



A technical and economic assessment of fuel oil hydrotreating technology for steam power plant SO₂ and NO_x emissions control

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ABSTRACT

This work presents a simulation approach to the design and economic evaluation of fuel oil hydrotreating processes for the control of SO₂ and NO_x emission in an Iranian steam power plant. The percent of fuel oil desulphurization was estimated from the SO₂ emissions standards for power plants. Based on two different scenarios according to (I) European and (II) Iranian standards, the design and simulation of hydrodesulphurization reactors, separation, heat recovery, and amine gas sweetening sections were performed and an economic assessment of the plant was investigated for each scenario. The results indicated the following: the cost price of each barrel of low sulfur fuel oil produced in such plants was estimated at 4.24 US\$ for scenario (I) and 3.17 US\$ for scenario (II); the levelized cost of desulfurization in power production was estimated at 0.618 c/kWh for (I) and 0.463 c/kWh for (II); and the social cost savings of replacing every barrel of low sulfur fuel oil instead of one barrel of high sulfur fuel oil were estimated at 26.7 US\$ for (I) and 13.6 US\$ (II).

1. Introduction

The control and reduction of power plant air emissions are central to the discussion of sustainable developments in fossil-fuel power stations. Sulfur-dioxide (SO₂) and nitrogen oxide (NO_x) emissions generated during fuel oil combustion in power plants have adverse effects on human health and the environment. In recent years, these adverse effects have attracted attention in Iran. The SO₂ generated by heavy oil are more important than other pollutants because exposure to them results in serious human health problems, in some cases leading to death; they also have negative effects on the environment. The residual fuel oils consumed in an Iranian power plant contain an average sulfur content of about 3 percent. According to a Statistical Report [1] for 2012 to 2013, the share of fuel oil in the power plant's total spent fuel was more than 25%. This led to a release of 890000 tonnes of SO₂ into the atmosphere and caused significant air pollution. It is important to mention that most of the Iranian fuel-oil-burning power plants are located near highly populated cities and exposes more of the population to the adverse effects of SO₂ emissions.

Hydrodesulphurization (HDS) is a catalytic chemical process widely used to remove sulfur from refined petroleum products such as gasoline or petrol, jet fuel, kerosene, diesel fuel, and fuel oils [2]. The HDS process was first employed by the Gulf Oil Company in Japan in the early 1960s for the production of low-sulfur fuel oil from atmospheric residue. In the decades that followed, various proprietary catalytic hydrodesulphurization processes have been commercialized in such a way that nowadays it is the most widely used method for the desulfurization of atmospheric and vacuum residues [3]. In addition, considerable efforts have been made to develop new alternative processes for the desulfurization of liquid fuels such as extractive desulfurization, extraction with ionic liquids, adsorptive desulfurization with solid adsorbents, bio-desulfurization, supercritical water based desulfurization, and electrochemical desulfurization [4, 5]. Although the above methods have comparative advantages when compared to the hydrotreating technology for light and medium petroleum distillates, applying these methods for heavy distillates such as residual fuel oil results in several drawbacks. Due to the high solvent volume relative to the

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extracted sulfur compounds, it is generally preferable to make use of a solvent that has a higher boiling point than the sulfur compounds. For heavy oil extraction, this is not an option and a lighter boiling solvent must be used. Thus, this increases the solvent recovery cost significantly for extractive processes like extractive desulfurization and ionic liquid extraction [5]. In the case of extractive desulfurization, several factors erode the cost effectiveness of extractive processes for the desulfurization of heavy distillates more than light and medium distillates. These factors include the following: the poor solubility of the heavy organosulfur compounds in the solvent which limits the efficiency of this method for heavy distillates; the high-viscosity of heavy distillates that leads to inefficient mixing [6]; the need to make use of a light solvent; and the potential loss of solvent by dissolution in such a complex matrix such as heavy distillates. In the case of ionic liquid extraction, the ionic liquids are high boiling solvents and their recovery is more challenging than with organic solvents, especially for heavy distillates like residual fuel oil. Other drawbacks such as the high cost and the water sensitivity of ionic liquids detract from its large-scale industrial application [7]. With regard to adsorptive desulfurization for heavy oils, the performance of even the most efficient of the adsorbents is still insufficient for industrial applications due to the poor accessibility of large molecules in the narrow pores of the adsorbent that reduces adsorption effectiveness [8]. In regard to bio-desulfurization, it is not commercially employed at the refinery because of the longevity of the method which restricts it from continuous application. Sanitary handling, shipment, storage, and use of microorganisms within the production field or refinery environment are the other drawbacks of this method [9]. Due to the complexity of the physical and chemical characteristics of fuel oil and the presence of complex sulfur compounds (benzothiophene, di-benzothiophene and even heavier thiophene derivative), widespread commercial use of oxidative desulfurization, supercritical water, and electrochemical desulfurization have encountered several drawbacks for the desulfurization of this fuel [4, 5, 10-13]. Considering all the difficulties caused by harsh operating conditions and the preparation of hydrogen, hydrotreating processes are still the preferred method of desulfurization from heavy liquid distillates and residual fuel oil. According to the statistical tables reported in the OGJ Worldwide Refinery Survey [14] the desulfurization capacity of atmospheric and vacuum residues was about 1.57 million barrels per day in 2012. This capacity was only related to HDS units which produced low-sulfur fuel oil (LSFO) as a final product in 28 refineries in Belgium, Canada, China, Taiwan, Japan, Kuwait, South Korea and the United States. The fuel oil which was used as feeds for hydrocracking and other refinery units were not included in these statistics [14, 15]. Also, the hydrotreating process has been licensed by several companies around the

world. The high capacity and continuity of this process has made it an option for reducing SO₂ emissions in Iranian power plants [16]. The objective of this study was to establish the economic feasibility of the hydrodesulfurization process for reducing the sulfur content of fuel oils spent in the Shazand power plant located in Arak, Iran [13]. To this end, the desired maximum sulfur content for an acceptable fuel oil was calculated with regard to international and national environmental laws. Based on the calculated sulfur content, the HDS plant and its major equipment designs were simulated. Finally, an economic assessment was conducted by estimating the investment, operating and maintenance (O&M), and levelized cost of the desulfurization process. These estimations were then compared with the social cost savings resulting from the SO₂ and NO_x removal by replacing the high sulfur fuel oil (HSFO) with LSFO.

2. Determination of the required amount of fuel oil desulfurization in order to achieve a permissible limit of SO₂ emissions

The minimum amount of required fuel oil desulfurization was calculated from the limit of SO₂ emissions from the flue gas of power plants according to national and international air quality standards regulations. According to the first Iranian Air Quality Legislation Act of 2000, the national emission standard for SO₂ emitted from both existing and new power plants.

Table 1. Iranian, European and US air quality standards for power plants SO₂ emission [17, 18]

	European Union	United States	Iran
New Plants (ppm)	70	56	800
Existing Plants (ppm)	140	56 (built after 2005) 225 (built between 1997-2005)	800

Due to the considerable differences between the Iranian standard and the developed world standards, this study investigated the design and economic assessment of the HDS process in two different scenarios based on current Iranian and European SO₂ emission standards for existing power plants (respectively, a limit of 800 ppm and 140 ppm for SO₂ emission). The theoretical concentration of SO₂ in the flue gas of fuel oil combustion can be estimated from the chemical equation for stoichiometric combustion, composition of fuel oil, and amount of excess air [19, 20] In Table 2, the elemental analysis of fuel oil for the Shazand power plant is presented. The air excess factor m , both the amount of real oxygen needed (M_{O_2}), and the amount of flue gas produced from combustion ($M_{flueGas}$) were obtained from the following set of combustion reactions:



and can be calculated by equations (4) and (5) [20] at $N_2:O_2$ molar ratio of 3.762, mole of N_2 correspond to the stoichiometric O_2 is $3.762 \times M_{O_2}$, and mole of O_2+N_2 related to the excess air is $4.762 \times (m-1) \times M_{O_2}$. By implementing these terms, the total mole of dry flue gas is calculated from Equation (5). By assuming that the flue gas is close to ideal at normal conditions (1 atm, 273K); the concentration of theoretical SO_2 in flue gas becomes [20]:

$$M_{Dry\ Flue\ Gas} = \left(\frac{10X_C}{12} + \frac{10X_S}{32} + (4.762 \times (m-1) + 3.762) \times M_{O_2} \right) \left[\frac{mol}{kg\ fuel} \right] \quad (5)$$

$$C_{SO_2} = \frac{\frac{X_S}{10} \times \frac{64}{32} \times 10^6}{22.4 \times 10^{-3} \times M_{Dry\ Flue\ Gas}} \left[\frac{mg\ SO_2}{Nm^3} \right] \quad (6)$$

$$C_{SO_2} = \frac{X_S \times \frac{10}{32} \times 10^6}{M_{Dry\ Flue\ Gas}} \quad (7)$$

By implementing these equations at standard conditions (1 atm, 273K and 16.7% excess air or $m=1.167$), the amount of fuel oil desulfurization required in order to achieve the permissible limit of Iranian and European SO_2 emission standards were calculated and are presented in Table (3).

Table 2. Elemental analysis of fuel oil for the Shazand power plant

Elements	Identifier	Mass percent%	Mole/kg fuel
Carbon	X_C	85	$X_C * 10/12$
Hydrogen	X_{H_2}	9.5	$X_{H_2} * 10/2.02$
Sulfur	X_S	3.0	$X_S * 10/32$
Oxygen	X_{O_2}	1.2	$X_{O_2} * 10/32$
Nitrogen	X_{N_2}	0.0	$X_{N_2} * 10/28$
Water	X_{H_2O}	0.2	$X_{H_2O} * 10/18.02$
Ash	X_{Ash}	1.1	-

Table 3. Required amount of fuel desulfurization to achieve the permissible limit of standards

Fuel	% in fuel	PPM SO_2 in flue gas	Desulfurization %
HSFO ¹	3%	1857	0%
LSFO (Scenario I: European standard)	0.23	140	92%
LSFO (Scenario II: Iranian standard)	1.29	800	57%

¹: High sulfur fuel oil, current fuel oil which was delivered to the Shazand power plant

3. Design and simulation of the process

In order to design equipment and evaluate the economic feasibility of the HDS process, the operational conditions and specifications of the process streams were determined. To this end, process simulation software can be used to simulate equipment and streams of the HDS process [21-23]. This research used SIM-Suite Process simulation software to perform the simulation of HDS and Amine Gas Sweetening (AGS) units by introducing the input streams flow rate, temperature and pressure and hydrotreating

reactions into the software [24]. The following section gives an overview of this simulation.

3.1. Simulation input data

Both of the HDS plants designed in this work have a production capacity of 5.4 million liters of low sulfur fuel oil (LSFO) per day. This amount of fuel is equivalent to the fuel requirement of the Shazand power plant for the generation of 1300 MW of electricity. Table 4 contains the information used for the simulation of the HDS units, including the hydrotreating reactor feed properties and the characteristics of the output products.

Table 4. Information used for the simulation of HDS units

Hydrotreating reactor feed properties			
Specification	Value	Specification	Value
H ₂ /C ratio	8.95	Initial boiling point	268°C
S%	3%	True Boiling Point (TBP) 10%	387°C
Kinematic viscosity at 100°C	72 cSt	TBP _{30%}	492°C
Specific gravity at 15.5°C	0.997	TBP _{50%}	574°C
Ni content	40 ppm	TBP _{70%}	638°C
Vanadium content	15 ppm	TBP _{90%}	731°C
API gravity	10.4	TBP _{99%}	815°C
Characteristics of the output products			
Production capacity of LSFO		5.4 million Litter/Day	
S% in produced LSFO	Scenario I : Base on Eu standard		0.23%
	Scenario II : Base on Ir standard		1.29%

3.2. Hydrotreating Reactions

Hydrotreating reactions are usually carried out under a high pressure of hydrogen in two or three trickle bed reactors with each consisting of two or three beds of catalyst particles with liquid and vapor flowing co-currently among the beds. Hydro desulfurization and hydrodenitrogenation (HDN), key reactions of the hydrotreating process, were modeled in the reactor simulation [24]. The present understanding of hydrotreating suggests that both the reacting hydrocarbon molecule and hydrogen must be adsorbed at an active site on the catalyst surface for the reaction to occur. This accounts for the inclusion of the concentrations of both the reacting specie and the hydrogen in the kinetic equations used in the model [24, 25]. For the HDS reaction, the order of the kinetics of HDS with respect to the sulfur compounds is one; the observed order of H₂ is more on the order of 0.5 to 1.0 due to diffusion limitations of the hydrogen reaching the catalyst surface. It is also possible that the H₂S competes with the reacting species in being adsorbed on the catalyst and slows down the overall HDS reaction rate [24]. In the simulation, the following form of the Langmuir-Hinshelwood reaction rate equation has been used for the HDS reaction:

$$r_{\text{HDS}} = k_{\text{HDS}}^S \times \frac{C_{\text{CN}}^S \times (P_{\text{H}_2})^{0.5}}{1 + K_{\text{H}_2\text{S}} \times (P_{\text{H}_2\text{S}})} \quad (8)$$

Where r_{HDS} is the reaction rate of HDS; K_{HDS} is the reaction rate constant of HDS; P_{H_2} and $P_{\text{H}_2\text{S}}$ are the partial pressures of Hydrogen and Hydrogen Sulfide; C_{CN}^S is the molar concentration of sulfur compounds; and $K_{\text{H}_2\text{S}}$ is the adsorption constant of H₂S on catalyst, which is 0.38 in the temperature of 650 °F [24]. The form of the HDN kinetic equations which were used in the simulation is in the following form [24]:

$$r_{\text{HDN}} = k_{\text{HDN}}^N \times C_{\text{CN}}^N \times (P_{\text{H}_2})^{1.6} \quad (9)$$

Where K_{HDN} is the reaction rate constant of HDN and C_{CN}^N is the molar concentration of the nitrogen compounds.

3.3. Process description

The flow-diagram of the hydrotreating unit used for the HDS process simulation is demonstrated in Figure 1. This process simulation is composed of five sections: (1) hydrotreating reactors, (2) heat recovery section which recovers heat from the hot fuel oil stream leaving the reactors, (3) hydrogen separation, (4) fractionation section which separate hydrocarbon fractions produced by hydro cracking reactions, and (5) AGS unit for the separation of H₂S [26, 24, 27]. At the beginning of the process, after setting the pressure of fuel oil feedstock to 24 bars, the fuel oil was blended with hydrogen stream and heated by passing through a furnace to reach the reaction temperature before entering the reactor (up to 380 C for the EU standard and 340 C for the Iranian standard). The reaction section of this process was comprised of two reactors, each having two catalytic beds. The catalyst used in these beds included Cobalt/molybdenum and Nickel/molybdenum, both supported by silica-alumina. The first bed was designed to provide high hydrodemetallization (HDM) reactions in order to remove the heavy metals of nickel and vanadium [28]. The second bed of the first reactor and the first bed of second reactor were filled by a Co-Mo catalyst which had good activity for the HDS reaction; the second bed of the second reactor was filled by a Ni-Mo catalyst which had good activity for the HDN reaction. Regarding the fact that hydrotreating reactions are exothermic, the temperature rose across the reactors beds. To compensate for the rise in temperature and to control the reactor temperature, cold quench hydrogen was introduced between the beds of the

reactors so that the temperature remained in the range of 400-410 °C (European standard) and 340-350 °C (Iranian standard). In the heat recovery section, the temperature of the products mixture leaving the reactor was reduced to lower than 100 °C using a collection of heat exchangers. This was performed for the purpose of condensing H₂S vapor,

steam and light hydrocarbons, and ammonia from the gas phase as well as increasing the purity of un-reacted hydrogen. Heat exchangers used in this section included 150 psi and 50 psi steam generators, pre-heater of recycle hydrogen stream, pre-heater of fractionator feedstock, and an air cooler.

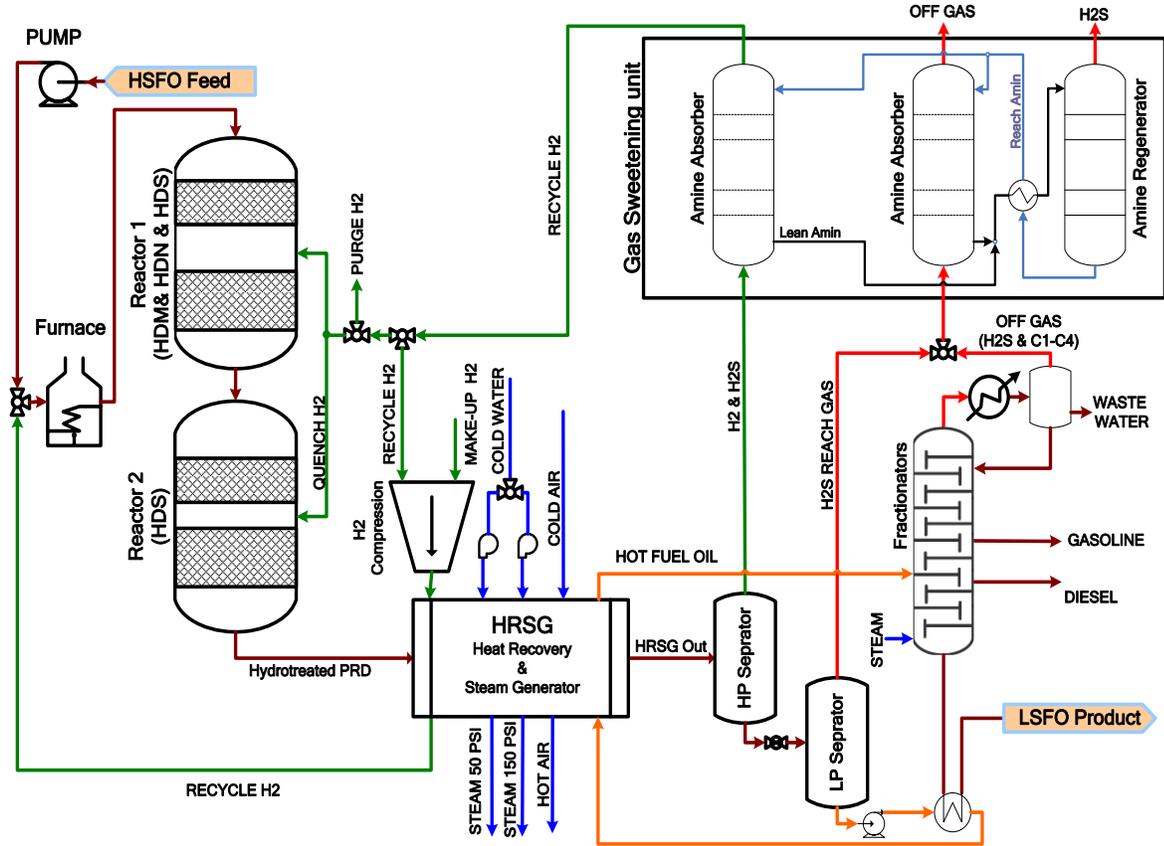


Fig. 1. Flow-diagram of the HDS unit

In the hydrogen separation section, the cold product was fed into the high pressure separator (HPS) in order to separate un-reacted H₂ and light gases from the liquid hydrocarbons at a pressure of 104 bar and a temperature of 70 °C. In the AGS unit, the separated H₂ was purified with the removal of associated H₂S gas. A small amount of H₂ stream was purged as off-gas to control the level of impurities in the recycled gas. Purified H₂ was mixed-up with makeup hydrogen and recycled to the process by a hydrogen compressor. The liquid outlet of the HPS passed to the low-pressure separator (LPS) where the H₂S and dissolved light hydrocarbons were separated from the liquid stream at the pressure of 4 bar and sent to the AGS section. The liquid outlet of LPS was passed through a series of heat exchangers to raise its temperature to 300 °C prior to entering the fractionation column to remove gasoline and diesel distillates. The low sulfur fuel oil from the bottom of the fractionating column was cooled by the inlet liquid of the fractionating column and sent to the power plant. The gas sweetening section was composed of two amine

scrubbers for the absorption of H₂S from H₂ and off-gas streams by a Diethanolamine (DEA) solution and one stripper for the regeneration of lean DEA [26].

3.4. Product Specifications

The presence of fuel oil, hydrogen, and catalyst at a high temperature and the pressure conditions led to the cracking of some part of the fuel oil into lighter hydrocarbon fractions through hydrocracking reactions. These fractions mainly included gasoline and diesel with boiling temperatures of 30-180 °C and 200-300 °C [29, 30]. In this research, by utilizing a fractionating column and injecting 150 psi steam at the bottom of this column, considerable amounts of these fractions were separated as the by-products of the process. Table 5 demonstrates the distribution of fractionating column fractions obtained from simulation for both the European and Iranian standards (separation efficiencies of gasoline and diesel fractions are considered 70%).

Table 5. Distribution of fractionating column fractions for European and Iranian standards

Fraction	Feedstock of fractionating column		Product	
	Volume percent		tonne/hr (m ³ /hr)	
	Iranian standard	European standard	Iranian standard	EU standard
Off Gas	0.4	1.0	0.9	2.5
Gasoline	0.5	1.5	0.8 (0.61)	2.6 (1.98)
Diesel	3.1	12.5	5.0 (4.12)	22.1 (18.10)
Residue	96	85	220	220

3.5. Equipments specifications

After having determined the operating conditions of the hydrotreating unit equipment, the major process equipment were designed using the methods and equations presented in the handbooks [31-33]. Some of the specifications obtained from these designs are presented in Table 6. Regarding the harsh operating condition of the

process and the presence of corrosive H₂S and hydrogen gases, choosing the appropriate material for the equipment and pipelines was of crucial importance. To this end, at critical locations, corrosive resistant material such as stainless steel 316L, A312, A358 and A213 was chosen for process equipment and pipes [34].

Table 6. Specifications of major hydrotreating equipment

Equipment	Specifications			
	Iranian standard		European standard	
Feedstock Pump	Q=250 m ³ /hr	Head: 213 m	Q=250 m ³ /hr	Head: 213 m
Reactor I	D:3.3 m	L: 16.8m	D:4.2 m	L: 21.15m
	Catalyst weight: (13+69)tonne		Catalyst weight: (27+146)tonne	
Reactor II	D:4.12 m	L: 19.6m	D: 5.3 m	L: 26.2m
	Catalyst weight: (78+112)tonne		Catalyst weight: (166+171)tonne	
H2 Compressor I	Δp= 60 bar	Q=326m ³ /hr	Δp= 60 bar	Q=1190 m ³ /hr
H2 Compressor II	Δp= 6 bar	Q=1213 m ³ /hr	Δp= 6 bar	Q=1317 m ³ /hr
HRSG	Exchanger fuel/fuel	Heat load: 2×11670 kW		Heat load: 2×20435 kW
	Exchanger fuel/steam 150	-		Heat load: 11200kW
	Exchanger fuel/H2	Heat load: 4594kW		Heat load: 3160kW
	Exchanger fuel/steam 50psi	Heat load: 6900kW		Heat load: 10959kW
	Air Cooler	Heat load: 24800kW		Heat load: 21472kW
	Pump 150 psi	-		Q=27.3 m ³ /hr
Pump 50 psi	Q=13.6 m ³ /hr	Head: 27.4m	Q=27.3 m ³ /hr	Head: 27.4m
HP separator	D:2.7 m	L:8.1m	D:2.9 m	L:8.7m
LP separator	D:2.7 m	L:8.1m	D:2.9 m	L:8.7m
Fractionating column	L:18 m	D:2.2 m	L:18	D:2.4 m
Gas sweetening absorber	L:5.35 m	D:1 m	L:6 m	D:1.38 m
Gas sweetening stripper	L:5.6 m	D:1.38 m	L:5.8 m	D:1.8 m

4. Economic Assessment

In this section, the economic feasibility assessment of the process was performed by estimating the overall annualized cost (OAC) as well as the final price of low sulfur fuel oil based on the two desulfurization standards of Iran and Europe. The OAC was obtained using the following equation [33].

OAC = Total Annual Capital Cost (TACC) + Total Annual Operating Cost (TAOC)

Total Annual Capital Cost (TACC) = Total Capital Cost (TCC) × ACCR

For an interest rate of *i* and a lifespan of *n*, the annual capital charge ratio (ACCR) can be calculated by Equation (10) [35]:

$$ACCR = \frac{i(1+i)^n}{(1+i)^n - 1} \quad (10)$$

The total capital cost (TCC) of the unit included the physical plant cost (PPC), design and engineering costs (PPC × 0.3), contractor fees (PPC × 0.05), and other contingency costs and were obtained using Equation (11) [31]:

$$TCC = 1.45 \times PPC \quad (11)$$

The physical plant costs (PPC) included the total capital investment of major equipment (TCI), piping (TCI×0.7), instrumentation, electrical equipment (TCI×0.1), building and structures (TCI×0.1), utilities (TCI×0.1), storages (TCI×0.15), and the site development costs (TCI×0.05) [31]. The results of the economic analysis are presented in Table 7. In this table, the approximate cost assessment of the major equipment including reactors, a fractionating column, a furnace, and compressors were determined on the basis of the construction cost of the reduced crude

desulfurization unit of the Shazand refinery in 2006; the cost of the remaining equipment were estimated through 2005 process economic evaluation handbooks [36, 32]. The Nelson-Farrar cost index was used for updating equipment costs to 2013 [37]. Consequently, the total annual capital cost (TACC) and total annual operating cost (TAOC) of the desulfurization process were calculated considering a capacity factor of $CF=0.9$ (330 days of operation per year). To estimate the operating costs, the following data were used: fuel cost- 33\$/barrel; water- 0.167\$/m³; electricity- 0.023\$/kWh; natural gas- 0.144\$/m³; and hydrogen

production and distribution costs-1\$/kg [38]. The net benefit of selling the by-products were calculated by multiplying the value of the price difference between gasoline and fuel oil (493 \$/tonne) and the price difference between diesel and fuel oil (351 \$/tonne) in mass flow of each of them [39]. In Table 7, the levelized cost of desulfurization in terms of cents/kW h was obtained using the following relation [6]:

$$\frac{\text{Net OAC}}{\text{MW}_e \times 1000 \times 8760 \times \text{CF}} \times 100 \left(\frac{\text{cents}}{\text{kWh}} \right) \quad (12)$$

Table 7. Economic analysis of hydrotreating process based on Europeans and Iranian standards

Component	Scenario I Europ. Std.	Scenario II Iranian Std.	Info.
	US\$		
Reactors	72,000,000	48,500,000	
Air Cooler	882,000	862,000	
H2 Compressors	3,416,000	3,172,000	
HRSX Exchangers	3,821,000	1,833,000	Hydrotreating Section
Fractionating column	1,393,000	1,149,000	
Pumps	63,000	30,000	
Separators	1,064,000	974,000	
Compressor	1,539,000	1,387,000	
Exchangers	1,707,000	1,197,000	
Pumps	116,000	86,000	Gas Sweetening Section
Stripper column	848,000	831,000	
Scrubbing column	296,000	285,000	
TCI *	87,145,000	60,306,000	
Piping	61,002,000	42,215,000	
Instrumentation	8,715,000	6,031,000	
Electrical	8,715,000	6,031,000	
Building, process	8,715,000	6,031,000	
Utilities	8,715,000	6,031,000	
Storages	13,072,000	9,046,000	
Site development	4,358,000	3,016,000	
PPC	200,437,000	138,707,000	
TCC	290,635,000	201,127,000	TCC= 1.45×PPC
TACC (US\$/year)	54,349,000	37,611,000	i=18%, n=20 year
VOC	US\$/Year		
Fuel	1,616,000	1,061,000	(I) : 147 BPSD & (II): 97 BPSD
Electricity	2,651,000	1,464,000	(I) 350,566 & (II) :193,641 [kWh/Day]
Cooling Water	1,330,000	734,000	(I) 4436 & (II) 2450 [GPM]
H ₂	38,103,000	10,422,000	(I)114 (II) 31 [tonne/day]
Catalyst	8,190,000	3,350,000	
FOC	US\$/Year		
Operating Labor	540,000	450,000	
Supervision	90,000	90,000	1 supervisor per shift
Maintenance	5,649,000	3,955,000	2% Of (TCC - Catalyst price)
Laboratories	100,000	100,000	
Taxes and insurance	2,906,000	2,011,000	1% TCC
TOAC	66,988,000	23,639,000	FOC+ VOC
OAC (Gross)	121,337,000	61,250,000	TOAC+ TACC
Gasoline Sales Revenue	7,715,000	2,372,000	
Diesel Sales Revenue	50,312,000	11,464,000	
Net OAC	63,309,000	47,413,000	Gross OAC - Revenues
Levelized cost (\$/barrel)	4.24	3.17	Net OAC /14944595
Levelized cost (c/kW-hr)	0.618	0.463	

*Budgetary estimate @ 2005 & 2006 and the Nelson-Farrar cost index was used to convert unit costs to 2013 dollars

5. HDS unit social costs

SO₂ and NO_x emissions cause various harmful effects on human health, agricultural crops, ecosystems and material corrosion, which in turn reduce the actual supply of labor, increase the user cost of capital and decrease agricultural productivity. The social costs of these emissions are an estimate cost of the monetized damages associated with their harmful effects [40]. Since the desulfurization of fuel oil reduces the social costs of both SO₂ and NO_x emitted

from power plants, it is important to calculate the reduction in social costs resulting from the implementation of the HDS unit. Based on the recent social costs of emissions from the energy sector (power plant) in Iran [41], the social cost of these pollutants were equal to 3.2 US\$/kg of SO₂ and 13 US\$/kg of NO_x. Considering the amount of pollutant emissions per kg of HSFO and changing the pollutant amount by replacing the LSFO, the reduced amount of social costs were calculated and are depicted in Table 8.

Table 8. Reduction of social cost resulting from the implementation of HDS unit

Fresh feed			
Sulfur, WT%		3	
Nitrogen, WT%		0.26	
Hydrotreater reactors & LSFO product Info.			
		European standard	Iranian standard
Desulfurization%		92%	57%
Denitrogenation%		72.40%	10.40%
Mass flow	tonne/hour	220	220
Sulfur removal	tonne /hour	6.072	3.762
Nitrogen removal	tonne /hour	0.414	0.059
SO ₂ removal	tonne /year	96,180	59,590
NO _x removal	tonne / year	7,028	1,010
Social cost savings of SO ₂ removal	\$/year	307,777,536	190,688,256
	\$/bbl	20.59	12.76
Social cost savings of NO _x removal	\$/year	91,368,469	13,124,752
	\$/bbl	6.11	0.88

Regarding Table 8, the total social cost savings of replacing HSFO with LSFO are calculated to be 399 million US\$ based on the European standard and 204 million US\$ based on the Iranian standard. Table 8 evinces that the social cost savings of replacing every barrel of LSFO instead of one barrel of HSFO is 26.7 US\$ based on the European standard and 13.6 US\$ based on the Iranian standard.

6. Conclusions

This study illustrated the engineering design simulation and the economic feasibility assessment of constructing large-scale industrial HDS and AGS sweetening units for fuel oil desulfurization which could be employed at the 1300 MW Shazand power plant in Iran. In sulfur removal from fuel oil, the degree of removal is an important variable affecting both economic and environmental burdens and benefits. In this study, the degree of fuel oil desulfurization was determined based on the SO₂ emission standards which comply with both the European (I) and the Iranian (II) standards. The obtained results showed that the fuel oil desulfurization percentages were 92% and 57% for scenarios I & II, respectively. Based on a simulation-based design of the process and its economic assessment, the gross costs of desulfurization of fuel oil in the power plant over a period of one year were estimated to be 121 million USD for scenario (I) and 61 million USD for scenario (II). If the revenue from selling by-products is considered, the net

costs of desulfurization of fuel oil over a one-year period are reduced to 63 million US\$ for scenario (I) and 47 million US\$ for scenario (II). Additionally, in the desulfurization standard of Europe, the share of the total annual operation cost (TAOC) is 55% of the overall annual cost (OAC) and the rest of it is related to the total annual capital cost (TACC). On the other hand, the share of the total annual operation cost according to the Iranian desulfurization standard (TAOC) is 39% of the overall annual cost (OAC). This is due to a greater degree of desulfurization and thereby, requires more amounts of hydrogen and catalyst. Furthermore, the harsh operating conditions along with the higher temperatures in the European desulfurization standard led to an increase in the production of by-products, so that the extra revenue obtained from selling such products compensate for the overall annual costs. In addition, the levelized costs of desulfurization in power production were estimated to be 0.618 c/kWh for scenario (I) and 0.463 c/kWh for scenario (II). From a social cost perspective and based on the European standard, the manufacturing of an HDS plant with a capacity of 225 tonne/hr will lead to a reduction of 399 million US\$/year in social cost expenditures. This reduction was estimated to be 204 million USD for Iran's standard. The results indicated that although the desulfurization cost per barrel of fuel oil based on the European standard was about 34% greater than the Iranian standard, the social costs

savings of this standard was about 96% which compensated for this extra cost.

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