $\pm$ 100 MW, along with well-streams, seawater and chemicals) and power (about 18.5 MW), and exits it along with streams of matter (separated crude oil, gas, produced water and seawater). The exergy inflows and outflows of the processing plant are dominated by the chemical exergy associated with the oil and gas streams (Table 1 and Figure 3).

The chemical exergy associated with pure components (Term $II$ in Equation 6) transits from the inlet to the outlet of the offshore plant, as the hydrocarbons are not consumed but are just separated. Mixing of the hydrocarbons, water and impurities results in a reduction of the chemical exergy (Term $III$ in Equation 6) between 0.11 and 0.23% for the well-streams, about 0.06% for the export oil, 0.01% for the produced water, 0.14–0.24% for the vented and flared gases, 0.24% for the export, lift and fuel gas streams.

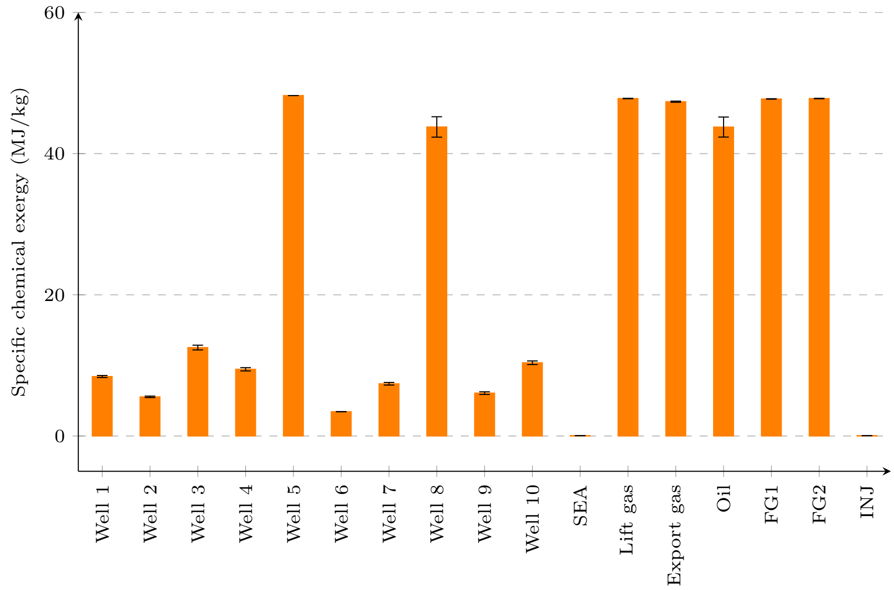


Figure 4: Specific chemical exergy of inflows and outflows. Chemical exergy is estimated from petroleum correlations used to estimate the lower heating value and from the works of Szargut [63–65] and Rivero [66]. Uncertainties are plotted considering the uncertainties on the hypothetical components, varying their specific chemical exergy from the lower heating value to the higher one. 'FG1' refers to the fuel gas sent to the power turbines, 'FG2' to the fuel gas sent to the water injection turbines, and 'SEA' to the cooling seawater, and 'INJ' to the injected water. Triethylene glycol entering the process is not shown, as it is processed as a heating loop.

The uncertainties related to the total chemical exergy of the well-streams and oil are higher than for the other streams (Figure 4). They are related to (i) the lumping of the heavy oil fractions into pseudo-components, (ii) the inaccuracies of the Peng-Robinson equation of state, and (iii) the use of empirical correlations for estimating the heating values. The choice of the hypothetical components affects the exergy calculations, in particular the estimations of the chemical exergy of the oil stream, and results in higher uncertainties in the flow rates. The latter may be high for each individual well-stream, but are generally low for the output flows. The overall picture stays nonetheless unchanged, as all exergy transformations taking place on-site are related to changes in the compositional and physical exergies.

The physical exergy of the well-streams is generally dominated by the thermal exergy, as a result of a high water content. The physical exergy associated with the lift, fuel and export gas is dominated by the mechanical exergy, while it is dominated by the thermal exergy for the discharged water.

*3.2.2. Exergy transformations*

No chemical transformations occur in the processing plant, since the gas turbines are not considered, meaning that the changes in chemical exergy are only related to the mixing and separation effects (*exergy of separation* [16]). Exergy is consumed within the offshore system to separate the crude oil mixture into its gas, oil and water components, resulting in an increase of the chemical exergy (Figure 5). The sum of the separation effects in all separators and scrubbers amounts to 320 kW, but the separation effect of the overall plant is only about 180 kW because of the effect of the mixing and recycling loops. This increase of chemical exergy between the inlet and outlets is performed at the expense of other forms of exergy [69], such as (i) thermal exergy (e.g. the crude oil mixture is heated at the 2nd separation stage) (ii) mechanical exergy (e.g. the pressure of the well-streams is decreased between the bottom head of the wells and the outlet of the separation train), (iii) potential exergy (e.g. oil and gas are separated by gravity in the 1st separation stage) or (iv) a combination of these three. The mixture velocity is also decreased, easing the phase separation.

Table 1: Exergy inflows and outflows of the processing plant of the studied platform. Physical exergy $\dot{E}^{\mathrm{ph}}$ is deduced from the enthalpy and entropy calculations for each material stream. Compositional exergy $\dot{E}^{\mathrm{mix}}$ is calculated by computing the enthalpies and entropies of the chemical compounds in pure and mixed forms, as shown in Equation 6.

| Stream | $\dot{m}$ ($10^3$ kg/h) | $\dot{E}^{\mathrm{ph}}$ (kW) | $\dot{E}^{\mathrm{mix}}$ (kW) | $\dot{W}$ (kW) | $e^{\mathrm{ph}}$ (kJ/kg) | $e^{\mathrm{mix}}$ (kJ/kg) |
|---|---|---|---|---|---|---|
| **In** | | | | | | |
| Wells | 398 | 12,338 | -5388 | | 31.0 | -13.5 |
| TEG | 255 | 11,258 | 0 | | 159 | 0 |
| Seawater | 2,302 | 0 | 0 | | 0 | 0 |
| Power | | | | 18,929 | | |
| **Out** | | | | | | |
| Wells | 55 | 5437 | -1760 | | 356 | -115 |
| Export wet gas | 10 | 774 | -322 | | 276 | -115 |
| Wasted gas | 0 | 25 | -11 | | 193 | -85 |
| TEG | 255 | 9,577 | 0 | | 135 | 0 |
| Fuel gas | 7 | 492 | -214 | | 261 | -113 |
| Injection water | 860 | 3,607 | 0 | | 15 | 0 |
| Discharged water | 1,441 | 637 | 0 | | 2 | 0 |
| Produced water | 1,126 | 6,156 | -1 | | 20 | 0 |
| Oil | 231 | 301 | -2,898 | | 5 | -45 |

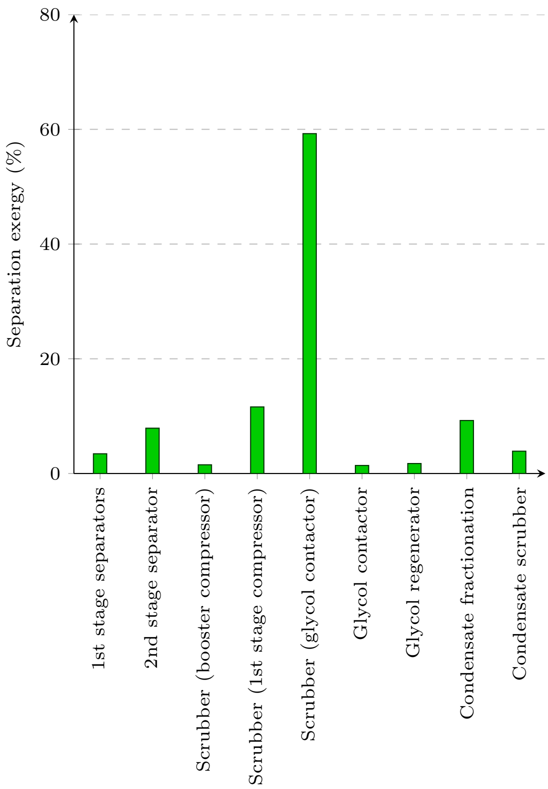


Figure 5: Separation exergy of the processing plant, expressed in relative values. Contributions of the condensate separator in the glycol regeneration process, of the 3rd stage scrubbers and of the hydrocyclones and degassers are not shown as they are smaller than 3 kW each.

### *3.2.3. Exergy destruction and losses*

The total exergy destruction on the processing plant amounts to 15,290 kW during a low-energy use production day (Figure 6), i.e. to about 79% of the total power consumption. It mostly takes place in subsystems where pressure is decreased (production manifold) or increased (gas compression) significantly. These thermodynamic irreversibilities are caused by the use of throttling valves, the inefficiencies of the gas

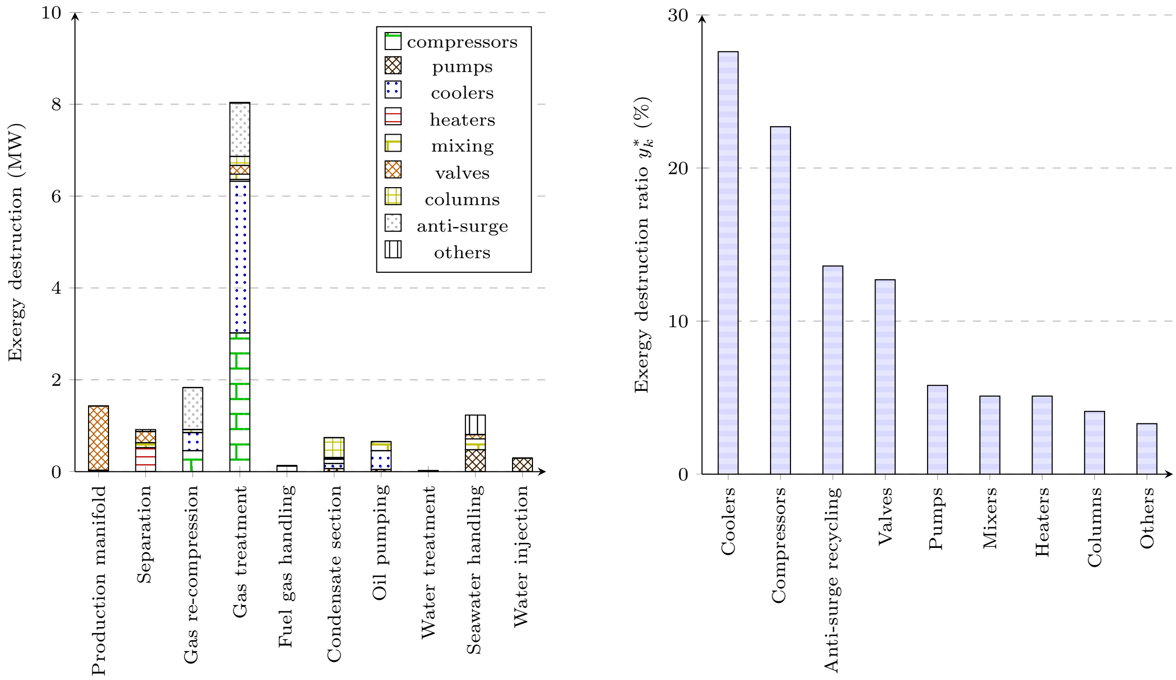


Figure 6: Exergy destruction share and ranking, sorted by processes (*left*) and components (*right*). Coolers refer to seawater coolers, and heaters to both fuel heaters and internal heat exchangers (glycol-glycol and condensate-condensate). Exergy destruction caused by pressure drops in the separators is lumped into the category 'valves' and irreversibilities resulting from pressure drops of the cooling medium in the heat exchangers are allocated to each sub-system. Others refer to exergy destruction taking place in pipelines, such as pressure drops and heat losses.

compressors and by heat transfer across the large temperature gap between the hot gases and cooling water. Anti-surge recycling results in a larger exergy destruction within the compression train, because (i) recycled gas is throttled to the pressure before the compressor it serves, (ii) the mixed gas streams have different temperature, pressure and chemical compositions and (iii) larger mass flows are processed through the heat exchangers. This exergetic analysis suggests that the amount of exergy destroyed on-site only varies slightly over a production year, with a maximum deviation of 5%. Exergy destruction in the fuel gas and produced water handling systems are of minor importance compared to the other studied systems.

The exergy destruction taking place during the oil loading process may be considered as equal to the power consumed in the pumping process. The overall storage and loading process destroys exergy, in the sense that oil was stored at the bottom and is brought again to the surface, where the potential energy and exergy are null. Similarly, the oil was brought in motion in the pumping process and is stored in shuttle tanks, where the kinetic energy and exergy are dissipated. The exergy destruction of the complete process operation is equal to 6640 kW, which is the additional power consumption induced by the oil loading. The actual configuration of the system can hardly be improved, unless if more efficient loading pumps are integrated, and if losses through the pipelines can be reduced, which is hardly feasible in practice.

Exergy losses of the processing plant system consist of (i) the exergy contained in the produced water rejected to the sea, (ii) the exergy associated with waste gas streams (flared and vented gases), (iii) the exergy carried with the lost cooling water, i.e. the exergy increase of the cooling water between the inlet and outlet of the processing plant. Physical exergy losses amount to 6150 kW with the produced water, 24.6 kW with flared and vented gases, 637 kW with lost cooling water, i.e. seawater that is not injected in the reservoir. Chemical exergy losses account for 4910 kW with flared and vented gases: this value is higher than the predicted energy losses from the operators, since the chemical exergy of hydrocarbons is generally higher than their lower heating value. The models of Szargut [64,65] and Kotas [16] estimate the chemical exergy of liquid water to 900 kJ/kmol and 3180 kJ/kmol. Considering the first model, the chemical exergy of the produced water effluent amounts to 15,630 kW, but using this exergy with current technologies is challenging. The exergy losses of the processing plant system do not vary with the operating modes, the volumes of flared and vented gas are, in theory, insensitive to the operation of the oil loading process, and the same reasoning applies for the exergy rejected with produced and cooling water.

Potential and kinetic exergies were not considered within this study, as no velocity or height measurements were available. In practice, kinetic exergy is destroyed, because the well-stream velocity is reduced between the wells and the separation plant. Similarly, potential exergy is destroyed when being converted into physical and kinetic exergies, when gas, oil and water flow through the processing plant.

#### *3.2.4. Exergy-based indicators*

The exergy intensity $\iota_{\hat{x}}$ and the exergy waste $\omega_{\hat{x}}$ amount to 2.5% and 0.17% for a *low energy use* production day, and similar values are found for the *high energy use* one. These values are sensibly similar to the energy intensity $\iota_{\hat{h}}$ and the energy waste $\omega_{\hat{h}}$ indicators, because the chemical energy and chemical exergy of hydrocarbons dominate the energy and exergy flows and differ by only ± 1.2%. The trends are similar for both types of production days, but the oil loading system is responsible for a significant amount of destroyed exergy when operated. It ranks as the second most exergy-destroying system after the gas treatment process and is followed by the gas recompression and production manifold.

The four gas compressors have an exergetic efficiency of 79%, 73%, 74% and 72%, and the corresponding polytropic efficiencies are 74%, 67%, 69% and 61%. These values were deduced from the pressures and temperatures at the inlets and outlets of the compressors, as well as from their power requirements. The heat exchangers display low exergetic efficiencies, in the range of 2–17% for the coolers, as exergy is transferred across a large temperature gap from the hot gases to the cooling seawater. The heaters and internal heat exchangers are more performant, with an exergetic efficiency of 35–65%, since exergy is exchanged at a higher heat transfer temperature.

No meaningful exergetic product can be defined for the valves, since they are operated above the dead state conditions and are dissipative by design. These components, along with the mixers, are used in the system to ease the processing of oil and gas. Some act as flow regulators, since they control the amounts of oil and gas entering each sub-system, compression or separation stage. Others act as pressure controllers, to limit the pressure in some parts of the plant and protect pumps and compressors.

## 4. Discussion

### *4.1. Comparison with literature*

#### *4.1.1. Comparison with general and generic studies*

The process simulations and the measurements of various platform parameters show that the gas compression ranks as the most power-consuming subsystem of the processing plant. The oil pumping and water injection processes follow, when they are in operation. These trends are similar to the ones observed by Svalheim and King [9], who conducted an energy performance assessment on four different oilfields on the Norwegian Continental Shelf. They also pointed out that the gas compression process may be responsible for a significant fraction of the power consumption over the field life cycle, even if gas export accounts for only a small fraction of the total oil and gas export. This is also observed in the present case study: the gas export represents only 5 to 10% of the total exergy export, and this number will decrease in the coming years, since gas will not be exported after 2015. Gas will only be used for lift and power generation. The findings of this work can also be compared to the generic study of Nguyen et al. [23], which also supports that (i) the gas compression and production manifold are the most exergy-destroying processes, (ii) a greater water cut results in a smaller thermodynamic performance of an oil and gas facility.

#### *4.1.2. Comparison with a North Sea platform*

There are two other thermodynamic assessments of real-case offshore processing systems presented in the archival literature: the exergy analysis of a North Sea oil platform, performed by Voldsund et al. [18,20,21] and the exergy analysis of a Brazilian oil platform, performed by Oliveira and Van Hombeeck [19].

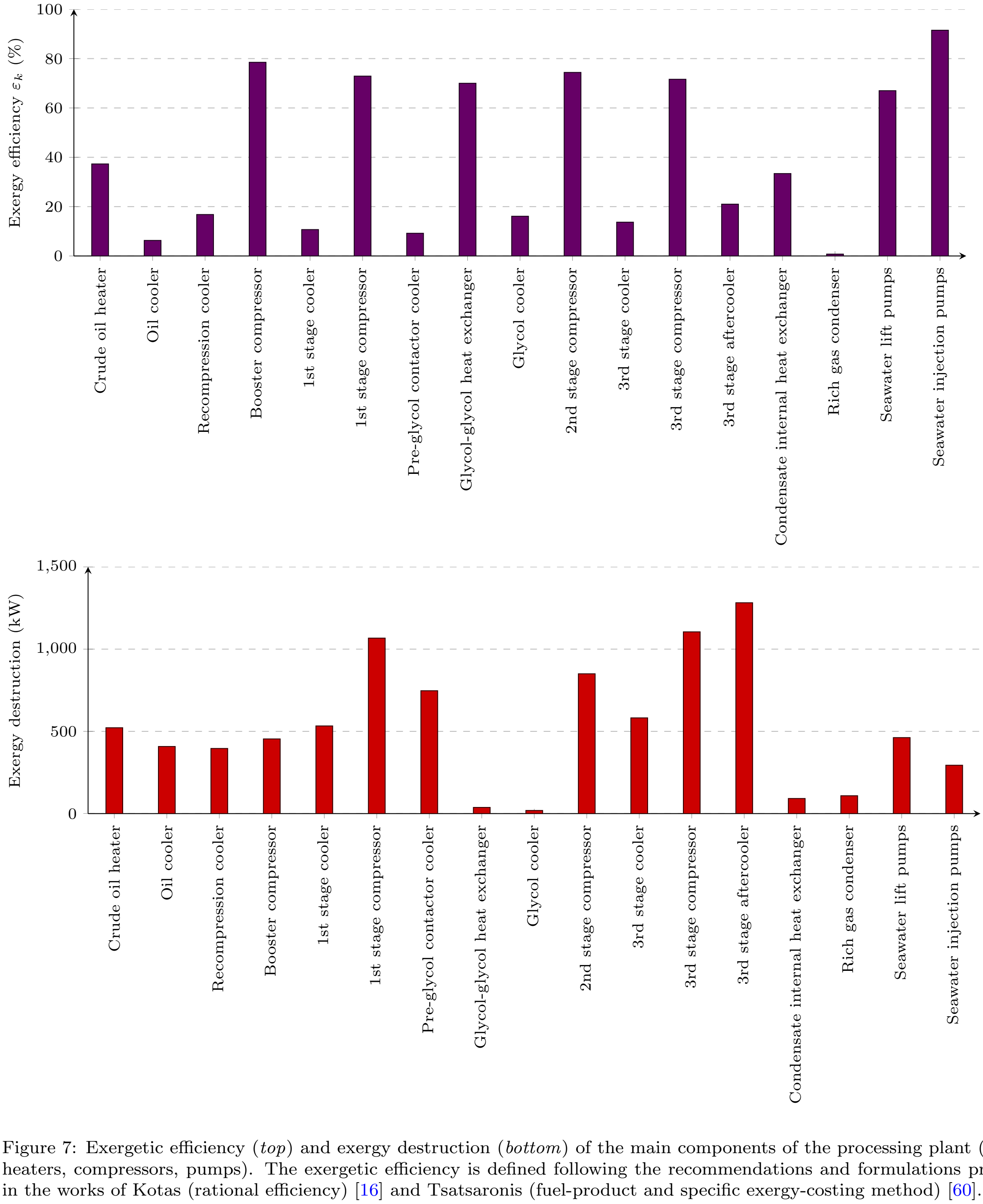


Figure 7: Exergetic efficiency (*top*) and exergy destruction (*bottom*) of the main components of the processing plant (coolers, heaters, compressors, pumps). The exergetic efficiency is defined following the recommendations and formulations presented in the works of Kotas (rational efficiency) [16] and Tsatsaronis (fuel-product and specific exergy-costing method) [60].

The platform investigated by Voldsund et al. [18,20,21] was started up during the last decade of the 20th century. The facility processes a reservoir fluid with a high methane content ($\simeq$ 78–82% on a molar basis) and the production was 132.5 $\pm$ 0.4 Sm$^3$/h of oil, 369 $\pm$ 17 kSm$^3$/h of gas and 67 $\pm$ 5 Sm$^3$/h of water for the studied day. Oil is pumped to 32 bar and is exported to the shore, gas is compressed to 236 bar and is reinjected into the reservoir for pressure maintenance, and water is discharged into the sea after cleaning. Water injection is not implemented. The power consumption, excluding utilities such as seawater lift and drilling, is about 23.8 $\pm$ 0.4 MW, which is in the same order of magnitude than the power demand of the platform studied in this work.

The two platforms are operated under very different natural (e.g. gas- and water-to-oil ratios) and technical conditions (e.g. export and injection specifications). However, the comparison between both works shows that the gas compression–treatment process is the most power-consuming step of the offshore plant. The absolute power consumption of the Voldsund's platform is in the same order of magnitude than of Draugen, but the specific power demand per unit of produced gas is about five times smaller. This large difference can be explained by (i) the larger propane and butane content of the produced gas, (ii) the lower efficiency of the compressors due to the off-design conditions and anti-surge recycling, and (iii) the different pressure levels. Similarly, the gas recompression and oil pumping sub-processes rank next. The significant losses of thermal energy with cooling water are also pointed out, as well as the technical difficulties for using this low-grade heat.

The exergy analyses show that the gas compression train is the most exergy-destroying sub-system ($\simeq$ 40–50%) for both offshore plants, because of the irreversibilities taking place in the compressors ($\simeq$ 15–30%) and coolers ($\simeq$ 10–15%). The production manifold is also one of the major exergy-destroying sub-system ($\simeq$ 20–30%). The platform investigated in this work presents a smaller exergy destruction in the production manifold, and this is because of (i) the smaller pressure difference between the upstream choke and the separation inlet, and (ii) the greater water content. At the same pressure level, water has a much smaller mechanical exergy than oil and gas.

A significant difference between both platforms is the exergy destruction caused by the anti-surge recycling. The gas *recompression* trains are, in both cases, run in severe off-design conditions: the anti-surge recycling fraction is in the range 69–92% for the Voldsund's platform and more than 65% for the one studied in this work. However, on the Voldsund's platform, the gas *compression* trains are run at maximum capacity, with no anti-surge recycling, while the recirculation flows amount to 15–30% in the present case. These differences can be explained by the production profiles of an oilfield. The oil peak is reached first, meaning that the oil production starts to decline while the gas production still increases, which is the case of the Voldsund's platform and of the one studied in this work. The amount of medium-weight hydrocarbons recovered in the separation system decreases, and the gas booster compressors start running in off-design conditions. At a further point of time, the gas peak is also reached, and the gas production starts decreasing, which is the case of the present platform. The gas compressors of the gas treatment section are operated far from their design point, and anti-surge recycling around these components starts taking place.

The exergy destruction associated with the separation system is larger for their plant, because of the greater pressure reduction (about 67 bar, compared to 6.2 bar) between the first and final separation stages. The produced water handling system and the exergy losses associated with wastewater are not investigated in their work, and there are neither a water injection nor a condensate treatment process on their platform.

The mapping of the exergy flows throughout the processing plant shows that, in both cases, the mechanical exergy term dominates the physical exergy of the gas streams. The separation work is mostly performed in the treatment sub-system because of the scrubbing prior to the compressors. The main difference is the contribution of the condensate treatment process, where the purity of the oil and gas flows is increased by a fractionation column.

#### *4.1.3. Comparison with a Brazilian Basin platform*

Oliveira and Van Hombeeck [19] studied a real-case oil platform located in the Brazilian Basin. The reservoir fluid has a high water content ($\simeq$ 79% on a molar basis). The platform produces about 285 t/h of oil, 29 t/h of gas and 136 t/h of water. Oil is pumped to 69 bar and is exported to the shore, gas is compressed to 174 bar and water is rejected into the environment. The power consumption of the processing plant is about 21 MW, which is in the same order of magnitude of the present case study, and in Voldsund et al. [18,20,21]. The significant power demand and exergy destruction of the gas compression sub-system are also pointed out, and a significant amount of separation work is performed on Oliveira's platform.

However, there are a few differences between their findings, and the conclusions of the present study and of Voldsund et al. [18,20,21]. First, they stressed the large exergy consumption and destruction associated with the crude oil heating in the separation sub-system. This heating demand is negligible on the platform studied by Voldsund et al. and minor on the present one, because the feed temperatures in the North and Norwegian cases are much higher ($\geq$ 50 °C) than in the Brazilian one ($\leq$ 10 °C). They suggested to focus on the development of new separation technologies such as centrifugal separation, and on a better matching of the stack and separation temperatures, rather than on an upgrading of the gas compression sections. Secondly, the production manifolds, which are pointed as the second most exergy-destroying sub-system in other case studies, are not considered: the inlet pressure at the 1st separation stage is about 10 bar, suggesting therefore that a significant amount of exergy is also destroyed in this process module.

### *4.2. Theoretical implications and practical applications*

The comparison of these different offshore platforms [18,19], the present findings and the generic analysis of Nguyen et al. [22] raise several points on the thermodynamic performance of oil and gas processing plants.

#### *4.2.1. Production manifolds*

Thermodynamic irreversibilities are likely significant in systems where pressure is substantially decreased. The exergy destruction within the production manifolds, where the pressure of the well-streams is reduced by throttling, represented 10% of the exergy destruction of the present case study.

Multiphase flow expanders would enable energy recovery from the depressurisation of the well-streams and reduce the quantity of exergy destroyed in the production manifolds. Multiphase flow ejectors would result in higher oil recovery in depleted wells, which is of particular interest for mature oilfields. There are in practice technical challenges in implementing these devices, as the fractions of oil, gas and water vary significantly from one well to another, and over their life cycle. The presence of impurities and sand in the reservoir fluids will also complicate the designing task.

Some platforms include manifolds operated at different pressure levels, both on the inlet and on the outlet side. The inlet pressure depends on the well pressure and on the necessary pressure drop between the wellhead and the production manifold to ease flowing of oil and gas. The outlet pressure depends on the requirements of the 1st stage separation process and on the lowest pressure of the mixed well-streams. Integrating production manifolds with more pressure levels would result in (i) smaller exergy destruction in this part of the plant, (ii) lower gas recovery at the 1st separation stage, (iii) smaller power demand of the gas compression and thus higher gas export, (iv) greater system complexity, and (v) possibly larger power demand of the recompression train, as the separation outlet pressure is constrained by the vapour pressure specifications. However, well-stream pressures decrease with time, and therefore these well-streams may be re-routed to a production manifold operated at a lower pressure.

The selection of possible improvements is therefore a compromise between capital (higher number of manifolds and pipelines) and operating costs (smaller power consumption and exergy destruction, and therefore higher export gas). It also points out the need for investigating the interactions between each sub-system, as the improvement of one process may have a negative feedback on another. Such improvements may not be effective at the end-life of a field, since lower pressures and higher water cuts result in a lower mechanical exergy of the well-streams.

#### *4.2.2. Separation*

Regarding the separation system, attention may be drawn to the gas fraction of each well-stream entering each separator. Some wells mainly process gas, while some mainly process oil and water. The mixed well-streams at the inlet of the 1st stage separator have different vapour fractions. The well-streams entering the 1st stage separator dedicated to the subsea wells contained more gas, and therefore gas could be recovered at a higher pressure. The possibilities of (i) integrating an extra phase separator or operating the 1st stage separator at a higher receiving pressure, (ii) increasing the number of separation stages, and (iii) by-passing the oil and gas mixing step and treating the gas separately, may be considered for wells processing a high amount of gas.

This would result in a smaller power consumption in the recompression and compression trains, and in a smaller exergy destruction in the separation section. Similarly, the operating benefits of adding more components or integrating more separation stages or parallel trains should be evaluated against their investment costs. The latter are likely to increase, as the footprint area and volume taken by the process would be higher. These benefits are also limited for platforms operated on mature fields, as the exergy destruction in the separation process is expected to be smaller than during a peak production.

The replacement of the throttling valves between each separation stage may be considered. Most water exits the separation sub-system at the 1st stage, even at high water cuts, and the oil composition in the separation system fluctuates less. Designing expanders for this section of the plant is therefore less problematic, and may result in significant exergy recovery if the pressure differential between the inlet and the outlet of the separation sub-system is high.

*4.2.3. Gas recompression and treatment*

The amount of exergy destroyed in the gas compression depends on parameters such as the processing gas composition, the need for dehydration and the efficiency of the compressors. Although the operating conditions differ from one platform to another, the exergy analyses of offshore platforms indicate that there is room for improving the performance of the gas compression processes. The exergy destruction in the gas recompression and treatment sections represented about two thirds of the total exergy destruction of the processing plant. The performance losses were mostly related to the existing compressors (20–25%), coolers (25–30%), and the anti-surge recycling loops (10–15%) in the present study. Similar trends were observed for other platforms, and this pinpoints the importance of improving the gas processing systems.

Considering the high variability of the gas flow rates over the life cycle of the facility, the following measures may be considered. Improving the performance of existing compression trains may be achieved by re-wheeling the compressors or by replacing them with new ones. When designing a new offshore compressing train, it may be interesting to implement compressors that exhibit an acceptable efficiency when they are operated at their maximum capacity and at part-load conditions, rather than ones that present a high efficiency at their design point only. The possibility of designing smaller but parallel trains, to delay the start of off-design operations, may likewise be considered. These measures would significantly improve the performance of the overall plant: (i) they would result in smaller power requirements of the major electricity consumers of the plant (compressors), and (ii) they would reduce the exergy destruction associated with two of the three major sources of performance losses (compression and anti-surge).

Substantial quantities of exergy are either destroyed in the heat transfer process in the coolers or wasted with the rejection of cooling water to the sea. The temperature of the gas streams flowing into the coolers varies between 40 and 160 °C, depending on the compression stage. Heat integration may be limited for these specific streams. The heating demand is generally small on North Sea and Norwegian Sea platforms, and the waste heat discharged from the gas streams is at a too low temperature for use in the condensate reboilers, glycol dehydrators or crude oil heaters. It may be used to satisfy the hot water and space heating needs, and possibly to replace the low temperature electric heaters. Power may be produced by using a low-temperature power cycle, such as an Organic Rankine Cycle, which may be space-consuming, or by using another heat engine such as a Stirling engine. These improvements would result in a smaller exergy destruction in the heaters and coolers of the plant. However, the irreversibilities in the heaters are not significant when put in prospective with the inefficiencies of the other components. Emphasis should therefore be on the ways to use heat from the gas streams. The final coolers of the gas treatment section may be worth investigating, because (i) they usually are responsible for a large amount of destroyed exergy, (ii) display a low exergetic efficiency, and (iii) there may be temperature requirements associated with the gas export pipelines.

*4.2.4. Flaring and venting*

Exergy losses with flared and vented gases are minimal, as gas is mainly flared during shut-down or emergencies, and vented in small quantities from the glycol regenerator. In general, continuous flaring is forbidden on Norwegian platforms and should be avoided, as it results in unnecessary losses. Integration of gas recovery systems may be recommended when building up new facilities.

### *4.3. Significance and limitations*

#### *4.3.1. Energy- and exergy-based analyses*

The present work may be used to support the applicability of energy and exergy-based assessments for evaluating the performance of petrochemical processes, as discussed in Rivero [17]. It was suggested that these thermodynamic analyses may, alongside simulations and optimization tools, indicate potentials for improvements and estimate economic costs of exergy destruction [60,70].

The combination of the findings presented in [18,19,22] and of the present ones (Section 4.1) may be used to predict and estimate qualitatively the main sources of thermodynamic irreversibilities, the extents of the exergy transformations and the exergy use on conventional oil and gas platforms. They also illustrate the similarities between the structural designs, as well as the operating differences from one platform to another. Caution should nevertheless be exercised in extending these conclusions, and further analyses may be conducted (i) to possibly confirm these trends for platforms operating in different oil regions, and (ii) to evaluate the relevance of applying thermodynamic analyses to such systems. An advanced exergetic analysis [71–73] may also be conducted to evaluate the interactions between the various components and sub-systems.

Regarding exergy losses, their economic costs and environmental impacts can also be discussed [74]. It should be emphasised that the amount of exergy losses is not directly related to the possible environmental damages [75,76] and is subject to several assumptions. The choice of the reference state has a direct impact on the numerical results: compositional exergy was determined by considering pure state concentrations, and the choice of another reference point (e.g. concentration in the Earth's crust) returns different numerical values. The choice of the reference temperature (air or water) has also an impact on the absolute values of the thermal exergy lost with cooling water.

Similarly, standard chemical exergies were computed using the model of Szargut [64], and others models (e.g. Kotas [16] and Ahrendts [77]) give different standard chemical exergies for chemical compounds such as water, as the reference humidity level of the atmosphere is different. The setting of the system boundaries also has an impact: considering pressure and temperature losses in the piping and well-head systems can result in a slightly different picture.

#### *4.3.2. Process integration opportunities*

This work gives hints on the behaviour of such systems at severe off-design conditions and on process integration opportunities (Section 4.2). The present findings (i) show differences between platforms operating on young or mature oilfields, (ii) indicate possibilities for achieving a higher efficiency, and (iii) suggest some alternative process routes for current and future offshore platforms. The benefits of such measures may be limited for the production manifold and separation sections, because of lower receiving pressures and higher water cuts, while they may be interesting when it comes to the gas treatment process.

The differences between the platform investigated in this work and in the studies of Oliveira [19] and Voldsund [18] illustrate the variety of the design setups and operating conditions in the oil and gas production sector. The implementation of a separate condensate treatment section resulted in an increased operation flexibility, a smaller power consumption but a greater heating demand, which is met by electric heating and waste heat recovery from the exhaust gases. Process integration is performed to some extent, as the internal heat exchangers between the condensate streams show. Further opportunities should be identified, by conducting a more detailed analysis of the heating and cooling needs, and of the integration of the different sub-systems.

The interactions between the utility system (gas turbines and waste heat recovery) and the processing plant could be studied to assess further possibilities for energy integration and site targets for energy recovery [78–81]. Improvement efforts would result in a higher performance of the processing plant, as the temperatures of the heat sources (e.g. glycol loop) and the heat receivers (e.g oil and gas) would be better matched. The trade-off between additional capital costs and smaller operating ones should be investigated carefully.

*4.3.3. Benchmarking and performance indicators*

These results may give hints for setting meaningful benchmarking and comparing different facilities, as proposed in Voldsund et al. [18]. The specific power consumption and energy efficiency are indicators already in use in the industry [8]. However, the use of the specific power consumption may be misleading because it is expressed on oil equivalent and volume bases. As operating conditions, reservoir properties and export specifications vary from one field to another, a comparison of different facilities using this metric may unfairly favour some plants [9].

Other indicators, such as the exergy efficiency and the specific exergetic waste, could be used along with adequate performance benchmarks. For instance, the exergetic efficiency shows, at the difference of the energy efficiency, the real consumption of the resources required to run the processing plant. The specific exergy consumption, at the difference of the specific power consumption, takes into account the exergy consumed on-site as heat, which may be significant for oil and gas platforms processing a low-temperature or highly viscous crude. The specific exergy waste reflects the losses of resources with the flared and vented gases, but also with the produced water, which is rejected in significant quantities for platforms operating on mature oilfields. It would also be relevant to compare the current power consumption with the power demand when integrating best-available-technologies, as this provides information on the maximum energy savings with the current technological limits [82].

## 5. Conclusion

The processing plant of an offshore platform on a mature oilfield was modelled based on reconciliated data, and its performance was assessed by conducting an energy and exergy analysis. The material and energy flows under two types of production days were derived and validated by comparison with the available measurements. In any case, electric power is mostly consumed to increase the pressure of the produced gas in the recompression and compression sections, while heat is used for enhancing the hydrocarbon separation in the condensate treatment and glycol regeneration systems. The total power consumption amounts to 18.6 MW and can reach 25.5 MW when oil is loaded and exported to the coast. The heating demand is minor in comparison: it is met by electrical heating, heat recovery from a glycol loop and internal heat exchanges.

As suggested in related studies, exergy is mainly destroyed in subsystems where the pressure is significantly reduced (throttling in well-heads and production manifolds) or increased (compression in gas treatment and recompression), because of the turbomachinery inefficiencies and the temperature gaps in the coolers. The exergetic efficiencies ranged between 2–17% for the coolers, exceeded 35% for the heaters, 60% for the compressors and 70% for the large pumps. Exergy is lost to the environment with flared and vented gases, and with produced and cooling water not used for water injection.

Mature conditions of an oilfield have a large impact on the platform performance. Separation work is mostly performed in the gas treatment and condensate sections: it is more significant, on a specific oil and gas equivalent basis, than on similar facilities, because of the higher amount of produced water. Anti-surge recycling around the compressors is practised at a ratio of 17 to more than 65% because of the smaller oil and gas flows entering the plant, leading to higher specific power consumption and greater exergy destruction. Higher water cuts result in a greater water production and to a larger discharge of thermal energy to the environment.

The present paper supports the results of other works, which enhance the usefulness of applying thermodynamic analysis methods on petroleum processes, and highlights the operational challenges of processing plants on mature oilfields. We suggest to (i) extend these analyses to other and similar systems, (ii) develop relevant performance parameters for comparing offshore facilities, and (iii) take into consideration such results when designing future platforms.

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### Nomenclature

| | |
|---|---|
| $T$ | Temperature, K |
| $\bar{e}$ | specific exergy, J/mol |
| $\dot{E}$ | Exergy rate, W |
| $e$ | specific exergy, J/kg |
| $i$ | chemical compound |
| $j$ | stream |
| $k$ | component |
| $p$ | pressure, Pa |
| $s$ | specific entropy, J/(kg·K) |
| $w$ | specific power consumption, W |
| $y$ | component/sub-system exergy ratio |
| *Abbreviations* | |
| API | American Petroleum Institute |
| EOS | Equation of State |
| PP | Processing Plant |
| TEG | Triethylene Glycol |
| UT | Utility Plant |
| *Greek letters* | |
| $\Delta$ | difference |
| $\delta$ | uncertainty (95% confidence level) |
| $\eta$ | energy efficiency |
| $\iota$ | intensity |
| $\omega$ | waste |
| $\sigma$ | standard deviation |
| $\varepsilon$ | exergy efficiency |
| *Superscripts* | |
| * | relative |
| ch | chemical |
| kn | kinetic |
| m | mechanical |
| ph | physical |
| pt | potential |
| t | thermal |
| *Subscripts* | |
| 0 | dead state |
| $\hat{h}$ | energy-based |
| $\hat{x}$ | exergy-based |
| $d$ | destruction |
| $f$ | fuel |
| $l$ | loss |
| $p$ | product |
| cool | cooling medium |
| feed | feed |
| in | inlet |
| mix | mixture |
| o.e | oil equivalent |
| out | outlet |
| tot | total |