FPSO fuel consumption and hydrocarbon liquids recovery optimization over the lifetime of a deep-water oil field

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A Floating, Production Storage and Offloading (FPSO) plant is a high-energy consumer (from a few to several hundreds of megawatts). Since a number of parameters have effects on the FPSO plant performance, screening analysis procedure could be used to select the most important parameters affecting a given output and an optimization procedure being applied to maximize/minimize an objective function. Thus, optimization procedures focused on fuel consumption and hydrocarbon liquids recovery can improve the energy efficiency, product recovery, and sustainability of the plant. In the present work, optimization procedures are used for an FPSO plant operating at three different conditions of the Brazilian deep-water oil field in pre-salt areas to investigate: (1) Maximum oil/gas content (Mode 1); (2) 50% BS&W oil content (Mode 2) and; (3) High water/CO2 content in oil (Mode 3). In order to reduce the computational efforts, we investigate the contribution of eight thermodynamic input parameters to the fuel consumption of the FPSO plant and hydrocarbon liquids recovery by using the Smoothing Spline ANOVA (SS-ANOVA) method. From SS-ANOVA, the input parameters that presented the major contributions (main and interaction effects) to the fuel consumption and hydrocarbon liquids recovery were selected for the optimization procedure. The optimization procedure consists of a Hybrid method, which is a combination of Non-dominated Sorting Genetic Algorithm (NSGA-II) and AFilterSQP methods. The results from the optimized case indicate that the minimization of fuel consumption is 4.46% for Mode 1, 8.34% for Mode 2 and 2.43% for Mode 3, when compared to the baseline case. Furthermore, the optimum operating conditions found by the optimization procedure of hydrocarbon liquids recovery presented an increase of 4.36% for Mode 1, 3.79% for Mode 2 and 1.75% for Mode 3 in total exportation oil.

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three different conditions. They concluded that the gas compression systems are responsible for the major energy consumption and their power consumption changes during the lifetime of an oil field. A similar work was presented by Barrera et al. [8]; their research focused on exergy performance analysis applied to a Brazilian FPSO process integrated with an Organic Rankine Cycle (ORC). The authors reported a saving of 15% in fuel consumption when compared to the baseline case. One of the most important processes in a petroleum production plant is the separation process. The primary objective of separation processes of a typical FPSO is to maximize the recovery of hydrocarbon liquids that might otherwise flow into the gas stream and to remove dissolved gases from hydrocarbon liquids, increasing liquids production as well as the API gravity [9,10].

It is well known that several operating parameters can affect the performance of a petroleum process plant. Recently, some researchers have used optimization methods to maximize/minimize the parameters of interest based on some input assumptions. Willersrud et al. [11] evaluated methods to maximize the total offshore oil production and concluded that pressure control is an important variable to increase the oil production. Ghaedi et al. [12] performed an optimization procedure of separator pressures in multistage production units. They used the Genetic Algorithm to optimize the separator pressure for a crude oil production unit with four separation stages and a gas condensate production unit with three separation stages. Compared to a baseline condition, the results showed improvements of 2.4% and 8.6% for crude oil and gas condensate productions, respectively. Liao et al. [13] used fuzzy logic control applied to the petroleum separation process. The authors commented that a crucial variable in quality control of the crude oil separation process is the gas-to-liquid ratio, and adjustments in that variable strongly affect the produced oil quality during the separation process. Tahouni et al. [14] showed that the optimum pressure drop in the condensate stabilization process can decrease the power consumption of the plant by about 20% in comparison with the baseline case. Hwang et al. [15] used an optimization procedure (Genetic Algorithm and sequential quadratic programming method) to find the optimal operating conditions of an LNG FPSO plant. The results indicated a 34.5% decrease in the FPSO required power. Allahyarzadeh-Bidgoli et al. [4] optimized the fuel consumption of a Brazilian FPSO for petroleum composition with maximum oil and gas content using the
Genetic Algorithm. The optimal solution presented a reduction of 4.6% in fuel consumption and an increase of 1.95% in oil production, when compared to a baseline condition.

The previous bibliographic review shows that there is a limited amount of work focused on optimization applied to FPSOs, especially those related to optimum operating conditions of the plant during the lifetime of the Brazilian deep-water reservoirs. Furthermore, there is a gap in the open literature about the strategy during the lifetime of the reservoir without adding new equipment or redesign of existing equipment.

The main goal of the present work is to find the optimum operating conditions of the separation train and main gas compression unit to improve the energy efficiency and products recovery. Thus, the present research aims to minimize fuel consumption and maximize the hydrocarbon liquids recovery of a typical Brazilian FPSO operating under different conditions of a deep-water oil field (Maximum oil/gas content, 50% BS&W oil content and high water/CO2 content in oil). Initially, eight thermodynamic parameters are submitted to a preliminary screening analysis via Smoothing Spline ANOVA (SS–ANOVA) method to reduce the number of input variables for the optimization procedure. After this analysis, the most important thermodynamic parameters are submitted to an optimization procedure, which uses Hybrid Genetic Algorithm to maximize/minimize the objective functions.

The eight thermodynamic parameters, also called input parameters, are: output pressure of the first control valve ($P_1$), output pressure of the second stage of the separation train before mixing with dilution water ($P_2$), input pressure of the third stage of separation train ($P_3$), input pressure of dilution water ($P_4$), output pressure of the main gas compressor ($P_c$), output petroleum temperature in the first heat exchanger ($T_1$), output petroleum temperature in the second heat exchanger ($T_2$) and dilution water temperature ($T_3$).

2. Methodology, assumptions and numerical simulation

FPSO units are applied to primary petroleum processing and they are normally composed by (i) deep-water oil and gas treatment, (ii) gas compression for exportation and injection and, (iii) CO2 and water treatment and injection [4]. The FPSO configuration evaluated herein is similar to an oil and gas processing platform used in a deep-water oil field of Brazilian pre-salt. In the present work, the FPSO consists of the main and utility plants. The main and utility plants are described according to Figs. 1 and 2.

The operating conditions are often determined by the features of the fluid reservoir, based on the composition of the hydrocarbons and the amount of impurities in the oil content. Three modes of fluid reservoir were considered for the numerical simulations: Mode 1 (maximum oil/gas content), Mode 2 (50% BS&W oil content) and Mode 3 (high water/CO2 content in oil). Fig. 3 shows the general scheme of the FPSO unit operating with these three different modes.

For operation mode 1, the separated gas in the first separation stage is directly sent to the main compression unit of the platform. The gas from the second and third separation stages is compressed in the vapor recovery unit. After this, the gas flows to the main compressor and is then sent to the gas dehydration unit and to the hydrocarbons dew point control unit. All the dehydrated gas goes through the CO2 removal unit and then goes to exportation. The separated gas is compressed in the exportation gas compressors and the separated CO2 is compressed by a section of the injection unit as shown in Fig. 3.

In operation mode 2, part of the dehydrated gas goes to the CO2 removal unit, while the remaining gas is processed directly in the injection compression unit for further compression and injection in the separated wells. The treated gas in the CO2 removal unit is mainly used for exportation purposes, and a small quantity is used as fuel in the power and heat generation unit. Finally, the gas to be exported is compressed in the exportation gas compression system.

In operation mode 3, most part of the gas is injected. The dehydrated gas bypasses the CO2 removal unit and then forwarded directly to the CO2 compressors to reach the required pressure for injection purposes. The fuel gas used for the power and heat generation units can be obtained from a slight percentage of the processed gas or from an external supplier source.

A typical FPSO processing plant is modeled and simulated by using ASPEN HYSYS [16], based on the Peng & Robinson Equation of State (EoS) [17], which is widely used for offshore simulations [4,18]. For the numerical simulations, the following assumptions were adopted:

- The operating pressures and temperatures and the number of stages of the separation train are simulated according to a typical FPSO condition [4,7] and based on the category of crude oil composition for separation processes considering the CO2 content [4,9];
- The processing capacity is 150,000 barrels per day for the separation train, in which 4.0%, 2.0% and 0.8% of the crude oil mass flow rate were considered for dilution water in Modes 1, 2 and 3, respectively;
- The reference temperature and pressure are 298.15 K and 101.3 kPa, respectively;
- As for the separation efficiencies of the dehydration and CO2 separation systems, the Gas dehydration system is projected to remove 99% of the water in the gas stream and the CO2 removal unit allows a maximum of 3% CO2 content in the gas;
- Adiabatic efficiency of 75% is assumed for all the centrifugal compressors with intercooler and pumps;
- An aero-derivative gas turbine (RB211G62 DLE 60Hz) used in offshore applications is considered for the Heat and power unit. The real performance data of the gas turbine is obtained from the GATECYCLE™ commercial software [19];
- Separators, pumps, mixers, splitters, and gas turbine are considered adiabatic;
- The exhaust gas is routed to the waste heat recovery. The exhaust gas transfers heat to the hot water system, and then the heated water is sent to the hot water consumers, such as the heat exchangers of the separation train;
- The pressure of gas exportation is set to 25 MPa and the injection of CO2 is equal to 55 MPa.

The crude oil composition from a Brazilian reservoir for the three operation modes is shown in Table 1.

Since the numerical simulations are considered in a steady-state condition, from the First Law of Thermodynamics, the energy balance can be written as:

$$\dot{Q} - \dot{W} = \sum \dot{m}_{\text{out}} h_{\text{out}} - \sum \dot{m}_{\text{in}} h_{\text{in}}$$

(1)

where $\dot{Q}$ and $\dot{W}$ are heat rate and power, respectively; $\dot{m}$ is the mass flow rate of a material stream and $h$ is the specific enthalpy. The mass balance and thermal efficiency of the gas turbine are, respectively,

$$\dot{m}_{\text{out,GT}} = \dot{m}_{\text{air}} + \dot{m}_{\text{fuel}}$$

(2)
\[ \eta = \frac{W_{net}}{m_{Fuel} \times LHV} \]  

where \( m_{out,GT} \) is the output mass flow of the gas turbine, \( m_{air} \) is the input mass flow of air in the combustion chamber and \( m_{Fuel} \) is the consumed fuel mass flow in the combustion chamber. \( \eta \), \( W_{net} \) and \( LHV \) are the thermal efficiency, the shaft power of the electric generation unit and the fuel's Lower Heating Value, respectively.

One of the main purposes of any oil and gas processing plant is to maximize the recovery of each group of hydrocarbons. Light hydrocarbons, such as methane, ethane, and propane, should be separated into the exported gas. Butane is dependent on the separation pressure (it can be gas or liquid) and pentane plus (C5, C6, C7 ..., and C20+) should be mixed into the oil. Therefore, according to these three types of hydrocarbons, the separation efficiency is calculated as follows:

\[ r_{light} = \frac{\sum_{C1} \hat{n}_i \cdot Separated_{gas}}{\sum_{C1} \hat{n}_i \cdot feed} \]  

\[ r_{medium-heavy} = \frac{\sum_{C5} \hat{n}_i \cdot Separated_{oil}}{\sum_{C5} \hat{n}_i \cdot feed} \]  

where \( r_{light} \) and \( r_{medium-heavy} \) are separation efficiencies of light hydrocarbons and medium-heavy hydrocarbons, respectively. \( n_i \) is the molar flow \( C1 \) is methane, \( C4 \) is butane and \( C5+ \) are pentane plus...
Finally, the separation efficiency \( \eta_{\text{sep}} \) is given by

\[
\eta_{\text{sep}} = \frac{r_{\text{light}}}{r_{\text{medium-heavy}}}
\]  

### 3. Sensitivity analysis and optimization

#### 3.1. Sensitivity analysis (screening analysis)

Screening analysis is often a preliminary step in any optimization procedure that uses a large number of input parameters. The main objective of the screening analysis is to identify the most important contributors to increase/decrease an output value [20,21,23]. This preliminary step is performed by the Smoothing Spline ANOVA (SS-ANOVA) model, already implemented in the ESTECO modeFRONTIER software [24]. The SS-ANOVA is a statistical algorithm based on functional decomposition similar to the classical analysis of variance (ANOVA). The SS-ANOVA is considered a suitable nonparametric technique for modeling or to estimate both univariate and multivariate functions.

#### 3.1.1. Smoothing Spline ANOVA method (SS-ANOVA)

The SS-ANOVA is associated with notions of main and interaction effects. For this reason, the interpretability of the results from SS-ANOVA is an additional benefit over standard parametric models.

Nonparametric regression methods can provide a better goodness-of-fit methodology when parametric assumptions are too restrictive, for instance, the linearity of the mean. The models such as SS-ANOVA can also suffer from lack-of-fit of the mean, which has become this method highly popular among the nonparametric models.

The framework for SS-ANOVA is a general multiple nonparametric regression model, with \( d \) independent variables \( (x_1, x_2, \ldots, x_d) \), continuous or discrete, and response variable \( y_i \) [25].

\[
y_i = f(x_{i1}, x_{i2}, \ldots, x_{id}) + \epsilon_i, \quad i = 1, 2, \ldots, n
\]  

where \( n \) is the number of designs (sample size), \( f \) is the unknown function of independent variables and \( \epsilon_i \) represents independent random errors.

Through the SS-ANOVA decomposition, an unknown mean function is decomposed as a sum of the main effects \( f_k(x_k) \) and interaction effects \( f_{ij}(x_i, x_j) \), for example:

\[
E(x_1, x_2, \ldots, x_d) = f(x_1, x_2, \ldots, x_d) = \mu + \sum_{k=1}^{d} f_k(x_k) + \sum_{i<j} f_{ij}(x_i, x_j) + \ldots
\]  

The SS-ANOVA is recognized as an extremely flexible class of additive models that allows selecting a parsimonious model from a large class of non-parametric additive models. It can also include linear terms for discrete variables (equivalent to ordinary ANOVA), linear or smooth terms for continuous variables and interaction terms between continuous and discrete variables [38].

The goodness-of-fit provided from the SS-ANOVA method can be assessed by four important parameters: (i) Collinearity index; (ii) Root Mean Squared Error (RMSE); (iii) Maximum Absolute Error (MAE) and; (iv) Pearson’s correlation. The collinearity index measures the orthogonality of the column vectors. Its value must be very close to the unity to guarantee that the column vectors are not parallel to each other. The Pearson’s correlation provides the

### Table 1

<table>
<thead>
<tr>
<th>Components</th>
<th>Mode 1</th>
<th>Mode 2</th>
<th>Mode 3</th>
</tr>
</thead>
<tbody>
<tr>
<td>H2O</td>
<td>0</td>
<td>0.83360</td>
<td>0.89774</td>
</tr>
<tr>
<td>N2</td>
<td>0.00490</td>
<td>0.00083</td>
<td>0.00023</td>
</tr>
<tr>
<td>CO2</td>
<td>0.16000</td>
<td>0.03009</td>
<td>0.05438</td>
</tr>
<tr>
<td>C1–C4*</td>
<td>0.63820</td>
<td>0.08990</td>
<td>0.03550</td>
</tr>
<tr>
<td>C5–C12</td>
<td>0.10290</td>
<td>0.02374</td>
<td>0.00569</td>
</tr>
<tr>
<td>C13–C19</td>
<td>0.04670</td>
<td>0.00982</td>
<td>0.00237</td>
</tr>
<tr>
<td>C20+</td>
<td>0.04730</td>
<td>0.01202</td>
<td>0.00409</td>
</tr>
</tbody>
</table>

*C4 and C5 include the nC4 and iC4, and the nC5 and iC5 hydrocarbons, respectively.*

![Fig. 3. General scheme of the FPSO with its main and utility plants.](image-url)
relationship, or curve adjustment, between the predicted function and the data from DoE. Values of Pearson’s correlation greater than 0.85 are considered satisfactory. Values of RMSE <5% and MAE <10% can be considered good enough for the most problems in engineering.

Both pressure and temperature in multistage crude oil separation processes play an important role in the efficiency of thermal systems and in the separation of hydrocarbons. However, it is important to quantify the impact of the input variables, which are all related to the operating pressure and temperature of the systems/devices to find the major contributors to the outputs and reduce the computational efforts for the optimization procedure. Thus, the input variables submitted to the SS-ANOVA analysis are: output pressure of the first control valve (P₁), output pressure of the second stage of the separation train before mixing with dilution water (P₂), input pressure of the third stage of the separation train (P₃), input pressure of the dilution water (P₄), output pressure of the main gas compressor (PC), output petroleum temperature in the first heat exchanger (T₁), output petroleum temperature in the second heat exchanger (T₂) and dilution water temperature (T₃). These parameters are shown in Fig. 4, in which the pressure parameters are indicated in the black circle and the temperature parameters are presented in the red circle.

The objective functions are the fuel consumption of the plant and hydrocarbon liquids recovery of the processed oil. Table 2 shows the eight input parameters, as well as their operating ranges (according to the real operating condition of an FPSO) and constraints used for the screening analysis. Furthermore, the technical constraints of the whole plant are assessed during the screening analysis, in order to avoid: (i) unfeasible separators performances; (ii) temperature cross in heat exchangers; (iii) decreasing of the volume of the oil and gas production (exportation) and; (iv) CO₂ compressors surge.

Before performing the screening analysis, a Design of Experiments (DoE) method used for deterministic computational experiments known as the Uniform Latin Hypercube sampling (ULH) algorithm is applied to create an initial sample based on data from Table 2. The ULH algorithm generates random candidates that meet the requirements to perform a uniform distribution, which maximizes the minimum distance between neighbouring points. The ULH method is detailed in Ricco et al. [25], Posadas et al. [26] and Nord et al. [27]. After the design samples are created, the screening analysis is run and the values of the outputs are saved for each run of the DoE. In the next step, the goodness-of-fit of the SS-ANOVA method is assessed by the collinearity index and Pearson’s correlations. If the values of these parameters are adequate, the contribution of each input variables to the outputs are reported; if the values of those two parameters are not suitable, new designs are generated by another DoE method, named Incremental Space Filler (ISF) algorithm, based on information from Table 2. The great advantage of the ISF is their ability to fill the gaps in the design space. Finally, the process continues until reliable values of

![Fig. 4. Input parameters for the screening analysis.](image-url)
collinearity indices and Person's correlation are achieved.

From the screening analysis, the major contributors to the overall variance of the fuel consumption and hydrocarbon liquids recovery of the FPSO unit are selected for the optimization procedure.

3.2. Optimization procedure

Optimization procedures coupled with process simulation tools have attracted the attention of different industries aiming at lower production costs and higher engineering system efficiencies. This approach is becoming an important tool to help engineers to find optimal operating conditions for complex problems. The growing demand of gas-petroleum and the maturing of existing oil fields have forced investments in new researches and technologies to optimize their production processes. Thus, the oil and gas industries have spent considerable time and efforts on the optimization of processing plants to improve the efficiency of the thermal systems and oil/gas productions.

An optimization problem to decrease cost with several input variables by using the Genetic Algorithm was evaluated by Ghorbani et al. [22], Nord et al. [27] performed an optimization procedure to minimize weight-to-power ratio design of steam bottoming cycle for offshore oil and gas installations. From the optimization algorithm procedure, Wood [28] identified high-performing portfolios, which are essential components of gas and oil application. From a multi-objective optimization algorithm, Liu et al. [29] investigated the characteristics of the oil—gas production process and the relationship among subsystems to maximize the overall oil production and to minimize the overall water production.

Recently, several works have successfully applied the Genetic Algorithm (GA) to optimize thermal systems [4,30–34]. The great advantage of the GA is that it tries to avoid the stalling problem through crossover and mutation operators. High values of crossover and mutation operators increase exploration and the exploitation of the design space.

In this research, a modified Genetic Algorithm (GA), which combines the global exploration capabilities of Genetic Algorithms with the accurate local exploitation of the Afilter of Sequential Quadratic Programming method (AfilterSQP), is used to run the multi-objective optimization.

The adaptive filter helps to achieve the convergence starting from a random point in the design space. AfilterSQP formulates a quadratic programming (QP) problem, a search line is performed along the computed direction and the adaptive filter checks whether the point can be accepted. This Hybrid Genetic Algorithm normally finds better optimized solutions and faster convergence than the classic GA. The AfilterSQP runs at the same time with the GA, exchanging information during the convergence procedure.

4. Results and discussion

4.1. Screening analysis procedure

The sensitivity analysis of the FPSO unit, operating in a deepwater Brazilian pre-salt area, is performed for three crude oil compositions (Modes 1, 2 and 3) to reveal the most important parameters affecting the fuel consumption (FUE) and hydrocarbon liquids recovery (HLR) of the plant. The flowchart in Fig. 5 shows the steps to perform the screening analysis. Design space, based on the operating range of the selected input parameters and constraints, is generated by the Uniform Latin Hypercubes (ULH) algorithm. From an initial distribution, Aspen HYSYS® software is the solver used to run the numerical simulations. After the initial sampling points are created, the automated numerical process runs until RMSE, MAE, R² and collinearity index has converged. If the convergence of those parameters is not achieved, new sampling points are generated to the design space and the process continues until reach the convergence. Then, the generated initial designs are used in SS-ANOVA and the results from the screening analyses are used for the optimization procedure.

4.2. Outcomes from the screening analysis

Fig. 6 shows the effect of the input parameters on the hydrocarbon liquids recovery with maximum gas/oil content (operation mode 1). P2 is the most important contributor (68%) to hydrocarbon liquids recovery, followed by P1 (13%) and T2 (5%). Moreover, a weak interaction effect is observed (P1×P2). The importance of P3 is related to the separation of heavy and pseudo-component hydrocarbon. Because of the proposed configuration of the separation train, the operating pressure of the third stage is also an important parameter to stabilize the volatile components in the liquid phase at a previous separation stage. P1 and T2 are important parameters since they are related to the removal of light hydrocarbons from the crude oil and the recovery of intermediate hydrocarbons in the processed oil. However, as the petroleum composition is volatile, the impact of T2, when compared to the pressure parameters, is negligible. Therefore, these pressures determine the amount of
liquid from the recovered volatile hydrocarbons.

According to Fig. 7, the major contributor to hydrocarbon liquids recovery for operation mode 2 is the parameter $P_3$ (about 50% of the total contribution), followed by $T_2$ (37%) and $P_2$ (6%). According to McCain et al. [36], the composition of operation mode 2 is within the wet gas group ($0.5 < Z_{C7} < 4.5$). Wet gas is a special two-phase flow (gas-liquid), which is often encountered in the oil and gas industry with the presence of hydrocarbons heavier than ethane, such as wet natural gas extraction from a condensate field [37]. Moreover, the wet gas is sensitive to both operating pressure and temperature in separation conditions. The effect of $P_3$ on the separation process is due to the presence of $C_5^+ + C_{20}^+$ in the crude oil composition and to the stabilization of those hydrocarbons in the last stage of the separation train. The effect of $T_2$ can be attributed to the importance of the temperature variation in separation processes for a given wet gas composition.

Fig. 8 shows the effects of the input parameters on hydrocarbon liquids recovery for operation mode 3 (Maximum water/CO$_2$ oil content). From that, the composition of hydrocarbon components for this operation mode is similar to operation mode 2; thus, the contribution of each parameter for operation mode 3 is very similar to that presented for operation mode 2. In this case, only $P_3$ and $T_2$ present a significant contribution on hydrocarbon liquids recovery, reaching about 90% of the total contribution.

Overall, by comparing the three operation modes, $P_3$ is the most important contributor to hydrocarbon liquids recovery and its order of magnitude is completely different for modes 1, mode 2 and 3. The second most important contributor for hydrocarbon liquids recovery is $P_1$ for mode 1 and $T_2$ for modes 2 and 3. In terms of $P_3$ and $T_2$ parameters, the contribution of $P_3$ is about 63% higher than $T_2$ for mode 1 and 13% higher for modes 2 and 3. It is also important to note that the contribution of $P_2$ can be considered negligible for mode 1, but important for modes 2 and 3 (for these 2 modes there are small interactions between $P_2$ and other parameters which must be added to the overall contribution).

From the previous discussion, the parameters that should be taken into account for an optimization procedure are: $P_3$ and $T_2$ (for all operation modes), $P_1$ (only for operation mode 1) and $P_2$ (for operation modes 2 and 3). Thus, by reducing the number of input parameters, the optimization procedure becomes simpler and faster, which is an important advantage to save time and computational resources.

Fig. 9 shows the contributions of the input parameters on fuel consumption for operation mode 1. The results indicate that parameters $P_3$, $P_1$ and $P_2$ are the major contributors to power consumption and, consequently, to fuel consumption, corresponding to about 81% of the total effect. Parameter $P_3$ is the major contributor and its effect on fuel consumption is about 10% higher than that of $P_1$ and $P_2$. $P_2$ is also an important parameter that impacts fuel consumption since its effect is about 6% of the total variance. Moreover, the main effects were more significant than the interaction effects on fuel consumption, similarly to that also observed for hydrocarbon liquids recovery.

The contribution of the input parameters on fuel consumption for operation mode 2 is presented in Fig. 10. The contribution of $P_3$ is the highest (63%), followed by $P_1$ (17%), $T_2$ (16%) and $P_2$ (2%). The effect of $P_3$ on fuel consumption is about 45% higher than $P_1$ and $T_2$. Relevant interaction effects between the input variables are not observed.
Fig. 11 shows that $P_c, P_3$ and $T_2$ are the most important contributors to fuel consumption for operation mode 3. The effect of the $P_c$ is verified to be about 50% of the total variance due to the highest CO₂ mass flow for this operation mode. Parameter $T_2$ is relevant for modes 2 and 3 and its contribution is practically similar for both operation modes.

Finally, an unusual interaction is observed between $T_2^*P_3$, corresponding to 10% of the total variance. This interaction is due to the recovered wet gas components on the third stage of the separation train ($P_3$).

Four input parameters ($P_1, P_2, P_3$ and $P_c$) correspond to 96% of the total contribution to fuel consumption for Mode 1. Similarly, three input parameters ($P_3, P_c$ and $T_2$) correspond to 97% of the total contribution to fuel consumption for Modes 2 and 3. In terms of the hydrocarbon liquids recovery, four input parameters ($P_3, P_2, P_3$ and $T_2$) correspond to 95% of the overall contribution for Mode 1. Similarly, three input parameters ($P_3, P_2$ and $T_2$) correspond to 97% and 98% of the total contribution to hydrocarbon liquids recovery for Modes 2 and 3, respectively. Thus, the optimization procedures for fuel consumption minimization and hydrocarbon liquids recovery maximization could be run with only these input parameters, which reduces the time and computational resources, when compared to an optimization procedure with those eight initial input variables.

### 4.3. Optimization procedure

From the screening analysis, a flowchart of the optimization procedure can be built, as observed in Fig. 12. To launch the optimization procedure, an initial population is generated by the Uniform Latin Hypercube method. In the next step, the integrator software (modeFRONTIER) calls the ASPEN HYSYS® and GATE-CYCLE™ software to solve the thermodynamic equations for each component of the FPSO unit and to calculate the gas turbine performance. After the thermodynamic convergence is achieved for the whole plant, the technical constraints are assessed. If the technical constraints are satisfied, the objective functions (fuel consumption or hydrocarbon liquids recovery of the plant) and other important outputs are picked up by the integrator software. The objective functions are then evaluated by the hybrid optimization algorithm until the objective functions reach the convergence. If the convergence criteria fail, new individuals are generated and the procedure is restarted.

The objective functions, their operating ranges and constraints are described below:

\[
\text{Obj} = \text{Function(Input parameters)}
\]

\[
\begin{align*}
\text{FUE} &= f(\text{the most important contributors for Modes 1, 2 and 3}) \\
\text{HLR} &= f(\text{the most important contributors for Modes 1, 2 and 3})
\end{align*}
\]
According to Rao [35], defining a design vector $I$, the optimization problem can be stated as follows:

$$ I = \{ \text{The most important contributors from the screening analysis} \} $$

(10)

Find $I = \{ \cdots \}$ which minimizes $FUE$ or maximize $HLR$ under constraint:

$$ P_i > P_{i+1}, \quad i = 1, 2 $$

Subject to all indicated technical constraints that are also considered in the sensitivity analysis

(11)

The current operating condition of a Brazilian FPSO, with its corresponding power consumption, fuel consumption and production, were taken as a baseline simulation (reference scenario) to be compared to the optimization solutions. For the objective function “FUE minimization“, the exportation oil was evaluated to guarantee that the amount of exportation oil could not be decreased. For the objective function “HLR maximization“, the total power consumption is checked to prevent it from increasing. It is important to emphasize that the separation performance of separators was verified during the optimization process. Finally, the results found by the optimization procedure are compared to a baseline case and the main conclusions are presented.

4.4. Fuel consumption optimization

Table 3 shows the fuel consumption and power demand of an FPSO unit for the baseline and optimized cases. From that, it can be observed that the optimized case for operation modes 1, 2 and 3 led to a mitigation of 4.46%, 8.34% and 2.43% in the fuel consumption of the FPSO, respectively, in comparison to the baseline case. This decrease in fuel consumption for all operation modes is due to the reduction of the overall power demand of the plant, as explained later. For operation mode 1, the comparison between the optimized and the baseline cases reveals a reduction of 6.4% in power demand, while there is a reduction of 10.1% in power demand for operation mode 2 and 2.9% for operation mode 3.

As observed in Fig. 13, the separation of the associated natural gas components in the third stage for operation mode 1 consists of 53.5% of C1-C4 and 46.5% of C5+ for the baseline while for the optimized case is verified 67.2% of C1-C4 and 32.8% of C5+ for the

![Fig. 12. General flowchart of the optimization procedure.](image-url)
optimized case. For Modes 2 and 3, the optimal set of the separation parameters leads to an improvement of 34.6% and 23.3% of the separated gas flow composition in the third stage of the separation train, respectively (Figs. 14 and 15). The improvement of hydrocarbons separation for Mode 1 is smaller than in Modes 2 and 3. However, as the composition of Mode 1 is richer in hydrocarbons than of the other modes, the effect of this improvement is important on gas compression and total power demand.

The values of the pressures and temperatures for the three operation modes for the baseline and optimized cases are depicted in Table 4.

The inlet pressure of the first pressure valve of the separation train is 2300 kPa and its output pressure ($P_1$) is 1300 kPa for the baseline case. The value of $P_1$ can also change the operating pressure of the separated gas in the second stage of the separation train which is directed to vapour recovery units. Thus, its variation influences the required shaft power of the compression system in the VRU. As shown in Table 4, for operation mode 1, the value of $P_1$ found by the optimization procedure is 1284 kPa (with a reduction of 16 kPa as compared to the baseline). This reduction in $P_1$ increases the light hydrocarbons separation at the medium pressure level and induces a better stabilization of the intermediate hydrocarbon separation in the liquid phase. Since the operation mode 1 refers to the oil composition with the highest GOR in comparison with Modes 2 and 3, the mass flow of the gas stream that follows to the compressors of MGC in Mode 1 is smaller than the other conditions, resulting in a reduction of the total power demand of the plant.

The optimum operating condition at the end of the second stage of the separation train ($P_2$) is important for stabilizing the separated liquid flow in the second stage. The optimization method found the value of 915 kPa for $P_2$ (with an increase of 475 kPa) in operation mode 1, as presented in Table 4.

With regard to parameter $P_3$ (pressure of the third stage of the separation train), the optimization procedure found an improvement in the separation of the intermediate and heavy hydrocarbons

<table>
<thead>
<tr>
<th>Operation mode 1</th>
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<tbody>
<tr>
<td>FUE (kg/h)</td>
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<tr>
<td>Baseline case</td>
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<td>Optimized case</td>
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<tr>
<td>FUE (kg/h)</td>
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<td>Optimized case</td>
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<table>
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<tr>
<th>Operation mode 2</th>
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</thead>
<tbody>
<tr>
<td>FUE (kg/h)</td>
</tr>
<tr>
<td>Baseline case</td>
</tr>
<tr>
<td>Optimized case</td>
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</table>

Table 3
Fuel consumption and power demand of the FPSO unit for the baseline and optimized cases.
in the third stage. Moreover, as the output pressure of the separated gas flow in the third stage of the separation train should be recovered by the compressor of the VRU to the initial feed pressure of the main compressor, any decrease in $P_3$ increases the shaft work of the compressors. The optimization procedure found $P_3 = 448$ kPa for operation mode 1; $P_3 = 449$ kPa for operation mode 2 and $P_3 = 447$ kPa for operation mode 3; in turn, the values for the baseline configuration were $P_3 = 244$ kPa for operation mode 1, $P_3 = 232$ for operation mode 2 and $P_3 = 229$ for operation mode 3. Thus, the compressors power demand is smaller for the optimized case of three operation modes and, as a consequence, a decrease in fuel consumption for the optimized case is observed. When parameter $P_c$ increases, a higher compressor shaft work is required to meet the necessary output pressure.

For this reason, the optimization procedure found $P_c = 7019$ kPa for Mode 1 and $P_c = 7020$ kPa for both Modes 2 and 3, i.e., the minimum feasible value to keep the stability of the whole plant (especially to guarantee the expected oil production).

Finally, the effect of $T_2$ in Modes 2 and 3 is due to the separation of the intermediate hydrocarbons, such as propane and butane in the gas flow, and heavy components of the intermediate hydrocarbons, such as pentane in the liquid phase. Since the shaft power of the oil pump can be influenced by the operating temperature, i.e., the required power of the oil pump trends to decrease as the oil temperature is decreased. Thus, Table 4 allows observing that the optimized case presents a value of $T_2$ that is lower than that of the baseline case.

According to McCain et al. [36], the compositions of operation modes 2 and 3 are included in the wet gas group. Wet gas is a special two-phase flow (gas-liquid) with volatile components often encountered in condensate compositions [37]. Thus, the optimal combination of $T_2$ and $P_3$ stabilizes the volatile components in the oil stream leading to a decrease of the compressor shaft power.

Hence, the optimization procedure found a maximum separation of light hydrocarbons in the gas flow and the maximum recovery of heavy hydrocarbons in oil stream, improving the stabilization of the volatile components leaving the last stage of the separation train.

### 4.5. Hydrocarbon liquids recovery

According to Abdel-Aal et al. [9], for separation processes, the three main goals of oil production facilities in terms of separation performance are: (1) separating the light hydrocarbons components (C1 and C2) from oil; (2) maximizing the recovery of the intermediate hydrocarbons (C3, C4, and C5) in the oil product and; (3) saving heavy components (C6+) in the oil product.

The light components contained in the liquid production (oil or/and condensate) flash to gas in the storage tank, because of processing issues, when the pressure is reduced within the export tanks or pipelines. The intermediate components, which are the major components in the crude oil, are released by undergoing a pressure drop in the separators. Therefore, one of the most important design aspects of an oil and gas processing offshore facility is the separation performance of light and intermediate components into gas and oil products to achieve the maximum oil recovery [5].

These important design parameters together are named hydrocarbon liquids recovery (stabilization and saving) and they are selected as an objective function for the optimization procedure for the three operation modes. In the optimization process, hydrocarbon liquids recovery is considered the objective function and besides monitoring two indicators of recovery of hydrocarbon liquids content (separation performance and C3-C5 content in oil product), the total power consumption for FPSO unit is also checked to avoid the cases with more demand of power consumption (as compared
with the baseline). Thus, the separation performance, saving C6+ and the recovery of intermediate hydrocarbon components (C3, C4, and C5) in exportation oil content are also important and monitored to analyze the quality of the exportation oil [9].

Table 5 shows the exportation oil rate, total power consumption, and percentage of propane, butane pentane of the exported oil of the FPSO unit for the baseline and optimized cases. From that, the exportation oil rate can be observed to increase by 4.36% for Mode 1, 3.79% for Mode 2 and 1.75% for Mode 3 in comparison with the baseline case. In addition, the C3–C5 content in exportation oil for the optimized case increases 10% for operation mode 1, 9% for operation mode 2 and 4% for operation mode 3 when compared with the baseline case. Moreover, as observed in Table 5, the power demand of the optimized case decreased by 4.9% for Mode 1, 7.7% for Mode 2 and 2.9% for Mode 3 when compared to the baseline case due to decrease of the gas mass flow to be compressed by MGC and EGC.

Table 6 shows the input parameters for the baseline and optimized configurations for operation modes 1, 2 and 3.

In order to analyze the behavior of hydrocarbon components for the three operation modes, some points in the separation train are selected. Fig. 16 shows the selected process points to investigate the phase stabilization (gas or liquid) in the optimization procedure: p0, p1, p2, p3, p4, p5, p6, and p7. Point p0 represents the input condition before the first pressure valve and the description of the further points is presented below:

- Output of the first control valve (p1);
- Output of the first heat exchanger (p2);
- Output of the second heat exchanger (p3);
- Input of pair separator and after mixing with the recycling liquid content of MGC (p4);
- Output of the first separator of the second stage of the separation train (p5);
- Output of the second stage of the separation train and before mixing with dilution water (p6) and
- Input of the third stage of the separation train (p7).

Note that the symbol P (uppercase) means the operating pressures and symbol p (lowercase) refers to the process points.

Fig. 17 shows the phase stabilization during the separation processes for the baseline (a) and optimized case (b) for operation mode 1.

The continuous line shows the mass flow rate of gas phase components while the dashed line represents the mass flow rate of the liquid phase components. The total gas mass flow rate is separated by the first stage of the separator (including pre-heater and heater units) up to point 3. The differences in the mass flow rates in both gas and liquid phases for the baseline and optimized cases between p2 and p3 are due to parameter T2. Parameter T2 impact on the separation of the intermediate hydrocarbons in gas phase and the separation of the heavy components of the intermediate hydrocarbons in liquid phase. The optimized stabilization related to T2 is shown in Fig. 17(b) (between p2 and p3) and it is obtained by a reduction of 53°C as compared to the baseline (Table 6).

The conditions of p3 and p4 are observed to be almost equal (Fig. 17), because of the very small mass flow rate of the recycling pipe from the MGC. Thus, the gas is separated after p4 at the medium pressure and, consequently, the mass flow rate of gas is zero at p5. The recovery oil between p5 and p6 is performed by an increase of 475 kPa in P2 (Table 6). At p6 and p7 (after pressure valves VLV-113 and VLV-100, as shown in Fig. 16), the separation of some hydrocarbon liquids to gas phase is pointed out in Fig. 17. The optimization method found the value of 448 kPa for P3 to maximize
the oil saving of the third stage of the separation train.

As shown in Fig. 17, from the same composition of the crude oil, an increase in oil recovery and stabilization phases can be noted for the optimized case when compared to the baseline case. The optimum operating condition resulted to an improvement in the separation performance of the gas and hydrocarbon liquids. Since the crude oil stream includes dissolved gas components, the operating pressures and temperatures affect the phase stabilization of the heavy components of the intermediate hydrocarbons, condensate components and pseudo-component of petroleum.

For example, at point 3, the optimized configuration provides the condition to maximize the stabilization of intermediate hydrocarbons in the liquid phase, especially propane and butane (466400 kg/h). In turn, for the baseline case, a large quantity of heavy components of intermediate hydrocarbons is separated in the gas phase (17000 kg/h). At point 7, the liquid content is increased by about 20173 kg/h by the improvement in the recovery of oil components for the optimized case compared to the baseline.

Fig. 18 shows the phase stabilization during the separation processes for the baseline and optimized cases of operation mode 2. As displayed in Fig. 18 (a) and (b) (between p2 and p3), the impact of T2 on the separation condition and on the oil composition is
significant. In this sense, the optimization procedure found a value of $T_2$ with a reduction of 53 °C in comparison to the baseline case, as observed in Table 6. Moreover, the effect of the input parameters $P_2$ and $P_3$ are shown in Fig. 18, between p5 to p7. Note that for the optimized case, the mass flow rate of oil is increased about 6150 kg/h (as compared to the baseline case) only by the stabilization of hydrocarbon liquids in p3. Overall, this increase is 12683 kg/h as compared with the baseline case.

Fig. 19 shows the phase stabilization during the separation processes for the baseline and optimized cases of operation mode 3. Similarly as discussed for Mode 2, the influence of $T_2$ on the separation condition of the oil for Mode 3 is relevant. Fig. 19 shows that the optimization procedure found a value of $T_2 = 49$ °C, which is 39 °C lower than the baseline configuration.

As the water content in the reservoir for Mode 3 is the maximum, the maximization of hydrocarbon liquids recovery is smaller than in the other two operation modes. According to Fig. 19, point 3 for the optimized case has about 1500 kg/h more oil content than the same point for the baseline configuration. Overall, an increase of about 2600 kg/h is observed in oil content for the optimized case. Moreover, with a decrease in the operating pressure $P_2$, fewer volatile components are separated in the gas flow of the optimization case (72 kg/h versus 781 kg/h of the baseline configuration).

As outlined in the previous paragraphs, the results from the optimization procedure indicated that the maximization of hydrocarbon liquids recovery is achieved by increasing the operating pressure in the second and third stages of the separation train and by decreasing the operating temperature in the oil heater of the separation train for all the operation modes.

5. Conclusions

A primary processing plant of a typical FPSO operating in a Brazilian deep-water oil field on pre-salt areas is modeled and simulated. Optimization procedures are used to minimize the fuel consumption (FUE) and to maximize the hydrocarbon liquids recovery (HLR) of an FPSO for three operating conditions of the oil field: (i) maximum oil/gas content (Mode 1), (ii) 50% BS&W oil content (Mode 2) and (iii) high water/CO2 content in oil (Mode 3). The contribution of eight thermodynamic parameters to fuel consumption and hydrocarbon liquids recovery of the FPSO unit is
investigated and the most important contributors are selected to reduce the computational costs of the optimization. For the screening analysis, a non-parametric regression model known as Smoothing Spline ANOVA method is used. From SS-ANOVA, the input parameters that presented higher main and interaction effects on fuel consumption and hydrocarbon liquids recovery were selected for the optimization procedure. The optimization procedure consists of the application of a modified Genetic Algorithm, which is a combination of Non-dominated Sorting Genetic Algorithm (NSGA-II) and Afflter SQP method. The results of the optimized case indicated that fuel consumption is minimized by increasing the operating pressure in the third stage of the separation train and by decreasing the operating temperature in the second stage of the separation train for all operation modes. The FPSO power demand was reduced by 6.4% for Mode 1, 10% for Mode 2 and 2.9% for Mode 3, in comparison to the baseline case. Consequently, the fuel consumption of the plant was decreased by 4.46% for Mode 1, 8.34% for Mode 2 and 2.43% for Mode 3, when compared to the baseline case. Moreover, for the maximization of the hydrocarbon liquids recovery, the optimized cases presented an increase of 4.36% for Mode 1 (10% of improvement for C3-C5 content in exportation oil), 3.79% for Mode 2 (an improvement of 9% for C3-C5 content in exportation oil), 1.75% for Mode 3 (an increase of 4% for C3-C5 content in exportation oil), when compared to the baseline operating conditions. A relevant feature of the optimization procedure proposed in the present work does not make any changes in the design of the heat exchangers. Overall, the optimization process showed to be a robust and promising tool to find efficient operating conditions for existing FPSO plants.

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