

Iranian Journal of Oil & Gas Science and Technology, Vol. 10 (2021), No. 1, pp. 40–65  
http://ijogst.put.ac.ir

## Techno-Economic Analysis of Heavy Fuel Oil Hydrodesulfurization Process for Application in Power Plants

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### Highlights

- Hydrodesulfurization is studied as an option for treating heavy fuel oil;
- Simulation and techno-economic analysis of HDS process;
- Sensitivity analysis of operational parameters for treating fuel oil;
- Parameters affecting net production cost and total capital investment are determined.

Received: October 26, 2020; revised: December 09, 2020; accepted: December 17, 2020

### Abstract

In Iran, power plants use liquid fuels such as heavy fuel oil (HFO) or mazut to prevent disruption in power generation. The high percentage of sulfur compounds in HFO and the lack of efforts to remove it cause significant damage to the environment. The purpose of this research is performing a techno-economic analysis of the hydrodesulfurization (HDS) process of HFO. The results show that for removing 85% of sulfur compounds from HFO with 3.5 wt % sulfur compounds at a volume flow rate of 250 m<sup>3</sup>/h, the total capital investment and the net production cost are 308.9 million USD and 114.5 million USD/year respectively. Moreover, the sensitivity analysis indicates that with a 100% increase in the catalyst loading, the mass percentage of sulfur compounds in the HFO decreases by more than 15%, which adds 6.4% and 32% to the total capital investment and net production cost respectively. With a 100% increase in the gas-to-oil ratio, the mass percentage of sulfur compounds in the HFO decreases by more than 15.3%, which adds 43.8% and 6% to the total capital investment and net production cost respectively. With a 100% increase in the pressure of the HDS process, the mass percentage of sulfur compounds in the HFO declines by more than 20.75%, which adds 43% and 6.75% to the total capital investment and net production cost respectively. Ultimately, with a 100% increase in the inlet temperature of the beds, the mass percentage of sulfur compounds in the HFO is reduced by more than 5%. Among the effective operational parameters, hydrogen consumption has the greatest impact on net production cost and payback period, and the pressure of the hydrodesulfurization process has the most significant influence on increasing the total capital investment of the process.

**Keywords:** Heavy Fuel Oil, Hydrodesulfurization, Net Production Cost, Payback Period, Power Plant

#### How to cite this article

Jafari M, Khalili Garakani A, *Techno-Economic Analysis of Heavy Fuel Oil Hydrodesulfurization Process for Application in Power Plants*. Iran J. Oil Gas Sci. Technol., Vol. 10, No. 1, p. 40–65, 2021.

DOI: 10.22050/ijogst.2020.254534.1569

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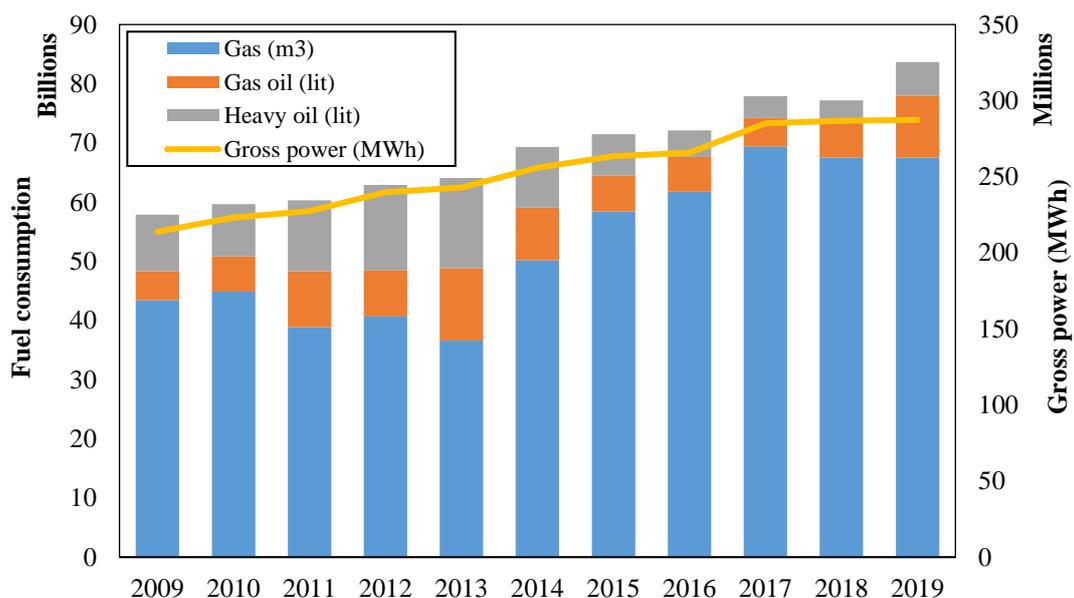
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## 1. Introduction

The production of environmentally friendly fuels meeting global pollutant emission standards is of paramount importance in today's world. Iran's power plants typically generate power using natural gas on most days of the year, but in winter, considering the priority of urban gas supply, the consumption of liquid fuels in power plants increases. Consequently, the rate of power plant pollution increases (Ameri et al. 2018). The consumption of gasoline and heavy fuel oil (HFO), also called mazut in power plants, results in the phenomenon of temperature inversion, especially in the cold seasons, which has a significant impact on air pollution (Kouravand et al. 2018). According to the Office of the General Directorate of Economic, Social, and Environmental of the Renewable Energy Organization, Iran has consumed 67 billion cubic meters of natural gas, 10 billion cubic meters of diesel fuel, and 5.7 billion cubic meters of HFO in its power plants in 2019 (Ebrahimi et al. 2019).

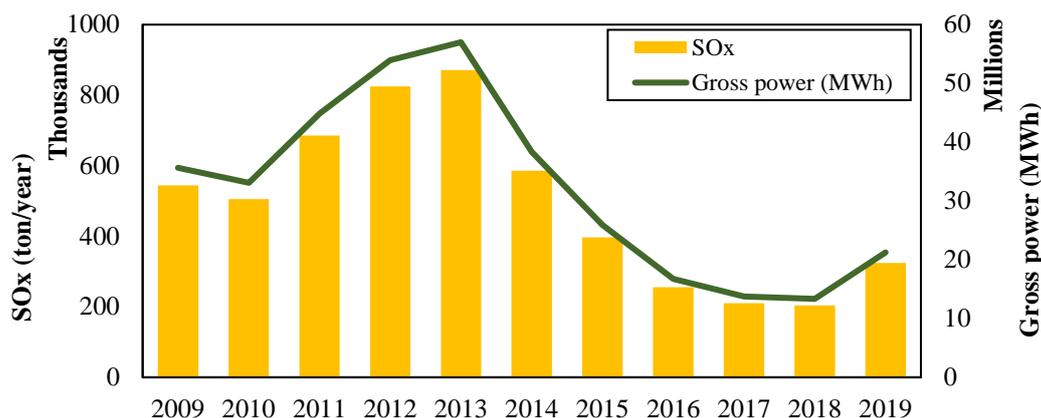
Figure 1 illustrates the fuel consumption and the total gross power generation by fossil fuel power plants. As can be seen, the amount of HFO consumption has increased by 63% in 2019 compared to 2018, which is due to the new rule constraining the consumption of HFO with more than 0.5% sulfur components in sailing ships and difficulties in exporting this fuel (Khosravi-Nikou et al. 2020). Therefore, the only consumer of this fuel in Iran is cement factories and power plants. Thirteen steam cycle power plants consumed HFO as fuel in 2019. Around 29% of the power generated in these power plants originates from HFO consumption. Regarding the statistics of the Air Quality Control Company, the share of power plants in the production of gas pollutions and particulate matter has been estimated to be 5% and 12% in 2018 and 2019 respectively (Shirzad et al. 2019).

Figure 2 presents the amounts of  $SO_x$  emitted due to HFO consumption by steam cycle power plants and the amount of power generated during 2009–2019 (Jafari et al. 2019). The simplest way to reduce exhaust emissions is to monitor the percentage of sulfur compounds in the fuel and refine and remove them before being used in power plants (Rashidi et al. 2015). It should be noted that the presence of heavier oil cuts is the reason for the increment of sulfur compounds; hence, the separation is more complicated and costly (Javadli et al. 2012). Moreover, as the boiling point increases, sulfur compounds become more resistant (Ghasemzadeh et al. 2016).



**Figure 1**

Fuel consumption in power plants of Iran during 2009–2019.



**Figure 2**

The amount of power generated by HFO consumption and  $\text{SO}_x$  emitted from steam cycle power plants during 2009–2019.

In light fuels such as naphtha and kerosene, most sulfur compounds include thiols and sulfides, but in heavier oil cuts such as vacuum gas oil, residual oil, or HFO, sulfur compounds are benzothiophenes (Marafi et al. 2019). In general, the desulfurization of compounds such as thiols and sulfides is much simpler than aromatic sulfur compounds like benzothiophenes. Various methods, including hydrodesulfurization (HDS), extractive desulfurization, absorption desulfurization, oxidative desulfurization (ODS), biodesulfurization, and desulfurization by supercritical water have been suggested for removing sulfur compounds from liquid fuels (Hosseini et al. 2017). It should be mentioned that choosing the appropriate method depends on the lightness or heaviness of oil cuts as well as the amount and type of sulfur compounds (Aramkitphotha et al. 2019). In the case of crude oils, the extractive and absorption desulfurization methods, as well as biodesulfurization, are not applicable due to the large molecules, high viscosity, and complexity of the compounds (Pouladi 2019). Generally, the most appropriate methods for the desulfurization of crude oils are HDS and ODS (Rajendran et al. 2020). Each of these methods has its benefits and disadvantages. Despite the recent developments in this field, HDS is still the most common industrial method (Treusch et al. 2020).

This work presents a simulation for the design and economic evaluation of HFO hydrodesulfurization processes to control  $\text{SO}_x$  and  $\text{NO}_x$  emissions in an Iranian steam power plant. It should be considered that in the economic evaluation of an HDS, expenses imposed by sulfur compounds on the environment, human beings, and the equipment corrosion are more substantial than the “Total Capital Investment” and “Net Product Cost”. The primary purpose of this work is to reach the world standards and remove 85% of sulfur compounds from HFO with 3.5 wt % sulfur compounds. The effective operating parameters that affect the removal of sulfur compounds from the HFO in this process are catalyst loading, the pressure of the HDS process, gas-to-oil ratio, and inlet temperature of the beds. Furthermore, since valuable by-products such as naphtha and diesel are obtained during the HDS process, the sensitivity analysis of the mass percentage of sulfur compounds in the product and the production of the by-products is carried out. Finally, after the simulation, the economic evaluation and economic sensitivity analysis of this process are carried out, which is one of the innovations of this work. The economic evaluation and analysis of the process, the net production costs, the total investment cost, and the return of investment are estimated. Economic sensitivity analysis is to find out which parameter is the least costly one in increasing sulfur removal. For example, is it better to increase the pressure or to increase the consumption of hydrogen or other parameters? Moreover, it can be

employed to find out which parameters have the most and least impact on the economics of this process, which is another innovation of this work.

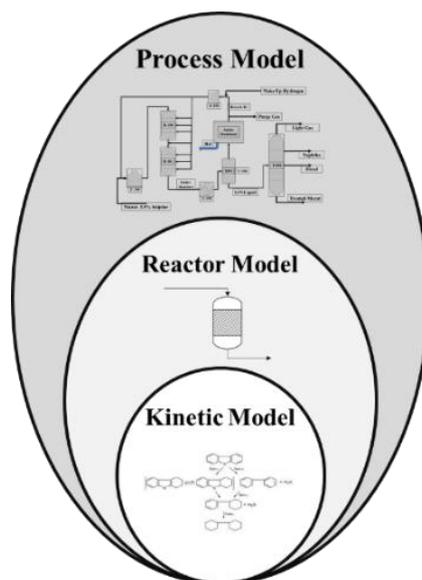
## 2. Materials and methods

Shazand Thermal Power Plant, with a capacity of 1300 MW, uses a significant amount of high-sulfur HFO for power generation. HFO is one of the heavy oil cuts obtained in the crude oil distillation column unit and is the cheapest fuel to use in power plants (Bayomie et al. 2019). The physical and chemical characteristics of the HFO used in Shazand power plant and most of Iran's power plants are presented in Table 1. The simulation of the HDS process for HFO is accomplished in Aspen HYSYS Petroleum Refining v.11 software (Rahimi, 2019), which is a supplementary tool added to Aspen HYSYS software. This software is appropriate for simulating refinery units such as hydrotreating, hydrocracking, fluid catalytic cracking (FCC), and catalytic units (Liu 2018). First, the component of HFO and its thermodynamic equation should be determined accurately. The application also has extensive library information, including various types of oil cuts and detailed thermodynamic equations for each of the refinery processes. In this simulation, the HCRSRK thermodynamic equations are applied (Wang et al. 2020). This thermodynamic equation is proper for hydrotreating processes such as HDS for heavy oil cuts. Moreover, the reactor model has used 97 reaction kinetics in Aspen HYSYS petroleum refinery (Chang et al. 2013), which belong to the six groups of light gas, paraffin, naphtha, aromatics, sulfur, and nitrogen compounds. Further, sulfur compounds are divided into eight groups containing thiophene, sulfide, naphthabenzothiophene, tetrahydrobenzothiophene, tetrahydro dibenzothiophene, dibenzothiophene, and tetrahydro naphthabenzothiophene. The Case Study tool is used to perform the sensitivity analysis and optimization of the effective operating parameters to remove sulfur compounds from the HFO. Figure 3 illustrates the range of the presented models of oil cut purification processes in a three-layer onion model.

**Table 1**

Physical and chemical properties of HFO (Shahsavan et al. 2020).

<b>Characteristic</b>	<b>Amount</b>
<b>API gravity</b>	10.40
<b>Specific gravity (60/60)</b>	0.9972
<b>Distillation type</b>	TBP
<b>0% Point (°C)</b>	268
<b>5% Point (°C)</b>	327
<b>10% Point (°C)</b>	387
<b>30% Point (°C)</b>	492
<b>50% Point (°C)</b>	574
<b>70% Point (°C)</b>	638
<b>90% Point (°C)</b>	731
<b>95% Point (°C)</b>	773
<b>100% Point (°C)</b>	815
<b>Nickel (ppm)</b>	40
<b>Vanadium (ppm)</b>	15
<b>Total nitrogen (ppmw)</b>	600
<b>Sulfur content (%)</b>	3.5
<b>Molecular weight</b>	611
<b>Flash point (°C)</b>	65
<b>Pour point (°C)</b>	32
<b>Ash (max) (wt %)</b>	0.15



**Figure 3**

Three-layer onion model of the simulation domain (Chang et al. 2013).

The core of the onion is a precise kinetic model, which investigates the type of the catalyst, the influence of the raw materials, and the reaction conditions under the current conditions. The middle layer is a reactor model that examines the product performance under various conditions such as changes in volume space, different temperatures, pressures, and changes in hydrogen consumption to find the optimal conditions. The last layer is the process model. It should be noted that a process model facilitates the optimization of the operating conditions, profit maximization, cost minimization, and safety enhancement. As a consequence, the simulation and sensitivity analysis and optimization are conducted with a high degree of accuracy using this model. In the next step, the economic evaluation of the process is accomplished. An appropriate way of economic evaluation for this process is to use IHS Markit resources, which provide a detailed technical and economic analysis of the intended processes and report the raw material, sales, equipment, utilities, and other prices in detail. The documents related to the process of hydrodesulfurization of HFO were published in January 1975 (George 1975). Furthermore, price indices of 2018 are used to update prices. The report's title on the IHS Markit is Petroleum Desulfurization. However, hydrodesulfurization of light oil cuts such as gasoline, naphtha, etc. can be performed. This report emphasizes the removal of sulfur compounds from heavy oil cuts such as residual fuel oil, which is very similar to HFO (George 1975).

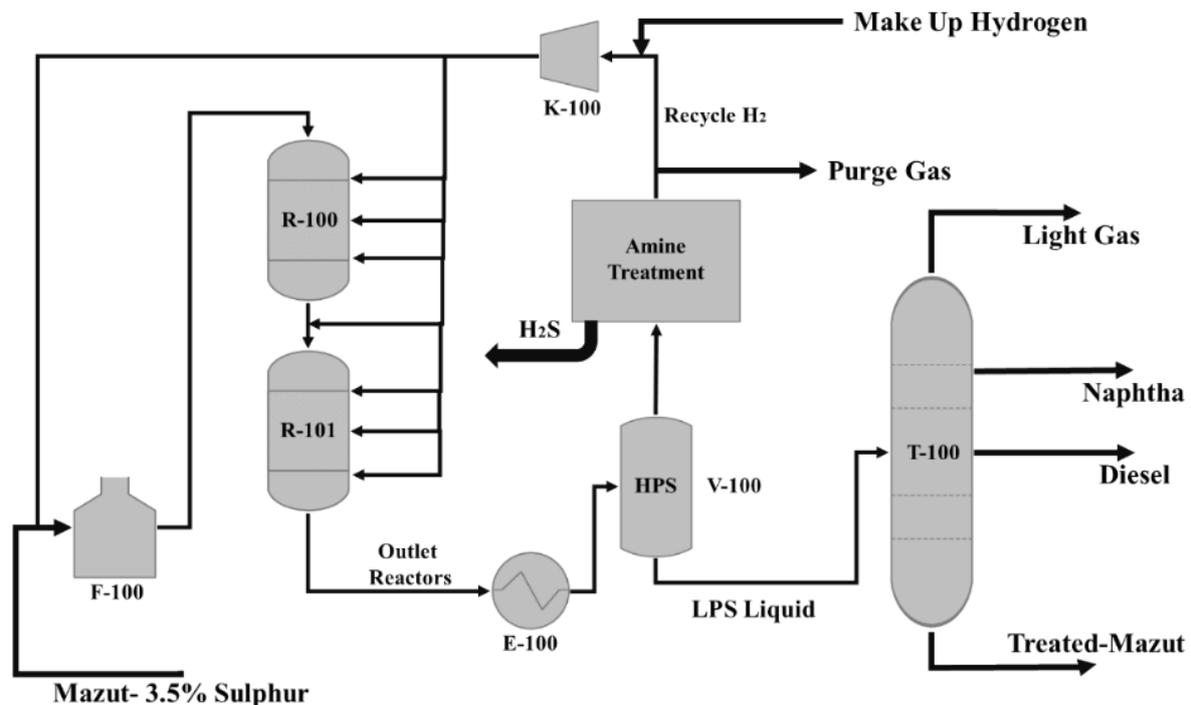
### 3. Simulation of HDS process

The purpose of this section is to design and simulate an HDS unit of the HFO with 3.5% sulfur and reduce its sulfur compounds to 0.5% in the final product (Valles et al. 2019). Since some hydrocracking products such as naphtha, diesel, etc. are produced in the process, the technical analysis of the sulfur compounds removal and the number of hydrocracking products and the amount of H<sub>2</sub>S produced in the process will be investigated. The HFO with the physical and chemical properties listed in Table 1 is mixed with the hydrogen stream after the pressure is increased and then heated by being passed through a furnace. Feed conditions, including volume flow, mass flow, temperature, and pressure are given in Table 2.

**Table 2**  
Conditions of HFO input to the process.

Feed	Value	Unit	Reference
Volume flow	250 (13.75)	m <sup>3</sup> /h (million bbl./year)	-
Mass flow	250 (549603)	Ton/h (lb./h)	-
Temperature	200 (401)	°C (F)	(George 1975)
Pressure	145 (2100)	bar (psia)	(George 1975)

Figure 4 shows the simulation flowsheet of the HDS unit of HFO. This simulated model is a one-stage type. A higher number of stages are used to increase the hydrocracking level of fuels (Calderón et al. 2019). This model may include a feed stream, one or more reactors, a high-pressure separator, a hydrogen recycling system, amine purification, a separation column, and a recycle stream. The amine purification model is a shortcut method in the purification of the HDS unit, which separates hydrogen sulfide from the high-pressure steam; thus, there is no need to simulate the amine unit individually (Chang et al. 2013). In this simulation, the amine purification model is ignored, and most of the hydrogen sulfide produced by the process is present in the Light Gas stream. A small portion of the hydrogen sulfide in the Purge Gas stream is sent to the sulfur recovery unit (SRU) (Jafari et al. 2018). The reaction section of this process consists of one stage with two reactors, each containing two catalytic beds corresponding to each reaction. Reactors include HDS, hydrogen denitrogenating (HDN), and hydrocracking. The first reactor is usually loaded with hydrotreating catalysts to remove sulfur and nitrogen compounds. It also converts HFO into a smaller range of more valuable products such as naphtha and diesel, with additional catalysts and hydrogen. In the second reactor, sulfur and nitrogen compounds are removed more intensely (Zhou et al. 2011).



**Figure 4**

A schematic of the HDS process as configured in the aspen HYSYS petroleum refining.

Table 3 presents the specifications of the internal diameter, catalyst loading, catalyst density, and bed voidage. In this HDS process, 41 tons of catalysts is needed per million barrels of heavy fuel oil for both reactors. Of this amount, 72% of the catalyst is used in the R-100 reactor and 28% in the R-101 reactor. Of course, the type of catalyst in the first and second reactors is different. As presented in Table 3, the catalyst density of the second reactor is different from that of the first one. The two most common catalysts for the HDS process are nickel-molybdenum/alumina and cobalt-molybdenum/alumina catalysts. In general, for the HDS process, due to the presence of more resistant sulfur compounds in the heavier fuels, nickel-molybdenum catalysts are more suitable. In the first bed of the first reactor, the separation reactions of vanadium and nickel occur, and in the second bed of the first reactor, the HDS and HDN reactions are carried out. In the two beds of the second reactor, reactions for further separation of sulfur and nitrogen occur; finally, the hydrocracking reaction of the HFO is partially carried out.

**Table 3**

Catalyst specifications (Bose et al. 2015).

	Internal diameter	Catalyst loading	Catalyst density	Bed voidage
<b>Reactor 1</b>				
<b>Bed-1</b>	4 m	163.8 ton	972 kg/m <sup>3</sup>	0.3
<b>Bed-2</b>	4 m	245.7 ton	972 kg/m <sup>3</sup>	0.3
<b>Reactor 2</b>				
<b>Bed-1</b>	4 m	73.7 ton	945 kg/m <sup>3</sup>	0.3
<b>Bed-2</b>	4 m	81.88 ton	945 kg/m <sup>3</sup>	0.3

Since the hydrocracking reactions are exothermic, the flow temperature increases as it passes the catalyst beds. A cool hydrogen flow or a quench flow is used to control temperature rise in beds (Gökçe et al. 2011). One of the most important parts of the simulation process is to enter the standard gas-to-oil volumetric ratio (Luo et al. 2018). For one barrel of HFO, approximately 900 SCF of hydrogen is required, which indicates approximately 40000 STD m<sup>3</sup>/h of hydrogen is required for 250 cubic meters of HFO. Thus, the gas-to-oil ratio will be 160 STD-m<sup>3</sup>/m<sup>3</sup> (George 1975). Table 4 tabulates the composition of the hydrogen makeup stream into the reactors. Both the compressor outlet pressure and inlet pressure drop to the reactor must be correct as they are used to calculate the inlet pressure to the reactor. Another important parameter in the simulation of this process is the input of the inlet temperature of the beds and the high-pressure separator (HPS) temperature. The information relevant to this and the other important parts of the simulation is summarized in Table 5.

**Table 4**

Composition of hydrogen makeup stream (Chang et al. 2013).

	Make-Up 1	Make-Up 2
<b>Composition mole fraction</b>		
<b>N<sub>2</sub></b>	0	0
<b>H<sub>2</sub>S</b>	0	0
<b>H<sub>2</sub></b>	0.85	0.85
<b>NH<sub>3</sub></b>	0	0
<b>Methane</b>	0.1	0.1
<b>Ethane</b>	0.05	0.05

**Table 5**

Parameters required for the simulation of the HDS process.

		Unit
<b>Gas-to-oil ratio (Reactor-1)</b>	160	
<b>Quench flow to Bed-1 (Reactor-2)</b>	500	
<b>Inlet temperature (Bed-1 and Bed-2)</b>	371	°C
<b>Temperature (HPS)</b>	49	°C
<b>Pressure (HPS)</b>	145	bar
<b>Purge fraction (Hydrogen Purge)</b>	0.03	
<b>Outlet Pressure (Compressor)</b>	145	bar
<b>Outlet Temperature (Compressor)</b>	49	°C
<b>Delta pressure to reactor inlet</b>	500	kPa
<b>Product temperature</b>	250	°C
<b>Products pressure</b>	100	kPa

#### 4. Results and discussion

The simulation results and sensitivity analysis of the HDS process to separate sulfur compounds up to 0.5 wt % and the production rate of the hydrocracking products are investigated in this section. The operating parameters that affect the removal of sulfur compounds from the HFO in this unit are the catalyst loading, the pressure of the HDS process, the gas-to-oil ratio, and the inlet temperature of the beds, which are listed in Table 6. In a sensitivity analysis of this process, by changing each operating parameter, the remaining operating parameters are assumed to be constant. The values of the changes in the effective operating parameters were also obtained using the Aspen HYSYS case study tool. It should be noted that the removal of sulfur and nitrogen compounds from the fuel oil, as well as the production of the hydrocracking products, is consistent and has no interaction. This indicates that by increasing the removal of the sulfur compounds from the HFO, the number of the by-products such as naphtha and diesel increases as well. Therefore, in the first part of the result analysis and discussion, the results of the simulation and validity of the model are examined, and in the second part, the economics of the process is analyzed.

**Table 6**

Effective operating parameters for the sensitivity analysis of the HDS process.

Operating parameters	Value	Down	Up	Unit
<b>Catalyst loading</b>	565	100	600	ton
<b>Gas-to-oil ratio</b>	160	50	300	STD-m <sup>3</sup> /m <sup>3</sup>
<b>Pressure</b>	145	30	150	bar
<b>Inlet temperature (Beds)</b>	371	250	390	°C

#### 4.1. Results of HDS process simulation

In the simulation of the HDS process, the connection between the flows in the main flow and sub-flows is well established. Then, by changing the effective operating parameters, the sensitivity analysis of this process is carried out. The simulation results of the HDS process are reported in Table 7. The first parameter to investigate is the catalyst loading. The distribution of the catalysts is as follows: 29% of the total catalyst in the first bed of the first reactor, 43% of the total catalyst in the second bed of the first reactor, 13% of the total catalyst in the first bed of the second reactor, and 15% of the total catalyst in the second bed of the second reactor (Wang et al. 2017).

**Table 7**

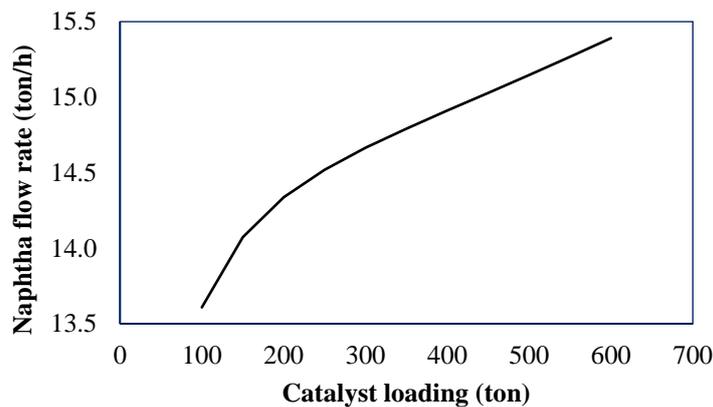
The simulation results of the HDS process.

HPS liquid	Simulation	Unit
<b>Sulfur content of HPS liquid</b>	0.50	Mass %
<b>Light gas</b>	9.02	(ton/h)
<b>Naphtha</b>	15.30	(ton/h)
<b>Diesel</b>	3.04	(ton/h)
<b>Hydrocracked products</b>	10.86	Mass %
<b>Bottoms (treated HFO)</b>	22.47	(ton/h)
<b>Bottoms (treated HFO)</b>	89.14	Mass %
<b>Purge</b>	4.98	(ton/h)
<b>Total H<sub>2</sub>S</b>	5.18	(ton/h)

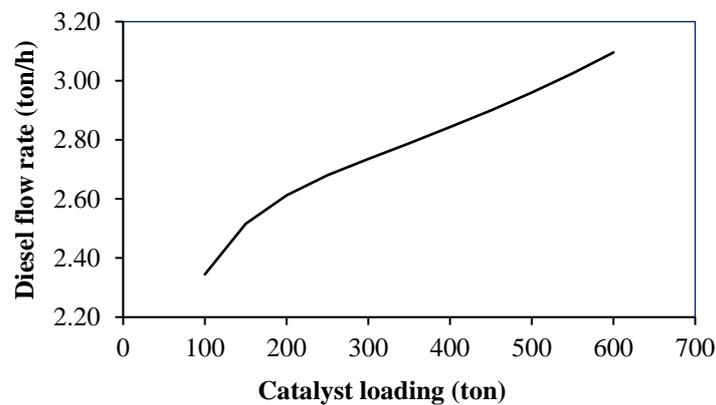
The HDS process is organized as a series of reaction sections, each having a differing loading (weight) of the catalyst. Typically, the inlet bed has the least amount of the catalyst, and the last bed has the most. This distribution of the catalyst loadings is typical in all the reformers and reflects that the highly endothermic reactions dominate the process during the initial beds of the reaction, which effectively slows down the reaction rate. Figure 5 shows delineates the diagrams of a) naphtha production, b) diesel production, c) H<sub>2</sub>S production, and d) sulfur content of the HFO (%) versus the catalyst loading (the volumetric space velocity of the feed). The volumetric space velocity of the feed is the ratio of the feed volume to the catalyst volume. Since the feed volume is constant, the spatial velocity varies with the volume of the catalyst correspondingly. As the catalyst density is almost constant, the changes are observed based on the mass of the catalyst consumed. The volumetric space velocity is chosen at the time of the design of the HDS process according to the mass percentage of the sulfur compounds in the feed and its lightness or heaviness. The level of the HFO treatment and the production of the by-products depend mainly on the feed volume space velocity. Reducing this speed increases the level of the HFO treatment and the production of the by-products. This figure clearly demonstrates that with increasing the catalyst loading (a reduction in the volumetric space velocity of the feed), the removal rate of sulfur, nitrogen, and hydrocarbons of the HFO increases. As a result of increasing the volume, the volumetric space velocity decreases, so the contact time and the conversion increase. Figures 5a and 5b illustrate the variations of naphtha and diesel productions versus the catalyst loading. These graphs show that with increasing the catalyst loading from 100 to 600 tons, the production of the diesel and naphtha increases from 2.34 and 13.61 ton/h to 3.10 and 15.39 ton/h respectively. This indicates that with a sixfold increase in the catalyst loading, the production of the diesel and naphtha rises by almost 32% and 13% respectively. Figure 5c shows the variation of the H<sub>2</sub>S production versus the catalyst loading. It is clear that with increasing the catalyst loading from 100 to 600 tons, the amount of H<sub>2</sub>S production

enlarges from 3.25 to 5.23 ton/h. Accordingly, with a sixfold increase in the catalyst loading, H<sub>2</sub>S production rises by 61%. Finally, Figure 5d plots the mass percentage of the sulfur content of the HFO versus the catalyst loading. It is obvious that by increasing the catalyst loading from 100 to 600 tons, the mass percentage of the sulfur content of the HFO decreases from 2% to 0.5%, which implies that with a sixfold increase in the catalyst loading, 75% of the sulfur content of the HFO is removed, and the international standard is reached.

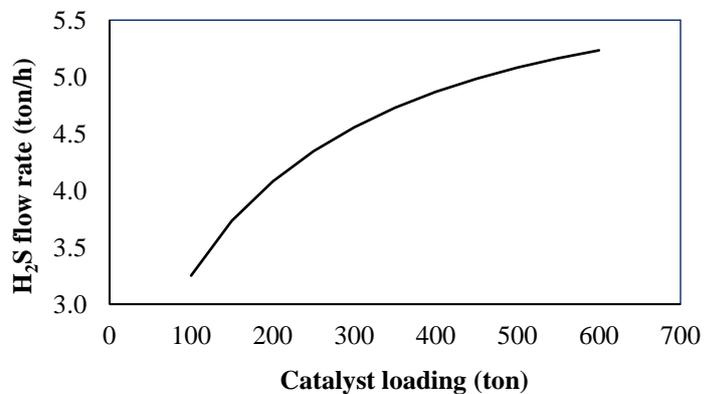
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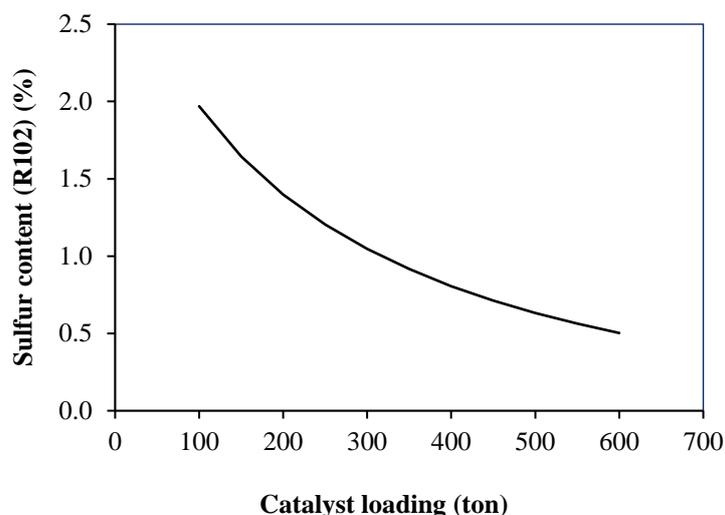
b)



c)



d)

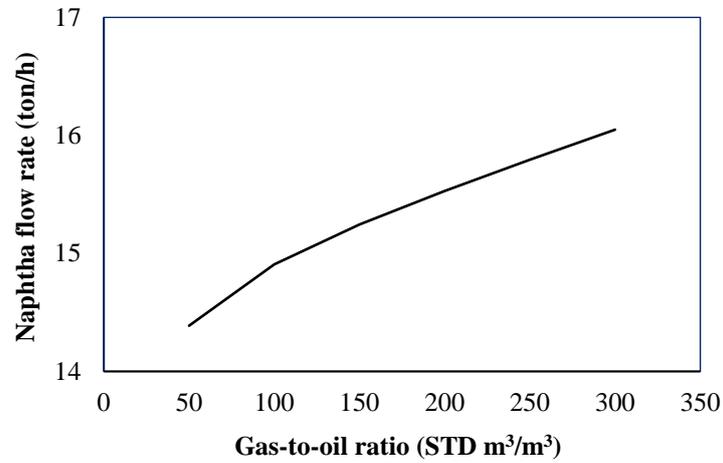
**Figure 5**

The variations of a) naphtha production, b) diesel production, c) H<sub>2</sub>S production, and d) sulfur content of the HFO versus the catalyst loading.

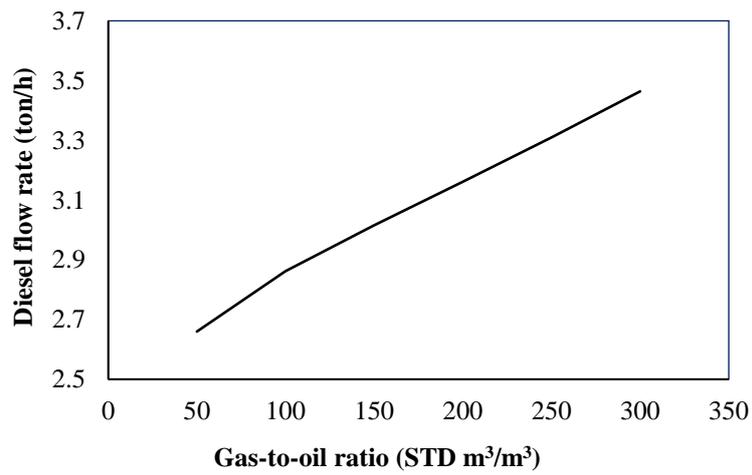
Increasing the gas-to-oil ratio (the volume ratio of the hydrogen consumed to the HFO fuel) raises the level of the HFO treatment and production of the by-products, which also depends on the general pressure of the system, the consumption of the hydrogen-containing gases, and the concentration of the hydrogen. The second operational parameter for the sensitivity analysis is the gas-to-oil ratio. Figure 6 delineates the variations of a) naphtha production, b) diesel production, c) H<sub>2</sub>S production, and d) sulfur content of the HFO (%) versus the gas-to-oil ratio. This figure shows that by increasing the gas-to-oil ratio, the rate of the removal of sulfur and nitrogen and the production of the hydrocracking by-products increase. Depending on the operating conditions, the amount of hydrogen consumed is different. Hydrogen is used in the hydrocracking reactions of paraffin and naphtha, in the saturation of olefins and aromatics, and in the decomposition of nitrogen and sulfur compounds. Most hydrogen consumption is related to the hydrotreating reaction of the HFO, that is, the removal of sulfur and nitrogen compounds, which can be calculated by the ratio of the average molecular mass of the feed and the products. The heavier the feed is, the higher the sulfur content of the fuel is, so it is more difficult to separate it; thus, the amount of hydrogen consumed increases. Moreover, as the hydrogen consumption rises, the production of the lighter products increases. It is important to note that the circulation of a considerable amount of hydrogen and the feed prevents the catalyst corrosion and extends its service life. Figures 6a and 6b illustrate the variations of naphtha and diesel production versus gas-to-oil ratio. It is clear that by enlarging this ratio from 50 to 300, naphtha and diesel production increases from 2.66 and 14.38 ton/h to 3.46 and 16.04 ton/h respectively. This indicates that with a sixfold increase in the gas-to-oil ratio, the diesel and naphtha production increases by 30% and 12% respectively. Figure 6c shows the variation of the H<sub>2</sub>S production versus the gas-to-oil ratio. This graph demonstrates that by increasing this ratio from 50 to 300 STD-m<sup>3</sup>/m<sup>3</sup>, the H<sub>2</sub>S production rises from 4.7 to 5.36 ton/h. Accordingly, with a sixfold increase in gas-to-oil ratio, the H<sub>2</sub>S production enlarges by 12%. Finally, Figure 6d presents the mass percentage of the sulfur content of the HFO versus the gas-to-oil ratio. It can be seen that by increasing this ratio from 50 to 300 STD-m<sup>3</sup>/m<sup>3</sup>, the mass percentage of the sulfur content of the HFO declines from 1.32% to 0.31%, that is to say, with a

sixfold increase in the volume of the hydrogen, the rate of the removal of the sulfur content rises by 425% times and reaches the international standard.

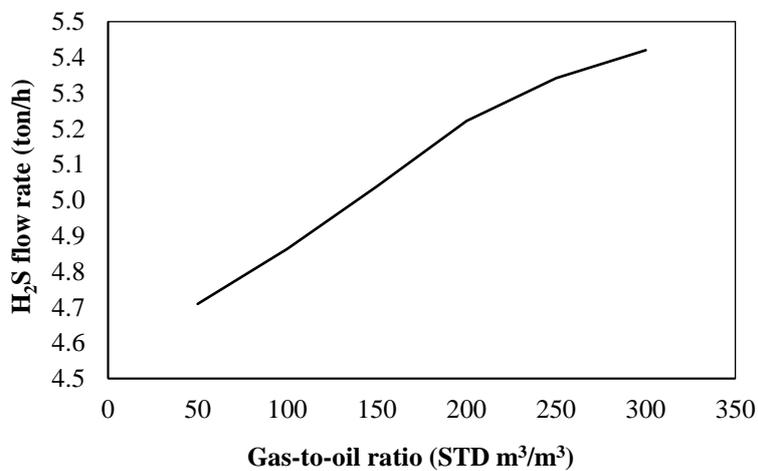
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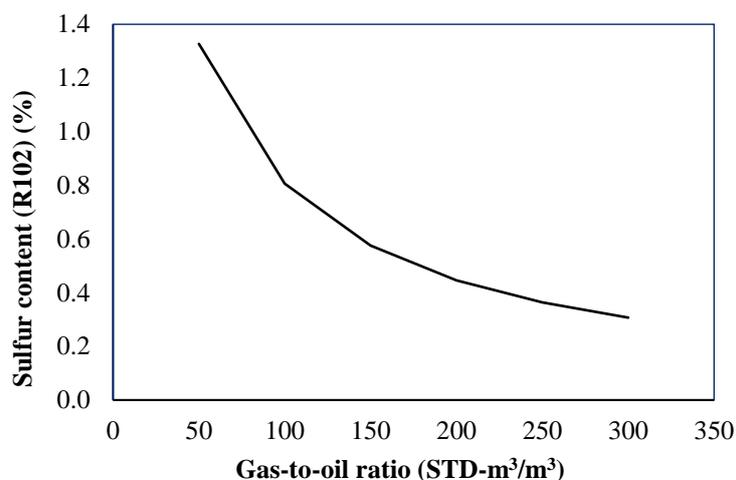
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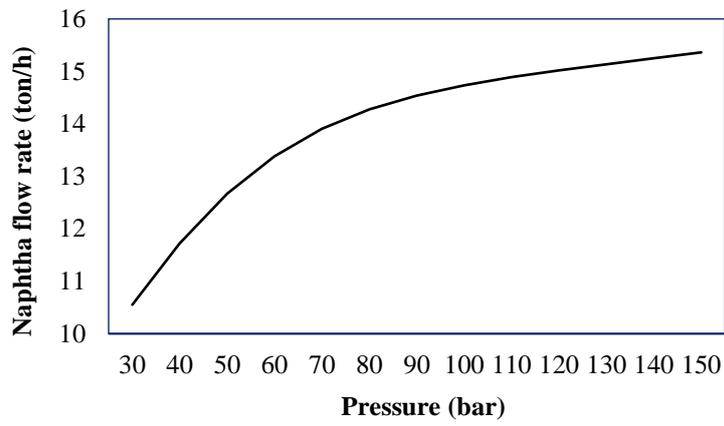
d)

**Figure 6**

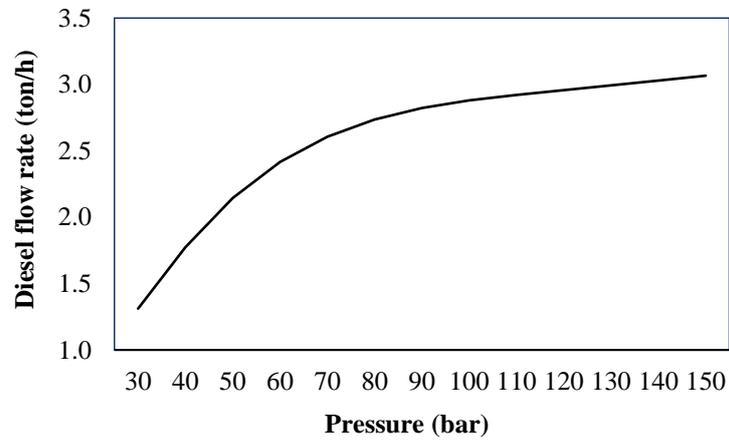
The variations of a) naphtha production, b) diesel production, c) H<sub>2</sub>S production, and a) sulfur content of the HFO (%) versus the gas-to-oil ratio.

Increasing the pressure of the HDS process raises the level of the HFO treatment and the production of the by-products. The third operational parameter for the sensitivity analysis is the pressure of the HDS process. Figure 7 shows the variations of a) naphtha production, b) diesel production, c) H<sub>2</sub>S production, and d) the sulfur content of the mazut (%) versus the pressure of the HDS process. This figure shows that by increasing the pressure, the rate of the removal of the sulfur and nitrogen and the production of the hydrocracking by-products increase. Figures 7a and 7b illustrate the variations of naphtha and diesel production against the HDS process pressure. It is obvious that with increasing pressure from 30 to 150 bar, diesel and naphtha production increases from 1.31 and 10.55 ton/h to 3.06 and 15.36 ton/h respectively. Accordingly, with a fivefold increase in the process pressure, diesel and naphtha production enlarges by 133% and 45% respectively. Figure 7c shows the variation of the H<sub>2</sub>S production against the pressure of the HDS process. It is obvious that by increasing the process pressure from 30 to 150 bar, the H<sub>2</sub>S production rises from 0.65 to 5.33 ton/h, which implies that with a fivefold increase in HDS process pressure, the H<sub>2</sub>S production enlarges by 720%. Finally, Figure 7d depicts the mass percentage of the sulfur content of the HFO versus the pressure of the HDS process. The graph shows that by increasing the pressure from 30 to 150 bar, the mass percentage of the sulfur content of the HFO declines from 3% to 0.5%. This indicates that with a fivefold increase in the process pressure, the removal rate of the sulfur content increases by 600% and reaches the international standard.

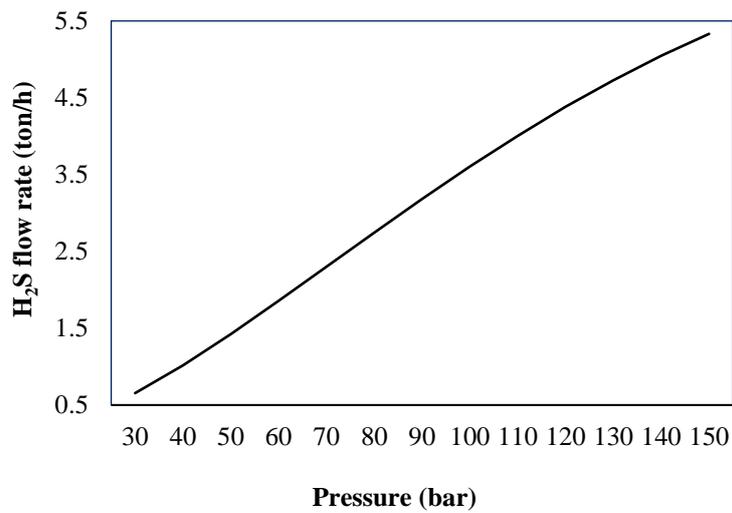
a)



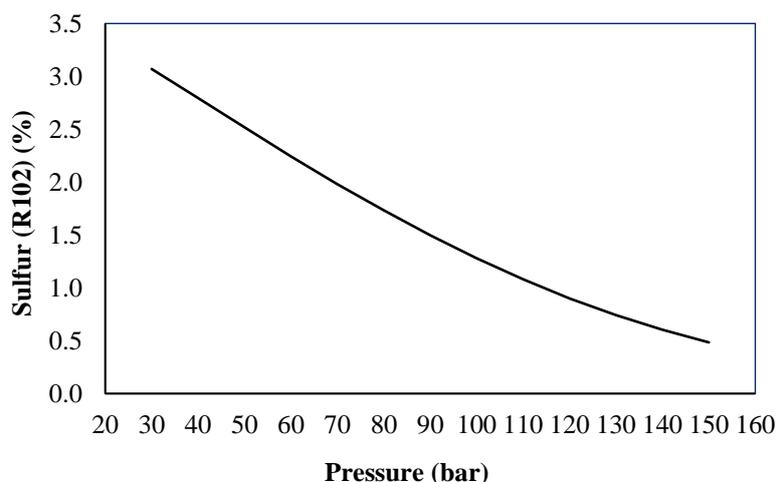
b)



c)



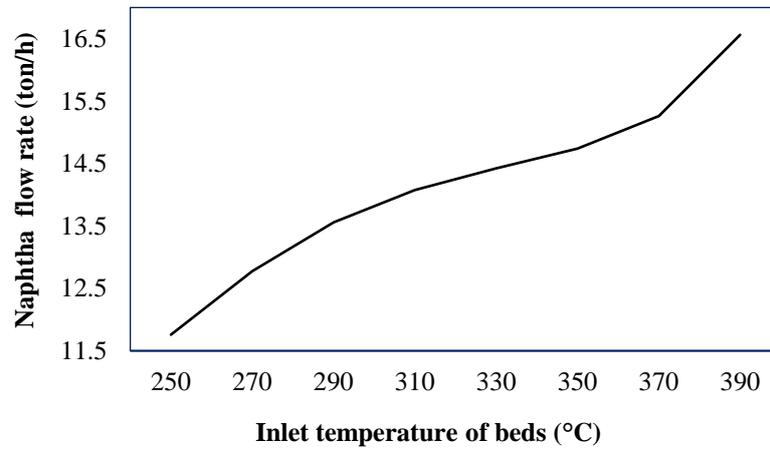
d)

**Figure 7**

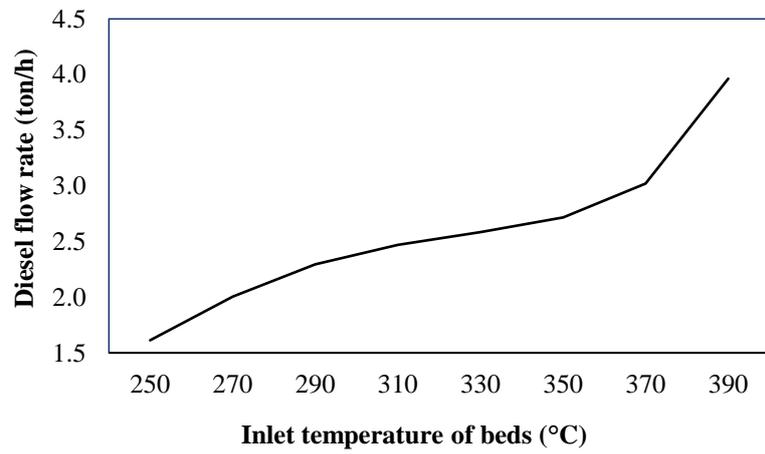
The variations of a) naphtha production, b) diesel production, c) H<sub>2</sub>S production, d) sulfur content of the HFO (%) versus the HDS process pressure.

Increasing the inlet temperature of the bed of the HDS process raises the level of the HFO treatment and the production of the by-products. The HDS process usually works at a temperature between 150 and 300 °C for light fuel oil and between 280 and 400 °C for heavy fuel oil. The inlet temperature of the beds can be increased if the catalyst activity decreases due to operation and does not respond to the demand. The fourth operational parameter for the sensitivity analysis is the inlet temperature of the beds. Figure 8 displays the variations of a) naphtha production, b) diesel production, c) H<sub>2</sub>S production, and d) the sulfur content of the HFO (%) versus the inlet temperature of the beds. This figure confirms that with increasing the inlet temperature of the beds, the rate of the removal of sulfur and nitrogen content and the production of the hydrocracking by-products increase. Figures 8a and 8b illustrate the mass flow rate of naphtha and diesel production against the inlet temperature of the beds. It is obvious that by increasing this temperature from 250 to 390 °C, the diesel and naphtha production improves from 1.61 and 11.75 ton/h to 3.96 and 16.56 ton/h respectively. Accordingly, with a 55% increase in the temperature, the production of the diesel and naphtha enlarges by 146% and 40% respectively. Figure 8c shows the changes in the H<sub>2</sub>S production versus the inlet temperature of beds. It demonstrates that as the inlet temperature of the beds increases from 250 to 390 °C, H<sub>2</sub>S production rises from 2.10 to 5.52 ton/h, which implies that with a 55% increase in the inlet temperature of the beds, the rate of H<sub>2</sub>S production enlarges by 262%. Finally, Figure 8d depicts the changes in the mass percentage of the sulfur content of the HFO versus the inlet temperature of the substrates. It is clear that by increasing the inlet temperature of the beds from 250 to 390 °C, the mass percentage of the sulfur content of the HFO declines from 2.7% to 0.35%, that is to say, with a 55% increase in the inlet temperature of the beds, the removal of the sulfur content of the HFO enhances by 770%, and the international standard is met.

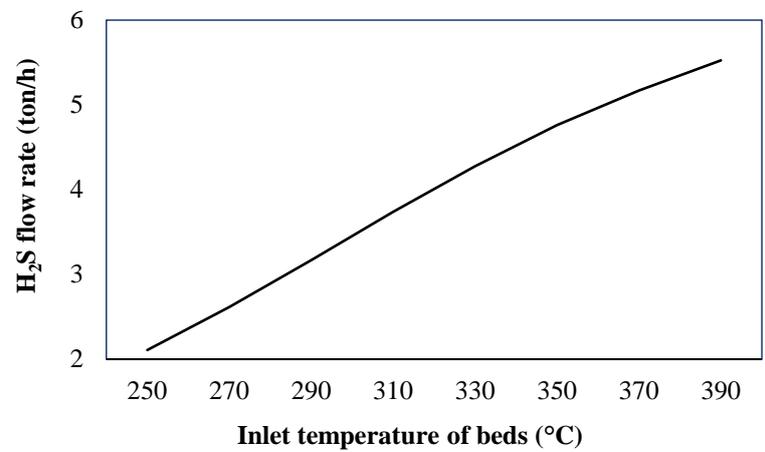
a)



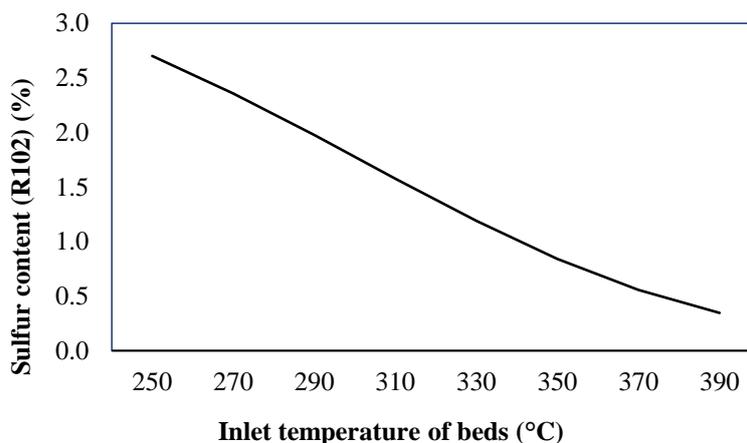
b)



c)



d)

**Figure 8**

The variations of a) naphtha production, b) diesel production, c) H<sub>2</sub>S production, and d) sulfur content of the HFO (%) versus the inlet temperature of the beds.

#### 4.2. Economic evaluation of HDS process

The economic evaluation of the HDS process of HFO with the estimates of the total capital investment and the net production cost is set to achieve the global standard for the sulfur content in the fuel. Tables 8 and 9 are used to obtain the total capital investment. The price of the entire process, the cost of the utilities, and the total investment cost of the HDS process are all reported in million USD/ (13.75 million bbl/year). It is inferred from Table 8 that 89% of the total process cost is for the reaction equipment, 9% for the compressors, and 2% for the product stabilizer. However, due to the different capacity exponent (CE) in each part of the process, these ratios will also change with increasing or decreasing the flow of the input fuel. Also, 75% of the whole cost of the cooling water is for the reaction, 18% for the compressors, and 7% for the product stabilizer. In order to use the values in Tables 8 and 9 to estimate the cost based on different dimensions and capacities, the power capacity law (Equation (1)) is used (Peters 1968).

$$C_E = C_B(Q/Q_B)^M \quad (1)$$

where  $C_E$  equals  $Q_B$ ,  $C_B$  equals  $Q_B$ , and  $M$  is the capacity exponent for each equation. Tables 10 and 11 are used to obtain the net production cost. In order to estimate the net production cost, the total direct operating cost, which is the sum of the raw materials, utilities, and labor costs, must be calculated. The total production cost is calculated from the total direct operating costs, plant overhead, taxes and insurance, depreciation, and interest on the working capital. Since some part of the oil is converted into valuable products such as naphtha and diesel, which will be sold, by subtracting the by-product sales from the total production cost, the net product cost, or the net production cost, is calculated (George 1975). Finally, by calculating the net production cost and the net profit, the sale price of oil on the market is determined. Since annual taxes and depreciation are included in net production cost, the net profit is equal to the selling price of the products minus the net production cost. The selling price of the sulfurized and refined petroleum is between \$200 and \$250 per ton, so the payback period is as follows:

$$\text{Payback period (year)} = \text{Total capital investment} / \text{Net profit}$$

**Table 8**

The cost of the total process and utilities of the HDS process (million USD/(13.75 million bbl/year)).

Section	Total		Reaction		Compressors			Product Stabilizer		
	Cost Million USD	Cost Million USD	Cost Up	CE Down	Cost Million USD	CE Up	CE Down	Cost Million USD	CE Up	CE Down
<b>Equipment</b>										
Reactors	47.9	47.9	0.5	0.5	0.0	-	-	0.0	-	-
Columns	0.2	0.0	-	-	0.0	-	-	0.2	0.7	0.6
Vessels and tanks	4.0	3.9	0.5	0.5	0.0	-	-	0.0	0.4	0.4
Exchangers	8.6	7.0	0.5	0.5	0.5	0.6	0.5	1.2	0.6	0.5
Furnaces	2.6	2.6	0.8	0.8	0.0	-	-	0.0	-	-
Compressors	11.4	0.0	-	-	11.4	1.0	0.8	0.0	-	-
Pump	2.6	2.1	1.0	0.9	0.0	-	-	0.5	1.0	0.8
<b>Total</b>	77.4	63.6	0.5	0.5	11.9	1.0	0.8	2.0	0.7	0.6
Feed filter	1.1	1.1	0.6	0.6	0.0	-	-	0.0	-	-
<b>Total process cost (TPC)</b>	206.0	156.1	0.5	0.5	44.5	0.9	0.7	5.4	0.7	0.6
<b>Utilities and tankage</b>										
Cooling water	4.8	3.6	1.0	0.9	0.9	1.0	0.9	0.3	1.0	0.9
<b>Total utilities cost (TUC)</b>	5.8	4.3	1.0	0.8	1.1	1.0	0.8	0.4	1.0	0.8
<b>Total process and utilities cost (TPUC)</b>	211.7	160.4	0.5	0.5	45.6	0.9	0.7	5.8	0.7	0.6

**Table 9**

The total investment of the HDS process (million USD/(13.75 million bbl/year)).

Total capital investment	Million USD	
Total process and utility cost (TPUC)	TPC+UC	211.7
General service facilities (GSF)	0.15 (TPUC)	31.8
Total fix capital (TFC)	TPUC + GSF	243.5
Interest on construction loan at 9%/year (ICL)	0.09 (21/24) TFC	19.2
Start-up cost (SUC)	0.06 (TFC)	14.6
Working capital (WC)	0.13 (TFC)	31.7
<b>Total capital investment, not including land</b>	<b>TFC + ICL + SUC + WC</b>	<b>308.9</b>

**Table 10**

The total direct operating cost of the HDS process (million USD/Year).

	<b>Basis or unit cost</b>	<b>USD/bbl</b>	<b>Million USD/Year</b>
<b>Labor</b>			
<b>Operating</b>	3 men/shift, 53 USD/man-hr.	0.08	0.98
<b>Maintenance</b>	3%/yr. of battery limits cost	0.40	4.81
<b>Control laboratory</b>	20% of operating labor	0.02	0.19
<b>Total labor cost (TLC)</b>		0.50	5.98
<b>Materials</b>			
<b>Catalyst</b>	7.4 USD/lb.	0.67	8.06
<b>Makeup hydrogen</b>	4.71 USD/thousand scf	4.24	50.99
<b>Maintenance</b>	3%/yr. of battery limits cost	0.40	4.80
<b>Operating</b>	10% of operating labor	0.01	0.10
<b>Total materials cost (TMC)</b>		5.32	63.95
<b>Utilities</b>			
<b>Cooling water</b>	0.17 USD/1000 gal	0.09	1.13
<b>Electricity</b>	0.1 USD/kwh	0.64	7.70
<b>Fuel oil</b>	7.4 USD/million Btu	0.23	2.76
<b>Total utilities (UC)</b>		0.96	11.59
<b>Total direct operating cost (TDOC)</b>	Labor + Materials + Utilities	6.78	81.53

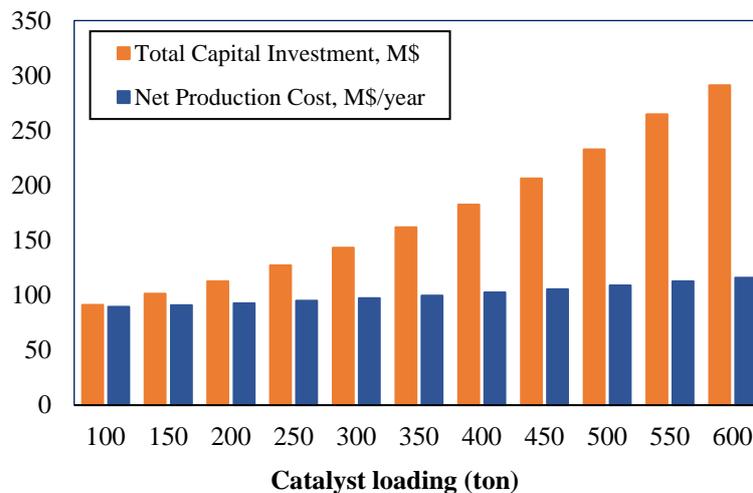
**Table 11**

The net product cost of the HDS process (Million USD/Year).

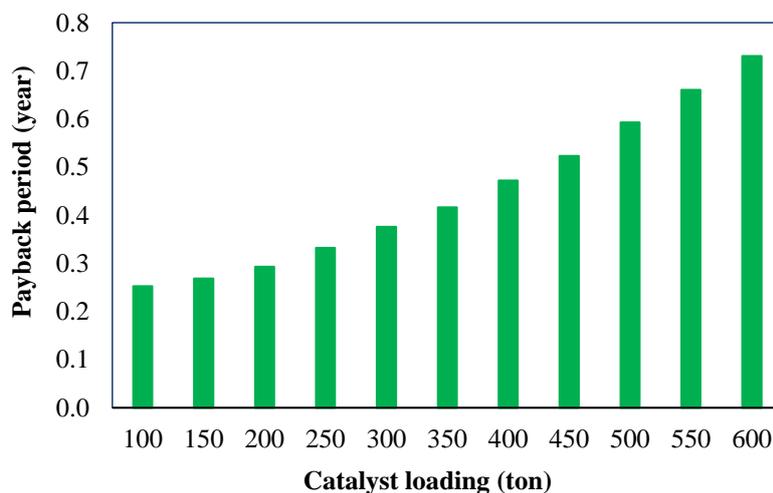
<b>Section</b>	<b>Basis or unit cost</b>	<b>USD/bbl</b>	<b>Million USD/Year</b>
<b>Plant overhead (PO)</b>	80% of total labor	0.40	4.79
<b>Taxes and insurance (TAI)</b>	2%/yr. of fix capital	0.35	4.26
<b>Plant cost (PC)</b>	TDOC + PO + TAI	7.53	90.57
<b>C&amp;A, sales, research (GS&amp;R)</b>	2%/yr. of fix capital	0.35	4.26
<b>Cash expenditures</b>	PC + (GS&R)	7.88	94.83
<b>Depreciation</b>	10%/yr. of fixed capital	1.77	21.31
<b>Interest on working capital</b>	10%/yr.	0.17	2.05
<b>Total production cost</b>		9.82	118.19
<b>By-product credit</b>			
<b>Unstabilized naphtha</b>		0.18	2.18
<b>Fuel gas and diesel</b>		0.12	1.49
<b>Net production cost</b>		9.52	114.52

Figures 9a and 9b show the changes in the total capital investment, the net production cost, and the payback period against the variation in the catalyst loading. It is obvious that as the catalyst loading increases from 100 to 600 tons, the total capital investment and the net production cost rise from 91 million USD and 89 million USD/year to 291 million USD and 116 million USD/year respectively. This implies that a sixfold increase in the catalyst loading raises the total capital investment and the net production cost by 219% and 30% respectively. The reason for the increase in the total capital investment is that 75% of the total cost of the process is related to the price of the reaction section, including reactors, separators, and converters. Since the volume of the reactors increases as the catalyst loading rises, it significantly impacts on the increase in the total capital investment. The payback period also extends from 3 to 8.7 months.

a)



b)



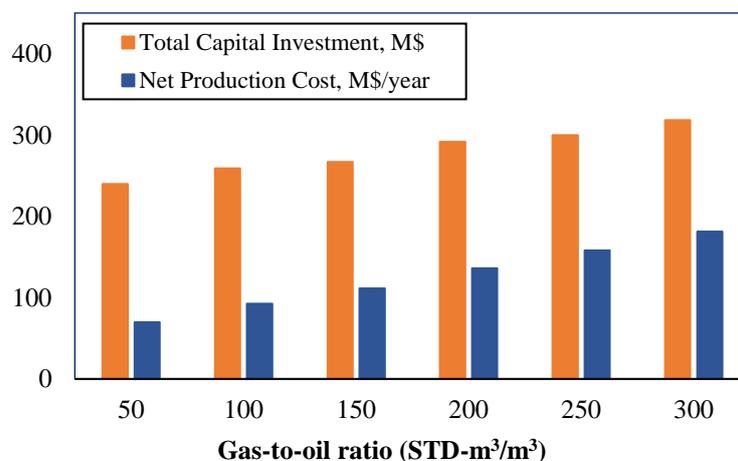
**Figure 9**

The variations of a) the total capital investment along with the net production cost and b) the payback period versus the changes in the catalyst loading.

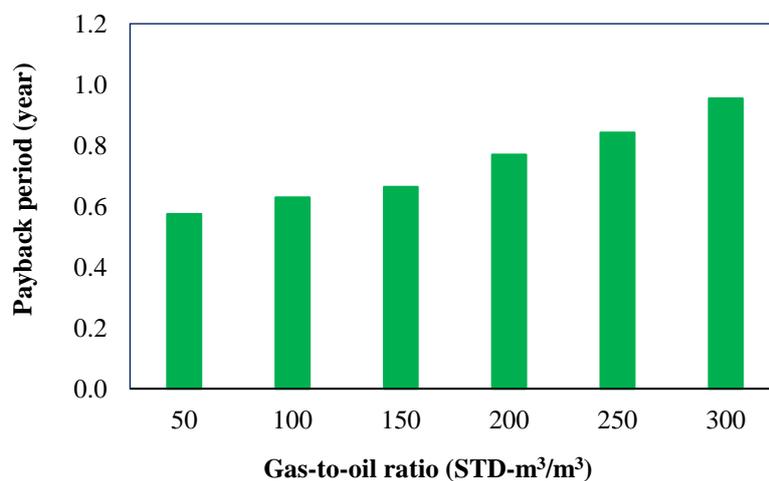
The ratio of the gas (hydrogen) to oil is a critical operating variable for the HDS process. It has two opposite effects on the product distribution and process profitability. A higher gas-to-oil ratio can

enhance aromatic hydrogenation, can increase the H/C of the products, and can extend the catalyst life by reducing coke precursors. Further, a higher gas-to-oil ratio leads to higher hydrogen consumption, which raises the processing cost. Figures 10a and 10b show the changes in the total capital investment, the net production cost, and the payback period against the variation in the gas-to-oil ratio. It is clear that as the gas-to-oil ratio increases from 50 to 300 STD- $\text{m}^3/\text{m}^3$ , the total capital investment and the net production cost rise from 240 million USD and 69.5 million USD/year to 319 million USD and 181 million USD/year respectively. Accordingly, with a sixfold increase in gas-to-oil ratio, the total capital investment and the net production cost rise by 33% and 160% respectively. The reason for this significant increase in the net production cost is that 62% of total direct operating costs is related to the hydrogen consumption. The payback period also extends from 7 to 11.5 months.

a)



b)



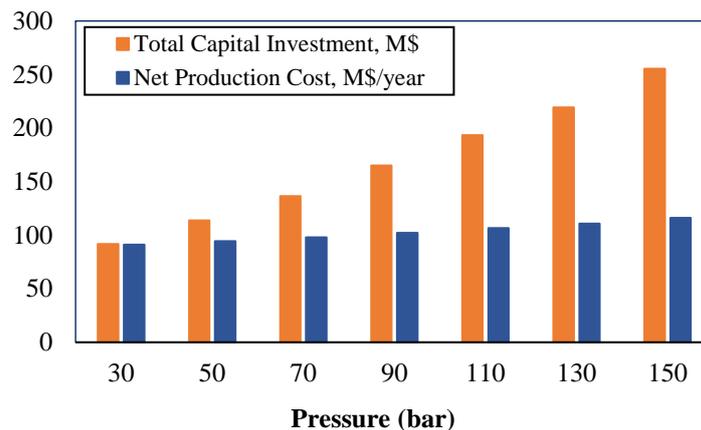
**Figure 10**

The variations of a) the total capital investment along with the net production cost and b) the payback period with the gas-to-oil ratio.

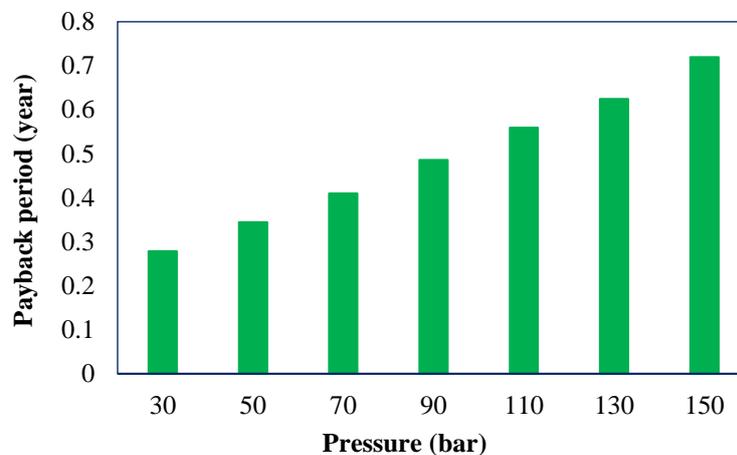
Figures 11a and 11b show the changes in the total capital investment, the net production cost, and the payback period against the pressure of the HDS process. These graphs demonstrate that as the pressure

of the HDS process increases from 30 to 150 bar, the total capital investment and the net production cost increase from 105 million USD and 91 million USD/year to 286 million USD and 116 million USD/year respectively. This implies that with a fivefold increase in the pressure of the HDS process, the total capital investment and the net production cost enlarge by 172% and 27% respectively. As the pressure increases, a factor of the multiplier is applied to the reactors and separators. Since over 80% of the equipment price is related to the price of the reactors and separators, the increase in the pressure of the process significantly impacts on the rise in the total capital investment. The payback period also extends from 3.3 to 8.7 months.

a)



b)



**Figure 11**

The variations of a) the total capital investment along with the net production cost and b) the payback period with the pressure of the HDS process.

## 5. Conclusions

The objective of the HDS process is to reach the permissible sulfur compound content of the HFO, that is, to reduce the sulfur compound content of the HFO from 3.5 to 0.5 wt %. The economic evaluation of this process was carried out, and the total capital investment and the net production cost of the process

were accurately estimated using the IHS Markit. In the simulation and economic evaluation of this process, the following results were obtained:

- With a 100% increase in the catalyst loading, technically the diesel production, the naphtha production, and the H<sub>2</sub>S production rise by 6.4%, 2.6%, and 12.2% respectively, which leads to 15% removal of the sulfur content of the HFO. Economically, the total capital investment, the net production cost, and the payback period increase by 6.4%, 32%, and 38% respectively.
- With a 100% increase in the gas-to-oil ratio, technically the diesel production and the naphtha production rise by 6.0% and 2.4% respectively; the H<sub>2</sub>S production decline by 6.2%. This increase in the gas-to-oil ratio also removes 15.3% of the sulfur content of the HFO. Economically, the total capital investment, the net production cost, and the payback period increase by 43.8%, 6%, and 12.8% respectively
- With a 100% increase in the pressure of the HDS process, technically, the diesel production, the naphtha production, and the H<sub>2</sub>S production rise by 32.25%, 11.25%, and 180% respectively. This also removes 20.75% of the sulfur content of the HFO. Economically, the total capital investment, the net production cost, and the payback period rise by 43%, 6.75%, and 40.75% respectively.
- With a 5% increase in the inlet temperature of the beds, technically the diesel production, the naphtha production, and the H<sub>2</sub>S production enlarge by 13.72%, 3.63%, and 15% respectively. This will also remove 7.9% of the sulfur content of the HFO. The negligible effects of changing the inlet temperature of the beds on the economic evaluation are neglected.

The reason for the more significant increase in the net production cost of the HDS process when the hydrogen consumption increases, compared to when the pressure of the process and the catalyst loading rise, is that hydrogen consumption is 6.33 times higher than the catalyst loading and 47 times more than the electricity consumption, affecting the direct operating costs. The results demonstrate that the net production cost must be reduced for the HDS process to be economically more profitable. One of the effective parameters to reduce the net production cost is to minimize the amount of hydrogen consumed because hydrogen has the most significant influence among the operating parameters affecting the increase in the net production cost; it is also possible to consider building a hydrogen production unit in the vicinity of the HDS process. In general, increasing the level of sulfur content removal and increasing the total capital investment or the net production cost are two opposing objective functions. Accordingly, while industries tend to reduce sulfur content of HFO, they tend to reduce the net production costs. Therefore, it is suggested that this process should be used with multi-objective optimization algorithms. Also, economically, the HDS process for removing the sulfur content of HFO will be compared with other processes suitable for the treatment of heavy hydrocarbon cuts such as the oxidative desulfurization (ODS) process.

## Nomenclature

API	American Petroleum Institute
CE	Cost index
FCC	Fluid catalytic cracking
HCRSRK	Hydrocracking modified Redlich Kwong
HDN	Hydrodenitrogenation
HDS	Hydrodesulfurization
HFO	Heavy fuel oil

HIS	Information handling services
HPS	High-pressure steam
MW	Megawatt
ODS	Oxidative desulfurization
PC	Plant cost
PO	Plant overhead
SRU	Sulfur recovery unit
TAI	Taxes and insurance
TLC	Total labor cost
TMC	Total material cost
TPC	Total process cost
TPUC	Total process and utility cost
TUC	Total utilities cost
USD	United States Dollar

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