Korean J. Chem. Eng., 35(1), 20-33 (2018)
DOI: 10.1007/s11814-017-0258-z

pISSN: 0256-1115
eISSN: 1975-7220

A new approach to developing a conceptual topside process design for an offshore platform

Yongheon Cho, Soojin Kwon, and Sungwon Hwang†

Graduate School of Chemistry and Chemical Engineering, Inha University, 100 Inha-ro, Nam-gu, Incheon 22212, Korea
(*Received 10 August 2017 • accepted 18 September 2017*)

Abstract−This study introduces a new approach for the conceptual design of an offshore topside process, satisfying environmental standards, saving utility consumption, and consequently, maximizing economic profit. Twelve individual processes are modeled as a case study, based on sets of combinations between four topside process configurations and three individual production scenarios (i.e., peak oil, peak gas, and peak water) over the life cycle of an oil reservoir. Then, the simulation results of these models are analyzed based on economic profit. In particular, the simulation program is integrated with a mixed-integer non-linear programming algorithm to optimize the design and operating variables (e.g., operating pressures of the multi-stage separators) in order to maximize the economic profit of the platform. Lastly, an economic feasibility study is performed for the design of a profitable and eco-friendly offshore platform.

Keywords: Modeling, Offshore, Topside Process, Optimization, Economic Analysis

INTRODUCTION

Energy consumption using primary energy resources, such as oil, coal, natural gas, and nuclear power, has increased with global economic growth, and the consumption of crude oil has recently grown by more than 30% [1]. Because of the steady increase in the oil price and the exhaustion of the oil reserves from onshore reservoirs, the development of offshore reservoirs has received much interest since the 1950s. Furthermore, the related industry has developed rapidly owing to the high oil price in the mid-2000s. However, under the new era of a low oil price, the high cost of oil and gas production from offshore platforms has caused a deficit, which has resulted in a significant decline in the offshore energy industry. At the same time, according to the IEO 2016 report published by the IEA [2], the oil price is expected to return to $80 per barrel within the next decade. With an oil price over $80 per barrel, offshore oil will be economically competitive, and the development of offshore oil wells is expected to be revitalized. In the meantime, researchers and engineers in the offshore industry have been developing new strategies to reduce the breakeven price below $50 per barrel [3]. Following this global trend, this study focuses on developing a new methodology for the design of the topside process to maximize oil production, minimize the capital and operating costs of a platform, and satisfy required product specifications.

Many researchers have modeled and optimized the topside process of offshore platforms and evaluated their economic performance over the past few decades. Bahadori et al. [4] analyzed the effect of a phase separator on the operating cost, such as the horse power of the compressor and the overall profit of the products in the topside process of offshore plants. They adjusted the pressure of the separator to maximize the profit of the process. As a result, they suggested an optimal pressure based on an analysis of the GOR (gas-to-oil ratio) and the API (American Petroleum Institute) specification that might change with variations in the pressure. However, the oil loss that is caused by gas stream in the middle of the process was not taken into account in the study.

Rasheed [5] worked on the optimization of the process to maximize an annual net profit. Here, the operating variables in a flash vaporization column, such as temperature and pressure, were optimized, satisfying the specified constraints (e.g., number of separator stages, minimum and maximum pressure of the separator). However, the RVP (Reid vapor pressure) and API specification were ignored in this study, under the assumption that these specifications are adjusted by blending a portion of crude oil from another oil field. For reference, the RVP is generally used as an index to indicate the volatility of the hydrocarbon mixture. Crude oil with a high value of RVP has a high potential for leakages of hydrocarbon from the process, causing environmental pollution. Therefore, the RVP has to be restricted under global or local standard specifications.

A recent study by Kim et al. [6] illustrated the optimization procedure of a particular process of an offshore plant to maximize economic profit, estimated based on multiple variables: oil and gas product sales and utility consumption, among others. The research conducted a simulation-based optimization, integrating two commercial software packages: a commercial process simulator, and a genetic optimization program. One of the advantages of this approach is that the vapor-liquid equilibrium and thermodynamic properties can be calculated easily, based on the selected EOS (equation-of-state) package embedded in the simulation program. However, this study used only one base case to determine the effect of the operating variables on profit, even though the feed conditions, such as flowrate, composition, and so on, change over the production period [7]. Normally, the oil or gas production rate is relatively

†To whom correspondence should be addressed.
E-mail: sungwon.hwang@inha.ac.kr

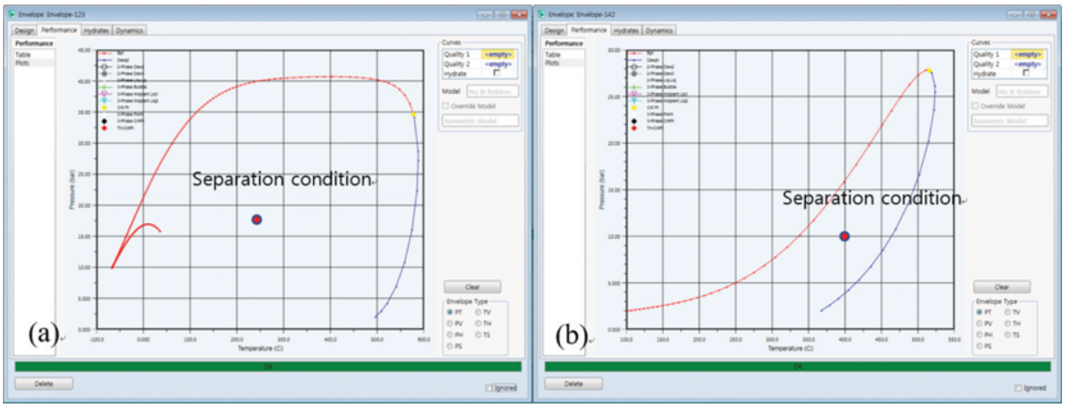

Fig. 1. Vapor-liquid phase equilibrium diagram of the first oil separator (a) and the third oil separator (b), taken from Aspen HYSYS®.

high in the early and middle stages. However, the water content in the feed stream increases as the operation proceeds to the late production stage, because increasing amounts of water are injected in order to maximize oil recovery.

Nguyen et al. [8] developed a model of an offshore process based on three production scenarios, and conducted an exergy analysis to analyze the change in energy consumption over the well's life cycle.

In this work, four different process configurations are considered that vary depending on the number of separators and the usage of condensate recycle streams. At the same time, three separate production scenarios (i.e., peak oil, peak water, and peak gas) are considered over the life cycle of the oil reservoir, because the production profiles, including the flowrate, temperature, and composition of the feed stream to the topside process, keep changing during the production periods. Therefore, 12 cases (3×4) are considered for the optimization of the process operating condition in order to maximize the economic profit. In practice, the operation strategy and economic profit are closely related, and always change as the operation proceeds. Therefore, it is critical to consider various production scenarios when developing a topside process configuration. The new approach adopted in this study considers various process configurations and representative production scenarios together. This approach leads to a more rigorous and systematic procedure for developing a conceptual topside process design, and provides detailed operation guidelines for the process throughout its life cycle.

To model the 12 cases, Aspen HYSYS® is used as a process simulator, integrated with Matlab using an ActiveX server to conduct the optimization [9]. For the MINLP (mixed integer non-linear programming) optimization algorithm, a GA (genetic algorithm) is adopted in this work because it has been widely used in MINLP simulation-based chemical process optimizations [10].

MODELING AND OPTIMIZATION

1. Modeling of Top-side Process

Oil FPSO (floating production storage and offloading) or offshore platforms for oil/gas production are mainly composed of two separate parts: sub-sea and topside processes. The sub-sea production system includes an oil or gas reservoir and submarine pipelines that transport the oil or gas to the topside process located on the platform. In the topside process, crude oil from the sub-sea reservoir is separated into oil, gas, and water phases. Then, the oil and gas products go through the treating unit to remove contaminants such as CO_2, sulfur, and nitrogen compounds, heavy metals, and so on. In this process, the operating conditions are controlled to decrease the loss of oil and gas products. For example, operating variables such as the pressure of the phase separators, differential head of a gas compressor, and temperature of a condensate scrubber affect the amounts of vapor and liquid products from the topside production system. Therefore, they need to be adjusted properly within specified operation ranges, thus maximizing the overall economic profit [11].

Fig. 1 shows pressure-temperature diagram of a phase separator. For example, the red dot indicates the vapor-liquid equilibrium under a certain pressure and temperature inside a vessel [12]. The equilibrium position is directly affected by the pressure and temperature inside a vessel. For example, as the pressure increases under a constant temperature, the vapor fraction of the oil decreases. Similarly, as the temperature increases under a constant pressure, the vapor fraction of the oil in the separator increases and, thus, more hydrocarbon components flash to the gas stream [13]. Therefore, the pressure and temperature inside a vessel are key factors affecting the amounts of vapor and liquid products from the topside production system, and need to be adjusted to maximize the economic profit.

Fig. 2 shows a series of phase separators that separate gas and oil products and produce water as a byproduct. Note that the pressure in the first separator is fixed, based on the oil well, and the pressure of the remaining separators downstream becomes lower.

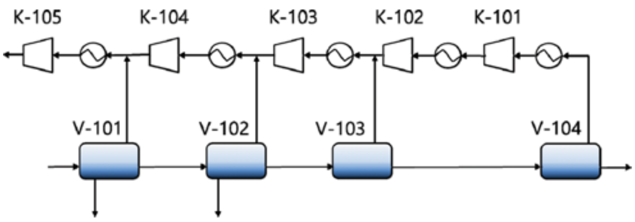

Fig. 2. Process configuration without condensate recycle.

Only light hydrocarbon components flash to the gas flow at the highest pressure of the first separator, and intermediate hydrocarbon components are vaporized at the lower pressure of the second separator. In the last stage separator, relatively heavy hydrocarbon components are normally vaporized. This multiple-stage separation system allows the efficient separation of light hydrocarbons from the heavy hydrocarbons, and prevents unnecessary loss of hydrocarbons. As a result, one of the gas product specifications, such as the RVP, can be satisfied by controlling the composition of the gas product [14]. RVP is an index used to indicate the volatility of crude oil, and the potential loss of hydrocarbons, causing environmental pollution. As lighter hydrocarbons are included in the oil product, the value of the RVP increases. In the refinery industry, an RVP value of 10 psi is normally considered acceptable [15].

The other effective way to decrease hydrocarbon loss is to use a condensate recycle stream for each compressor loop to recover heavy components from the gas product, as shown in Fig. 3.

The process configurations in Figs. 2 and 3 are modeled using the commercial process simulation software, Aspen HYSYS®. In this work, the oil production process includes an oil separation unit, gas compressing, and TEG dehydration units. To model the system, two individual fluid packages are adopted: Peng-Robinson is used for the oil and gas process, and the glycol package is used for the TEG (Triethylene Glycol) dehydration unit [11,16]. The completed models are shown in Fig. 4. The oil product produced through the oil separators has to satisfy the RVP specification of 10 psi (maximum) and BS&W (basic sediment and water) content of 0.2 vol% (maximum) [15,17]. The gas product from the TEG dehydration has to be dry enough not to form a hydrate and cause equipment corrosion. In this work, the water dew point is used to indicate the water content of the gas product.

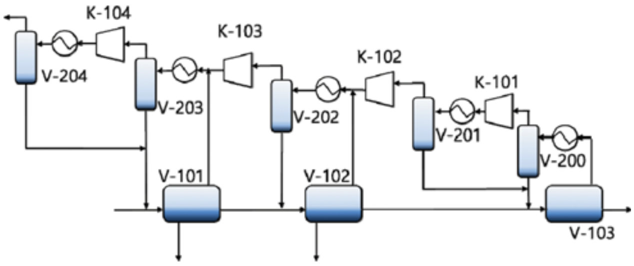

Fig. 3. Process configuration with condensate recycle.

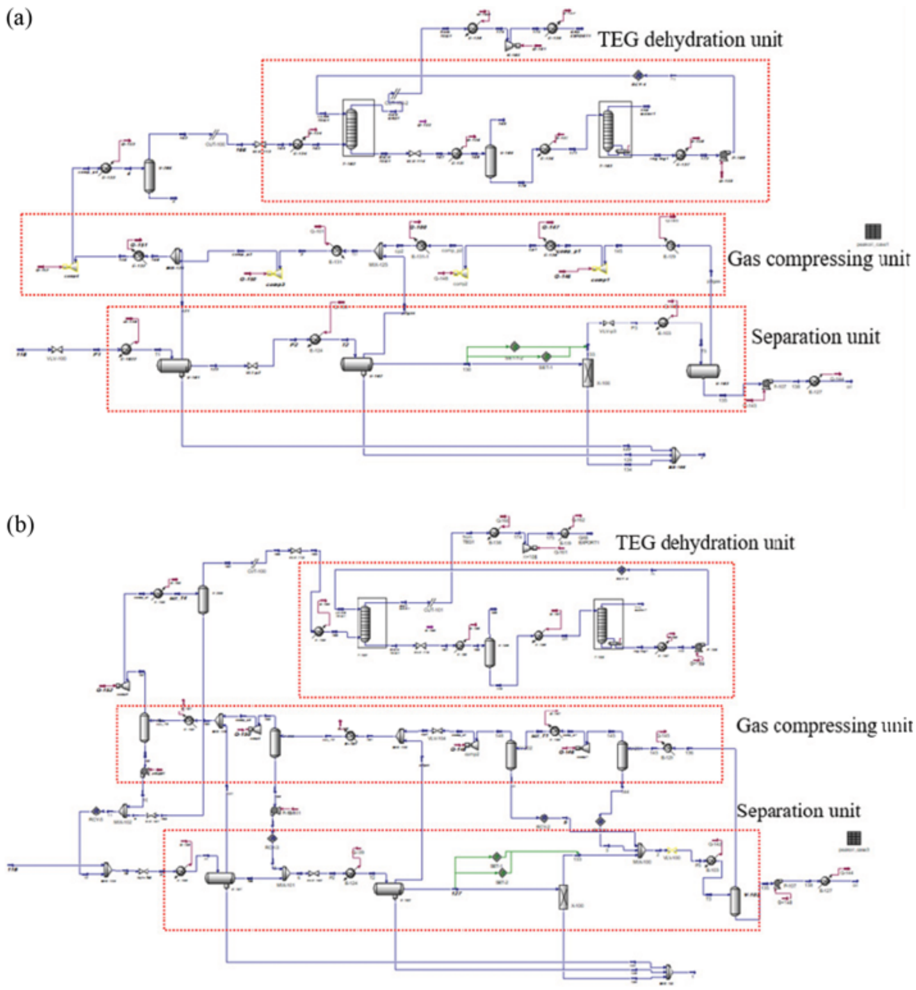

Fig. 4. Model of oil production process using Aspen HYSYS®; (a) process without condensate recycle streams, (b) process with condensate recycle streams.

2. Optimization Algorithm

Aspen HYSYS® makes it possible to conduct simple and visible simulations as a sequential modular simulator. However, for this sequential approach, it is difficult to obtain the gradient information required for a deterministic optimization. Therefore, for the sequential process model, considered as a black box, a stochastic algorithm is more appropriate than a deterministic algorithm, especially for a complicated process model [18]. The main advantage of using a stochastic algorithm is that the black box model can be used for optimization. Therefore, the unit operations of the sequential simulator, developed using rigorous thermodynamic packages, can be used for optimization. In this study, the GA (genetic algorithm) in Matlab, which is commonly used for MINLP problems in stochastic algorithms, is integrated with Aspen HYSYS® for the optimization [19].

An overall optimization procedure is depicted in Fig. 5. Initial values of the optimization variables are sent from Matlab to Aspen HYSYS®, and the simulation results are sent back to Matlab. Then, the objective value is estimated in Matlab based on the simulation results, and compared with that from the previous run. If the difference between the objective values of the current and previous runs is smaller than an allowable criterion, the optimization algorithm terminates. Otherwise, the GA suggests new values for the optimization variables, which Matlab sends to Aspen HYSYS® for another simulation. This procedure iterates until either the maximum number of iterations is reached (as specified by the user), or the difference in the objective values between runs satisfies the termination criterion, ε [20].

Table 1. Product and utility prices [21-23]

Item	Price
Gas product	\$2.75/MMBtu
Oil product	\$48.18/BBL
Steam (100 psig)	\$7.42/1,000 lb
Electricity	\$0.13/kWh

In this study, the overall profit of the process is estimated as an objective function, as shown in Eq. (1). The function mainly considers the sales prices of gas and oil, utility costs, and so on. For reference, the price of cooling water is not considered in this work, because its impact on the objective function is negligible. In addition, it should be noted that the capital cost is not considered during optimization because estimation of the capital cost increases computational burden for optimization. However, once the operating condition is optimized, the capital cost is analyzed afterwards. The unit prices of gas, oil, steam, and electricity are provided by the U.S. Energy Information Administration, as shown in Table 1 [21-23].

Then, the GA optimization algorithm is adopted to maximize the objective function, satisfying the product specifications described in Eqs. (2) and (3). For the optimization variables, the operating conditions for the main equipment are shown below. In order to increase the optimization efficiency, a penalty variable is added to the objective function. For example, if the estimated RVP becomes higher than required maximum value of 10 psi, a significantly high number is applied to the objective function as a penalty variable, so that the profit become very low such as a negative value.

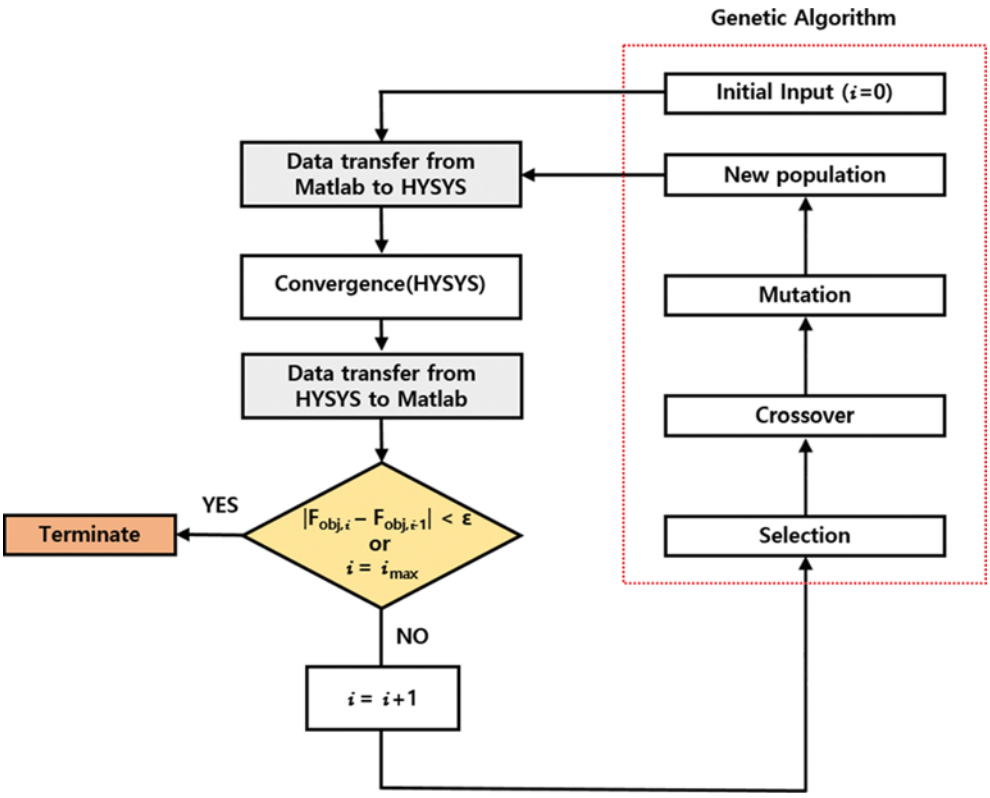

Fig. 5. Optimization procedure with Aspen HYSYS® and Matlab.

In such a way, input variables that leads to violation of product specification are easily excluded for consideration during optimization. This approach leads to a more efficient optimization [24].

Objective: Maximize $f_{profit}(x)$

Constraints:

- $f_{profit}(x)=Sales_{oil}(x)+Sales_{gas}(x)-Cost_{utility}(x)-Penalty(x)$ (1)
- $RVP\leq 10$ psi (2)
- BS&W contents$\leq$0.2 vol% (3)

Variables: $P_{1st\ separator}$, $P_{2nd\ separator}$, $P_{3rd\ separator}$, $P_{1st\ compressor}$, $P_{2nd\ compressor}$, $P_{3rd\ compressor}$, $P_{4th\ compressor}$, $T_{1st\ scrubber}$, $T_{2nd\ scrubber}$, $T_{3rd\ scrubber}$, $T_{4th\ scrubber}$, $T_{5th\ scrubber}$

CASE STUDY

In this study, 12 cases are considered, based on combinations of four process configurations and three individual production scenarios (i.e., peak oil, peak gas, and peak water) in the well life cycle, as shown in Fig. 6. As described in the previous section, the process configurations are classified by the number of phase separators (i.e., three or four phase separators) and the usage of recycle streams, as shown in Fig. 7. The processes in cases 1 and 2 do not have recycle streams, while those in cases 3 and 4 do have recycle streams. The processes in cases 1 and 3 use three separators, and the processes in cases 2 and 4 have four separators.

In general, the process configuration with higher number of separators shows more effective separation performance, because it can achieve sharper separation under more subdivided temperature and pressure conditions. On the other hand, the capital and operating costs increase accordingly. Therefore, it is important to have the appropriate number of phase separators.

As shown in cases 3 and 4 in Fig. 7, heavy hydrocarbon components in the gas stream are compressed, cooled, liquefied, and finally recycled back to the oil separator through the condensate recycle process. This enhances the separation efficiency, reduces the oil and gas product loss, and helps the process to satisfy the product specifications effectively, such as the RVP and hydrocarbon dew point [25]. Once again, adding recycle streams increases the capital and operating costs and, thus, we need to consider carefully whether the process should include recycle streams.

A typical production profile of an oil well over its life cycle is shown in Fig. 8. The flow rate of oil and gas from the reservoir increases rapidly at the start of production, and gradually decreases after the maximum point. On the other hand, the water production rate steadily increases, and exceeds the oil and gas flow rate after a certain period. Therefore, if the topside process is designed on the basis of a specific production period, the continuous changes in the production rate and feed compositions to the topside process might cause significant inefficiency in terms of the process design (e.g., improper equipment sizing) and operating strategy. Therefore, various process configurations and production scenarios need to be considered together when designing the process and preparing the operating strategy. In this study, three production scenarios based on the condition of a feed stream from subsea system to topside process are adopted: peak oil, peak gas, and peak water production scenarios. The condition and composition of each feed stream is summarized in Tables 2 and 3, respectively. As shown in Fig. 8, the composition of a feed stream is similar for both peak oil and peak gas production scenarios. However, because temperature of a feed stream in peak gas production scenario is

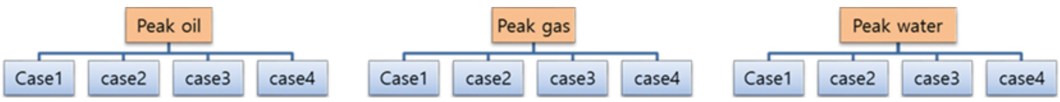

Fig. 6. Three cases in terms of production scenario over the well life cycle.

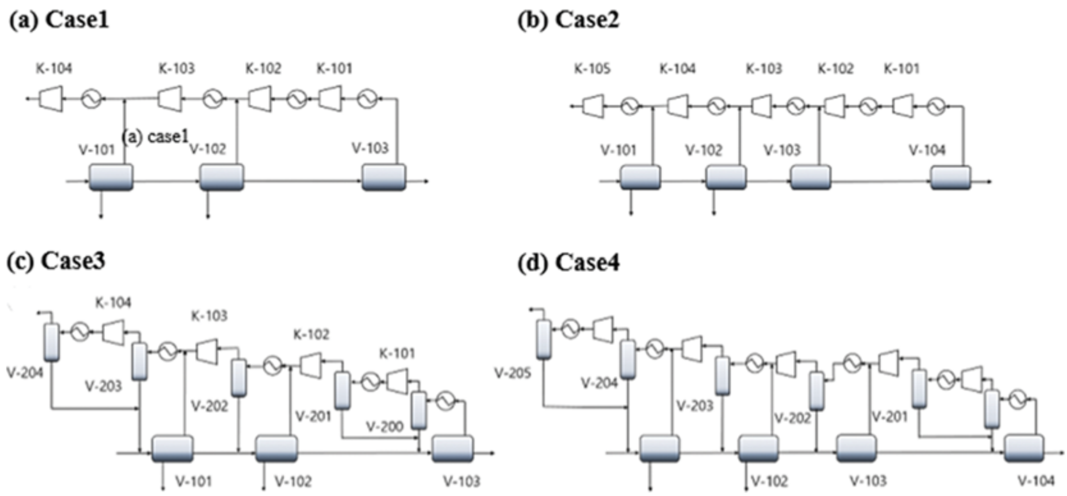

Fig. 7. Graphical illustration of four process configurations, (a) Case 1, (b) Case 2, (c) Case 3, (d) Case 4.

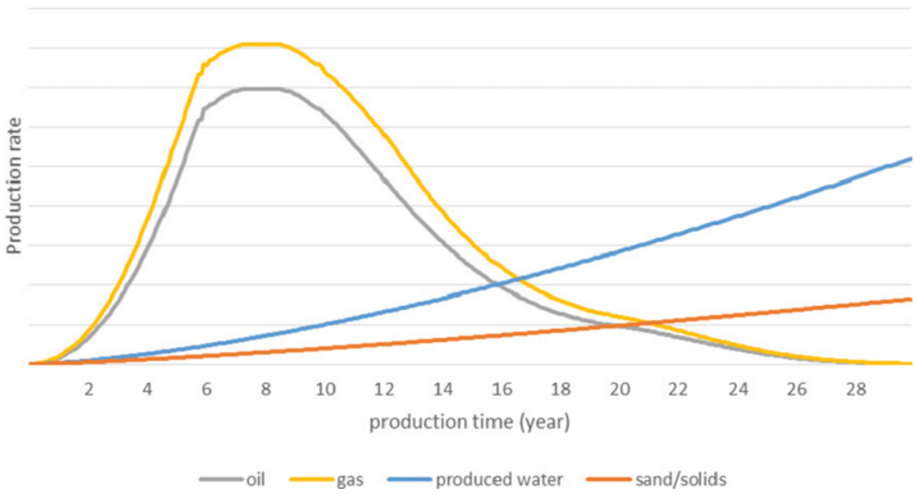

Fig. 8. Typical production profile of an oil/gas well over the life cycle.

Table 2. Feed conditions of each production scenario

Production scenario	Peak oil	Peak gas	Peak water
Description	Maximum oil flow rate	Maximum gas flow rate	Maximum water flow rate
Feed temperature	29.8 °C	59.2 °C	29.9 °C
Feed rate	80,620 kmole/h	49,380 kmole/h	116,880 kmole/h
Feed pressure	30 bar	30 bar	30 bar
Feed vapor fraction	0.096	0.162	0.052

Table 3. Compositions of a feed streams for each production scenario

Component	Peak oil	Peak gas	Peak water
Nitrogen	14.03	18.68	18.29
CO_2	324.69	395.30	552.90
Methane	8800.91	8494.45	5694.52
Ethane	41.35	36.18	24.76
Propane	11.94	9.66	8.50
i-Butane	6.23	7.10	3.33
n-Butane	6.21	6.99	3.17
i-Pentane	8.03	9.12	2.44
n-Pentane	17.82	24.48	5.63
C6*	39.26	52.77	8.91
C7*	208.90	266.63	35.12
C8*	320.22	374.48	45.45
C9*	493.27	504.69	60.61
C10*	1548.58	1617.66	194.27
CN1*	6702.15	5915.80	744.11
CN2*	2552.38	4224.00	566.92
CN3*	875.96	142.86	28.50
CN4*	842.16	1436.64	151.64
CN5*	863.46	1471.75	155.92
H_2O	56942.46	24370.76	108575.02

Note) unit: kmole/hr

higher than the peak oil production scenario, the vapor fraction becomes higher as indicated in Table 2.

The optimization variables and their minimum and maximum operating bounds are summarized in Table 4. For reference, the pressure of the first separator is fixed at 30 bar in all cases. The minimum and maximum bounds of the variables are determined based on actual operating conditions, covering all feasible ranges in order to find a global optimum point.

RESULTS AND DISCUSSION

1. Peak Oil Production Scenario

The peak oil production scenario represents the periods in which the oil production rate is maximized across the overall life cycle of the offshore platform. The results of the peak oil production scenario considered as the base case are summarized in this section. According to Fig. 9, case 3 and case 4, which have recycle streams, produce more oil products than case 1 and case 2 do, because intermediate hydrocarbon components in the gas stream are recovered to the oil stream through the recycle processes. At the same time, the number of separators has a relatively small influence on the oil and gas sales values. In fact, the process with four separators has lower oil sales than that of the process with three separators.

For gas product sales, cases 3 and 4, which have recycle streams, still produce more gas products than cases 1 and 2 do, and it is shown that having more separators produces more gas products. Furthermore, it is known that the optimization maximizes oil products rather than gas products because the oil price is higher than the gas price.

Table 4. Minimum and maximum bounds of optimization variables

Variable	Notation	Peak oil	Peak gas	Peak water
Case 1				
Second separator pressure	P2 (bar)	12-30	12-30	12-30
Third separator pressure	P3 (bar)	1-6	1-11	1-6
Discharge pressure of compressor 1	Pcomp1 (bar)	1-7	1-11	1-7
Discharge pressure of compressor 2	Pcomp2 (bar)	7-20	11-20	7-20
Discharge pressure of compressor 3	Pcomp3 (bar)	20-30	20-30	20-30
Case 2				
Variable	Notation	Peak oil	Peak gas	Peak water
Second separator pressure	P2 (bar)	12-30	12-30	12-30
Third separator pressure	P3 (bar)	6-20	10-20	6-20
Fourth separator pressure	P4 (bar)	1-6	1-11	1-6
Discharge pressure of compressor 1	Pcomp1 (bar)	1-9	1-11	1-9
Discharge pressure of compressor 2	Pcomp2 (bar)	9-20	11-20	9-20
Discharge pressure of compressor 3	Pcomp3 (bar)	9-20	11-20	9-20
Discharge pressure of compressor 4	Pcomp4 (bar)	20-30	20-30	20-30
Case 3				
Variable	Notation	Peak oil	Peak gas	Peak water
Second separator pressure	P2 (bar)	12-30	12-30	12-30
Third separator pressure	P3 (bar)	1-6	1-11	1-11
Discharge pressure of compressor 1	Pcomp1 (bar)	1-7	1-11	1-11
Discharge pressure of compressor 2	Pcomp2 (bar)	7-20	11-20	11-20
Discharge pressure of compressor 3	Pcomp3 (bar)	20-30	20-30	20-30
Scrubber temperature 1	T1 (°C)	38-60	38-60	38-60
Scrubber temperature 2	T2 (°C)	38-60	38-60	38-60
Scrubber temperature 3	T3 (°C)	38-60	38-60	38-60
Scrubber temperature 4	T4 (°C)	38-60	38-60	38-60
Case 4				
Variable	Notation	Peak oil	Peak gas	Peak water
Second separator pressure	P2 (bar)	12-30	12-30	12-30
Third separator pressure	P3 (bar)	6-20	10-20	10-20
Fourth separator pressure	P4 (bar)	1-6	1-11	1-11
Discharge pressure of compressor 1	Pcomp1 (bar)	1-9	1-12	1-11
Discharge pressure of compressor 2	Pcomp2 (bar)	7-20	12-20	11-20
Discharge pressure of compressor 3	Pcomp3 (bar)	7-20	12-20	11-20
Discharge pressure of compressor 4	Pcomp4 (bar)	20-30	20-30	20-30
Scrubber temperature 1	T1 (°C)	38-60	38-60	38-60
Scrubber temperature 2	T2 (°C)	38-60	38-60	38-60
Scrubber temperature 3	T3 (°C)	38-60	38-60	38-60
Scrubber temperature 4	T4 (°C)	38-60	38-60	38-60
Scrubber temperature 5	T5 (°C)	38-60	38-60	38-60

Note that the overall cost of utility consumption is also considered for the economic profit evaluation after the optimization is completed for each process configuration. For example, the total consumption of steam, and its cost, are taken into account when the crude oil is heated before entering each separator. In addition, the electric power consumption is added to the total utility cost because of the power consumption of the gas compressors. It is shown that cases 2 and 4 consume more in terms of utilities because of the operation of the additional phase separator and preheater. On the other hand, recycle streams do not have much impact on the overall utility consumption, because the compressor is already being used, even without the recycle streams. Thus, cases 3 and 4 consume slightly more steam than the other cases do. Note that the overall gas flow rate from the compressor in cases 3 and 4 be-

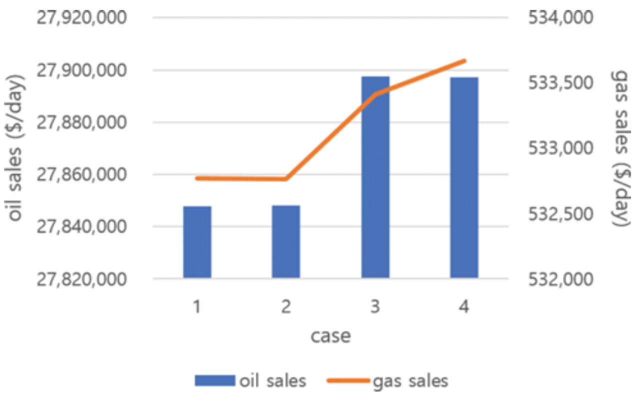

Fig. 9. Oil and gas product sales in the peak oil production scenario.

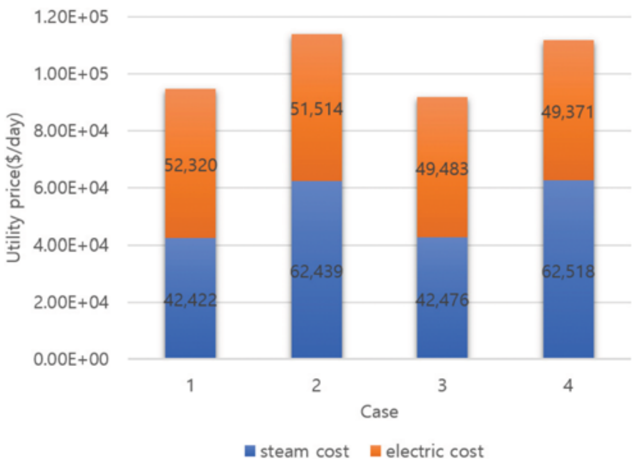

Fig. 10. Utility consumption in the peak oil production scenario.

comes less than in the other cases. Thus, the power consumption of the compressor becomes lower in this case than in the others. The utility consumptions of the four cases under the peak oil production scenario are summarized in Fig. 10.

The overall profits of the four different cases are compared in Fig. 11. In order to compare the profits in each case, the overall profit of case 1 is assumed to be zero, and then the profits and losses of the other cases are compared with that in case 1 (see Fig. 11). According to Fig. 11, case 3 produces the largest profit, followed by case 4, while case 2 produces the lowest profit. As a result, using recycle streams is recommended, minimizing the number of phase separators to three in the peak oil production scenario.

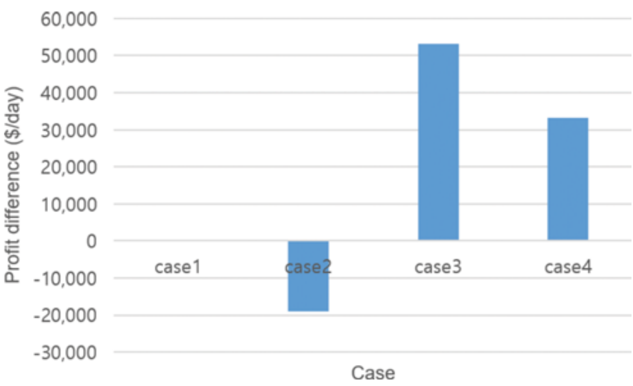

Fig. 11. Profit index in the peak oil production scenario.

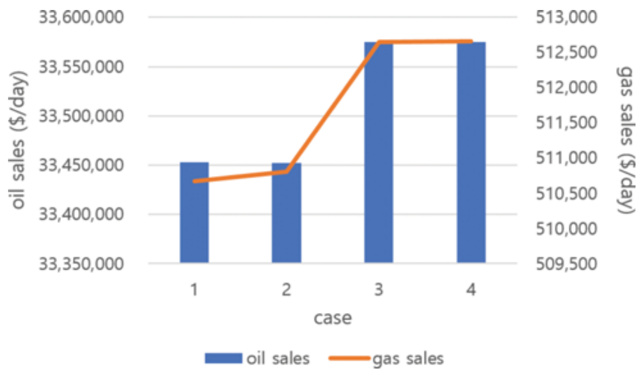

Fig. 12. Oil and gas product sales in the peak gas production scenario.

2. Peak Gas Production Scenario

The peak gas production scenario represents the period in which the gas production rate is maximized across the overall life cycle of the offshore platform. From Fig. 12, which shows the oil and gas product sales, we can easily see that the gas product sales of the four cases are lower than those under the peak oil production scenarios, while it shows similar trends to those in Fig. 9. Once again, the processes with recycle streams produce higher product sales.

The total utility consumptions under the peak gas production scenario seem to be lower than those of the peak oil production scenario, as shown in Fig. 13. In particular, far less of the heating utility is required for the peak gas production scenario, owing to the significantly lower heat duty required for pre-heating the heat exchangers. Note that the feed stream under the peak gas production scenario contains a higher portion of light hydrocarbons and, therefore, more gas is separated from the three-phase separator located at the beginning of the process. Therefore, a higher utility cost is incurred from the compressors than that of the steam from the oil pre-heaters. Case 3 in this scenario shows the lowest utility

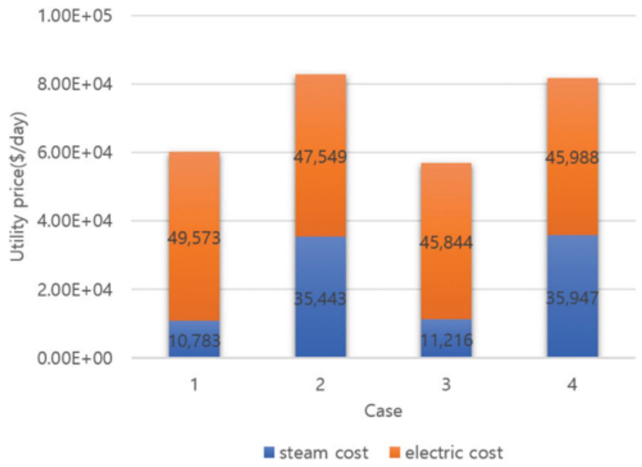

Fig. 13. Utility consumption in the peak gas production scenario.

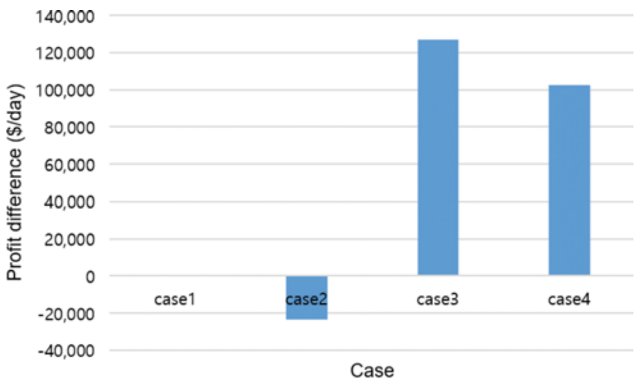

Fig. 14. Profit index in the peak gas production scenario.

consumption because of the lower amounts of gas (i.e., maximizing oil production through recycle streams) than in case 1 and the lower oil heating (i.e., fewer number of oil pre-heaters) than in case 4.

Owing to the higher oil and gas production rates and the lower consumption of utilities, case 3 in this scenario shows the highest economic profit, followed by case 4, while case 2 shows the lowest profit, as illustrated in Fig. 14, which shows merely the profit difference between the cases.

3. Peak Water Production Scenario

The peak water production scenario represents the periods in which the feed stream contains the maximum amounts of water (e.g., 87 mole%) across the overall life cycle of the offshore platform. In this scenario, cases 1 and 2, which do not use recycle streams, produce more gas product sales than oil, while the other cases (i.e., cases 3 and 4) produce more oil product sales. The use of recycle streams increases the recovery of heavy hydrocarbon components to the oil flow. However, no significant difference is noticed when different number of separators are used in the process. The products sales are summarized in Fig. 15.

In terms of economic profit, this scenario shows a similar trend to the other scenarios. In other words, case 3 still shows the highest profit, while case 2 shows lowest profit, as shown in Fig. 17. However, the absolute profit values are much lower than in the other two scenarios.

Table 5 summarizes the optimization results of the four process

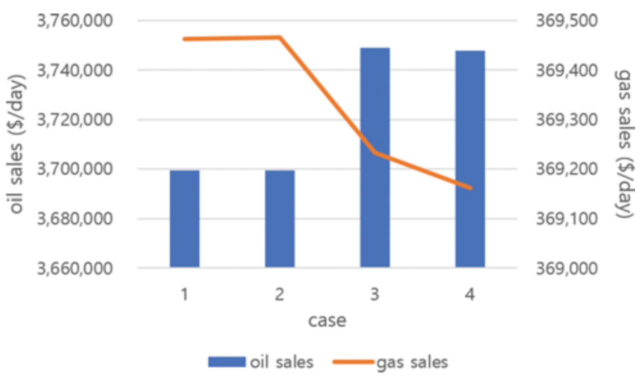

Fig. 15. Oil and gas product sales in the peak water production scenario.

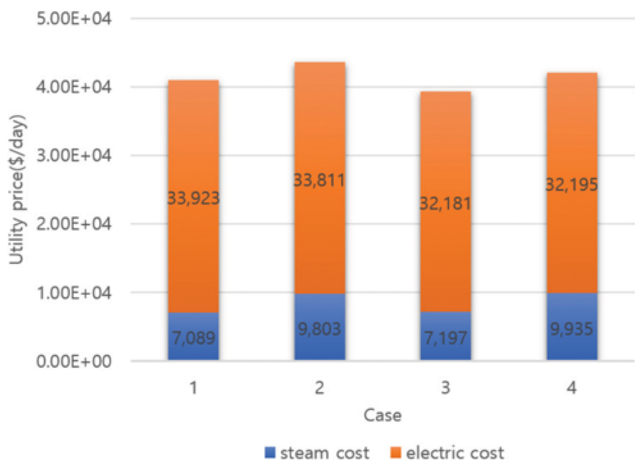

Fig. 16. Utility consumption in the peak water production scenario.

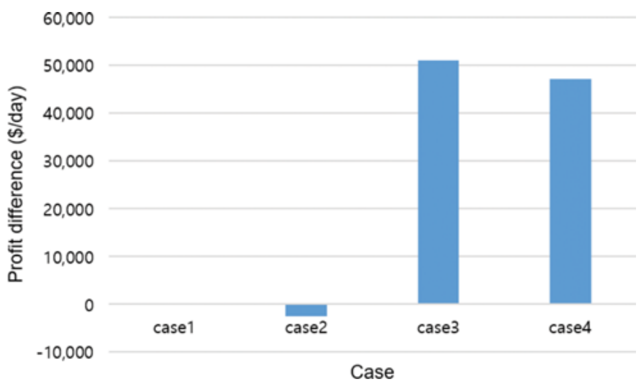

Fig. 17. Profit index in the peak water production scenario.

configurations under the three different scenarios (i.e., peak oil, peak gas, and peak water). Note that the type and number of optimization variables are different for each case and scenario. For example, the optimization of the scrubber temperature is considered only in cases 3 and 4, because the scrubbers are used for the recycle streams.

4. Optimization of Operating Conditions

For the design of the topside process on an offshore platform, we need to consider not only the process design, but also the process operation, including the feasible operating range of the major equipment. For example, the flowrate of the feed stream to the topside process varies along the life cycle, but the composition also changes as the operation proceeds. Therefore, the process design needs to be considered under various production scenarios. This approach provides an efficient strategy for the conceptual design of the process and its operation.

Fig. 18 compares the optimized operating conditions of a specific major item of equipment under different production scenarios for each case. The variables on the x-axis show the pressure values of the multi-stage separators as independent variables, except for the first separator, because its pressure is fixed. For example, Fig. 18(a) shows the optimum pressure of the second and third separators. This indicates that the pressure of these separators needs to be maintained at a high level in the peak gas production scenario,

Table 5. Optimization results in the peak oil production scenario

	Notation	Case 1	Case 2	Case 3	Case 4
Second separator pressure	P2 (kPa)	1,835	1,865	1,769	1,869
Third separator pressure	P3 (kPa)	515	695	513	671
Fourth separator pressure	P4 (kPa)	-	549	-	538
Compressor P1	comP1 (kPa)	532	682	561	567
Compressor P2	comP2 (kPa)	782	1,131	1,676	1,343
Compressor P3	comP3 (kPa)	2,252	1,864	3,000	1,865
Compressor P4	comP4 (kPa)	-	2,935	-	2,988
Scrubber temperature 1	T1 (°C)	-	-	38	42
Scrubber temperature 2	T2 (°C)	-	-	38	38
Scrubber temperature 3	T3 (°C)	-	-	38	38
Scrubber temperature 4	T4 (°C)	-	-	58	38
Scrubber temperature 5	T5 (°C)	-	-	-	57
Steam cost	$/day	42,422	62,439	42,476	62,518
Electric cost	$/day	52,320	51,514	49,483	49,371
Oil product sales	$/day	27,847,787	27,847,988	27,897,460	27,897,298
Gas product sales	$/day	532,769	532,762	533,411	533,667
Overall profit	$/day	28,285,814	28,266,798	28,338,912	28,319,075
Profit difference (against Case 1)	$/day	0	−19,016	+53,098	+33,261
Oil RVP	kPa	68	67.8	67.8	66.6

Table 6. Optimization results in the peak gas production scenario

	Notation	Case 1	Case 2	Case 3	Case 4
Second separator pressure	P2 (kPa)	2,444	2,410	1,977	2,476
Third separator pressure	P3 (kPa)	951	1,368	938	1,324
Fourth separator pressure	P4 (kPa)	-	990	-	987
Compressor P1	comP1 (kPa)	985	991	960	1,198
Compressor P2	comP2 (kPa)	1,996	1,425	1,635	1,474
Compressor P3	comP3 (kPa)	2,967	1,956	3,000	1,762
Compressor P4	comP4 (kPa)	-	2,995	-	2,977
Scrubber temperature 1	T1 (°C)	-	-	39	43
Scrubber temperature 2	T2 (°C)	-	-	38	42
Scrubber temperature 3	T3 (°C)	-	-	38	41
Scrubber temperature 4	T4 (°C)	-	-	60	38
Scrubber temperature 5	T5 (°C)	-	-	-	59
Steam cost	$/day	10,783	35,443	11,216	35,947
Electric cost	$/day	49,573	47,549	45,844	45,988
Oil product sales	$/day	33,453,293	33,452,472	33,575,109	33,575,180
Gas product sales	$/day	510,671	510,803	512,648	512,659
Overall profit	$/day	33,903,608	33,880,282	34,030,697	34,005,905
Profit difference (against Case 1)	$/day	0	−23,326	+127,089	+102,297
Oil RVP	kPa	68	67.6	68	68

as compared with the other scenarios, for cases 1 and 2 (i.e., the process configurations without recycle streams). The feed stream in the peak gas production scenario contains more gas components. Therefore, the pressure of the separators has to be higher in order to minimize the vaporization of the hydrocarbon components, which produces more in terms of oil products. On the other hand, the pressure of the second separator for cases 3 and 4 (i.e., process configurations with recycle streams) needs to be maintained at a high level in the peak water production scenario.

5. Capital Cost Analysis

After the optimization of each case is completed, the required capital cost is estimated for each case, and the results are compared, as shown in Tables 8, 9, and 10. Each table shows the details of the capital costs for various process configurations under the peak oil, gas, and water production scenarios, respectively. The equipment and installation costs are initially estimated using the Aspen Pro-

Table 7. Optimization results in the peak water production scenario

	Notation	Case 1	Case 2	Case 3	Case 4
Second separator pressure	P2 (kPa)	1,363	2,233	2,425	2,492
Third separator pressure	P3 (kPa)	464	909	819	1,061
Fourth separator pressure	P4 (kPa)	-	501	-	873
Compressor P1	comP1 (kPa)	569	701	1,075	1,045
Compressor P2	comP2 (kPa)	1,330	1,997	1,800	1,378
Compressor P3	comP3 (kPa)	2,978	1,253	2,999	1,704
Compressor P4	comP4 (kPa)	-	3,000	-	2,995
Scrubber temperature 1	T1 (°C)	-	-	40	54
Scrubber temperature 2	T2 (°C)	-	-	39	51
Scrubber temperature 3	T3 (°C)	-	-	38	44
Scrubber temperature 4	T4 (°C)	-	-	56	39
Scrubber temperature 5	T5 (°C)	-	-	-	55
Steam cost	$/day	7,089	9,803	7,197	9,935
Electric cost	$/day	33,923	33,811	32,181	32,195
Oil product sales	$/day	3,699,292	3,699,305	3,748,973	3,747,785
Gas product sales	$/day	369,464	369,455	369,233	369,161
Overall profit	$/day	4,027,744	4,025,146	4,078,828	4,074,817
Profit difference (against Case 1)	$/day	0	−2,598	+51,084	+47,073
Oil RVP	kPa	67.9	67.9	66.5	67.1

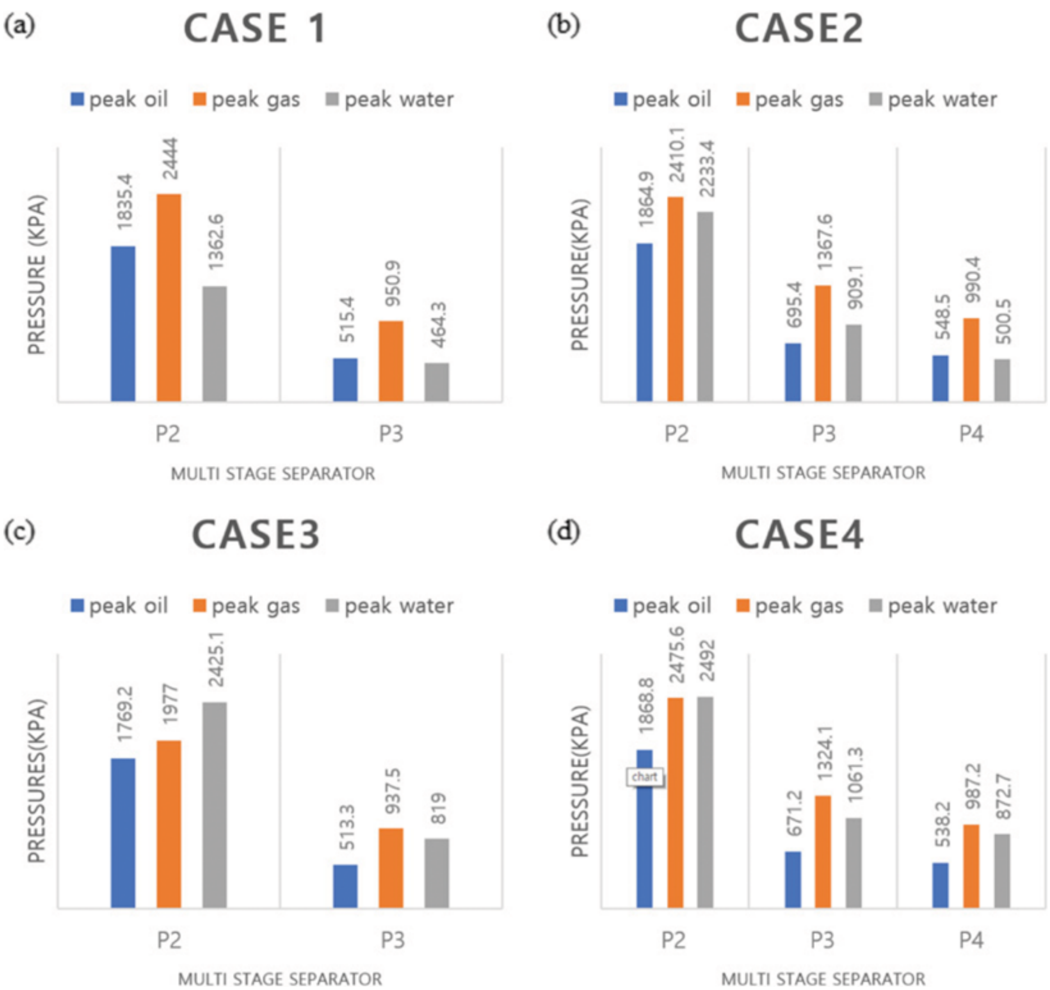

Fig. 18. Optimum pressure of multi-stage separators under various production scenarios, (a) case 1, (b) case 2, (c) case 3, (d) case 4.

Table 8. Summary of capital cost: Peak oil production scenario (optimized)

Peak oil production scenario				
Case / Cost items (USD)	Case 1	Case 2	Case 3	Case 4
Equipment cost	12,099,700	13,022,100	11,708,500	12,868,800
Total installed cost	17,337,000	18,881,300	17,149,100	19,018,100
ISBL	29,436,700	31,903,400	28,857,600	31,886,900
OSBL	11,774,680	12,761,360	11,543,040	12,754,760
Engineering & contingency cost	8,242,276	8,932,952	8,080,128	8,928,332
Working capital	16,484,552	17,865,904	16,160,256	17,856,664
Total capital cost	65,938,208	71,463,616	64,641,024	71,426,656

Table 9. Summary of capital cost: Peak gas production scenario (optimized)

Peak gas production scenario				
Case / Cost items (USD)	Case 1	Case 2	Case 3	Case 4
Equipment cost	12,423,200	13,684,000	12,161,400	13,609,000
Total installed cost	17,913,400	20,016,600	17,920,300	20,293,500
ISBL	30,336,600	33,700,600	30,081,700	33,902,500
OSBL	12,134,640	13,480,240	12,032,680	13,561,000
Engineering & contingency cost	8,494,248	9,436,168	8,422,876	9,492,700
Working capital	16,988,496	18,872,336	16,845,752	18,985,400
Total capital cost	67,953,984	75,489,344	67,383,008	75,941,600

Table 10. Summary of capital cost: Peak water production scenario (optimized)

Peak water production scenario				
Case / Cost items (USD)	Case 1	Case 2	Case 3	Case 4
Equipment cost	8,023,800	8,485,600	7,805,400	8,543,200
Total installed cost	11,710,700	12,590,000	11,790,600	13,061,100
ISBL	19,734,500	21,075,600	19,596,000	21,604,300
OSBL	7,893,800	8,430,240	7,838,400	8,641,720
Engineering & contingency cost	5,525,660	5,901,168	5,486,880	6,049,204
Working capital	11,051,320	11,802,336	10,973,760	12,098,408
Total capital cost	44,205,280	47,209,344	43,895,040	48,393,632

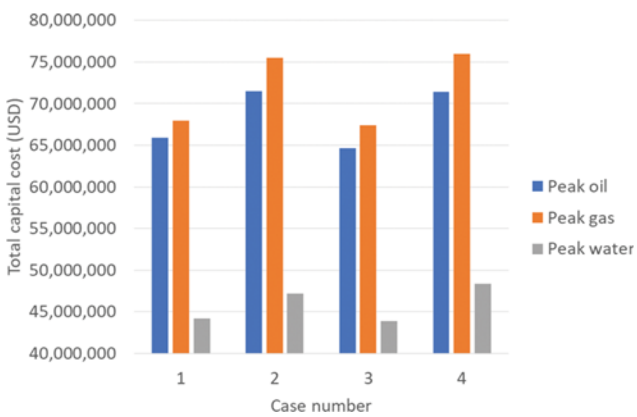

Fig. 19. Summary of capital costs.

cess Economic Analyzer (APEA), after which the ISBL (Inside Battery Limit), OSBL (Outside Batter Limit), engineering cost, contingency cost, and working capital are estimated [26-28].

The estimated total capital costs of the four different cases under the three production scenarios are compared in Fig. 19. The total costs for each process configuration show similar trends, in that the peak gas production scenario requires the highest capital costs, while the peak water production scenario requires the lowest capital costs. Furthermore, the process configuration of case 3 requires the lowest capital costs, while cases 2 and 3 require the highest capital costs owing to the additional phase separator. In addition, we can assume that the recycle process reduces the total flowrate of the gas compressing unit and, thus, the heat duty of the coolers. Even though the condensate scrubbers for the recycle process are added to case 3, their cost is negligible compared to that of the gas compressors, which have a substantial impact on the total equipment cost. In conclusion, if the topside process is built based on the configuration in case 3, it can cover all feasible production scenarios with minimum capital costs.

CONCLUSION

In this work, four different process configurations were modeled using the commercial simulation software Aspen HYSYS®, and the results were compared based on capital and operating costs. In addition, three representative production scenarios were considered over the life cycle of the oil well (i.e., peak oil, peak gas, and peak water) together with the aforementioned process configurations. Therefore, 12 individual cases were simulated, and their results compared.

- For peak oil production scenario, the process configuration of case 3 that has a recycle stream with three multi-stage separators produces the highest economic profit, followed by the process configuration of case 4 that has a recycle stream with four multi-stage separators. It is noticed that the use of a recycle stream enables producing more oil products that leads to higher economic profit.
- For peak gas production scenario, the process configuration of case 3 shows the highest economic profit, followed by the process configuration of case 4, which is similar to the results of peak oil production scenario. Meanwhile, the total operating cost is estimated to be lower than the costs of the peak oil production scenario due to significantly lower heat duty required for pre-heating the heat exchangers. The impact of using a recycle system on economic profit is shown to be higher than the one in peak oil production scenario.
- For peak water production scenario, the results were similar to the other two production scenarios. However, the economic profit in this scenario is significantly lower than the other scenarios.
- Optimum operation range of major equipment are compared over the well's life cycle. For example, the change of optimum operating pressure of the multi-stage separators are identified over the well's life cycle. It is interesting to note that the optimum point changes significantly, depending on selection of production scenario and process configuration.
- Lastly, capital costs of four different process configurations were compared for each individual production scenario. The results show that the process configuration of case 3 requires minimum amounts of capital cost, followed by the process configuration of case 1. It concludes that addition of multi-stage separation requires higher capital cost increase, compared with addition of a recycle stream. Among three production scenarios, the peak gas production scenario requires the highest capital cost, followed by the peak oil production scenario, while the peak water production scenario requires the lowest capital cost.

It should be noted that the results might change depending the location of reservoir, condition of a feed stream, sales prices of products, utility costs, etc. For example, price of natural gas in Japan is about \$7.85/MMBtu that is about 3 times more expensive than US [29]. In this case, the optimization results might show that more gas product needs to be produced to maximize an economic profit. For this, operating pressure of multi-stage separators becomes lower to produce more gas product, while operating temperature of a scrubber will be higher, compared with the case that we discussed earlier. As a result, economic evaluation results and operation strategy might be totally different from the previous study.

This is the first study to consider both process configurations and production scenarios, and we believe this approach provides a robust methodology to develop a conceptual topside process design for an offshore platform and a reliable strategy for operation.

ACKNOWLEDGEMENTS

This work was supported by an INHA UNIVERSITY Research Grant.

REFERENCES

1. KEMCO, http://www.energy.or.kr/web/kem_home_new/energy_issue/mail_vol19/pdf/publish_02_201506.pdf (2015).
2. J. Conti, *International Energy Outlook 2016*, U.S. Energy Information Administration, U.S.A. (2016).
3. Oil price, http://oilprice.com/Energy/Energy-General/Gulf-Of-Mexico-Breakeven-Falls-Below-50-Per-Barrel.html (2017).
4. A. Bahadori, H. B. Vuthaluru and S. Mokhatab, *ASIA-PACIFIC J. Chem. Eng.*, **3**, 380 (2008).
5. R. S. Al-Ameeri, *Middle East Tech. Conf. Exhib.* (2013).
6. I. H. Kim, S. Dan, H. Kim, H. R. Rim, J. M. Lee and E. S. Yoon, *Ind. Eng. Chem. Res.*, **53**, 8810 (2014).
7. IFP school, http://www.ifp-school.com/upload/docs/application/pdf/2015-02/3_main_steps_oil_gas_field_development.pdf (2015).
8. T. Van Nguyen, T. G. Fülöp, P. Breuhaus and B. Elmegaard, *Energy*, **73**, 282 (2014).
9. O. K. Nordvåg, Master's Thesis, *NTNU*, Norway (2012).
10. R. Malhotra, N. Singh and Y. Singh, *Comput. Inf. Sci.*, **4**, 39 (2011).
11. I. Ashour, N. Al-Rawahi, A. Fatemi and G. Vakili-Nezhaad, *Applications of Equations of State in the Oil and Gas Industry*, InTech (2011).
12. Ø. Kylling, Master's Thesis, *Institutt for teknisk kybernetikk*, Norway (2009).
13. A. M. Fulgueras, J. Poudel, D. S. Kim and J. Cho, *Korean J. Chem. Eng.*, **33**, 46 (2016).
14. N. Moghadam and M. Samadi, *Int. J. Chem. Eng. Appl.*, **3**, 461 (2012).
15. T. Lidderdale, *Fuel Technol. Manage.*, **6**, 47 (1999).
16. M. A. Isa, U. Eldemerdash and K. Nasrifar, *Chem. Eng. Res. Des.*, **91**, 1731 (2013).
17. L. Vafajoo, K. Ganjian and M. Fattahi, *J. Pet. Sci. Eng.*, **90**, 107 (2012).
18. D. Huang, T. T. Allen, W. I. Notz and N. Zeng, *J. Glob. Optim.*, **34**, 441 (2006).
19. A. Aspelund, T. Gundersen, J. Myklebust, M. P. Nowak and A. Tomasgard, *Comput. Chem. Eng.*, **34**, 1606 (2010).
20. S. Park, Y. Lee, G. Kim and S. Hwang, *Korean J. Chem. Eng.*, **33**, 3417 (2016).
21. IEA, http://www.iea.org/media/statistics/surveys/prices/mps.pdf (2017).

22. IEA, https://www.eia.gov/dnav/ng/hist/rngwhhdd.htm (2017).
23. IEA, https://www.eia.gov/electricity/monthly/pdf/epm.pdf (2017).
24. O. Yeniay, *Math. Comput. Appl.*, **10**, 45 (2005).
25. F. S. Manning and R. E. Thompson, *Oilfield processing of petroleum: Crude oil (Vol. 2)*, Pennwell books, 167 (1995).
26. EIA, *Engineering Economic Analysis Guide : Liquid Fuels Technologies*, U.S. Energy Information Administration, U.S.A. (2015).
27. AspenTech, *Aspen Capital Cost Estimator User's Guide*, Aspen Technology, Inc., U.S.A. (2012).
28. A. Permatasari, P. Fasahati, J. Ryu and J. Liu, *Korean J. Chem. Eng.*, **33**, 3381 (2016).
29. YCHARTS, https://ycharts.com/indicators/japan_liquefied_natural_gas_import_price (2017).