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Energy Recovery from LNG (Part 2): Evaluation of Process Flowsheets for the Retrofit of an LNG Regasification facility with a Propane ORC for Power Production

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Abstract

This article is the second article in a series that will examine the opportunities for LNG cold energy recovery. This article uses Key Performance Metrics to evaluate process flowsheets for a power recovery plant containing a propane ORC. DWSIM V9.0.2. was used to simulate the plant using the Peng Robinson 1978 Advanced thermodynamics and Steam tables (IAPWS-IF97) models. Heat exchanger and economics calculations were performed using Python in DWSIM. Seven process flowsheets, with at most one heat integration level in and across each process system, were evaluated against the Key Financial Performance Metrics (KFPMs) and Key Technical Performance Metrics (KTPMs). It is shown in this article that case 5 with the expansion of the LNG after the ORC has the highest Total Annualized Earnings (TAEs). Case 5 has a TAE of 3.55 MUSDy^{-1} and an LNG ORC Value (LOV) of 3.43 x 10^{-3} USD per Sm^3 LNG. It also has a Plant Electrical Energy Efficiency (PEEE), Plant Exergy Utilization Efficiency (PEUE) and Plant Power Production Density (PPPD) of 14.8 %, 25.4 % and 0.0223 kWh per Sm^3 LNG. Case 6 which uses a 2-stage turbine to increase the net power has a TAE and LOV of 3.17 MUSD y^{-1} and 3.06 x 10^{-3} USD per Sm^3 LNG. While the PEEE, PEUE and PPPD are 15.2 %, 26.6 % and 0.0236 kWh per Sm^3 LNG. It is also shown that amongst the flowsheets, expanding the LNG stream results in the greatest recuperation of exergy with a PERR of approximately 0.544.

Keywords: ORC; LNG; TEA; Cold Energy Use; Process Design; Optimum inter-stage pressure and pressure ratio

1. Introduction

Natural gas is employed for energy production and is also employed in the chemical and plastics production process. LNG plays an important part of this as it is mainly used to ship natural gas internationally. There are currently 25 LNG major projects under construction or in advanced development globally representing 390 MTPA of new and expanded liquefaction capacity [1]. The exergy of methane calculated at 112.15 K and 113kPag at reference conditions of 300.15 K and 0 kPag is 1090 kJ/kg or 0.303 kWh/kg.

This is the second article of a series that will examine the different uses for the cold energy of LNG. The first article [2] contains exergy and sensitivity analyses of the cycle where the Total Annualized Earnings (TAE) and the Annualized Interest Rate of Return (AIRR) for the new equipment are Key Financial Performance Metrics (KFPM). It should be noted that an old version of one of the textbooks used in Part 1 resulted in a very large overestimate in the calculation of the seawater pump cost and different cost values for the heat exchanger calculation compared to the new textbook. Using a different pump cost calculation [3] results in a translational shift of + 99.9 kUSD in the TAE (Addendum D). Therefore, the simulation now produces a base case with a TAE of 400.3 kUSD and AIRR of 18.1 %. Since the alternate case is apart from the base case, minor adjustments were also made to keep it within the design constraints. The simulation now produces an alternate case TAE of 936.6 kUSD and an IRR of 27.6%. Some Key Technical Performance Metrics (KTPMs) were also employed in the analysis of the cycle and these KTPMs are employed in this article to perform an analysis of the plant.

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Different methods of improving the energy usage of the facility by employing different technologies are examined in this paper via developing different process flowsheets. S.B. Sutra and M. Sudibandriyo [4] examined the use of direct expansion of the LNG stream and found it to be economical, producing a NPV of USD 695,032. J. Park et al [5] evaluated an LNG regasification power plant that included ORCs with a recuperator and DE turbines and found that compared to the scheme that wasted energy and to the scheme that also had an integrated cryogenic energy storage, it produced the highest NPV.

Therefore, this article builds on the first article and presents the use of DWSIM for the simulation and modelling of process flowsheets using the single stage propane ORC with one heat integration level and / or direct expansion of the LNG stream and an inherently safer design (i.e. Stainless Steel). A hypothetical facility is presented in the article. The objective of this article is to present the development of the process flowsheets and the analysis of the process flowsheets with respect to the KFPMs and KTPMs. This work is motivated by the need for alternative energy sources, reduction of carbon emissions and the study of the development, design and optimization of process flowsheets.

2. Methodology

2.1. Introduction

DWSIM V9.0.2 was used to simulate the plant and heat exchanger modelling and economic modelling were also performed. The Peng-Robinson 1978 (PR78) Advanced and Steam tables (IAPWS-IF97) were the property packages employed for hydrocarbons and water respectively. The thermodynamic calculations were found to be inaccurate at the higher pressures for the mixed LNG stream. Hence, the LNG stream was assumed to be only methane. The phase envelope for the mixed stream and for the methane stream are shown in Appendix A. The exergy losses in the plant were then examined and simple modifications to the flowsheet were completed to improve on the efficiency and profitability of the facility.

It was shown in the first article that several variables affected the profitability of the cycle including the cycle pressure and mass flowrate. When the heat exchanger attained its pinch temperature, the ORC was at its maximum mass flowrate as anticipated. Therefore, pinch technology was employed to overcome the pinch in the cycle and to therefore increase the power output and hence profitability of the cycle by using a recuperator for heat recovery.

Additionally, the LNG stream was expanded to increase the energy output. This assumed that there were neighboring customers that required a supply pressure of 25 barg. Therefore the 3 main process systems were: 1) the LNG system, 2) the ORC system and 3) the seawater system. The seawater system is used to heat the inlet stream of each turbine. At most one level of heat integration was considered in each system and across each system. Further heat and equipment integration of the different process systems were not considered. Additionally, the turbine was examined and an inter-stage was included in the design to optimize the power output. In my first article [2], I show that the largest amount of exergy destroyed is in the ORC heat exchangers. Complex heat exchangers to reduce exergy destruction by better matching the heat curves in the heat exchangers were not studied [6] [7] [8].

2.2. Process Flowsheets

The above modifications were employed to develop different process flowsheets. Each flowsheet considers a different aspect of the modification to permit comparisons between the modifications. The alternate route of heating the LNG stream is not included in the flowsheets below. The process flowsheets are shown below:

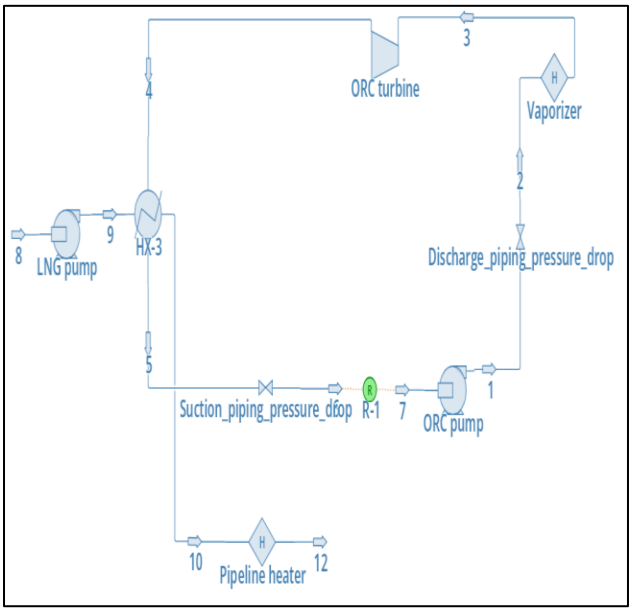


Figure 1 Simple ORC

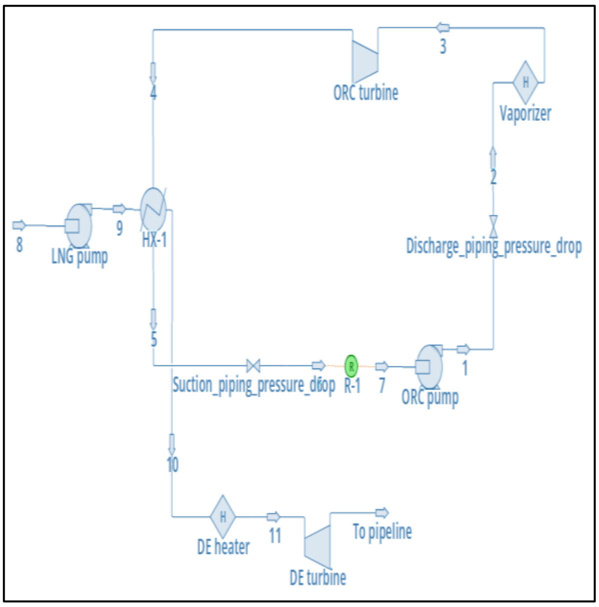


Figure 2 ORC With Direct Expansion

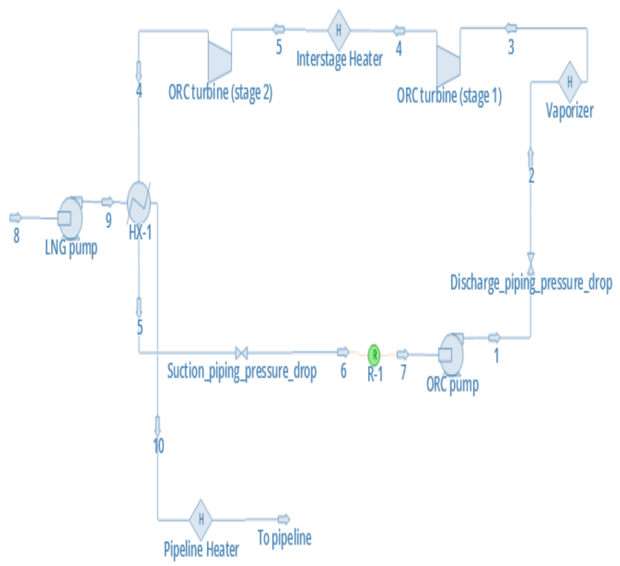


Figure 3 2-Stage Turbine ORC

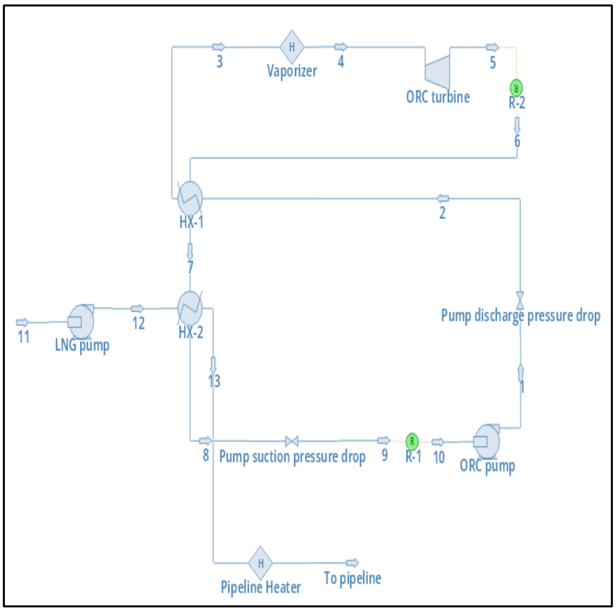


Figure 4 Heat Integrated ORC

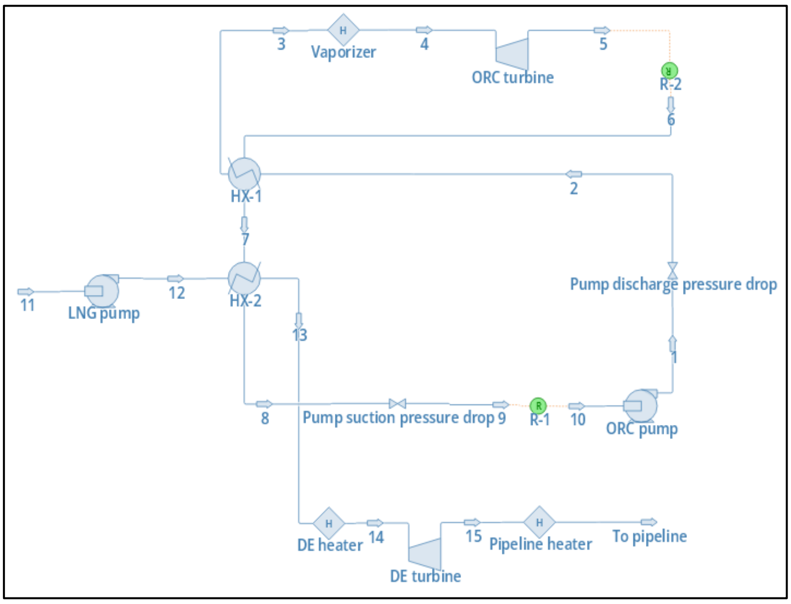


Figure 5 Heat Integrated ORC With Direct Expansion

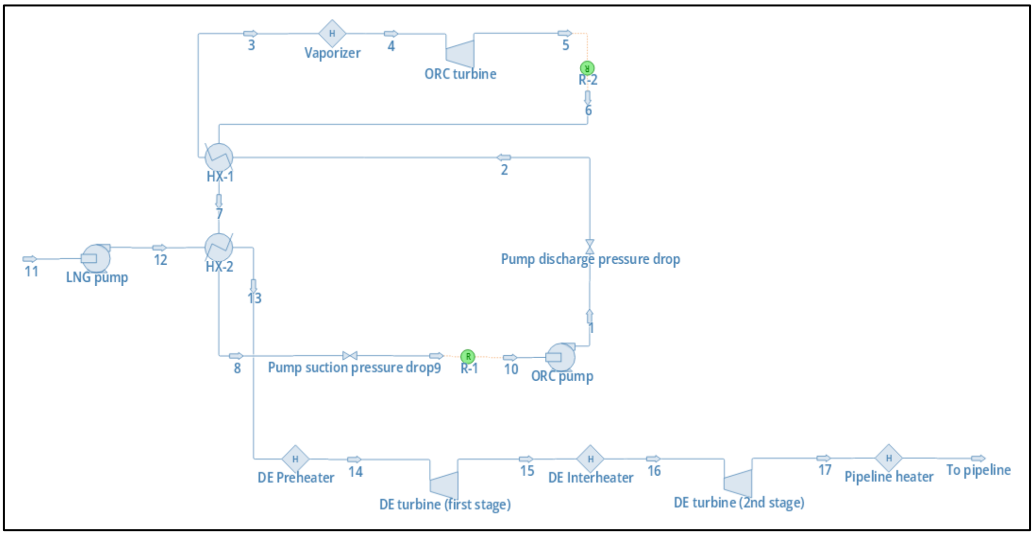


Figure 6 Heat Integrated ORC With 2-Stage Direct Expansion

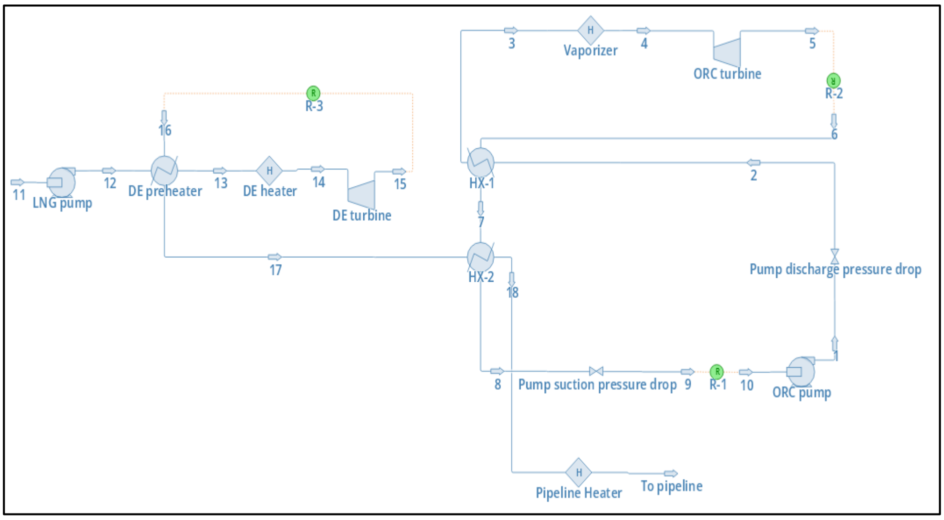


Figure 7 Heat Integrated ORC With Direct Expansion Before the ORC

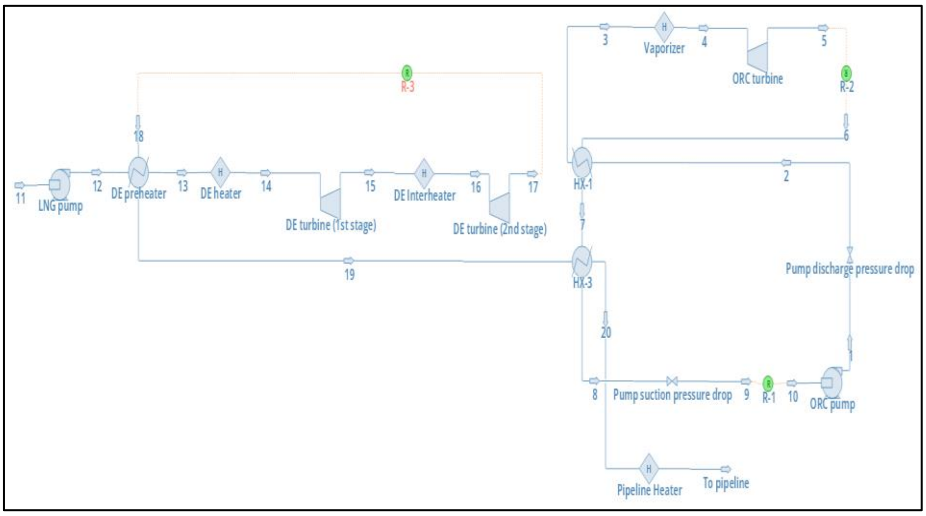


Figure 8 Heat Integrated ORC With 2-Stage Direct Expansion Before the ORC

2.3. Process Design Basis

The process design basis is shown below:

Table 1 Ambient or Reference Conditions

Parameter	Value
Temperature	300 K
Pressure	101325 Pa

Table 2 Process Design Basis [9] [3] [10] [2] [11]

Parameter	Value
Existing Facility Design Details:	
Pipeline heater heat exchange area	52.1 m^2
Seawater mass flowrate	2.06 x 10^6 kgh^{-1}
Pipeline LNG temperature	0 ^oC
Turbomachinery:	
Pump efficiency	82 %
Turbine efficiency	90 %
Turbine mechanical efficiency	97 %
Generator Efficiency	98 %
Gearbox Efficiency	97 %
ORC Turbine pressure ratio	0.375
ORC pump suction pressure	34.5 kPag
LNG pump discharge pressure	7 MPa_g
LNG pump suction pressure	1.13 barg
LNG pump suction temperature	-161 ^oC

LNG pump flowrate	22.9 kgs^{-1}
Heat Exchanger	
Heat Exchanger Minimum Approach Temperature	2 K
Minimum discharge seawater temperature	293 K
Pressure Drop:	
Seawater stream	100 kPag
Hydrocarbon (HC) liquid stream	80 kPag
Gas stream	70 kPag
Condensing HC stream	21 kPag
Boiling HC stream	14 kPag
Overall Heat Transfer Coefficient:	
HC vaporization – water	4.0 kW/m^2 K
HC vaporization – HC condensation	3.8 kW/m^2 K
HC condensation – HC liquid	4.0 kW/m^2 K
HC vapor - water	3.0 kW/m^2 K
Facility life	20 years
Pump suction and discharge pressure drop	5 kPa_g
Natural gas to pipeline temperature	≥ 0 ^oC

2.4. Assumptions and Risks

Some of the assumptions and technology risks in the calculations are listed below:

- The LNG heat exchanger can be purchased.
- No heat loss to the environment (i.e perfect insulation of the equipment)
- The existing facility LNG vaporizer and the seawater pump can be reused.
- Only 4 additional operators are required.
- The cost is for a new facility.
- Stainless Steel is conservatively employed in the calculations for the cycle and natural gas flow path.
- The material specification for the seawater pump is bronze.
- The number of maintenance days is approximately 10

2.5. Economics Model and Key Performance Metrics

The modelling work from R. Rauceo [2] was used for equipment costing and for the installed green field facility costing with the exception of the seawater pump where the model from M. Astolfi [3] was employed. New model parameters from G. Towler and R. Sinnott [11] were employed. The cost equation for a plate and frame heat exchanger was used in place of a brazed aluminum heat exchanger. A cost overdesign factor of 7 % was included for the equipment [11].

New financial data below based on a hypothetical facility are presented below

Table 3 Key financial data

Parameter	**Value**
Price of power	USD 0.2 per kWh
Interest rate	7.5 %
Operator salary	USD 16,500

A new KFPM, the LNG ORC Value that is the money earned from a standard volume of LNG is defined below:

$$\text{LOV} = \frac{\text{TAE}}{\text{Standard LNG volume flow}} \qquad \text{Equation 1}$$

KTPMs were defined accordingly for the respective process flowsheet. General definitions are given below. The Plant Electrical Energy Efficiency (PEEE) was calculated to evaluate the efficiency in converting the energy added to the plant to electrical energy.

$$\text{PEEE} = \left(\frac{\text{Useful work from the plant}}{\text{Energy or Work inputted into the plant}}\right) * 100 \qquad \text{Equation 2}$$

The Plant Exergy Utilization Efficiency (PEUE) is a measure of how efficient the plant is in converting plant exergy (maximum work potential of the cycle) into useful work by examining the plant exergy input (PEI) and the plant exergy output (PEO) including from process streams and equipment

$$\text{PEUE} = \left(\frac{\text{Useful work from the plant}}{\text{PEI} - \text{PEO}}\right) * 100 \qquad \text{Equation 3}$$

The Plant Exergy Recuperation Ratio is the maximum work the plant can produce to the exergy input into the plant.

$$\text{PERR} = \frac{\text{PEI} - \text{PEO}}{\text{PEI}} \qquad \text{Equation 4}$$

The Plant Power Production Density is the net power that can be obtained from the plant per standard volume of LNG

$$\text{PPPD} = \frac{\text{Net plant power}}{\text{Standard LNG volume flow}} \qquad \text{Equation 5}$$

This study examines various simple modifications of the cycle via different process flowsheets to improve the efficiency and profitability of the plant. Formulae are contained in Appendix B. The axial turbine cost equation was used in place of the radial turbine cost equation [12].

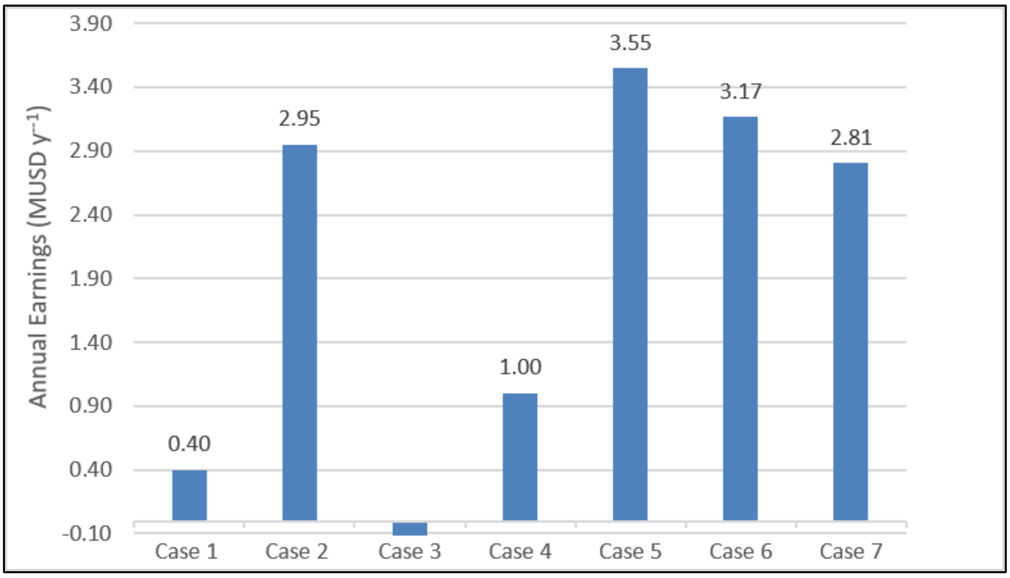
The following equation which was employed is a good starting point to obtain the optimum inter-stage pressure of the turbine to give maximum work (Addendum C):

$$P_i = \sqrt{P_{in} P_{out}} \qquad \text{Equation 6}$$

Ideally, a sensitivity analysis will give a very accurate value for the maximum work and TAEs.

3. Discussion

3.1. Profitability Comparison Analysis



LEGEND:	
Case 1	Simple ORC
Case 2	ORC with direct expansion
Case 3	2-stage turbine ORC
Case 4	Heat integrated ORC
Case 5	Heat integrated ORC with direct expansion
Case 6	Heat integrated ORC with 2-stage direct expansion
Case 7	Heat integrated ORC with direct expansion before the ORC

Figure 9 Comparison of the TAE

Case 5 yields the highest TAE with a value of 3.55 MUSD y^{-1}. Case 1 is the new simple ORC case. It has a TAE of 0.402 MUSD y^{-1} which means it is profitable. Case 2 includes direct expansion of the LNG stream in the plant and produces a TAE that is more than 7 times Case 1 with a value of 2.95 MUSD y^{-1}. This indicates direct expansion of the LNG stream is immensely profitable and increasing the LNG pressure and then expanding it to the required pressure will increase the profitability of the plant.

Case 4 includes a recuperator to recover cold energy in the cycle therefore allowing for an increase of mass flowrate of the working fluid until the pinch point is reached again. When this is performed a TAE value of 1.0 MUSD y^{-1} is obtained. This is over double the base case value but is less than Case 2's value. It was also found that including a 2nd-stage for the turbine does not improve the TAE. While the power produced from the LP-HP turbine increases, the resultant TAE is less for the comparable case. This was performed for an inter-stage heat exchanger pressure drop of 70 kPag and 6.9 kPag and in both cases there was no improvement in the TAE (this is also seen with case 3, where additionally the net power produced is higher than case 4, a profitable case). This indicates a costing equation for the multistage turbine is required and improvements in its design and manufacturing are required. It is also anticipated that the power output increase will increase for a staged expansion compared to a single turbine for increases in the inlet pressure at the same outlet pressure.

Placing the LNG expansion turbine before the ORC increases the backpressure on the turbine and also lowers the enthalpy of the stream that will pass through the ORC. A comparison of the direct expansion case with the direct expansion before the ORC shows that TAE is less than for the case where it is placed after the ORC unit but there are excellent opportunities for further heat integration.

The analysis also shows that the AIRR which is equivalent to IRR in this analysis produces the same ranking profile as the TAE ranking profile with the exception of case 6 hence AIRR and IRR are not good metrics for the ranking of projects. It should be noted advanced economics such as depreciation or salvage value are not considered in this article. Another insightful graph that shows the money earned from a standard cubic meter of LNG is also shown below.

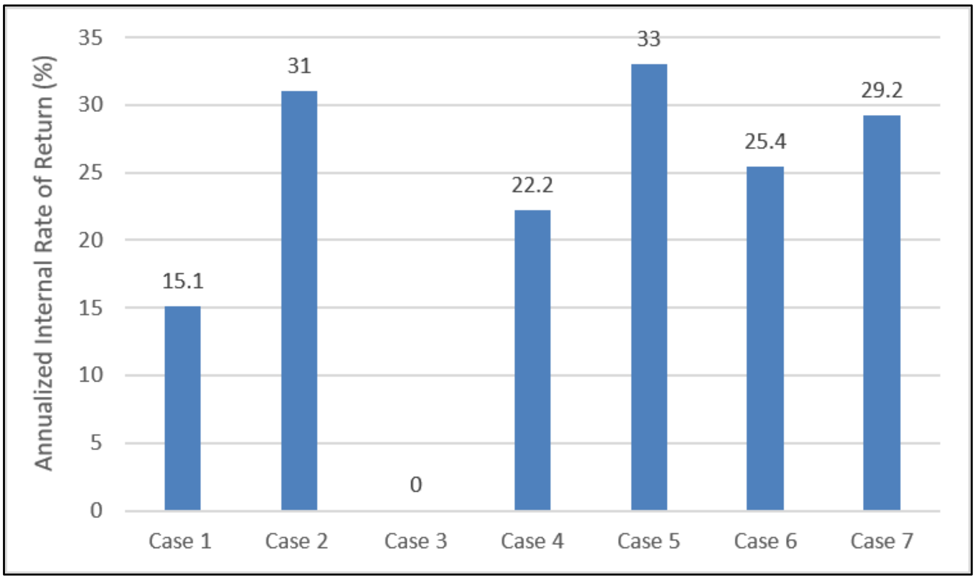


Figure 10 Comparison of the AIRR Cases

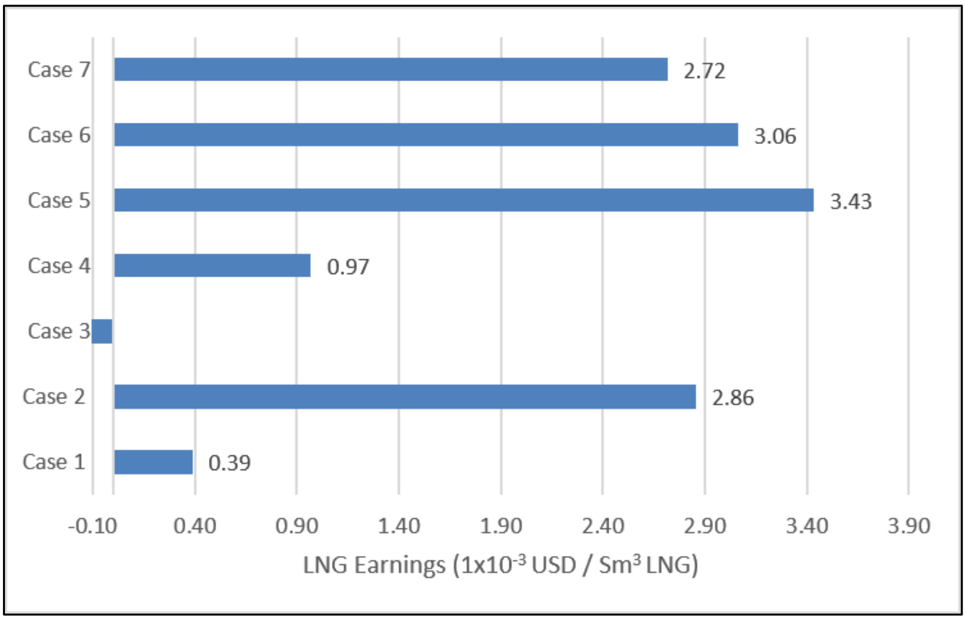


Figure 11 Comparison of the LOV

3.2. Plant Technical Performance Comparison

The PEEE and PEUE follow the trend of the TAE for most cases except case 6 where the cost of the additional turbine stage outweighs the additional power produced. Case 6 has a PEEE and PEUE of 15.2 % and 26.6 % respectively, the highest values of the cases. It should also be noted that seawater is low grade heat and the main operating cost considered in the seawater system is the power required to pump the seawater.

The PERR shows that direct expansion process flowsheets are able to absorb the greatest exergy from the input streams as more power is produced as the gas is expanded to a lower pressure compared to the other process flowsheets. They each have a value of approximately 0.544. This graph is shown below:

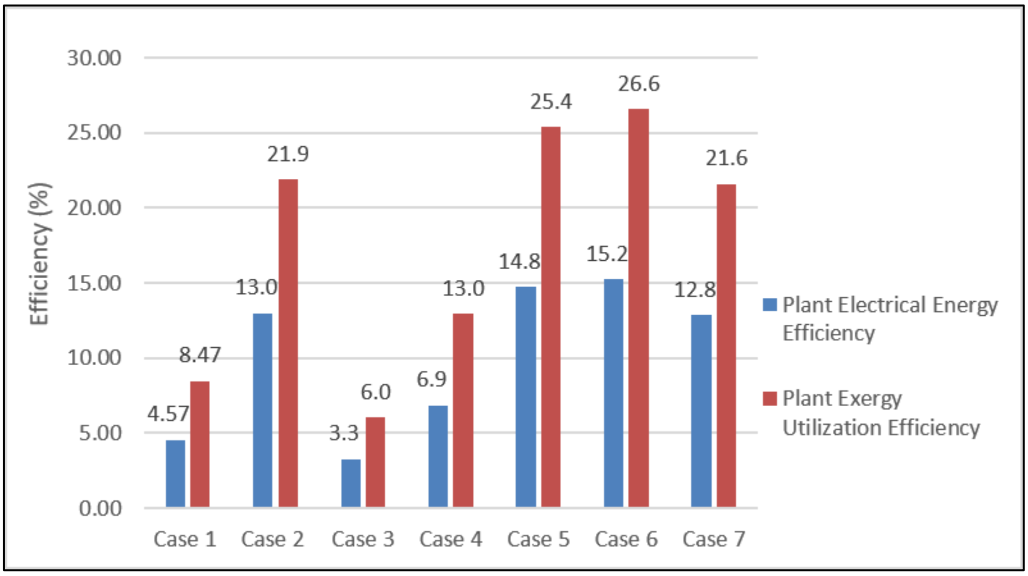


Figure 12 A Comparison of the Plant Electrical Energy and Exergy Utilization Efficiency

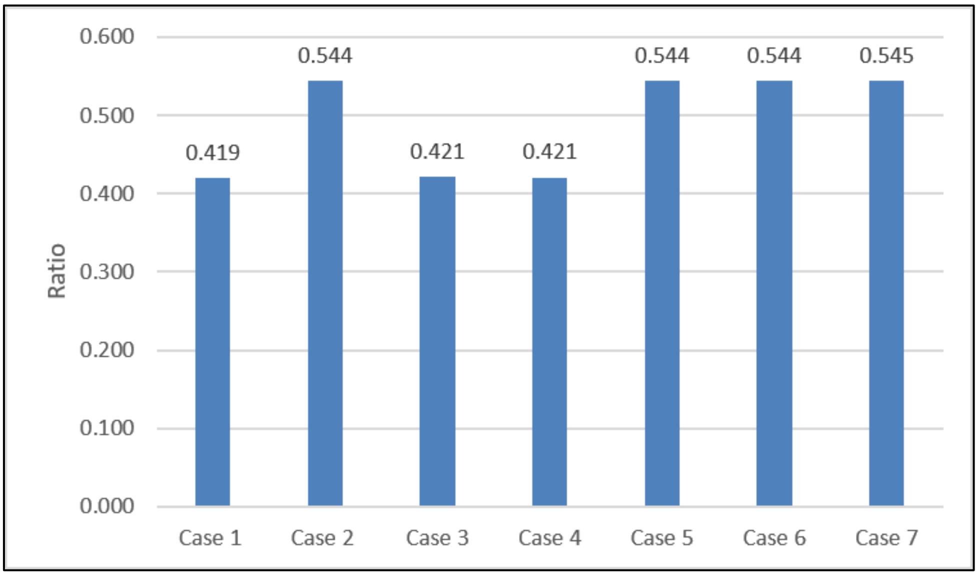


Figure 13 A Comparison of the Plant Exergy Recuperation Ratio

Finally, the PPPD is the net power that can be produced per standard volume of LNG by connecting the plant to a LNG regasification facility. Case 6 has the highest value at 0.0236 kWh / Sm^3 LNG but it has the second highest TAE. The PPPD graph is shown below:

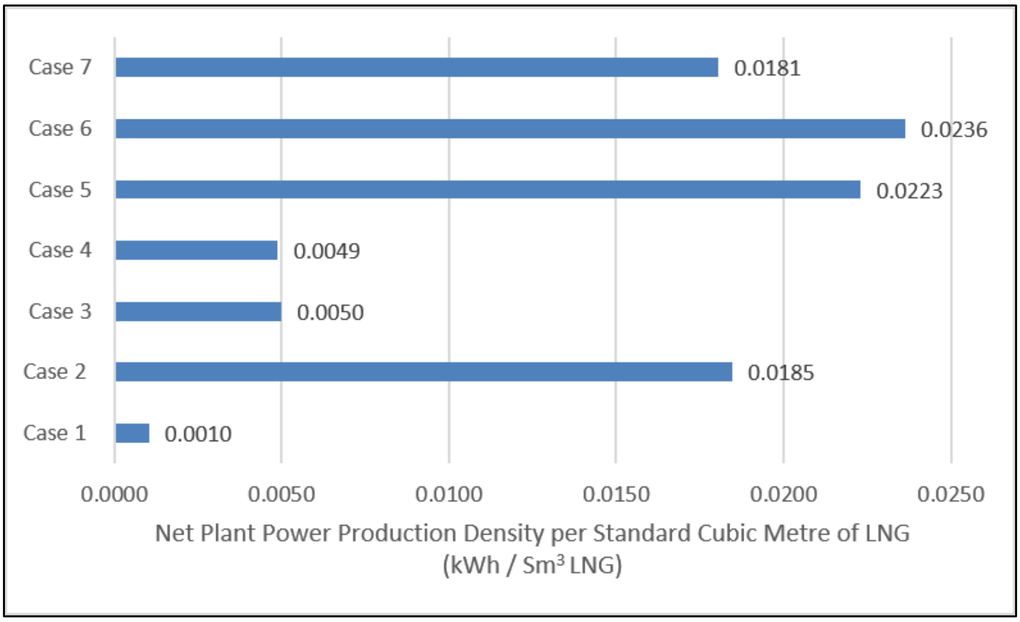


Figure 14 A Comparison of the Plant Power Production Density

Case 8 was not simulated since it was already shown that with the current methodology there is no increase in the TAE with a turbine 2nd stage. Additionally, it was noticed that most of the lowest TDRs were attained in the vaporizer where the propane underwent phase change. This and the effect of the type of fluid on the turbine are some of the things that will be explored in more depth in another article.

4. Conclusion

The analysis above evaluates different process flowsheets for a power plant that contains an ORC unit. It employs the KFPMs and the KTPMs to analyze the different configurations. It is shown that the simple ORC is profitable with a TAE of 0.402 MUSD y^{-1}. The ORC profitability more than doubles to a value of 1.0 MUSD y^{-1} when a recuperator is included and it increases by more than 7 times the base case value to a value of 2.95 MUSD y^{-1} when the LNG stream is expanded. The combined effect of these changes is a power plant that produces a TAE of 3.55 MUSD y^{-1}. It is also shown that having a 2-stage turbine doesn't improve the TAE and the AIRR is not a good metric for ranking projects. Case 5 has a TAE of 3.43×10^{-3} USD per Sm^3 LNG and this can be used to determine the additional earnings from a tank of LNG.

A turbine 2nd stage improves the power output, this results in a maximum plant efficiency of 15.2 % and PEUE of 26.6 % but it doesn't improve the TAE. It is shown that the maximum exergy recovery is obtained when the gas is expanded with an approximate PERR value of 0.544. Case 6 has a maximum PPPD of 0.0236 kWh / Sm^3 LNG which can be employed to determine the cold power capacity of an LNG tank. Therefore, increasing the pressure of the LNG and then expanding it is very profitable. It is shown that simple improvement modifications to the process flowsheet redounds in significant benefits.

Compliance with ethical standards

Disclosure of conflict of interest

The author has no potential conflicts of interest to disclose

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Authors short biography

Raphael Rauceo is currently a freelance general engineer that conducts research on ORCs. He has accumulatively over 9 years of experience encompassing process engineering and, in recent years, product engineering. He has a BSc from the University of the West Indies, an MSc from the University of Manchester and an MS from Drexel University.

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Appendix A

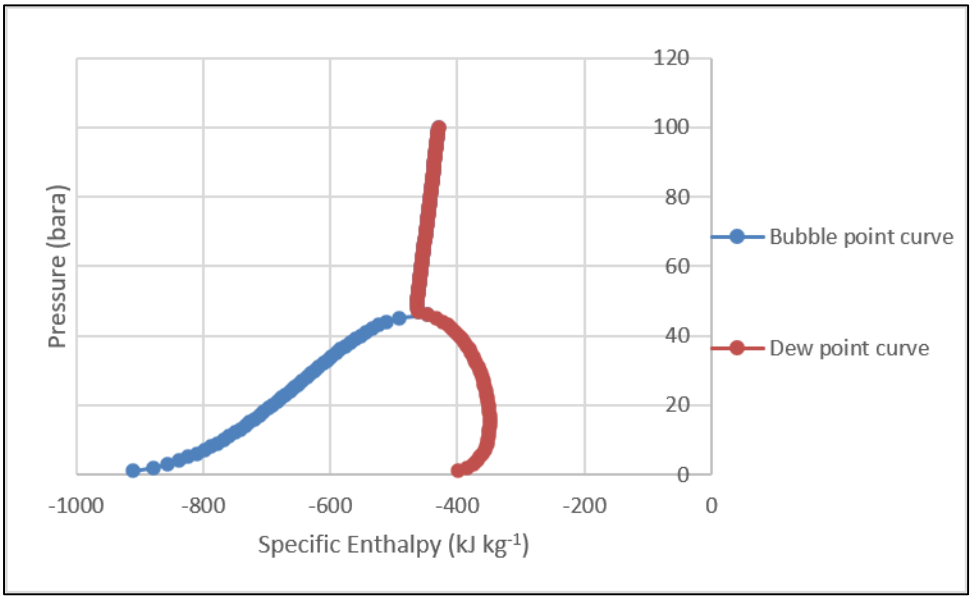


Figure 15 Methane Phase Envelope

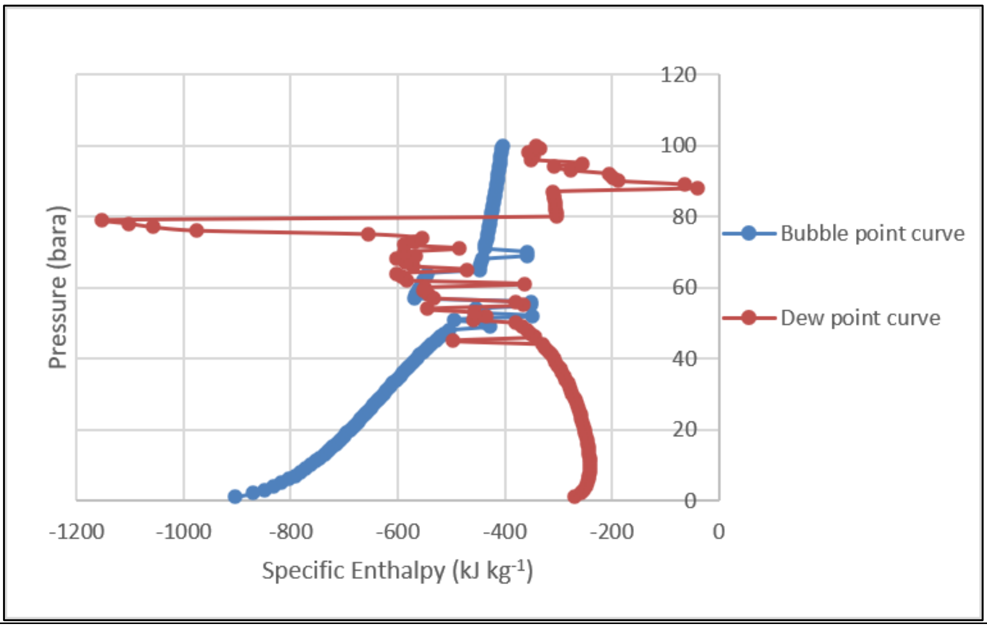


Figure 16 Mixture Phase Envelope

Appendix B

Table 4 Plant efficiency and exergy equations

Equation	Equation Number
Net plant power production = Generator efficiency x turbine mechanical efficiency x gearbox efficiency x ∑Turbine power output - ∑ Pump motor power	Equation 7
Plant electrical energy efficiency= $\left(\frac{\sum W_{G_i}}{\Delta H_i + W_{PL} + W_{P,ORC} + \sum W_{P,SW_i}}\right)*100$	Equation 8

Where W_{G_i} is the work produced by generator i, ΔH_i is the heat input into the plant, W_{PL} is the motor power for the LNG pump, W_{ORC} is the motor power for the ORC pump and W_{SW_i} is the motor power for each seawater pump.	
Plant exergy utilization efficiency = $\left(\frac{\sum W_{G_i}}{\Delta B_{Wi}+\Delta B_{Li}+\sum W_{P_i}-\sum \Delta B_{Wo}-\Delta B_{Lo}}\right)$*100 Where ΔB_{Wi} exergy of the inlet water stream, ΔB_{Li} is the exergy of the LNG stream, exergy of the water streams out, ΔB_{Lo} is the exergy of the LNG stream out, W_{P_i} is the motor power of the individual pumps	Equation 9
Plant exergy recuperation ratio = $\frac{\Delta B_{Wi}+\Delta B_{Li}+\sum W_{P_i}-\sum \Delta B_{Wo}-\Delta B_{Lo}}{\Delta B_{Wi}+\Delta B_{Li}+\sum W_{P_i}}$	Equation 10

The equations above were modified from reference [2] of this article for the plant calculation

Addendum C

The optimum specific work produced [11] and inter-stage pressure for gas expansion are derived below:

$$\text{Work produced, } W = \alpha_1\left[1-\left(\frac{P_i}{P_{in}}\right)^{\beta_1}\right] + \alpha_2\left[1-\left(\frac{P_{out}}{P_i}\right)^{\beta_2}\right], P_{losses} \approx \text{negligible} \quad \text{Equation 11}$$

$$\alpha_j = z_j \frac{RT_j}{M_w}\frac{E_p}{\beta_j} \quad \text{Equation 12}$$

$$\frac{1}{\beta_j} = \frac{n_j}{n_j - 1} \quad \text{Equation 13}$$

Taking alpha, beta, P_{in} and P_{out} as constants and solving for $\frac{dW}{dP_i} = 0$,then

$$\frac{dW}{dP_i} = -\alpha_1\beta_1 P_{in}^{-\beta_1} P_i^{\beta_1 - 1} + \alpha_2\beta_2 P_{out}^{\beta_2} P_i^{-\beta_2 - 1} \quad \text{Equation 14}$$

$$P_i = \left[\left(\frac{\beta_2\alpha_2}{\beta_1\alpha_1}\right) P_{in}^{\beta_1} P_{out}^{\beta_2}\right]^{\frac{1}{\beta_1+\beta_2}} \quad \text{Equation 15}$$

Assuming $\beta_1 = \beta_2$ & then $\alpha_1 = \alpha_2$ produces:

$$P_i = \sqrt{\left(\frac{\alpha_2}{\alpha_1}\right) P_{in}P_{out}} \text{ reduces to } P_i = \sqrt{P_{in}P_{out}} \quad \text{Equation 16}$$

Equation 16 is then substituted in the pressure ratio equations to give:

$$\frac{P_i}{P_{in}} = \frac{P_{out}^{1/2}}{P_{in}^{1/2}} \quad \text{Equation 17}$$

$$\frac{P_{out}}{P_i} = \frac{P_{out}^{1/2}}{P_{in}^{1/2}} \quad \text{Equation 18}$$

Equation 16 is a simple and good approximation for the optimum inter-stage pressure for the 2-stage expansion or compression and has been derived and pervasively employed [13]. Equation 15 is a more complex and informative equation that was thought to be more accurate but nevertheless yielded a good result. Intuitively from the last 2 equations the optimum work is produced when the 1st and 2nd stage pressure ratios are equal and hence the work produced from each turbine is the same [9] [11]. In reality, they may be different as shown in the other equations.

Addendum D

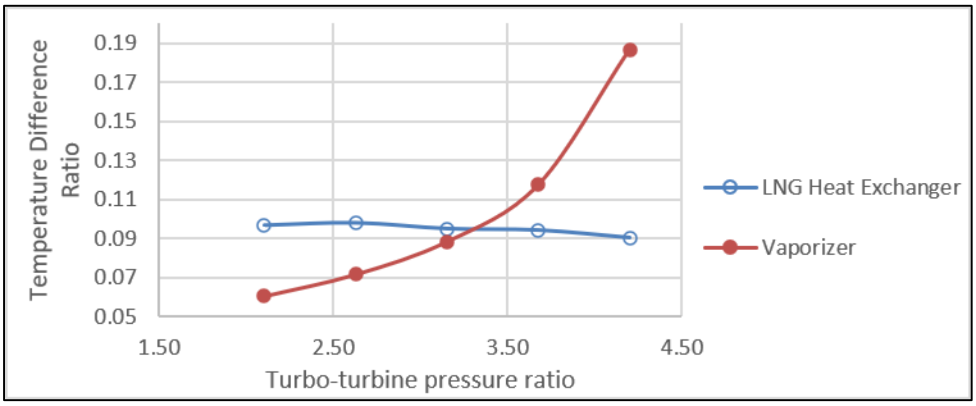


Figure 17 The updated Figure 11 from reference [2]

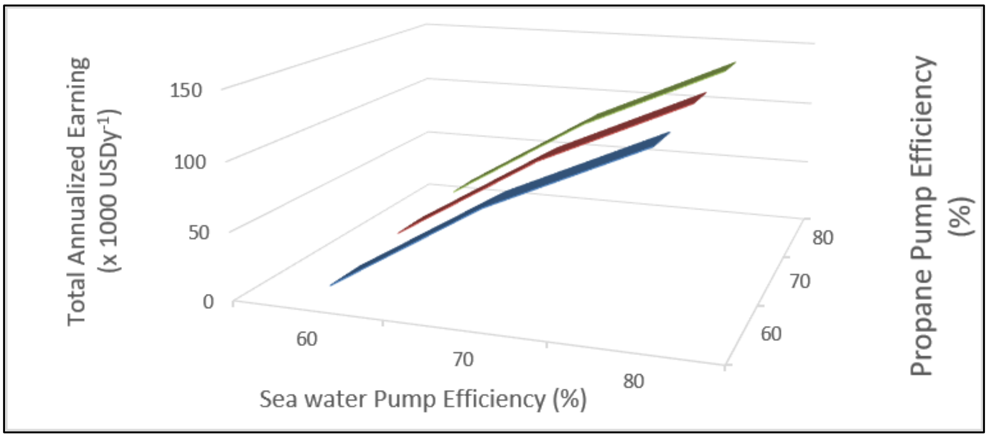
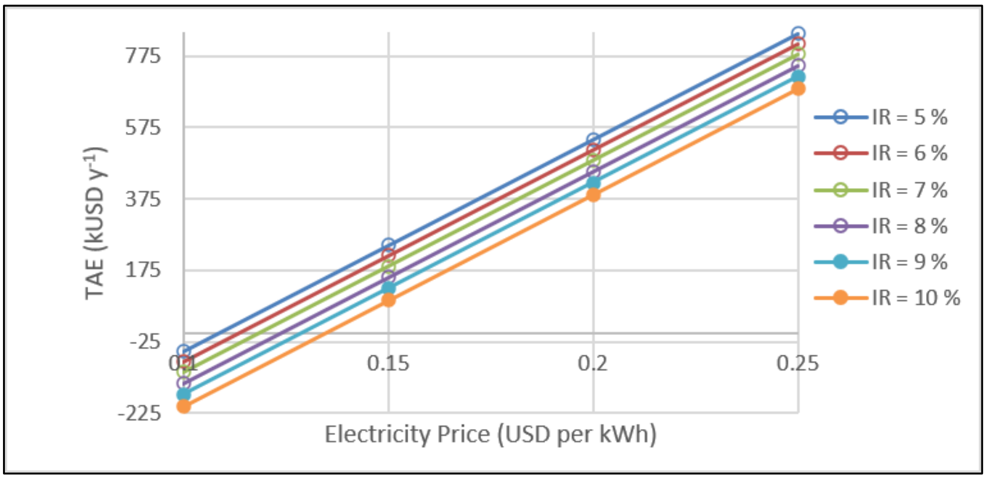


Figure 18 The updated Figure 14 from [2]



Note: The updated graphs only include the results for the adjusted calculation for the new seawater pump cost equation

Figure 19 The updated Figure 15 from [2]