Steam Consumption Minimization Using Genetic Algorithm Optimization Method: An industrial case study

Abdullah Alabdulkarem[a] and Nejat Rahmanian[b,*]

[a]Department of Mechanical Engineering, King Saud University, PO Box 800, Riyadh 11421, Saudi Arabia
[b]Department of Chemical Engineering, Faculty of Engineering and Informatics, University of Bradford, BD7 1DP, UK
*Corresponding Author Email: N.Rahmanian@bradford.ac.uk

Abstract

Condensate stabilization is a process where hydrocarbon condensate recovered from natural gas reservoirs is processed to meet the required storage, transportation and export specifications. The process involves stabilizing of hydrocarbon liquid by separation of light hydrocarbon such as methane from the heavier hydrocarbon constituents such as propane. An industrial scale back-up condensate stabilization unit was simulated using Aspen HYSYS software and validated with plant data. The separation process consumes significant amount of energy in form of steam. The objectives of the paper are to find the minimum steam consumption of the process and conduct sensitivity and exergy analyses on the process. The minimum steam consumption was found using genetic algorithm optimization method for both winter and summer conditions. The optimization was carried out using MATLAB software coupled with Aspen HYSYS software. The optimization involves six design variables and four constraints, such that realistic results are achieved. The results of the optimization show that savings in steam consumption is 34% as compared to the baseline process while maintaining the desired specifications. The effect of natural gas feed temperature has been investigated. The results show that steam consumption is reduced by 46% when the natural gas feed temperature changes from 17.7°C to 32.7°C. Exergy analysis shows that exergy destruction of the optimized process is 37% less than the baseline process.

Keywords: *GA, Optimization, Condensate Stabilization Unit, HYSYS, MATLAB*

Highlights

- Back-up condensate stabilization unit (BCSU) was optimized using genetic algorithm.
- Savings in optimized BCSU steam consumption is 34% as compared to the baseline BCSU.
- Effect of the gas feed temperature and heat exchanger have been investigated.

Abbreviations

BCSU: Back-up condensate stabilization unit

GA: Genetic algorithm

MEG: Monoethylene glycol

RVP: Reid Vapor Pressure

LPG: Liquefied petroleum gas

$\dot{Q}$: Heat rate

$\dot{m}_{Sat.Steam}$: Saturated steam mass flow rate

h_{fg}: Latent heat

HP: High pressure

LP: Low pressure

LHV: Low heating value

$\dot{X}_{destroyed}$: Exergy destruction

$\dot{s}_{gen}$: Entropy generation

$h_1(\vec{x})$: Equality constraint

$g_j(\vec{x})$: Inequality constraint

$\vec{x}$: Design variables

1. Introduction

Condensate stabilization is a key unit operation to separate low molecular weight hydrocarbons such as methane and ethane from the heavier hydrocarbon constituents. In general, in this process the amount of intermediates (C_3 to C_5) and heavy fractions (C_{6+}) in the condensate is increased. Recovery of hydrocarbon condensates from a natural gas occasionally is a finishing process for the liquid treatments and is simply a stabilization process for blending the condensate with crude oil to improve quality of crude oil before exporting. In case of raw condensate, there are no stringent requirements for the product beside process specifications such as Reid Vapor Pressure (RVP). The process is conducted to decrease the condensate vapor pressure so that vapor is not generated upon flashing the liquid into atmospheric storage tanks or injection into the export pipelines. The hydrocarbon condensate stabilization is also necessary to reduce the hydrocarbon losses from the storage tank (Benoy and Kale, 2010). Light products such as Liquefied Petroleum Gas (LPG) and gasoline are produced in oil refinery cracking processes, which utilize heavier constituents (Gary and Handwerk, 2001).
RVP test is used to measure the vapor pressure of condensate (ASTM D323-99a, 2012). Gasoline volatility is represented by the RVP which can also be predicted by using an algorithm (Esparragoza et al., 1992). Rahmanian et al. (2015) set RVP as a measure for off-specification conditions of the product where 68.94 kPa was set to be the maximum summer and 82.73 kPa for winter conditions. Under real plant operating conditions, condensate produced over the RVP

range is delivered to an off-spec tank for temporary storage and additional treatment. The off-spec tank should have normally the ability to store 24 h of off-specification products.

Several papers in the literature utilize process simulation software to optimize systems performance by optimizing operating parameters. For example, Alabdulkarem et al. (2011) optimized a natural gas liquefaction facility using HYSYS software coupled with Matlab software, and they were able to achieve power savings of 10%. Sunny et al. (2019) optimized the reforming process for syngas production in regasified liquefied natural gas based ammonia plant using HYSYS software. Jalali et al. (2020) optimized a reformate stabilizer tower of a petrochemical plant using HYSYS software where they were able to reduce energy consumption by 12.6%. Ramadan et al. (2019) optimized heat exchanger for heat recovery applications for three optimization criteria, which were maximum water temperature, maximum recovered energy, and minimum gas temperature threshold. Shafiee et al. (2018) optimized distillation tower of a pyrolysis gasoline hydrogenation unit and compared the optimized unit vs the baseline unit using pinch analysis.

Optimization methods can be gradient-based or heuristic. Genetic Algorithm (GA) is classified as heuristic and global optimization technique which simulates natural evolution through altering individual solutions population where chromosomes represent design variables (x). GA arbitrarily chooses individual designs from current population to be parents and uses them to produce children for the next generation. The population reaches an optimal solution because "good parents" produce "good children" while "bad points" are eliminated over successive generations. Several optimization problems in different areas that are not suited for deterministic optimization methods could be solved using GA, including objective functions that are non-differentiable, discontinuous or non-linear (Deb, 2001). If GA did not stick at a local optimal solution, it has the ability of reaching the global optimal solution. Hence, it can't be ensured that the resulted solution is the global optimum when using GA method which is considered as a disadvantage of GA. Various designs could be produced by various runs since GA is a probabilistic method. To improve the confidence that optimal designs have been achieved, several runs are required. Global optimum designs cannot be guaranteed unless the entire solution is evaluated using exhaustive search, which is computationally expensive.

Common software used in the hydrocarbon process simulation are Aspen Plus, Aspen HYSYS and PRO/II. For instance, Bahmani et al. (2016) simulated an industrial liquefied petroleum gas and natural gas liquid production unit using ASPEN Plus software. Khazini et al. (2014) simulated a sulfur recovery unit of a petroleum refinery using HYSYS software. Alabdulkarem et al. (2012) simulated CO_2 capturing and sequestration integrated with natural gas liquefaction plant using HYSYS software. Al-Sobhi et al. (2015) simulated and optimized natural gas processing and production network using ASPEN Plus software. Bonyadi et al. (2014) proposed a new method for determination of condensate stabilization fluid compositions. Aspen HYSYS was used to investigate debottlenecking of a condensate stabilization unit using Pinch Analysis in the gas plant in Assaluyeh (Iran) by Tahouni et al. (2014). Rahmanian et al. (2016) simulated condensate stabilization unit using ASPEN HYSYS software where they investigated the effect of several process parameters on the performance of the condensate stabilization unit. Kazerooni et al. (2016) used artificial neural network to model an industrial scale condensate stabilization plant.

An industrial scale Back-up Condensate Stabilization Unit (BCSU) was simulated using Aspen HYSYS software by Rahmanian et al. (2015). The simulation of BCSU results were validated with the plant data and a good agreement was observed. In addition, their HYSYS model results were confirmed with results obtained using PRO/II® software. According to their simulation results, the BCSU consumes 11.93 MW of energy. The design of BCSU, which consume so significant energy, is complicated and involves many variables. Adjusting those variables with conventional approaches would affect the product quality which makes optimization methods of interest in improving the design of such complex process while maintaining the desired products quality.

As can be seen from the literature, there is no available study which used a novel approach where genetic algorithm method was used to optimize back-up condensate stabilization unit using a validated model. In addition, no study optimized the BCSU when the heat exchanger performance varies or when the feed gas temperature varies. Therefore, the objectives of this work are (1) to optimize the BCSU using genetic algorithm, (2) to conducted parametric studies of the critical parameters on the process performance, (3) to carry out exergy analysis on the BCSU.

2. Condensate Stabilization Process Description

The flow chart of the whole gas plant consisting of a CSU and a BCSU located at Assaluyeh port, Assaluyeh, Bushehr province, in the southern region of Iran is shown in Figure 1. This paper focuses on the BCSU of this plant, which is modeled using HYSYS software as shown in Figure 2 and will be discussed in the Model Development and Optimization Approach section. As shown in Figure 1, feed reservoir composing of gas, water and condensate are produced and pre-treated at offshore platforms where free water is removed and the remaining is delivered to the on-shore gas facility. Gas mixture can form gas hydrates in the presence of water, which hinders the smoothness of gas flow. Therefore, Monoethylene Glycol (MEG) is injected to the exit stream from the offshore platform so that it helps to avoid gas hydrates formation. At the onshore arrival facilities, the reservoir fluids are splitted into gas and liquid streams using a slug catcher, a multi finger sloped pipes. This is used to separate large amount of liquids (condensate, water and MEG) that have accumulated in subsea pipelines especially during pigging operations. The exit gas stream from the slug catcher is collected in a major pipeline manifold for distribution to the six parallel gas processing trains. The liquid stream containing MEG, water and some heavy hydrocarbons exit at the outlet of slug catcher by using a large manifold and further separated to a condensate stream and aqueous MEG. The heavier liquid phase containing MEG and water is distributed to the six parallel MEG regeneration units where MEG is separated by evaporation of water and recycled back to the offshore plant using three 4’’ MEG piggyback lines (see Fig. 1). The hydrocarbon condensate contain partial water and MEG is fed to CSU or BCSU. A BCSU is designed to treat the condensate when CSU process is in shutdown. The stabilized condensate is then delivered to storage tanks set for exporting purposes.

Figure 1: Process diagram of the gas plant including Condensate Stabilization Unit and Back-up Condensate Stabilization Unit.

3. BCSU Process Description

The purpose of BCSU is to support the steady-state operation of the gas plant when the normal CSU is under maintenance. This is possible in the case that CSU is in shut down mode due to mal-operation of flash gas compressors in CSU. The process involves multi-stage cooling and heating of the hydrocarbon condensate through flash vaporization.

BCSU utilizes the equilibrium between liquid and vapor phases, which occurs at the separation pressure, and temperature. As shown in the process flow diagram (see Fig. 2), firstly, the feed, specifications shown in Table 1, from the slug catcher is fed to the Pre-flash Drum for removal of light hydrocarbons where the light acid gases are also separated in this drum. The temperature of the condensate is increased in two series of independent heat exchangers by high pressure steam to 80°C and 143°C, respectively. Finally, the main stream fluid crosses the Pre-flash Exchanger and degasses in the Degassing Drum before it is pumped to the floating roof storage tanks. The feed condensate coming from the upstream at the exit of Slug Catcher is at high pressure of 78 bara and then it goes to let down valve station which its pressure drop to 13 bara. Then, MEG/Water and condensate splitted into two main streams (see Fig.1) where condensate is further processed in BCSU or CSU. The pressure and temperature of the condensate arrives the battery limit of BCSU is 1252 kPa and 17.7 °C, respectively. The Pre-flash drum and Flash drum operates at pressures of 1200 kPa and 401 kPa, respectively. The BCSU process is akin to multi-stage flash with only difference that there are heat exchangers between each stage of the process here. This helps flashing of large quantities of lighter ends that join the sour vapor stream after recompression. The flashed vapor can either be released into HP or LP flare or more recently it is planned to divert these to the main gas stream in the plant. Efficient condensate separation is achieved through degassing the condensate in the Degassing Drum at the lowest acceptable pressure of 150 kPa before it goes to the storage tanks at 40°C. The function of degassing drum is to minimize excess condensate flashing in the storage tank as well as minimization of the required gas blanket pressure.

Table 1: Specifications and compositions of feed stream.

Phase Fraction	0.25372
Temperature (°C)	17.7
Pressure (kPa, Absolute)	1300
Molar Flow Rate (kgmole/h)	4645
Mass Flow Rate (kg/s)	84.179
Std Ideal Liq. Vol. Flow Rate (m^3/h)	462.763
Mass Fraction (X 100)	
Methane	5.362

Ethane	2.507
Propane	3.501
i-Butane	1.683
n-Butane	3.466
i-Pentane	2.542
n-Pentane	2.858
Mcyclopentan	0.424
Benzene	0.268
n-Hexane	5.016
Cyclohexane	0.594
Mcyclohexane	1.862
Toluene	0.537
n-Heptane	7.178
n-Octane	9.477
p-Xylene	3.281
n-Nonane	9.097
Cumene	1.004
n-Decane	8.118
n-C11	21.031
Nitrogen	0.113
CO_2	0.811
H_2S	0.531
H_2O	3.569
M-Mercaptan	0.010
E-Mercaptan	0.161
COS	0.001
NPMercaptan	0.173
NBMercaptan	0.070
1Pentanthiol	0.174
EGlycol	4.581

The major components of flash gas include low molecular weights hydrocarbons such as C1 to C3 hydrocarbons and hydrogen sulphide. The liquid stream include aqueous phase of MEG which is fed to MEG regeneration units for the further processing. The other exit stream from Degassing Drum which include mainly water and partially some mercaptans are fed to the off-specification product storage tank. The off-spec storage tank will be recycled back to CSU or BCSU and sometimes are transferred to the waste treatment unit depending on the routine sample analysis of these components in the off-spec tank.

Figure 2: Process flow sheet of BCSU using Aspen HYSYS simulation (Rahmanian et al., 2015).

4. Model Development and Optimization Approach

To analyze the performance of BCSU, process simulation was conducted using Aspen HYSYS (v8.8). It is important to run a model-based approach because some process data are not accessible or not measured in the plant under normal operation, but can be produced only from the model. Those unavailable process data are steam temperature and flow rate, and process temperature and pressure. The simulation results were compared to the actual plant data obtained previously by Rahmanian et al. (2015).
The simulation is executed based on a reference BCSU in operation (Behbehani and Atashrouz, 2011) with convergence set to $1x10^{-4}$. The equation of state of Peng-Robinson (PR) (1976) was selected for prediction of thermodynamic properties of the fluid components. Fig. 2 illustrates the flowsheet of the BCSU created in the HYSYS software. The product quality are dependent on the composition and RVP before exporting.

The HP heater is a multi-stream heat exchanger that has the condensate on one side and saturated steam on the other side. The heat supplied, $\dot{\mathrm{Q}}_{HP}$, through the heat exchanger is heat content in the saturated steam which is calculated by multiplying the saturated steam mass flow rate, $\dot{\mathrm{m}}_{Sat.Steam}$, by the latent heat of the steam, h_{fg} which is function of the steam temperature.

$$\dot{\mathrm{Q}}_{HP} = \dot{\mathrm{m}}_{Sat.Steam}\,\mathrm{h}_{fg}$$

There are many variables such as pressure, temperature, flow rate, etc involved in design and operation of BCSU, therefore the optimization problem is computationally intensive. Therefore, the most important design parameters that affect the performance of the BCSU were selected. Matlab (version 2014) optimization tool has variety of optimization techniques such as gradient and probabilistic techniques. GA was selected since it can handle discrete as well as continuous problems. Furthermore, global optimum could be achieved with GA. The design of a BCSU is a non-linear problem with local optima. HYSYS Component Object Model (COM) is accessed using Matlab code. The Matlab “actxserver’’ command is used to create the HYSYS COM server. HYSYS simulation variables are read and altered by MATLAB depending on GA population. Matlab-HYSYS integration approach is shown in Figure 3. Each cycle took about 48 hours to solve on an Intel Core 7 processor (3.4 GHz) with 16 GB of RAM. Table 2 shows GA tuning parameters implemented in this work.

Table 2: Typical GA tuning parameters.

Tuning parameters	Value
Population size	120
Reproduction count	50% of the population size
Maximum number of generations	100
Crossover function	Two points
Crossover fraction	0.8
Selection method	Tournament
Tournament size	8
Fitness scaling method	Top
Number of crossover points	1
Mutation method	Adaptive feasible
Stopping criteria	Fitness Limit

The BCSU optimization problem formulation is given as:

Minimize fn_x: $f(\vec{x}) = f(x_1, x_2, \ldots, x_6)$

Subjected to

$$\vec{x}_L \leq \vec{x} \leq \vec{x}_U$$
$$h_1(\vec{x}) = 0$$
$$g_j(\vec{x}) \leq 0, j = 1, \ldots, 3$$

The optimization objective function, $f(\vec{x})$, is to minimize energy consumption which is the steam flow rate. The HYSYS BCSU model is treated as a black box which means the objective function value is provided by HYSYS which is the HP Steam Heater shown in the model. The optimization variables, $\vec{x}$, are listed in Table 3 with limits taken to be ±20% of the baseline values. The equality constraint, $h_1(\vec{x})$, is fixed feed temperature of 17.7°C in summer and 12.7°C in winter. The inequality constraints, $g_j(\vec{x})$, are as follows:

$T_{\text{Pinch Pre-Flash HX}} \geq 3$°C
$T_{\text{Pinch HX}} \geq 3$°C
RVP ≤ 10 psia (68.9 kPa) in summer and 12 psia (82.7 kPa) in winter

Figure 3: MATLAB-HYSYS integration approach.

Table 3: List of the BCSU unit optimization variables.

Variable	Baseline Value	Lower Bounds	Upper Bounds
Pressure after valve 1 (kPa, absolute), x1	1200	1150	1250
Pressure after valve 2 (kPa, absolute), x2	450	365	535

Pressure after valve 3 (kPa, absolute), x3	200	104	296
Temperature after pre-flash Exchanger (°C), x4	39	35	53
Temperature after heat exchanger (°C), x5	80	72	88
Temperature after HP steam heater (°C), x6	143	129	158

5 Results and Discussions

5.1 Baseline Optimization

The optimization was carried out for the both summer and winter conditions. For the summer case, the optimized BCSU unit's steam consumption is 7.87 MW or 3.68 kg/s of saturated steam, which is 34% less than the baseline unit's steam consumption of 11.93 MW or 5.58 kg/s of saturated steam. The optimized unit has lower pressure drop in the valves than the baseline unit and higher temperature than the baseline unit after the heat exchangers as shown in Table 4. In addition, the temperature after the HP steam heater of the optimized BCSU is 10.6°C lower than the baseline unit which means the lower steam quality can be used. Lower steam temperature can be provided using the waste heat from the steam generation unit in the plant.

As for the winter case, the optimized BCSU unit's steam consumption is 7.51 MW or 3.51 kg/s of the saturated steam, which is 37% less than the baseline unit's steam consumption. The optimized unit has the lower pressure drop than the baseline unit in valve 1 only and a higher temperature than the baseline unit after the heat exchangers. In addition, the temperature after the HP steam heater of the optimized BCSU (129.5°C) is 13.5°C lower than the baseline unit (143°C). The product compositions for the baseline unit and optimized units are shown in Table 5. As can be seen in Table 5, the optimization has not affected the composition of the product stream significantly.

Table 4: The results of optimization of BCSU unit.

Case	Baseline (Feed at 17.7°C)	Optimized (Summer at 17.7°C)	Optimized (Winter, Feed at 12.7°C)
Objective fn: Min. steam consumption (MW)	11.93	7.87	7.51
Steam mass flow rate (kg/s) at 150°C	5.58	3.68	3.51
Pressure after valve 1 (kPa, Absolute)	1200	1170	1189
Pressure after valve 2 (kPa, Absolute)	450	442.3	513.5
Pressure after valve 3 (kPa, Absolute)	200	180.8	274.3
Temperature after pre-flash exchanger (°C)	39	49.6	52.6

Temperature after heat exchanger (°C)	80	89.3	87.7
Temperature after HP steam heater (°C)	143	132.4	129.5
RVP at (37.8 C) (Psia)	8.92	9.94	11.87
RVP at (37.8 C) (kPa, Absolute)	61.5	68.53	81.84
Mass flow rate in the "To Storage Tank" (kg/s)	55.17	57.77	59.43

Table 5: Mass fraction (multiplied by 100) of product into the "Storage Tank" stream.

Case	Baseline (Feed at 17.7°C)	Optimized (Summer at 17.7°C)	Optimized (Winter, Feed at 12.7°C)
Methane	0.027	0.031	0.041
Ethane	0.117	0.129	0.166
Propane	0.587	0.669	0.820
i-Butane	0.570	0.660	0.776
n-Butane	1.426	1.653	1.913
i-Pentane	1.648	1.875	2.070
n-Pentane	2.031	2.298	2.509
Mcyclopentan	0.430	0.461	0.479
Benzene	0.273	0.293	0.304
n-Hexane	4.986	5.382	5.605
Cyclohexane	0.634	0.673	0.693
Mcyclohexane	2.262	2.328	2.349
Toluene	0.664	0.681	0.685
n-Heptane	8.655	8.942	9.033
n-Octane	12.740	12.753	12.654
p-Xylene	4.512	4.482	4.432
n-Nonane	12.972	12.728	12.508
Cumene	1.428	1.402	1.378
n-Decane	11.943	11.582	11.324
n-C11	31.472	30.315	29.559
Nitrogen	0.000	0.000	0.000
CO_2	0.014	0.016	0.021
H_2S	0.035	0.040	0.050
H_2O	0.001	0.002	0.003
M-Mercaptan	0.003	0.003	0.004

E-Mercaptan	0.086	0.099	0.111
COS	0.000	0.000	0.000
nPMercaptan	0.170	0.184	0.192
nBMercaptan	0.082	0.085	0.087
1Pentanthiol	0.233	0.234	0.232
EGlycol	0.000	0.000	0.000

5.2 Effect of Feed Temperature on Optimized Designs

The natural gas feed temperature could vary with seasons, different reservoirs or length and conditions of delivery pipelines. To investigate the effect of natural gas feed temperatures, the optimization was carried out with the feed temperature starting from 12.7°C to 32.7°C. The results are tabulated in Table 6. RVP constraint was taken to be the summer value in all feed temperatures above 12.7°C. In terms of the design variables, pressure changes slightly but more significant changes are observed on the temperature. Steam consumption is reduced from 7.87 MW to 4.25 MW (or by 46%) when the natural gas feed temperature changes from 17.7°C to 32.7°C (or by 85%). Figure 4 shows steam consumption variation with various gas feed temperatures.

Figure 4: Steam consumption at different natural gas feed temperatures.

Table 6: The results of optimization of BCSU unit with different feed temperatures.

Feed temperature (°C)	12.7°C	17.7°C	22.7°C	27.7°C	32.7°C
Objective fn: Min. steam consumption (MW)	7.51	7.87	5.77	4.76	4.25
Steam mass flow rate (kg/s) at 150°C	3.51	3.68	2.7	2.23	1.99
Pressure after valve 1 (kPa, Absolute)	1189	1170	1144	1144	1144
Pressure after valve 2 (kPa, Absolute)	513.5	442.3	367.1	371	378.8
Pressure after valve 3 (kPa, Absolute)	274.3	180.8	81.5	85.2	92.9
Temperature after pre-flash exchanger (°C)	52.6	49.6	48.71	42.2	42.8
Temperature after heat exchanger (°C)	87.7	89.3	76.8	79.4	81.6
Temperature after HP steam heater (°C)	129.5	132.4	110.3	106.7	105.9
RVP at (37.8 C) (Psia)	11.87	9.94	9.88	9.97	9.97

RVP at (37.8 C) (kPa, Absolute)	81.84	68.53	68.12	68.74	68.74
Mass flow rate in the "To Storage Tank" (kg/s)	59.43	57.77	60.15	60.6	60.77

5.3 Required Steam Temperature

Saturated steam is supplied to the HP steam heater. Steam generation is an energy intensive process where the temperature of the steam represents how valuable the steam is. From Carnot cycle efficiency, steam value decreases with its temperature. Therefore, it is desirable to use steam that is at low temperature. In addition, low steam temperature can be provided using the waste heat from the power generation condenser, exhaust gas or other processes. The minimum required steam temperature is a function of heat exchanger pinch temperature. If the heat exchanger pinch temperature is assumed to be 7°C, then the required minimum steam temperature would be 7°C above the Stream 5 temperature. The required steam flow rates of the optimized process at the minimum required temperature are shown in Table 7. The required steam flow rates can be found by dividing the HP steam consumption over the latent heat of the steam which is the difference between saturated vapor and saturated liquid enthalpy at the required temperature. The optimized BCSU unit's steam consumption, at 17.7°C feed temperature, is 3.63 kg/s of the saturated steam, which is 35% less than the baseline unit's steam consumption of 5.58 kg/s of the saturated steam.

Table 7: Steam consumption at minimum steam temperature.

Case	Baseline (Feed at 17.7°C)	Optimized (Summer at 17.7°C)	Optimized (Winter, Feed at 12.7°C)	Optimized (Summer at 22.7°C)	Optimized (Summer at 27.7°C)	Optimized (Summer at 32.7°C)
Minimum required steam temperature (°C)	150	139	137	117	114	113
Steam mass flow rate (kg/s)	5.58	3.63	3.45	2.59	2.13	1.89

5.4 Effect of heat exchanger performance

To investigate the effect of heat exchanger performance on the steam consumption, the optimization was conducted at the various pinch temperatures (i.e., by adjusting the second optimization constraint). The lower pinch temperature heat exchangers will be more efficient, thus more expensive, because they will have the higher UA values to compensate for the reduction in LMTD values. Table 8 shows the effect of heat exchanger pinch temperature on the steam consumption at feed temperature of 17.7°C. Increasing the pinch temperature from 3°C to 7°C resulted in reduction in the steam consumption by 16%. The effect was mainly reflected on the required steam temperature where the temperature after HP steam heater was reduced from

132°C to 109°C. The products mass fractions for 3°C and 7°C pinch temperatures are listed in Table 9. It can be seen that the methane content was reduced in the 7°C case as compared to the 3°C case.

Table 8: Effect of heat exchanger pinch temperature.

Feed temperature (°C)	17.7°C		
Minimum allowed heat exchanger pinch temperature	3	5	7
Heat exchanger UA (kJ/ C-h)	465	382	326
Objective fn: Min. Steam consumption (MW)	7.87	6.71	6.59
Pressure after valve 1 (kPa, Absolute)	1170	1144	1134
Pressure after valve 2 (kPa, Absolute)	442.3	363.9	375
Pressure after valve 3 (kPa, Absolute)	180.8	78.1	79.2
Temperature after pre-flash exchanger (°C)	49.6	49.8	48.6
Temperature after heat exchanger (°C)	89.3	73.7	70.9
Temperature after HP steam heater (°C)	132.4	112.8	109.5
RVP at (37.8 C) (Psia)	9.94	9.73	9.95
Mass flow rate in the "To Storage Tank" (kg/s)	57.77	59.76	60.36

Table 9: Mass fraction (multiplied by 100) of product into the "Storage Tank" stream.

Heat exchanger pinch temperature	3°C	7°C	Difference
Methane	0.031	0.012	-62%
Ethane	0.129	0.117	-9%
Propane	0.669	0.762	14%
i-Butane	0.660	0.781	18%
n-Butane	1.653	1.959	18%
i-Pentane	1.875	2.150	15%
n-Pentane	2.298	2.613	14%
Mcyclopentan	0.461	0.492	7%
Benzene	0.293	0.312	7%

n-Hexane	5.382	5.774	7%
Cyclohexane	0.673	0.709	5%
Mcyclohexane	2.328	2.379	2%
Toluene	0.681	0.693	2%
n-Heptane	8.942	9.170	3%
n-Octane	12.753	12.697	0%
p-Xylene	4.482	4.432	-1%
n-Nonane	12.728	12.453	-2%
Cumene	1.402	1.372	-2%
n-Decane	11.582	11.222	-3%
n-C11	30.315	29.209	-4%
Nitrogen	0.000	0.000	-86%
CO_2	0.016	0.011	-29%
H_2S	0.040	0.039	-1%
H_2O	0.002	0.001	-29%
M-Mercaptan	0.003	0.004	18%
E-Mercaptan	0.099	0.115	17%
COS	0.000	0.000	12%
nPMercaptan	0.184	0.198	8%
nBMercaptan	0.085	0.088	3%
1Pentanthiol	0.234	0.233	0%
EGlycol	0.000	0.000	6%

6. Exergy Analysis

Exergy analysis is used to compare the baseline BCSU against the optimized BCSU at 17.7°C feed temperature. Exergy is the maximum work potential (Cengel and Boles, 2014). Exergy destruction, on the other hand, represents the lost work potential which can be used to locate where the losses occur in the process components. Tahouni et al. (2014) showed that 2088 kW out of 10400 kW could be saved after improving the heat exchangers. In this paper, exergy destruction was calculated to see which part of the BCSU components contribute to the highest losses in the process. Exergy destruction ($\dot{X}_{destroyed}$) is associated with entropy generation ($\dot{s}_{gen}$) which occurs due to entropy transfer with mass (s) and heat ($\dot{Q}_{in}$). The following equations were used to calculate exergy destruction in each component of the BCSU:

$$\dot{X}_{destroyed} = T_o\, \dot{s}_{gen}$$

$$\dot{s}_{gen} = \sum_{out} \dot{m}s - \sum_{in} \dot{m}s - \frac{\dot{Q}_{in}}{T_{source}}$$

Where, T_o is the ambient temperature and T_{source} is the steam temperature.

The exergy destruction values are shown in Table 10 and Figure 5. The highest exergy destruction is in the heat exchangers and HP steam heater which means they can be improved. Total exergy destruction of the optimized BCSU is 37% less than the baseline BCSU which confirms that the optimized BCSU does not only consumes less steam, but also more efficient. It is to be noted that other parameters could affect the performance of the BCSU, baseline and optimized, with time such as heat exchangers fouling.

Figure 5: Exergy Destruction in (a) Baseline BCSU and (b) Optimize BCSU in (kW).

Table 10: Exergy destruction in BCSU Components.

Component	Baseline BCSU Exergy Destruction (kW)	Optimized BCSU Exergy Destruction (kW)	Difference (%)
Valve 1	71.9	94.9	32
Valve 2	402.4	304.6	-24
Valve 3	8.3	10.2	23
Pre-flash Exchanger	461.6	303.3	-34
Heat Exchanger	698.4	426.8	-39
HP Steam Heater	803.4	402	-50
Total	2446	1542	-37

Since flare gases carry exergy, and to find the exergy transfer with flare gases, the same equation was applied but with extending the system boundary so that it includes the entire BCSU. The exergy entering the system is the exergy of the feed stream and the exergy transfer with the heat of the HP steam heater. The exergy leaving the system is the exergy of the feed LP flare, HP flare, To MEG Regeneration and To Storage Tank. Other streams have zero mass flow rate under steady state conditions. The total exergy destroyed is then the difference between exergy coming and exergy leaving the system, which is identical to what was obtained when adding all exergy destroyed in different components. The difference between the baseline and optimized BCSU is 37% which also confirms the superiority of the optimized BCSU. The results are tabulated in Table 11 and plotted in Figure 6. The results show that high exergy is leaving the system through the flare, and thus, should be utilized.

$$\dot{X}_{destroyed} = \sum_{out} \dot{X} - \sum_{in} \dot{X}$$

In addition, the heat content in the fuel and flare gas were calculated to compare the baseline BCSU against the optimized BCSU based on the low heating value (LHV). From an energy point of view, flare gas contains energy and it should be utilized. LHV represents the energy content when the fuel is burned. The heat rate in the fuel when it’s burned is calculated as follows:

$$\dot{Q}_{Fuel} = \dot{m}_{Fuel}\, LHV$$

The results are tabulated in Table 12. The optimized system LP Flare stream contains less heat (27%) than the baseline while it contains more heat (11%) in the HP Flare steam. The heat rate in the storage tank of optimized system is 4% higher than the baseline which also confirms the superiority of the optimized system.

Figure 6: Exergy transfer in different streams of BCSU.

Table 11: Exergy transfer in different streams of BCSU.

Stream	Baseline System (kW)	Optimized System (kW)	Difference
Feed	41994	41994	0%
LP Flare	12658	9470	-25%
HP Flare	18585	19962	7%
To Storage Tank	21080	19165	-9%
To MEG Regeneration	296	480	62%
Heat Stream	8179	5541	-32%
Total	2446	1542	-37%

Table 12: Heat rate in different streams.

System	Baseline	Optimized	Difference (%)
LP Flare Heat Rate (MW)	584	425	-27
HP Flare Heat Rate (MW)	392	436	11
Storage Tank Heat Rate (MW)	2451	2560	4

7. Conclusions

An industrial BCSU was simulated using Aspen HYSYS software. Rahmanian et al. (2015) already validated the simulation against the plant data. The simulation results show that 11.93 MW of heat is consumed in the BCSU. This work was focused mainly on optimization of BCSU as an industrial case study. Optimizing BCSU design is complex and involves six variables, with four constraints and non-linear objective function. Therefore, genetic algorithm optimization approach was employed to find the optimum design and operating conditions.

The results of the optimization show that the optimized BCSU unit's steam consumption in the summer conditions is 7.87 MW, which is 34% less than the baseline unit's steam consumption. As for the winter conditions, the optimized BCSU unit's steam consumption is 7.51 MW, which is 37% less than the baseline unit's steam consumption. The product compositions have not been affected by minimizing the steam consumption.

The minimum required steam temperature has been found so that the waste heat can be utilized in the BCSU. The effect of heat exchangers performance has been studied. In addition, the effect of gas feed temperature has been investigated. The results show that steam consumption is reduced by 46% when the natural gas feed temperature changes from 17.7°C to 32.7°C. Exergy analysis shows that exergy destruction of the optimized BCSU is 37% less than the baseline BCSU and most exergy destruction takes place in the heat exchangers.

The results show that the followed approach where system simulation software is coupled with optimization software can be used to effectively optimize complex cycles such as any gas processing cycles or oil refineries. However, global optimum design cannot be ensured due to the probabilistic nature of GA. Investigation of critical variables on the optimum design, such as environmental conditions or heat exchanger performance, is essential to evaluate the impact on the robustness of the optimum design for the selected range of variations.

A possible future work that was not addressed here is when the feed gas compositions varies, such that the BCSU processes gases from different gas fields. In order to solve this type of optimization problems, robust optimization methods must be used because they are able to handle optimization problems with uncertain variables.

References

- Alabdulkarem A., Mortazavi A., Hwang Y., Radermacher R., and Rodgers P., 2011, “Optimization of Propane Pre-cooled Mixed Refrigerant LNG Plant” Applied Thermal Engineering Journal, 31 (6), pp. 1091-1098.
- Alabdulkarem A., Hwang Y., Radermacher R., 2012, Energy Consumption Reduction in CO_2 Capturing and Sequestration of an LNG Plant through Process Integration and Waste Heat Utilization, International Journal of Greenhouse Gas Control Journal, 10, pp. 215-228.
- Al-Sobhi, S.A., Elkamel, A. 2015, Simulation and optimization of natural gas processing and production network consisting of LNG, GTL, and methanol facilities, Journal of Natural Gas Science and Engineering, 23, pp. 500-508.
- Aspen Technology, Inc. Aspen and HYSYS Software. www.aspentech.com
- ASTM D323-99a, 2012. Historical Standard: ASTM D323-99a Standard Test Method for Vapour Pressure of Petroleum Products (Reid Method). www.astm.org.
- Bahmani, J., Shariati, A., Rouzbahani, A. 2016 “Simulation and optimization of an industrial gas condensate stabilization unit to modify LPG and NGL production with minimizing CO_2 emission to the environment” Accepted in Chinese Journal of Chemical Engineering, Available online 15 July 2016
- Behbehani, R.M., Atashrouz, E., 2011. Fundamentals of Natural Gas Processing and Transmission. Publisher, Ayej, Tehran, Iran.

- Benoy, J., Kale, R.N., 2010, Condensate Stabilization, in Offshore World. Chemtech Foundation, India.
- Bonyadi, M., Esmaeilzadeh, F., Mowla, D., Nematollahi, M., 2014 Theoretical and experimental determination of initial reservoir fluid in a lean gas condensate reservoir, Journal of Petroleum Science and Engineering, 114, pp. 74–81.
- Cengel, Y. and Boles, M., 2014, Thermodynamics: An Engineering Approach, 8th Edition, McGraw-Hill Education
- Deb, K., 2001, Multi-objective Optimization Using Evolutionary Algorithms, 1st Edition, John Wiley & Sons Inc.
- Esparragoza, J.J.V., Iglesias-Silva, G.A., Michael, W., Hlavinka, M.W., Jerry, A., Bullin, J.A., 1992, How to Estimate Reid Vapour Pressure (RVP) of Blends, ydrocarbon Processing, 1992. Bryan Research & Engineering, Inc, USA, pp. 135-138.
- Gary, J.H., Handwerk, G.E., 2001, Petroleum Refining: Technology and Economics, fourth ed. CRC Press.
- Jalali, A., Shafiee, M., Iranshahi, D., Mohammadi, A., 2020, Simulation and energy optimization of a reformate stabilizer unit in a petrochemical plant, Energy Sources, Part A: Recovery, Utilization, and Environmental Effects, 42:1, 104-112.
- Kazerooni, N., Adib, H., Sabet, A., Adhami, M., Adib, M., 2016 Toward an intelligent approach for H_2S content and vapor pressure of sour condensate of south pars natural gas processing plant, Journal of Natural Gas Science and Engineering, 28, pp. 365-371.
- Khazini, L., Fatehifar, E., Fouladpanjeh, B., Ebrahimzadeh, M., 2014, A Simulation of a Claus Tail Gas Treatment Unit in a Petroleum Refinery, Energy Sources, Part A: Recovery, Utilization, and Environmental Effects, 36:22, 2431-2442.
- Kim, W.S., Yangb, D.R., Moona, D.J., Ahna, B.S., 2014, The process design and simulation for the methanol production on the FPSO (floating production, storage and off-loading) system, Chemical Engineering Research and Design, 92 (5), pp. 931-940.
- Lee, Y.J., Hong, S.-I., Moon, D.J., 2011, Studies on the steam and CO_2 reforming of methane for GTL-FPSO applications, Catalyst Today, 174 (1), pp. 31-36.
- Liao, B., Lei, Z., Xu, Z., Zhou, R., Duan, Z., 2001, New process for separating propylene and propane by extractive distillation with aqueous acetonitrile, Chemical Engineering Journal, 84 (3), pp. 581-586.
- MathWorks, MATLAB Software. www.mathworks.com.
- Mokhatab, S., Poe, W.A., Speight, J.G., 2006, Handbook of Natural Gas Transmission and Processing. Gulf Professional Publishing, UK.
- Peng, D.Y., Robinson, D.B., 1976. A new two-constant equation of state, Industrial Engineering and Chemistry Fundamentals, 15 (1), pp. 59-64.
- Rahmanian, N., Ilias, I., Nasrifar, K., 2015, Process simulation and assessment of a back-up condensate stabilization unit, Journal of Natural Gas Science and Engineering, 26, pp. 730-736
- Rahmanian, N., Jusoh, S., Homayoonfard, M., Nasrifar, K., Moshfeghian, M. 2016, Simulation and optimization of a condensate stabilisation process, Journal of Natural Gas Science and Engineering, 32, pp. 453-464.
- Ramadan, M., Khaled, M., Jaber, H., Faraj, J., Bazzi, H., Lemenand, T., 2019, Numerical simulation of Multi-Tube Tank heat exchanger: optimization analysis, Energy Sources, Part A: Recovery, Utilization, and Environmental Effects

- Shafiee, M., Jalali, A., Mohammadi, A., Amirhamzeh, M., 2018, Simulation and energy optimization of the stabilizer tower of a pyrolysis gasoline hydrogenation unit, Energy Sources, Part A: Recovery, Utilization, and Environmental Effects, 40:14, 1714-1720
- Sunny, A., Gazliya, N., Aparna, K., 2019, Optimization of regasified liquefied natural gas based reforming process for syngas production in an ammonia plant, Energy Sources, Part A: Recovery, Utilization, and Environmental Effects
- Tahouni, N., Khoshchehreh, R., Panjeshahi, M., 2014, "Debottlenecking of condensate stabilization unit in a gas refinery" Energy, 77, pp. 742-751
- www.invensys.com, PRO/II comprehensive process simulation.