Thermoeconomic analysis of an oil separation process on offshore platform

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Abstract:
Petroleum is one of the major sources of energy in the world and its consumption is destined to increase in the next years. At the same time, the reduced availability of conventional reservoirs leads oil companies seeking for offshore reservoirs, with an increase of the environmental impact, costs and risks associated with oil exploration and exploitation.

Oil platforms are complex systems where different processes occur and energy production is required. This work presents the analysis of a crude oil separation process and is divided into two parts. In the first part, a simplified model of a separation plant is described: the system is constituted of a pre-separator unit, an Electrostatic Oil Heater Treater (EOT) unit and one post-separator unit. Configuration and data are obtained from a real plant and simulated in Aspen Plus.

In the second part, Exergy and Thermoeconomic analyses of each unit are performed. The two analyses take into account the irreversibility of the process and the economic costs.

The exergy analysis shows that there is a great amount of heat discharged from the EOT to the environment, indicating that the process is far from being thermodynamically perfect, the Thermoeconomic analysis that the investment and O&M costs have a higher impact compared to exergy destruction.

As a novelty within the paper, the whole set of thermoeconomic variables are here evaluated, i.e. unit thermoeconomic cost of resource and product, relative cost difference and exergoeconomic factor. Their study leads to a possible optimization of the units able to reduce fuel consumption, but results show however that the improved exergy efficiency remains under the 10% for the EOT and the 20% for the overall system, with a relatively small cost reduction. Nonetheless, the reduction of fuel in the optimized configuration can be relevant in terms of avoided CO_2 emissions.

Therefore, this work brings out the importance of exergy analysis as a tool that leads to several system improvements, even beyond the thermodynamic point of view. In this perspective, exergy-based analyses can support plants refinements and optimization both from the economic (i.e. reduction of costs) and the environmental perspective (i.e. reduction of emissions).

Keywords:
Exergy cost, Thermoeconomic analysis, Oil platforms.

1. Introduction

Petroleum is one of the major sources of energy in the world and its consumption is destined to increase in the next years [1]. At the same time, the reduced availability of conventional reservoirs leads oil companies seeking for offshore reservoirs, with an increase of the environmental impact, costs and risks associated with oil exploration and exploitation. An efficient use of resources can drive to a reduction in the cost of oil treatment plants and to less significant environmental impacts, through the study of inefficiencies and irreversibility of the process [2].

In most of oil fields worldwide, the oil produced is accompanied by water, dissolved salts and other components. These components can cause considerable operational problems, therefore, desalting, dehydration and demulsification plants are installed in production units to remove water and inorganic particles from oil streams [3]. According to literature [4], the components in crude oils, such as inorganic particles, natural surfactants and asphaltenes, contribute to define the W/O ratio (water in oil) of the oil [5]. The separation of these emulsions is crucial in the process, since their presence is critical to satisfy the quality requirements of the product.

Several techniques ([5] and [6]) are used in order to achieve these requirements: chemical demulsification ([7] and [8]), pH adjustment [6], gravity or centrifugal settling [10], filtration [6],

heating treatment [5]. One of the most effective and utilized methods in industry is electrostatic-demulsification ([10], [11] and [12]).

The system analyzed in this work is a dual polarity Electrostatic Oil Treater (EOT). The application of electrical methods for dehydrating crude oil emulsions is already present in literature and has been followed up ([13] and [14]).

The EOT analyzed is equipped with two firetubes for partially heating the crude oil and separate the components. A typical Electrostatic Oil Treater is shown in

Fig. 1: it can be conceptually divided into two parts, the heating section and the coalescing section. Initially, the crude oil emulsion enters the Electrostatic Oil Treater and flows downward over a hood baffle. Free water is not heated and separates immediately, while oil emulsion is heated as it goes in contact with the firetube. As temperature increases, larger water droplets coalesce and separate, while smaller water droplets flow with the oil into the coalescing section. Remaining water droplets are subjected to the AC/DC electrostatic fields, causing them to coalesce and settle to the bottom of the vessel [15].
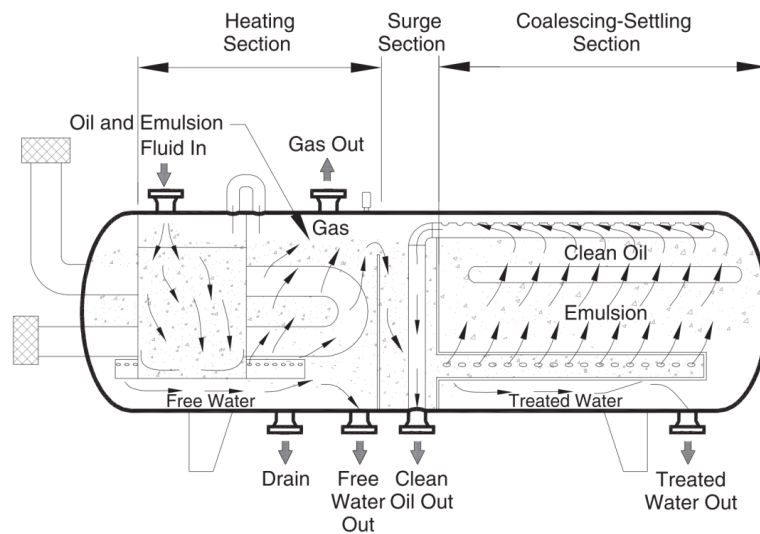

Fig. 1 Electrostatic Oil Treater [14]

The dual polarity electrostatics provide a complete dehydration, and consequently it can process at high viscosities, which means that less heat is required to lower the viscosity of the oil at processing conditions [15]. Therefore the dual polarity treater is ideal for offshore production platforms.

2. Model of the crude oil separation process

In this work, the dual polarity Electrostatic Oil Treater (EOT) is modelled by two separated components: a part related to the heating process of crude oil (BURNER) and a second part related to the separation process (OILTR). The model of the system is represented in Fig. 2. A private company provided the input set of values of mass flow rates, temperatures and pressures for each stream involved in the process. Aspen Plus® has been used to simulate the model and calculate the remaining thermodynamic properties of the each stream.

The main stream of crude oil (MIX1) enters into a pre-separator (SEP1) and is divided in a first flow of gas (GAS1) and in a second main stream of oil (MIX2). From SEP1, a heat stream (QSEP1) exits. The pre-separated crude oil (MIX2) enters the separation section (OILTR) in the electrostatic treater. From the separation section, a heat flux (QOILTR), and two streams, one composed by gas (GAS2) and the other by waste water (WWAT), are generated.

In the upper section of the system, the stream GAS2 is mixed in the mixer (MIXER) with the stream GAS1, coming from the pre-separator unit, to obtain a non - compressed gas stream (GAS3). This stream, passing through the compressor (C2), is compressed and it finally generates the final compressed gas stream (GAS). The compressor receives a workflow indicated by the label WC2.

In the lower section of the system, instead, the stream of water (WWAT) enters into the post-separator (SEP2) to obtain a stream of water (WATER) separated from the stream of waste (WASTE). From the post separation (SEP2), a heat flow (QSEP2) exits.

The heating section (BURNER) simulates the combustion that takes place into the firetube and therefore requires a mass flow rate of fuel (BIN) and an air mass flow rate (AIR).
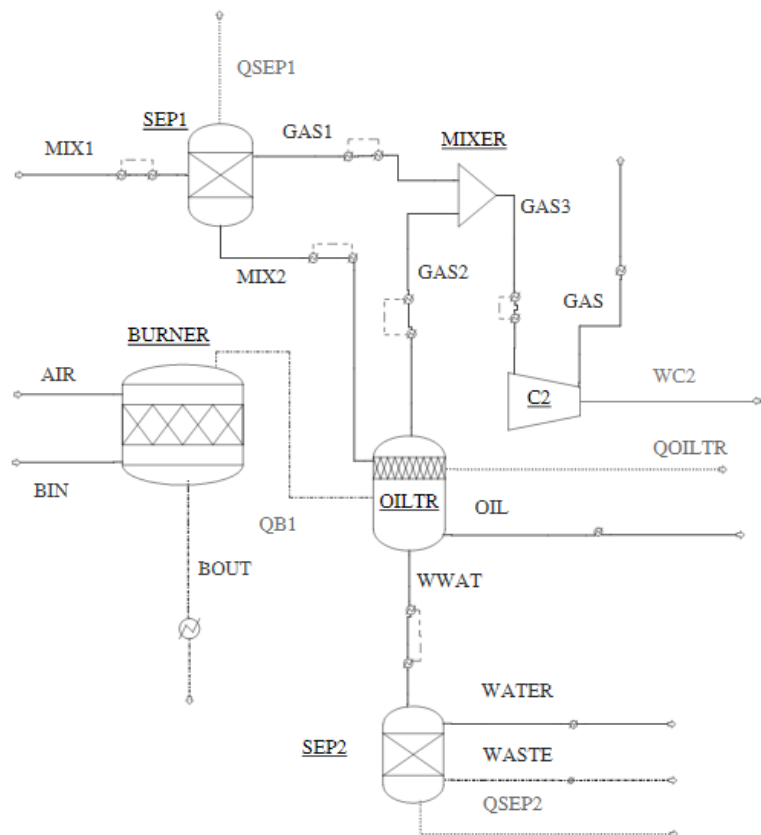

Fig. 2 Separation of Crude Oil in Aspen Plus® simulation.

The outputs of the BURNER are combusted gases (BOUT), that flow in the stack of the firetube, and the heat flow (QB1), that is supplied to the separation section and simulates the portion of the heat that is transferred from the firetube to the stream MIX2. Heat fluxes QSEP1, QSEP2 and QOILTR are assumed to be exchanged with the environment and therefore they are not reused.

Target of the simulation is to find the composition, temperature and mass flow rate of the streams of fuel (BIN), combusted gases (BOUT) and air (AIR).

To complete the set of data for the simulation, the following assumptions are made:
- The fuel (BIN) is assumed to be composed by methane (CH_4) at temperature and pressure of the environment.
- The air stream (AIR) is considered for simplicity constituted for 21% of O_2 and for 79% of N_2, with the same inlet temperature and pressure of the fuel.
- For the excess air a_{air}, a range of values between 10% and 18%, adopted for natural gas, is taken.
- The combustion is supposed to be complete, with negligible content of NO_x in combusted gas (BOUT).
- The temperature of flue gas in the stack ($T_{out,stack}$) is taken in the range of 450-550 ° C.

The simulation is therefore in two variables (a_{air} and $T_{out,stack}$).

The thermodynamic properties and the composition of streams obtained are shown in Appendix.

For the aim of precision, it is possible that other unknown streams are present in the system, but since they are not classified as products nor fuel, they are considered as inert or waste flows.

3. Exergy analysis of the crude oil separation process

As emphasized by Rivero [18], the application of the concept of exergy in the petroleum industry provides more detailed and consistent information on the performance of petrochemical systems. Unlike enthalpy, exergy is rarely handled by process simulators, hence several solutions have been developed ([20]-[21] and [22]). Hinderink [20] produced ExerCom, a calculator of exergy material streams for Aspen Plus®. Montelongo-Luna [22] produced an open-source tool for exergy calculation of material streams for the software Sim42.

Recently, Querol [23] developed a Microsoft Excel-based exergy calculator for Aspen Plus®, which facilitates the thermoeconomic and exergetic analysis. It calculates exergy of work, heat and material flows, while the exergy of mixing process is included in the calculation of the physical exergy [24]. This work follows the methodology implemented by Querol for the exergy and thermoeconomic analysis. A Microsoft® Excel workbook is implemented for all the operations and results, using Microsoft® Excel formulae.

The expression for the exergy balance of the system is derived by applying the energy and entropy balances for open systems [26]:

$$\dot{E}_d = \sum \dot{E}_{in} - \sum \dot{E}_{out} \equiv \sum \dot{m}_{in} e_{in} - \sum \dot{m}_{out} e_{out} + \sum \left(1 - \frac{T_0}{T_k}\right)\dot{Q}_k - \dot{W} \qquad (1)$$

In (1), $\sum \dot{E}_{in}$ is the sum of all the exergy streams entering the system, $\sum \dot{E}_{out}$ is the sum of all the exergy streams leaving the system and $\dot{E}_d$ is the exergy destroyed inside the system. The symbol $\dot{m}$ denotes the mass flow rate of a stream, $\dot{Q}_k$ and $\dot{W}$ the rates of energy transfer by heat and work and e the specific exergy. The symbols T_0 and T_k denote the environmental temperature and the local temperature where heat transfer takes place. The subscripts *in* and *out* denote the inlet and outlet of the system and k the boundary of the system. The exergy destruction rate can be calculated from the Gouy-Stodola theorem:

$$\dot{E}_d = T_0 \dot{S}_{gen} \qquad\qquad\qquad (2)$$

In (2) the entropy generation rate inside the system is indicated by $\dot{S}_{gen}$.

Physical exergy accounts for temperature and pressure differences between the actual state of the system and the environment, and is defined as:

$$e^{ph} = (h - h_0) - T_0(s - s_0) \qquad (3)$$

Where h and s are the specific enthalpy and entropy of a stream of matter, respectively.

Chemical exergy takes into account for the difference in chemical composition of the system from the reference substances present in the environment. In this work, chemical exergy is calculated using the reference environment defined by Szargut [24]. The molar chemical exergy of a given mixture e^{ch}_{mix} is expressed as [37]:

$$e^{ch}_{mix} = \underbrace{\sum_i x_i e^{ch}_{i,mix}}_{\textbf{I}} = \underbrace{\sum_i x_i e^{ch}_{i,0}}_{\textbf{II}} + \underbrace{\left(\sum_i x_i \left(h_{i,mix} - h_{i,0}\right)\right) - T_0\left(\sum_i x_i \left(s_{i,mix} - s_{i,0}\right)\right)}_{\textbf{III}} \qquad (4)$$

where the molar fraction, the chemical component and the mixture are denoted by x, i and mix, respectively. The molar exergy of a given chemical component is indicated as $e_{i,\mathrm{mix}}^{\mathrm{ch}}$ when it is part of the mixture, $e_{i,0}^{\mathrm{ch}}$ when it is a pure component. The term I illustrates the chemical exergy of each individual chemical component in the mixture, the term II the chemical exergy of these components in an unmixed form and the term III the reduction in chemical exergy due to mixing effects. If no chemical transformations are taking place within a separation process like the one analyzed, the terms related to the chemical exergy of pure components are null and the change of chemical exergy corresponds to the exergy used to perform the separation work [19].

Following Querol's approach [23], the term III is considered part of the physical exergy e^{ph} [29], since the value of entropy of the stream calculated with Aspen Plus® already includes the mixing effect. On the consequence, the term I simplifies between input and output streams.

To calculate the thermodynamic properties of the streams involved in the combustion process, the expression used is the generic combustion reaction of a typical hydrocarbon:

$$C_n H_m + \left(n + \frac{m}{4} \right) O_2 \rightarrow nCO_2 + \frac{m}{2} H_2 O \qquad (5)$$
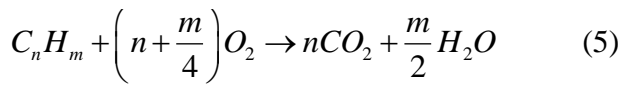
To perform the calculations, the conventional reference environment ($T_0 = 298.15$ K and $P_0 = 1$ atm) is considered. Therefore, the thermodynamic properties of the different components of the streams are those tabulated by Szargut in [25] and are shown in Appendix.

Finally, work is considered pure exergy, therefore the value calculated with Aspen Plus® is used directly as the exergy of each work stream:

$$\dot{EX}_W = \dot{W} \qquad (6)$$

Instead, the exergy of a heat flow is given by (7):

$$\dot{EX}_q = \dot{Q} \left(1 - \frac{T_0}{T} \right) \qquad (7)$$

In this work, kinetic and potential contributions on the flow exergy are assumed to be negligible compared to physical and chemical exergy.

The expression for the exergy efficiency used, denoted by η_{ex}, is therefore given by:

$$\eta_{ex} = \frac{\dot{E}_p}{\dot{E}_f} = \frac{\sum_k \dot{EX}_{k,u} - \dot{EX}_{feed}}{\dot{EX}_{comb} + \dot{EX}_W} \qquad (8)$$

where $\dot{EX}_{feed}$ is the exergy of the stream entering the system, and $\dot{EX}_{comb}$ is the exergy of the fuel mixture (air plus combustible) provided to the combustion process.

Hence, the exergy efficiency of the separation plant of crude oil simulated becomes:

$$\eta_{ex} = \frac{\dot{EX}_{OIL} + \dot{EX}_{GAS} + \dot{EX}_{WATER} - \dot{EX}_{MIX1}}{\dot{EX}_{BIN} + \dot{EX}_{AIR} + \dot{EX}_W} \qquad (9)$$

A difficulty that emerges from the application of the exergy efficiency when evaluating petroleum separation processes originates from the fact that these formulations include the chemical exergy of

hydrocarbons, that is usually two or three (or even more) orders of magnitude bigger than the molar physical exergy.

To solve this problem, if no chemical transformations occur in the system, the chemical exergy of a stream is simplified between input and output, since the exergy of mixing effect is already included in the physical exergy calculated. This occurs for the part of separation. In the heating section this consideration does not apply instead, since there is the combustion reaction.

This approach is taken from the concept of transit exergy defined by Brodyanksky [31]. According to this concept, the transit exergy is the part of the exergy supplied to the system that flows through it without undergoing any physical or chemical transformation. This is one of the approaches present in literature, since at the level of the process, a unique formulation may not be available, as explained by Voldsund [32].

From this consideration, physical exergy of streams is calculated only in those components of the system where chemical transformations do not occur (i.e. no chemical transformation occur in all the components but in the combustion chamber).

4. Thermoeconomic analysis of the crude oil separation process

To evaluate the cost required to produce a general stream of exergy j in the system, a thermoeconomic cost, $\dot{\Pi}_j$, can be assigned. Following the theory of Thermoeconomics [27,34-37], the total cost of the outgoing streams (i.e. products) from a process, $\dot{\Pi}_P$, is equal to the total cost of the incoming streams (i.e. resources), $\dot{\Pi}_R$, plus the value of the fixed costs flow, $\dot{Z}$, that takes into account the capital investments and operation and maintenance (O&M) costs of the process equipment.

Applying the cost balance to each piece of equipment, the equation (10) is obtained:

$$\dot{\Pi}_{P,i} = \dot{\Pi}_{R,i} + \dot{Z}_i \ (10)$$

$\dot{Z}$ can be divided into its components to obtain:

$$\dot{Z}_j = \dot{Z}_j^I + \dot{Z}_j^{OM} \ (11)$$

Where $\dot{Z}_j^I$ is the investment costs flow and $\dot{Z}_j^{OM}$ is the operation and maintenance costs flow. For the economic analysis, the total production cost in each process component has to be evaluated, obtained by the calculation of:

1. Total investment cost (TCI): the TCI takes into account all the investment costs of the plant, such as the purchase cost of equipment (PCE), the installation costs and other type of cost as contingencies and financial charges. The TCI can be estimated through the Lang Factor coefficient, if the PCE of each piece of equipment is known [37]. The PCE of each component is shown in Table 1 and has been provided by an oil company.

Table 1: PCE of each component.

	€
Sep1	4.45E+05
Mixer	-
C2	2.87E+04
Sep2	4.45E+05
EOT	1.15E+06
Total	2.06E+06

It is possible to find in literature the Lang Factor coefficient relative to a similar plant [37]. The Lang Factor is defined as the ratio of PCE and TCI:

$$LF = TCI\big/PCE \qquad (12)$$

In this work, LF has been assumed equal to 3, from Bejan and Moran [27].

2. Annuity (A): once TCI is calculated, the annuity A in which the TCI is amortized over the useful life of the plant is calculated:

$$A = TCI \cdot \underbrace{\frac{a(1+a)^y}{(1+a)^y - 1}}_{RF} \qquad (13)$$

where a is the interest rate, y the numbers of years and RF the capital recovery factor. The value assumed for a and y are 10% [38] and 20 years [39], respectively.

3. O&M costs: if operation and maintenance costs are not known, they can be estimated by the factor f_{OM} taken from the literature data [37]:

$$O\&M = f_{OM} \cdot TCI \qquad (14)$$

The value of f_{OM} adopted is 1,40% [40].
The cost of the methane has been estimated equal to 0.02 $/m³, the cost of the oil mixture at the inlet of system 30$/barrel.

4. Annual cost flow (FCⱼ): the annual cost flow FC_j of each piece of equipment is obtained knowing the O&M costs, A and the TCI, and weighing them on the factor f_{PCE}:

$$FC_j = (A + OM) \cdot \underbrace{PCE_j \big/ \sum_j PCE_j}_{f_{PCE}} \qquad (15)$$

where j indicates the j-th piece of equipment.

Finally, the fixed cost flow of each piece of equipment $\dot{Z}_j$ is obtained as:

$$\dot{Z}_j = \frac{FC_j}{OT} \qquad (16)$$

where OT is the annual operating time (8760 h/y).

In a thermoeconomic evaluation and optimization, the information on stream costs is used as information to calculate the thermoeconomic variables associated with each component of the system [27]:

1. The unit thermoeconomic cost of the resource, that expresses the average cost at which the resource (i.e. the inlet exergy streams) is supplied to the component:

    $$C_R = \dot{\Pi}_R / \dot{R} \qquad (17)$$

2. The unit thermoeconomic cost of the product, that expresses the average cost at which the component generates the product (i.e. the outlet exergy streams):

    $$C_P = \dot{\Pi}_P / \dot{P} \qquad (18)$$

3. The relative cost difference, that expresses the relative increase in the average cost per unit of exergy between input and output of the component:

    $$r = \frac{C_{P,i} - C_{R,i}}{C_{R,i}} \qquad (19)$$

4. The exergoeconomic factor, which represents a measure of the contribution of fixed costs $\dot{Z}_i$ to the absolute cost increase originated in the equipment

    $$f_i = \frac{\dot{Z}_i}{\dot{\Pi}_{c,i}} \qquad (20)$$

5. Results and discussion

In this section, the results of the calculations are shown. The input and output flows for each piece of equipment are tabulated in Table 2.

Table 2: Input-output table of all flows of the plant.

	Input	Output
SEP1	MIX1	MIX2. GAS1, QSEP1
SEP2	WWAT	WATER, WASTE, QSEP2
MIXER	GAS1, GAS2	GAS3
C2	GAS3, WC2	GAS
EOT	MIX2, BIN, AIR	WWAT, OIL, GAS2, QOILTR
Plant	MIX1,BIN, AIR, WC2	OIL, GAS, WATER, WASTE, QSEP1, QSEP2, QOILTR

Results of Exergy analysis

In Table 3 the results of the exergy evaluated for each stream is represented. From the exergy balance formulated in (1), the exergy destruction for each component is evaluated and shown in Table 4.

Table 3 Total exergy of the different streams.

	Units	AIR	BIN	BOUT	GAS	GAS1	GAS2	GAS3	MIX1	MIX2	OIL	WWAT	WATER	WASTE
e^{ph}	J/kmol	0.0E+00	0.0E+00	7.4E+06	5.2E+06	3.9E+06	2.3E+06	2.3E+06	3.4E+06	6.8E+05	7.6E+05	2.2E+05	2.2E+05	2.3E+05
e^{ch}	kJ/kmol	1.4E+03	8.4E+05	2.8E+03	1.3E+06	8.4E+05	1.3E+06	1.3E+06	5.4E+06	5.4E+06	6.1E+06	3.1E+03	3.1E+03	3.2E+03
e^{tot}	kJ/kmol	1.4E+03	8.4E+05	1.0E+04	1.3E+06	8.4E+05	1.3E+06	1.3E+06	5.4E+06	5.4E+06	6.1E+06	3.4E+03	3.3E+03	3.4E+03
E_{tot}	W	2.3E+04	1.2E+06	1.8E+05	1.3E+07	7.0E+05	1.2E+07	1.3E+07	7.5E+08	7.4E+08	7.3E+08	2.4E+04	1.9E+04	4.9E+03

Table 4: Exergy destruction for each component.

$Ex_{destr.}$	W	%
Sep1	3.67E+05	26.0
Mixer	1.51E+03	0.1
C2	5.20E+03	0.4
Sep2	1.02E+01	0.0
EOT	1.04E+06	73.8
Total	1.41E+06	100.0

From Table 4 it can be evinced that the EOT has the highest share of exergy destruction, close to the 74% of the total exergy destruction. The pre-separator SEP1 contributes for the remaining 26%, while the MIXER, the compressor and the post separator SEP2 have negligible contribution. These values of exergy destruction affect the exergy efficiency. For the EOT, indeed, the result of exergy efficiency calculation gives a value of 1,7%, indicating that the exergy destruction in the component is extremely relevant. For the whole system the exergy efficiency is 8,34%, a value in line with other studies present in literature [32].

Results of Thermoeconomic analysis

The results of the Thermoeconomic analysis are shown in Fig. 3 and are expressed in €/h. The EOT presents the highest cost flow since its PCE is the highest and it represents the 55% of the entire PCE_{tot}.
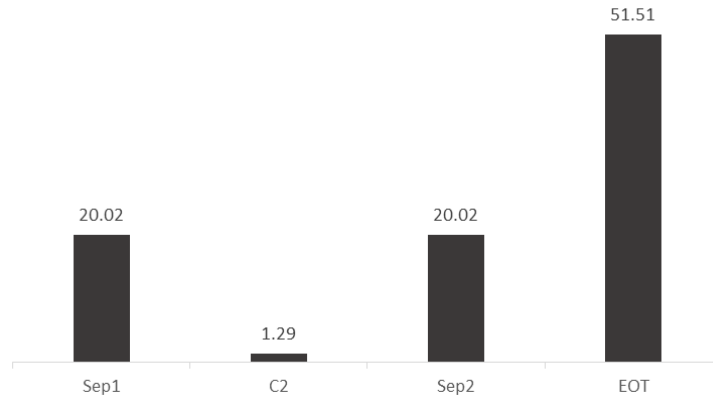
Fig. 3 Cost flow Zj [€/h] of each component

Table 5 summarizes the thermoeconomic variables calculated for each component of the oil separation process. The variables include the cost flow $\dot{Z}$, the specific thermoeconomic cost of resource c_R, the specific thermoeconomic cost of product c_P, the relative cost difference r, and the exergoeconomic factor f.

Table 5: Thermoeconomic variables for each piece of equipment in all the configurations.

	$\dot{Z}$ [€/h]	Π_R [€/s]	Π_P [€/s]	c_r [€/MWh]	c_p [€/MWh]	r_i	f
Sep1	2.00E+01	2.61E+00	2.62E+00	20281.90	97634	3.81	2.68E-03
C2	1.29E+00	1.07E-03	1.43E-03	112.43	177	0.57	6.88E-01
Sep2	2.00E+01	3.56E-02	4.12E-02	79876.98	115427	0.45	4.38E-01
EOT	5.15E+01	7.10E-04	1.50E-02	2.06	2510	1219.91	9.53E-01

From the results of the analysis, some considerations can be developed:

- the exergy analysis shows that there is a great amount of heat discharged from the EOT to the environment, indicating that the process is far from being thermodynamically perfect. Among internal exergy destruction there are the heat transformation from high temperature zone (radiant section in the firetube) to lower temperature zone (convection section) and the combustion. This exergy destruction cannot be reduced through simple exergy analysis. On the contrary, the combusted gas BOUT outgoing the firetube is an external exergy loss.
- the Thermoeconomic analysis, instead, helps to understand that the EOT, the pre-separator SEP1, and post-separator SEP2 have the highest values of exergoeconomic factor f and relative cost difference r. Consequently, they are the important components from the point of view of thermoeconomic analysis. In particular, the EOT has the highest values. The high values of these variables, and in particular of f, mean that the investment and O&M costs have a higher impact compared to exergy destruction. Since the investment cost is far greater than the O&M, the PCE has the greatest influence in the thermoeconomic evaluation.

Starting from these considerations, Authors investigated how reducing the external exergy losses and improve the efficiency of the system. It is then proposed to modify the process by reducing the mass

flow rate of the fuel (BIN) in input in the firetube (QBURNER), so that the heat flow QOILTR discharged to environment by the EOT is reduced to zero and external exergy losses are minimized. Performing with Aspen Plus® an optimization on the fuel mass flow rate BIN, with the condition of QOILTR equal to zero, the value decreases from 2.3E-02 kg/s to 4.8E-03 kg/s. This value of fuel mass flow rate represents therefore a lower boundary of fuel in input.

The new process configuration obtained by reducing the fuel flow rate is named Optimized Configuration, to differentiate it from the configuration previously described, named, from now on, Design Configuration.

In the Optimized Configuration, the exergy destruction in the EOT decreases to the value of 1.96E+05 W, one order of magnitude smaller compared to the initial value of 1.04E+06 W. This leads to an increase of the exergy efficiency of the EOT and of the overall system, as shown in Table 6. The increased exergy efficiency obtained represents therefore an upper boundary, since all the other variables of the process do not change and they are fixed by the conditions imposed for the separation.

Table 6 Exergy efficiency and comparison between Design Configuration and Optimized Configuration

	Design Conf.	Opt. Conf.
EOT	1.73	8.45
Overall System	8.34	19.28

It is interesting to notice that the possible improvement of the exergy efficiency remains however under the 10% for the EOT and the 20% for the overall system. This remarks the relevance of internal exergy destruction of the processes considered. Concerning the Thermoeconomic analysis of the Optimized Configuration, there is a little sensitivity in the value of the exergoeconomic factor f, due to the high PCE. The reduction of fuel leads f to increase from the 95,3% to the 99,2%. This consequence is logical, since the reduction of fuel allows reducing the term $\dot{\Pi}_{c,\mathrm{EOT}}$ in (20), thus increasing the final value of f.

A further consideration can be done regarding the absolute cost of equipment of the EOT, $\dot{\Pi}_{c,\mathrm{EOT}}$.

Indeed, it is possible to identify a gain by shifting from the Design Configuration to the Optimized Configuration. The absolute cost decreases of the EOT from 54,02 €/h to 51,99 €/h, with a difference of about 2 €/h, corresponding to 3,7%. This is equal to 355.045 € along the operation life of the processing plant, as a result of less consumption of fuel. The annual cost saving relative to this optimization, however, is very small compared to the annual fixed costs of the overall plant. The relative difference, inedeed, is only 2,18%. Nonetheless, the reduction of fuel between the Optimized Configuration and the Design Configuration can be relevant from the environmental point of view. Indeed, the total amount of CO_2 emissions avoided can be calculated by means of (5) and results 32.220 ton, corresponding to 79,5% of the CO_2 emissions in the EOT. This consideration reveals the opportunity to reduce also the cost of externalities (i.e. CO_2 emissions) by thermodynamic optimization.

6. Conclusion

In this work, the process of crude oil separation is analyzed from two different points of view: the thermodynamic one, by means of exergy analysis, and the economic one, through a Thermoeconomic analysis. In particular, the focus is on the electrostatic oil treater, a component mainly used to separate crude oil on offshore platforms. The electrostatic oil treater results to be the most important component of the process in terms of low efficiency and high costs.

The exergy analysis indicates that the exergy efficiency of the oil separation process is very low when the chemical exergy of streams is not considered and only physical exergy is measured. Results highlight the relevance of internal losses, since the improvement the efficiency that can be realized

by minimizing external losses is less than 20%. From the Thermoeconomic analysis it emerges that the capital costs are the main drivers in the economic assessment: in the case of the electrostatic oil treater, the exergoeconomic factor, that represents the relevance of fixed capital costs versus variable costs, is higher than 95%. The incidence of capital costs is visible also in the economic gain of only 2% obtainable from the optimization of variable costs.

However, an interesting result is obtained by assessing the CO_2 emissions that can be avoided by thermodynamic optimization. Results show that the reduction obtainable in the heating section is close to the 80%.

In conclusion, this work highlights the relevance of the possible effects obtainable by thermodynamic improvement of processes relatively to economics and emission reduction, and the usefulness of exergy and Thermoeconomic analyses to assess such effects.

Acknowledgments

Appendix

Table 7 Selected data of reference environment from Szargut [32].

T_0	25 °C
P_0	101325 kPa
Ref substance	Partial pressure
CO_2	3.35E-02
H_2O	2.20E+00
N_2	7.58E+01
O_2	2.04E+01
Ar	9.00E-01
D_2	3.42E-04
He	4.85E-04
Kr	4.85E-04
Ne	1.77E-03
Xe	9.00E-08

Table 8 Thermodynamic properties of streams[1] [14].

	Units	AIR	BIN	BOUT	GAS	GAS1	GAS2	GAS3	MIX1	MIX2	OIL	WWAT	WATER	WASTE
Mole Flow	kmol/s	1.6E-02	1.5E-03	1.8E-02	1.0E-02	8.3E-04	9.2E-03	1.0E-02	1.4E-01	1.4E-01	1.2E-01	7.2E-03	5.8E-03	1.4E-03
Mass Flow	kg/s	4.7E-01	2.3E-02	5.0E-01	2.8E-01	1.3E-02	2.6E-01	2.8E-01	1.6E+01	1.6E+01	1.6E+01	1.3E-01	1.0E-01	2.6E-02
Temp.	K	3.0E+02	3.0E+02	8.1E+02	4.0E+02	3.3E+02	3.4E+02	3.4E+02	3.3E+02	3.3E+02	3.4E+02	3.4E+02	3.4E+02	3.4E+02
Pressure	N/sqm	1.0E+05	1.0E+05	1.0E+05	6.0E+05	4.8E+05	2.4E+05	2.4E+05	4.8E+05	4.8E+05	2.4E+05	2.4E+05	2.4E+05	2.4E+05
Molar Enthalpy	J/kmol	-1.3E+04	-7.5E+07	-5.5E+07	-9.4E+07	-7.4E+07	-1.0E+08	-9.8E+07	-2.5E+08	-2.5E+08	-2.6E+08	-2.8E+08	-2.8E+08	-2.8E+08
Enthalpy Flow	W	-2.1E+02	-1.1E+05	-9.9E+05	-9.4E+05	-6.1E+04	-9.2E+05	-9.8E+05	-3.4E+07	-3.4E+07	-3.1E+07	-2.1E+06	-1.6E+06	-4.1E+05
Molar Entropy	J/kmol-K	4.2E+03	-8.1E+04	3.2E+04	-1.4E+05	-9.0E+04	-1.5E+05	-1.4E+05	-8.1E+05	-8.1E+05	-8.9E+05	-1.6E+05	-1.6E+05	-1.6E+05
h_0	J/kmol	-1.3E+04	-7.5E+07	-7.8E+07	-1.0E+08	-7.5E+07	-1.0E+08	-1.0E+08	-2.4E+08	-2.6E+08	-2.7E+08	-2.9E+08	-2.9E+08	-2.9E+08
s_0	J/kmol-K	4.2E+03	8.1E+04	-2.0E+04	-1.4E+05	-8.1E+04	-1.5E+05	-1.4E+05	-7.6E+05	-8.3E+05	-9.3E+05	-1.7E+05	-1.7E+05	-1.7E+05
$h-h_0$	J/kmol	0.0E+00	0.0E+00	2.3E+07	6.3E+06	1.2E+06	3.1E+06	2.8E+06	-1.1E+07	7.7E+06	1.1E+07	3.4E+06	3.4E+06	3.4E+06
$s-s_0$	J/kmol-K	0.0E+00	0.0E+00	5.2E+04	3.7E+03	-9.1E+03	2.6E+03	2.0E+03	-4.7E+04	2.4E+04	3.5E+04	1.1E+04	1.1E+04	1.1E+04

Table 9 Composition of streams provided by the private company

Molar fraction (kmol/s)	AIR	BIN	BOUT	GAS	GAS1	GAS2	GAS3	MIX1	MIX2	OIL	WASTE	WATER	WWAT
METHANE	0.00E+00	1.00E+00	0.00E+00	3.75E-01	1.00E+00	3.19E-01	3.75E-01	3.00E-02	2.41E-02	3.20E-03	1.20E-05	0.00E+00	2.40E-06
TOLUENE	0.00E+00	0.00E+00	0.00E+00	8.08E-03	0.00E+00	8.81E-03	8.08E-03	6.00E-02	6.04E-02	6.79E-02	2.19E-07	0.00E+00	4.38E-08
WATER	0.00E+00	0.00E+00	1.63E-01	9.51E-02	0.00E+00	1.04E-01	9.51E-02	1.00E-01	1.01E-01	4.65E-02	1.00E+00	1.00E+00	1.00E+00
C6H14-01	0.00E+00	0.00E+00	0.00E+00	1.25E-02	0.00E+00	1.36E-02	1.25E-02	3.00E-02	3.02E-02	3.32E-02	1.11E-10	0.00E+00	2.22E-11
C8H18-01	0.00E+00	0.00E+00	0.00E+00	9.33E-03	0.00E+00	1.02E-02	9.33E-03	1.40E-01	1.41E-01	1.59E-01	0.00E+00	0.00E+00	3.52E-14
C9H20-01	0.00E+00	0.00E+00	0.00E+00	1.21E-02	0.00E+00	1.32E-02	1.21E-02	4.40E-01	4.43E-01	5.02E-01	0.00E+00	0.00E+00	1.54E-15
C11H2-01	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
C15H3-01	0.00E+00	0.00E+00	0.00E+00	1.90E-05	0.00E+00	2.07E-05	1.90E-05	1.20E-01	1.21E-01	1.37E-01	0.00E+00	0.00E+00	1.37E-21
C20H4-01	0.00E+00	0.00E+00	0.00E+00	6.63E-08	0.00E+00	7.22E-08	6.63E-08	2.00E-02	2.01E-02	2.29E-02	0.00E+00	0.00E+00	2.29E-22
C30H6-01	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
C3H8	0.00E+00	0.00E+00	0.00E+00	5.19E-02	0.00E+00	5.66E-02	5.19E-02	1.00E-02	1.01E-02	7.13E-03	4.86E-07	0.00E+00	9.71E-08

[1] *Values of mass flow rate, pressure and temperature of each stream have been provided by the private company. All the other properties have been calculated by means of Aspen Plus®.*

C2H6	0.00E+00	0.00E+00	0.00E+00	4.36E-01	0.00E+00	4.75E-01	4.36E-01	5.00E-02	5.03E-02	2.10E-02	2.09E-05	0.00E+00	4.18E-06
CO2	0.00E+00	0.00E+00	8.17E-02	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
N2	7.90E-01	0.00E+00	7.25E-01	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00

Nomenclature

EOT Electrostatic Oil Treater
BURNER heating section of the Electrostatic Oil Treater
OILTR separation section of the Electrostatic Oil Treater
MIX1 main flow of crude oil
MIX2 second main flow of oil
BIN fuel
BOUT flue gases
QB1 heat flow from burner
SEP1 pre separator
SEP2 post separator
WWAT waste water flow
C2 compressor
$T_{out,stack}$ temperature of flue gas in the stack
a_{air} excess air
η_{ex} functional exergy efficiency
$\dot{\Pi}_j$ thermoeconomic cost of exergy stream j
$\dot{\Pi}_P$ total cost of the outgoing exergy streams from a process
$\dot{\Pi}_R$ total cost of the incoming exergy streams in a process
TCI total cost of investment
PCE purchase cost of equipment
LF Lang Factor
A annuity
RF recovery factor
FC annual cost flow
OT annual operating time
$\dot{R}$ incoming streams in a process
$\dot{P}$ outgoing streams from a process

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